ARMY CONSTRUCTION
ENGINEERING RESEARCH LAB

U.S. ENVIRONMENTAL
PROTECTION AGENCY

University of Pittsburgh

PROCEEDINGS:

FIRST NATIONAL SYMPOSIUM/WORKSHOP ON ROTATING BIOLOGICAL CONTACTOR TECHNOLOGY

HELD AT CHAMPION, PENNSYLVANIA
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EDITED BY ED D. SMITH, R. D. MILLER, AND Y. C. WU
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FOREWORD

As our population increases, and our costs for treating wastewater increase, greater emphasis is being placed upon the development of simple to operate, effective, reliable, energy conservative, and efficient pollution abatement technology. Rotating Biological Contactor (RBC) technology meets these requirements. Too often, however, RBC technology has not even been considered as a viable treatment technology wherever wastewater pollution abatement problems exist.

The objective of this Symposium/Workshop was to provide a forum and focal point in which an interdisciplinary group of scientists, engineers, planners, academicians, researchers, consultants, and sewage treatment plant operators could exchange ideas, present technical information, define the problems, assess the state of knowledge and identify the research needs regarding rotating biological contactors.

More than 345 participants representing a wide variety of experiences and viewpoints attended the symposium. Many of these attendees expressed satisfaction that for the first time, they had had the opportunity for extended face-to-face communication with persons with widely different perspectives about what the RBC problems are and how they might be resolved.

This event provided a unique national platform for the presentation of new knowledge and the most advanced thinking on all aspects of RBC technology. The fact that the Symposium/Workshop included 68 papers and that participants traveled from Japan, Canada, Switzerland, Italy, Belgium, Sweden, Denmark, France, and Norway, is testimony that significant global interest exists in RBC technology; that RBC technology is being applied to a broad range of functions; that application of RBC technology is increasing; and that RBC applications are important solutions to wastewater pollution abatement efforts.

The symposium focused mainly upon municipal and industrial RBC application with speakers presenting papers ranging from the theoretical to the highly practical. In particular, the intent was to:

1. Provide a history, overview, and perspective,

2. Present research results,

3. Present practical experiences,

4. Provide a forum for discussing the problems of compliance with pollution abatement through use of the RBC process,

5. Encourage information exchange and the transfer of technology, and

6. Identify research needs.

To accomplish these goals, the symposium was designed to provide information concerning various aspects of the theory, design, operation, and evaluation of the RBC treatment system. This information should significantly improve the state-of-the-art understanding of the RBC process, thus optimizing treatment performance. Moreover, it is hoped that through the definition of specific research needs, a large portion of interested research talents in the environmental engineering profession will be diverted to RBC scrutiny. More importantly, it is hoped that RBC technology will be considered as a waste treatment option whenever a wastewater pollution problem exists. It was calculated that these 68 papers represent a doubling of the state of knowledge for RBC technology.

No attempt has been made to edit, reformat or alter the material provided except for printing production requirements or where obvious errors or discrepancies have been detected. Any statements or views here presented are totally those of the speakers and are neither condoned nor rejected by the Symposium/Workshop co-sponsors.

Ed D. Smith, Ph.D.
Yeun C. Wu, Ph.D.
Roy D. Miller, Ph.D.
Co-editors

ABSTRACT

This document is a compilation of 68 papers presented at the First National Symposium/Workshop on Rotating Biological Contactor (RBC) Technology sponsored by the University of Pittsburg in cooperation with the US Army Construction Engineering Research Laboratory (Champaign, IL) and the USEPA Office of Reseach and Development's Municipal Engineering Research Laboratory (Cincinnati, Ohio). The Symposium/Wokshop was held 4-6 Feb 80 at Champion, Pennsylvania.

The papers presented in the three-day Symposium and the findings of the research needs Workshop comprise the major portion of these proceedings. Question and answer sessions preceeding each paper and a list of participants are provided.

The Symposium/Workshop proceedings will document pesent knowledge regarding RBC technology. The papers are divided into 11 major topic areas:

1. Perspective, Overview, History
2. Process Variables and Biofilm Properties
3. Municipal Wastewater Treatment
4. Biokinetc Studies
5. Air Drive and Supplemental Aeration
6. Industrial Wastewater Treatment
7. Concepts and Models
8. Upgrading Primary and Secondary Waste Treatment Systems With RBC's
9. Design and Operation
10. Nitrification and Denitrification
11. Selections and Economics

The Research Needs Workshop discussions were taped and are presented as an appendix. These proceedings document present knowledge regarding RBC's and are disseminated as a definition and establishment of priorities for research.

ACKNOWLEDGEMENTS

Because of the space limitation, the organizing committee cannot list all of the persons who had contributed to this Symposium/Workshop, but we wish to mention that Dr. R. K. Jain, Chief of the Environmental Division, USACERL, Dr. M. L. Williams, Dean of the School of Engineering, the University of Pittsburgh, were among them. Their support and advice are gratefully appreciated. In addition, the organizing committee deeply appreciates keynote speaker, Dr. R. L. Bunch of the U.S. Environmental Protection Agency for his effort and time, and also all session chairmen who successfully monitored the paper presentations. Special thanks is given to Ed Opatken of the Wastewater Research Division (Municipal Environmental Research Laboratory, Office of Research and Development, USEPA, Cincinnati, Ohio 45268) for his help in selecting a Symposium site and for looking after numerous logistic details before, during, and after the symposium. His cooperation was invaluable to the success of this Symposium/Workshop.

Finally, the organizing committee would like to thank all of the participants for their interest and enthusiasm that assured success of the Symposium. The Symposium assistants from the Seven Springs Mountain Resort and Department of Civil Engineering of the University of Pittsburgh did most of the typing and clerical work. Their pleasant and efficient assistants were instrumental in getting us through the entire period of this Symposium. To all these people we express our greatest appreciation. The proprietors who provided RBC exhibits are also thanked. Many others contributed ideas, time and effort; though not all can be named, they are warmly and sincerely thanked.

SYMPOSIUM CONCEPT AND DESIGN

The symposium was intended to provide a mechanism for:

1) Getting to know other people, including meeting and talking informally with other people working on similar projects. This enabled the formation of a resource base which can increase and improve the information available.

2) Sharing experiences, which included learning who is doing what, how they're doing it, what problems they've encountered, and how these problems have been overcome or are being approached.

3) Examining differences in perspective in such areas as mathematical modeling of the RBC process, and the economics associated with the process. It is believed that some progress toward resolving some of these differences occurred.

4) Providing input to State and Federal agencies relative to action deemed desirable to accelerate utilization of RBC technology.

The symposium was designed to facilitate the exchange of information and ideas. To this end, there were "keynote" speakers for each of several topics. The symposium featured one general session, several concurent sessions and one workshop session.

The workshop session began with opening remarks by one person, followed by a brief sharing of experiences by the panel members. At that point, the floor was opened up for general discussion with anybody free to make a contribution.

It is believed that the Symposium will have, as a final result, the effect of accelerating the rate at which RBC technology is utilized as an economically viable treatment technology in the United States.

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PART I: GENERAL SESSION

Keynote Address

ROTATING BIOLOGICAL CONTACTORS - ARE ALL SYSTEMS GO

By

Robert L. Bunch
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It is a real pleasure for me to be here this morning to "kick off" the first symposium ever devoted to the rotating biological contacting process (RBC). I am happy to see the excellent turnout for this symposium. Most of the experience and competency in this field are represented here today. The agenda indicates you will be quite busy in the next three days, but I am confident it will be a pleasant and productive experience for you. I am certain the proceedings, which will be published from this meeting, will provide very good technical guidance to those who could not attend.

The discovery of the wheel about 5000 years ago was one of the most important steps in man's development. Wheeled carts and wagons were much easier and faster to pull than sledges. The wheel soon found use in mechanics in controlling the flow of power. The three power sources used in the Middle Ages, animals, water and wind, were all exploited by means of the wheel. For example, waterwheels, windmills and beasts of burden were all used to drive millstones for grinding grain. Today's civilization would not be possible without the wheel. In the next three days, we will be discussing the wheel and how it can further be used to benefit mankind. It is a strange coincidence that the first wheels for carts and RBC process were made of wood. In the latter case, sheets of plywood served as discs. In fact, Dr. Buswell back in 1929 referred to his process as the biological wheel.

One of the few pleasures of getting old is that you are able to reminisce. You never realize how far you have progressed until you look back. It was in the middle 60's that I visited the University of Stuttgart and saw Dr. Popel's pilot plant and discussed the features of the RBC. Much progress has been made in our understanding of the process. We have come a long way since I was Project Officer for the first full-scale application of the RBC process in the United States on municipal wastewater at Pewaukee, Wisconsin. At first thought, it seems strange that with all the interest displayed in the RBC systems in the early thirties that the process should lay dormant until the middle sixties. Factors that probably prevented the early adoption were the deep depression in the thirties, World War II in the forties, and the popularity of the trickling filter process in the fifties. Lastly, most of man's technical progress has been achieved in the last 62 years. If we divide man's last 50,000 years of existence in lifetimes of 62 years, we would have about 800 lifetimes. Fully 650 of these were spent in caves. Only during the last two lifetimes was the electric motor used. The majority of all the material goods we use today were developed in the present lifetime or the last 62 years.

Enough of the reminiscing, let's look at the future. According to the 1978 EPA needs survey of treatment facilities, there were 59 RBC facilities in operation, 68 under construction and 305 required but not funded. The growth rate is increasing geometrically. We did not come here today to praise the RBC system nor to criticize it, but to decide where the voids are in our knowledge on design criteria and operating conditions. If we are to take full advantage of all the good features of this process, we need to know these voids. Let's briefly look at some of the design considerations.

SIZING AN RBC PLANT

How do we size an RBC plant? Should the design of an RBC plant be based on hydraulic flow or organic loading? Should contact time be a consideration? The early developments of wastewater treatment processes in England and the USA were based on domestic sewage. Since all domestic sewage, in those days, contained about the same organic strength and the K-factors were similar, it was easier to report data in terms of hydraulic flow and percentage reduction. This parameter is convenient because for a given physical system, hydraulic loading will be inversely proportional to detention time. If the first order reaction kinetics are applicable, percentage removal will be independant of organic concentrations.

As our country became more industrialized, the domestic wastes became mixed with industrial wastes. Today we refer to this mixture as municipal wastewater. To account for this change, the sanitary engineers introduced a design factor called population equivalent. This was supposed to compensate for the differences in organic strength of wastes. This factor, however, did not take into consideration the treatability of the wastewater nor the K-factor. Unfortunately today we still refer to the efficiency of wastewater treatment systems in terms of percentage reduction. Percentage reduction has little meaning unless related to the strength and type of wastes. If a plant is heavily loaded organically, it is possible to have a 90% reduction and still produce a very poor effluent as opposed to one normally loaded achieving a very low BOD with the same percentage reduction. Percentage reduction can be misleading. For example, it is much easier to achieve a 90% reduction of phosphorus on a raw wastewater that has 10 mg P/l than one that has 5 mg P/l. An effluent standard of 1 mg P/l can be met with mineral addition but to achieve 0.5 mg P/l would probably require filtration. In the future we will be more concerned with residuals in the effluent rather than about percentage reduction.

The measure of unit processes in the future will be cost vs residual. In other words, how much does it cost to obtain a certain residual. For example, to obtain an effluent containing less than 1 mg P/l costs about $50 per million gallons (3785 m^3). To achieve a residue of less than 0.5 mg P/l increases the cost twofold and for a limit of 0.05 mg P/l the cost soars twentyfold over the cost for 1 mg P/l.

Since the early development work on the RBC process was done on domestic sewage, it was natural that the hydraulic flow was first used to size the plants. In that the basis for the RBC system is biological degradation, should not the controlling factors be based on microbiological principles? If so, then organic loading on the discs and oxygen mass transfer efficiencies of the system will be the controlling factors. Many investigators have concluded that the RBC system follows first order kinetics, but with varying reaction rates with various stages.

DISC CONFIGURATION

Most RBC plants designed today have equal disc surface areas for each successive stage of treatment. Dissolved oxygen (DO) profiles follow a pattern of rapid initial decline and slow recovery in successive stages. Under heavy organic loadings, the liquor from the first stage can be distinctly anaerobic. Thus, increasing the hydraulic loading and/or organic concentration can stress the system. Increasing the number of stages will increase the total treatment potential of the system, reducing the stress. The excess organic material left untreated by the first stage can be treated by the second and third stage, etc. Having several stages can dampen out hydraulic surge and organic slugs.

For a given disc area and disc speed, the amount of oxygen transferred is fixed. An overstressed system will reduce the DO below the critical concentration and the efficiency of the system will be drastically reduced. It is not the relative disc size or oxygen transfer efficiency, but the absolute oxygen transfer capability of the system which determines whether the DO will be reduced below the critical concentration.

How do we alleviate the stress on the first stage? Should the first stage contain more disc area than the successive stages? Is the addition of liquid aeration a better design? Would step feeding of the system with part of the load added to the second stage be more cost effective? The addition of aeration would give the plant more flexibility in handling different types of wastes. Aeration may be the answer in situations where the waste characteristics are changed significantly by the addition of new industrial wastes after the plant has been constructed. Most certainly during the next three days disc configuration should be high on the list of topics to be discussed.

DISC SIZE AND ROTATIONAL VELOCITY

Much of the data upon which mathematical models and full-scale designs are based were obtained from small pilot plants. The assumption was made that a unit area on a small disc is equivalent in organic removal capacity to that of a larger diameter disc. Because peripheral velocity (tip speed) is directly proportional to the disc diameter, both rotational velocity (RPM) and peripheral velocity cannot be simultaneously scaled. Tip speed has been used in most cases as the scale-up parameter.

Evidence has been accumulating in the literature that larger discs have poorer oxygen transfer characteristics than small discs at the same tip speed. These studies would indicate that the possibility exists of introducing variable size discs in plants with smaller discs in the first stage where there is a high oxygen demand. Are further considerations of the diameter of disc and rotational velocity in order?

COSTS

The cost of constructing an RBC plant for a small treatment plant is almost as great as that of an equivalent activated sludge plant. The RBC process, being a unit modular process, does not have the scale-up advantages that other systems do. Ways need to be found to lower construction costs without increasing the service rate. Can cheaper material for the discs be found? Bacteria can grow on practically any material. All that needs to be done is rotate it. In the milder climatic portions of the USA a more open system could be designed. Lessons can be learned from the petrochemical industry for they have reduced cost by eliminating expensive structures. Less costly protection of the discs from elements can be designed.

In closing, I make the plea that each one of you try to make this symposium a workshop where new ideas and unexplored needs of the rotating biological contact process are discussed informally. The conference will be considered a success if we can clearly set forth the present knowledge on design criteria and define the research and development needs to fill the voids in our knowledge. The adaptation of wheels as gears was a conceptual leap. Engaging wheel rims to transmit or modify motion was not obvious. Quantum improvements in the RBC system are not obvious, but I am confident that there are among us today many who will continue to improve the biological wheel.

Keynote Address

TECHNOLOGY AND PUBLIC POLICY

by

Ravi Jain*

Mr. Chairman, ladies and gentlemen; it is indeed an honor for me to be here. I hope to listen to your presentations this morning and learn from you about the RBC technology. After looking at the roster of attendees, it is clear that you are indeed a distinguished group of participants.

I should tell you that for this keynote address I did receive advice from a number of sources. Dr. Ed Smith, who is one of the organizers of this conference, sent me a note and he said, "I know your keynote address will be thought-provoking, clever, and dynamic," and then he proceeded to attach an example keynote address from another conference, ostensibly to assist me with preparing my remarks. The example keynote address would have taken about 37 minutes to deliver. I asked Dr. Wu, Chairman of the Symposium Organizing Committee, as

*Chief, Environmental Division, USA-CERL, Champaign, IL. For the 1979-80 academic year, Dr. Jain was on leave from CERL to study Public Administration and Policy at Harvard.

to how much time I have for my presentation. His response was: "not much."

So, with the charge of making my presentation thought-provoking and dynamic, and with the requirement of not using much time, I have decided to share a few ideas with you on the characteristics of a professional scientific community like the one you represent here and ideas on technology and public policy.

As you know, a group of scientists, researchers, and other professionals, like yourselves, form a unique community. This community, as Daniel Bell has stated,[1] is such where the sovereignty is not coercive and the conscience is individualistic and sharing. As an imago (or an image), it comes closest to the ideal of the Greek polis. Robert Merton (a philosopher and author of Sociology of Science), has stated that ethos of a scientific community has four elements.[2] Two of these elements, that are particularly relevant to this conference, deal with sharing of knowledge and participating in organized skepticism. This conference is an example of a scientific community where many of you are willing to share your knowledge and scientific discoveries and at the same time participate in an organized skepticism in an effort to scrutinize and learn from the discoveries of others. So this sets the

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1. Bell, Daniel, The Coming Of Post Industrial Society, Basic Books, New York, 1973 (P 380).

2. Merton, Robert K., The Sociology Of Science, The University of Chicago Press, Chicago, IL, 1973 (P 270).

stage and the environment for this symposium.

Next, let me briefly comment on technology and public policy. In the context of this symposium, two aspects of technology: technology development and technology transfer might be of interest.

Technology Development: It is clear that as scientists and researchers you are involved in developing new technology. Technology developed has to accommodate many conflicting requirements on our resources such as energy and other material and human resources. The RBC technology may well provide answers to some of these issues.

Technology Transfer: The process of technology transfer, however, is difficult to understand if one were to look at science and technology in the context of the 19th century when science dealt primarily with machines and physical tools, i.e., hardware. Today technology consists increasingly of "software". This software deals with procedures, methodologies, and systematization of ways of doing things as opposed to merely specifications for things. For the research organizations, technology transfer would have two distinct components. The scientific information which the organization obtains from the outside scientific community, this could be referred to as an "input" and the "output", would be the scientific concepts, procedures and methodologies, developed by the organization for the potential user. You would agree with me that any research organization represents only a very small portion of the total scientific community;

therefore, interaction with the wider scientific community is essential. This conference can serve as a vehicle for technology transfer both on the input side and the output side. For instance, on the input side by helping us relate to the wider scientific community as represented by this group, and also on the output side by documenting research results for the use of practitioners who are also represented here.

As you know, tied to technology development and transfer is national productivity. It is interesting to note that the United States is far ahead of other industrialized countries in major technological innovations. However, if one were to look at trends in productivity represented by output per person hour, it is quite a different story. The U.S. productivity gain between 1960 - 76 was the smallest of the other major industrialized countries.[3] Some figures for the U.S. for 1979 show a decline in productivity which is quite alarming. It is possible that a lack of sufficient investment for transferring and implementing new technology accounts for, to some degree, this decline in productivity.

I would suggest that your effort here would help immensely towards bridging this gap between technology development and technology transfer.

The last item I would like to discuss with you is public policy as it relates to the environmental issues. While most of us

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3. Science Indicators - 1976, National Science Board, National Science Foundation, 1977 (P 35).

understand that public policy affects the extent to which resources are available for clean water, clean air and environmental protection and also resources for developing technology necessary for achieving these goals, let me suggest, we often ignore the effect technology has on policy. If clean water, clean air and other environmental amentities are important, then, it is essential that when the trade-offs between conflicting demands on resources are made and public policy decisions reached that people like yourselves -- who are knowledgeable not only in the environmental issues but also in economic and social issues -- be involved in these deliberations. This is essential if your knowledge of not only existing technology but also emerging technology is to positively affect national policy. I believe more can be done in this area.

I would simply like to close my address by commending the many sponsors and organizers of this symposium. Organizing a symposium like this requires considerable effort. Many of you who have organized similar activities know exactly what I am talking about. I hope to listen to your papers and get an opportunity to exchange ideas with you. As Dr. Wu mentioned, many of our participants have come here from other countries; all of you have come here leaving behind other important commitments. Your willingness to share your ideas and participate in this symposium are commendable.

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A HISTORY OF THE ROTATING BIOLOGICAL CONTACTOR PROCESS

By

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Introduction

In comparison to many other sewage treatment technologies (e.g., activated sludge and trickling filters) relatively few dollar and manpower research resources have been spent studying Rotating Biological Contactor (RBC) technology. The resultant lack of knowledge is due, in part, to the fact that the RBC process is relatively new in the United States. In fact, only a few RBC plants have been operational for more than a few years. Most of these are utilized for secondary sewage treatment with a few used for upgrading existing sewage treatment plants (nitrification and denitrification) or industrial waste treatment applications. Although millions of dollars have been spent by American industries and municipalities for RBC process equipment, the latest wastewater treatment guidance documents reveal a conspicuous lack of information regarding the RBC unit process. For instance, many excellent documents which provide design and operation and maintenance criteria/guidelines are readily available for traditional technologies such as the activated sludge and trickling filter processes. An example of such a publication is the excellent EPA report - Process Control Manual for Aerobic Wastewater Treatment Facilities (1). The purpose of the publication is to provide guidance to optimize the performance and to help establish process control techniques for trickling filter and activated sludge systems. There is no comparable manual for RBC technology. Other examples which demonstrate the novel nature of RBC technology in the United States are two excellent EPA documents - (1) Upgrading Trickling Filters (2) and (2) Process Design Manual for Upgrading Existing Wastewater Treatment Plants (3). They do not mention RBC technology. In addition, commonly used "state-of-the-knowledge" documents which are designed as guidance for the selection of wastewater treatment systems based upon economic considerations either do not have RBC cost curves (capital, O&M, energy, etc.) or the curves are dated. Because of the relative newness of RBC technology in the U.S., guidance is scarce with regard to RBC applicability, design, O&M and economic considerations.

The lack of empirical data and guidelines is complicated by the fact that there is no well-defined theory of design and operation accepted by all RBC manufacturers. Activated sludge, trickling filter and most other wastewater treatment processes may be designed and constructed without significant dependence upon equipment proprietors. This is not the case with RBC technology. Design engineers who have selected RBC technology are extremely dependent upon proprietors' design curves. This situation is compounded by the fact that the various RBC proprietors/manufacturers have differing philosophies of design and varying media densities, structure, etc.

Status of RBC Technology Today

RBC technology has been very popular in Europe for many years, and recently, in the U.S., has become increasingly popular for both municipal and industrial utilization. The extent and magnitude of interest regarding RBC technology becomes immediately evident when one contemplates the recent number of publications reporting RBC related research and operations experience. This symposium is further evidence of the interest the various sectors (private, academic, research, government agency, regulatory, A/E, professional organization, design engineer, industrial, and plant operators) have regarding RBC technology. All of the above and other professionals involved in wastewater treatment and management are represented at this symposium. Other manifestations of interest with RBC technology include the following:

1. The American Society of Civil Engineers (ASCE) has formed a "Rotating Biological Contactor Task Committee." (A report from this committee is scheduled in this symposium).

2. The U.S. Army Construction Engineering Research Laboratory and the U.S. Army Medical Bioengineering Research and Development Laboratory is investigating RBC technology applicability for upgrading existing Army sewage treatment plants.

3. The Federal Highway Administration, U.S. Department of Transportation has cooperated in research associated with the RBC process as a treatment method for wastewaters of roadside packs (4).

4. An ad hoc committee has been formed to evaluate the applicability of RBC technology for the People's Republic of China.

5. Proprietors/manufacturers of RBC equipment have increased dramatically during the last few years. Many of them are represented at this symposium.

Legislative Requirements

In 1972, Congress initiated a comprehensive program to restore and maintain the quality of the nation's rivers and lakes by passing amendments to the Federal Water Pollution Control Act (P.L. 92-500). The 1977 Clean Water Act (P.L. 95-217) reaffirmed this commitment through additional amendments which strengthened a number of the provisions of P.L. 92-500. These two laws require that industrial and municipal waste treatment operations constrain their point source wastewater effluents within prescribed limits of quality. In fact, certain mandatory penalties are stipulated and are enforced by the EPA.

Several wastewater technologies are available as candidate mechanisms for meeting these secondary or even more stringent NPDES permit stipulations.

Each of these technologies exhibits its own inherent technical and economic attributes (advantages/disadvantages, etc.). There has been a recent tendency among consulting firms to choose the more capital- and energy-intensive, and the more complex technology should be used when it is applicable to particular wastewater problems. However, it is more sensible to choose simple to operate, economical and reliable technology whenever possible. RBC technology is conducive to meeting these requirements. In particular, if one evaluates (and compares to other processes) RBC energy scenarios, operational/maintenance requirements, efficiency, reliability under various environmental and loading conditions, it becomes evident that RBC technology should be considered as an option whenever municipal and industrial pollution abatement is required.

Personnel interested in considering the RBC process as an option are faced with finding answers to the following questions:

1. How can I insure that RBC technology is right for my particular situation?

2. How much does it cost?

3. Are the RBC units easy to install and start up? What about site preparation?

4. Can we obtain the process and install it into a tight compliance schedule.

5. What are the RBC's operational and maintenance problems/costs?

6. How does RBC technology compare with other technologies?

7. Is the process reliable and effective under a variety of climatic conditions and under hydraulic, organic, and ammonia loadings?

8. What are the appropriate design criteria?

9. What are the system's land requirements?

10. What are its skill and manpower requirements?

11. What are the process advantages/disadvantages?

12. Can the process be retrofitted to existing secondary equipment to meet biochemical oxygen demand (BOD), suspended solids (SS), and ammonia requirements?

13. What industrial pollutants will RBC technology successfully treat?

14. What about nuisances (odors, filter flies)?

15. How does energy consumption compare to other processes?

16. What are the sludge characteristics?

17. What is the need for clarification prior to disinfection and discharge, and what design criteria are appropriate for clarification?

18. What are the life expectancies of major control components?

19. What new developments are anticipated for RBC technology?

20. What is the effect of extremely low temperatures?

21. What are the safety considerations?

22. What information is available?

23. What are the opinions of RBC plant operators? What kind of problems can I expect?

Need for an RBC State-of-the-Knowledge Definition

It is anticipated that this symposium will answer many of these questions by providing a state-of-the-knowledge definition of RBC technology. At the very least, the symposium will provide a forum for identifying problems. This problem definition will be interpreted into a prioritized list of research needs. As the research is performed, the design and O/M problems may be solved with a resultant increased popularity of RBC technology.

History of the Rotating Biological Contactor

According to a recent EPA report (5):

> The RBC concept of treating waste streams biologically has been known for many years, but it was not until strong, lightweight plastics become available that significant interest in the technique began to develop. The treatment technique is to grow biologically active masses on a series of discs that slowly rotate, alternately exposing the biomass to the air and to the wastewater.

In early models, the discs were made of metal and were heavy, cumbersome, and subject to corrosion. Recent models have discs fabricated of polyethylene or polystyrene. Many investigators have found advantages for the RBC over activated sludge or other conventional treatment systems based on specialized circumstances...

Historical information is also provided in a Civil Engineering article (6) titled "Behind the Rapid Rise of the Rotating Biological Contactor":

> The rotating biological contactor goes back to the 1920s. Investigators in both the U.S. and Germany experimented with using rotating wood surfaces. But wood surfaces were impractical to manufacture and deterioriated, and in those days, few communities were putting in secondary treatment.
>
> Not much more happened until the 1950s. In that decade, investigators at Stuttgart University, West Germany, attempting to improve the secondary treatment process, experimented with wooden and plastic flat disks rotating in wastewater.
>
> In 1959, J. Conrad Stengelin began to manufacture 2 and 3 meter diameter expanded polystyrene disks in West Germany. The first commercial installation went on stream there in 1960. But the rotating disk process was not cost competitive with the activated sludge process; initial capital costs were considerably more than for

activated sludge plants. Nonetheless, many small plants were installed in Germany in the 1960s -- most serving less than 1000 people. These small municipalities were willing to pay more in initial cost to get a plant requiring little maintenance and low energy consumption.

After 1960, further development of the rotating biological contactor stopped in Europe. But between 1960 and 1965 in the U.S., Allis-Chalmers did much development of rotating disks.

In 1970, Allis-Chalmers sold its rotating biological contactor technology to the Autotrol Corp. (Milwaukee, Wisc). At that time, the polystyrene disks were still not competitive with the activated sludge process. Even as late as 1972, Autotrol had sold only a few RBC installations for sewage treatment. The capital cost of the polystyrene disks was simply too high.

Breakthrough Sparks Growth of RBCs

Then, in 1972, came an important breakthrough: the development of a more compact disk, one with much more surface area for a given volume. Until then, the RBC unit consisted of a series of parallel, flat 0.5-in.-thick expanded polystyrene sheets, each separated by a 0.75-in. space. Now, Autotrol came out with an arrangement of 1/16-in.-thick polyethylene sheets with a 1.2-in. space separating them filled with a honeycombed polyethylene configuration. Whereas the standard polystyrene RBC unit was 10 ft in diameter and 17 ft long with 21,000 ft^2 of surface area, the new polyethylene RBC unit was 12 ft in diam., 25 ft long, with 100,000 ft^2 of surface area. In recent years, Autotrol has developed a still more compact arrangement for nitrification applications -- the distance between adjacent polyethylene sheets being only 0.6 in., with total surface area of a standard RBC being 150,000 ft^2...

The use of RBC technology in Europe (particularly in Germany) has been quite extensive, and over 700 installations (some with more than 25 years of experience) are presently in operation. Most of them are small, but the municipal plant at Ponavischigen, West Germany, serves about 100,000 persons (7).

Since 1972, the number of wastewater treatment facilities in the United States utilizing rotating biological contactors has increased more than 300, with another 300 now in the planning stages (8). An excellent historical review of the RBC process can be found in Ph.D. thesis of C. G. Grieves submitted to Clemson University.

RBC Literature Review

The following literature review provides information concerning various aspects of theory, design, and operating experience associated with RBC systems.

Historically, rotating biological contactors have been used to remove organic carbon from wastewater. This process was later expanded to include nitrification and denitrification of wastewater. One of the earliest reports of RBC application in the United States is by Welch (9), who successfully treated highly concentrated wastes using an RBC system installed at Allis-Chalmers, West Allis, WI. In terms of chemical oxygen demand (COD), as much as 800 lb/1000 cu ft/day (1.28 kg/m^3/day) removal was recorded. Torpey, *et al*. (10) reported a 10-stage RBC with aluminum disks which decreased BOD from 124 mg/l in the influent to 9 mg/l in the effluent after 5 months. Nitrification also occurred, which reduced the ammonia nitrogen content of the effluent (NH_3-N) from 14.2 mg/l to 5.7 mg/l and correspondingly increased the nitrate from zero to 10.4 mg/l in the effluent. Antonie (11), in his study of the RBC process response to fluctuating flow, reported significant chemical oxygen demand (COD) removal when the hydraulic residence time of wastewater was approximately 60 minutes. Hydraulic surge, which reduced the residence time to 30 minutes or less, resulted in low COD reductions. In a later report, Antonie (12) noted successful applications of the RBC process for treating various food and nonfood processing wastes. In an EPA demonstration project using the RBC system as a full-scale secondary treatment plant, Antonie (13) reported good BOD removal and some nitrification. In the winter, the system was placed in an enclosure to protect the biomass from freezing temperatures. In a pilot study conducted by LaBella, *et al*. (14) it was reported that the RBC process at a hydraulic loading of 1 gal/sq ft/day (0.04 m^3/m^2/day) could remove BOD from winery wastes at an efficiency comparable to that of an activated sludge process. However, the yearly operating cost of the RBC process was found to be $6000 per year less than the activated sludge process for a flow of 0.34 to 0.44 MGD (1290 to 1665 m^3/day). Chittenden, *et al*. (15) also used the RBC system to treat anaerobic lagoon effluents. At a hydraulic loading of 4.0 gpd/sq ft/day (0.16 m^3/m^2/day), increasing the rotating speed of the first stage to 6 rpm produced a 79.5 percent BOD reduction and an overall BOD reduction of 83.2 percent from an influent having an average of 225 mg/l BOD. Higher hydraulic loading and lower rotational speeds resulted in poor efficiency of BOD removal and little or no dissolved oxygen in the system. Using a synthetic wastewater for an RBC process study, Stover, *et al*. (16) reported that more than 90 percent COD removal was possible as long as the organic loading was kept below approximately 400 lb/1000 cu ft/day (0.64 kg/m^3/day). Using the same RBC system for slaughterhouse waste treatment, only 70 percent COD removal was achieved, even though the organic loading was low at 100 lb/1000 cu ft/day (0.64 kg/m^3/day). Increasing the loading to 400 lb/1000 cu ft/day (0.64 kg/m^3/day) reduced removal efficiency to 15 percent. Expressed in lb COD/day/1000 sq ft of disk surface area, the maximum COD removal for slaughterhouse waste was approximately 4.0 lb COD/day/1000 sq ft (19.5 g/m^2/day) at loadings of 8 lb COD/day/1000 sq ft (39 g/m^2/day) or higher. An investigation (17) to determine the efficiency of the RBC process on raw wastewater from a liquid detergent manufacturing plant was performed.

Selected parameters were chosen for measurement, including (COD), COD, BOD, MBAS and DO.

Applications of the RBC process for nitrification of wastewater or sludge supernatant have been reported. Weng, et al. (18) evaluated various parameters affecting the process performance and showed that among influent loading, flow rate, rotational disk speed, detention time, effective disk surface area, and submerged disk depth, only influent loading, flow rate, and effective disk surface area were important in determining nitrification efficiency (temperature steady at $20^{o}C$ and disk rotating speed at 10.5 or more rpm). In effect, NH_3-N loading was the only controlling factor.

Antonie (19) reported that at various treatment plants using the Bio-surf RBC process, as much as 0.8 lb NH_3-N/day/1000 sq ft (3.9 g/m^2/day) could be removed. Generally, 90 to 95 percent nitrification was obtainable. A pilot plant study conducted by Hao, et al. (20) showed excellent NH_3-N removal at the Columbus, Indiana, sewage treatment plant. In January and February, when cold temperatures prevailed, 50 to 60 percent NH_3-N removal was obtained at a hydraulic loading of 2.5 gpd/sq ft (0.013 m^3/m^2/day) and 90 to 95 percent NH_3-N removal at 1.5 gpd/sq ft (0.06 m^3/m^2/day). When high strength ammonia wastewater (780 mg/l NH_3-N on the average) was applied to a four-stage RBC system, Lue-Hing, et al. (21) found that at an overall NH_3-N loading of 15.6 lb of NH_3-N/day/1000 cu ft (25 g/m^3/day) and a wastewater temperature of $10^{o}C$, 99.4 percent of the NH_3-N was removed; at an overall loading of 43.5 lb of NH_2-N/day/1000 cu ft (70 g/m^3/day) and a wastewater temperature of $20^{o}C$, 99.8 percent of the NH_3-N was removed. The maximum removal rates in the first stage ranged from 95 lb of NH_3-N/day/1000 cu ft (152 g/m^3/day) at $9^{o}C$ to 170 lb of NH_3-N/day/1000 cu ft (272 g/m^3/day). Recirculation of effluent in the RBC process showed insignificant improvement of nitrification.

Temperature sensitivities of the RBC system have been evaluated by Murphy, et al. (22) over a range of 5 to $25^{o}C$. For both nitrification and denitrification, RBC temperature sensitivities were reported to be similar to those of suspended growth systems having long sludge retention times.

With more than 4 months of RBC nitrification study at the Belmont Wastewater Treatment Plant at Indianapolis, Indiana; Reid, Quebe, Allison, Wilcox and Associates, Inc. (23) reported that although the RBC process appeared a feasible alternative nitrification process for waste containing relatively consistent NH_3-N loadings, the process was unable to consistently maintain low (less than 1.0 mg/l) NH_3-N levels in the effluent when the influent NH_3-N load varied. In the same study, it was found that the RBC system could reduce the total BOD_5 (carbonaceous portion only) in the clarified activated sludge effluent from 8 to 18 mg/l to 6 to 13 mg/l. The percentage of BOD_5 removal was low (0 to 57 percent) compared to the secondary treatment process. However, soluble carbonaceous BOD_5 removal was more successful (1 to 10 mg/l to 1 to 3 mg/l, or 0 to 80 percent removal). By removing a portion of the treated effluent total suspended solids (TSS), an effluent with BOD_5 less than 10 mg/l can be obtained with no difficulty.

Other studies also indicate the inability of the RBC system to remove total BOD_5. Reh, et al. (24), Lagnese (25), and Sullivan, et al. (26) collected and analyzed operational data from various full-scale RBC plants and concluded that design of RBC systems should be based on soluble BOD_5 loading, rather than on total BOD_5 loading. In using the RBC system for upgrading existing secondary treatment plants and for tertiary treatment, it is important to recognize the inability to remove particulate BOD_5, particularly when the particulate portion of the total BOD_5 is high. When the RBC unit is operated in series and following secondary treatment, a less efficient performance can be expected, since the wastewater contains a higher fraction of refractory organics. Finally, nitrified effluent from the RBC unit contains nitrogeneous oxygen demand (NOD) which can be a significant portion of the effluent BOD_5. Lagnese (27) suggested that a nitrification inhibitor be used in the BOD_5 analysis to eliminate NOD from the analysis. However, this approach may require some revision or clarification of the NPDES permit. Important RBC design considerations include the characteristics of wastewater to be treated and the degree of treatment desired. These considerations dictate such system parameters as number of stages, speed of RBC rotation, reaction tank volume, media density, and pretreatment.

According to a literature search performed by Griffith, et al. (28), systems treating municipal wastewater usually provide for two to four stages for secondary treatment and up to 10 stages if further treatment is required. Disk rotation velocities of 1 fps (peripheral velocity) are common for initial stages, with lower velocities (0.5 fps) used in later stages as the oxygen demand in the wastewater is reduced. Disk reaction tank volumes which provide 0.12 gal/sq ft (4.89 l/m^2) of disk (including disk volume), or 1-hour detention time, at a hydraulic loading rate of 0.06 $m^3/day/m^2$ (1.5 gpd/sq ft) of disk area, are common. A wide range of hydraulic and organic loading rates has been reported for systems treating domestic wastewater. Hydraulic loading rates ranging from 0.004 to 0.17 $m^3/day/m^2$ (0.09 to 44.1 gpd/sq ft) of disk surface area and organic loading rates of 0.20 to 6.0 lb BOD per day/1000 sq ft (0.98 to 2.93 g/m^2) of disk surface area are documented. Systems having disks aligned parallel to the direction of flow and perpendicular to the direction of flow have been described. The disk reaction tank is generally contoured to the shape of the disks, which improves mixing of the wastewater within each stage. Documented disk materials include aluminum, polystyrene, polyethylene, and plexiglas. Desirable properties in a disk material are low density and rigid shape. Disk diameters range from 6 in. to 12 ft (15.2 cm to 365.8 cm), with spacing between disks ranging from 3/8 to 3/4 in. (0.96 cm to 1.9 cm). The disk is generally immersed in the wastewater to between 40 and 50 percent of its diameter, with the only criterion being that its entire surface becomes wet.

Final solids removal facilities are generally incorporated into the total treatment scheme. A biomass generation of approximately 0.4 lb (.16 kg) of dry solids per pound of BOD removal has been reported. Systems used to transport the settled biological solids to storage and treatment facilities include screw conveyors, scraper/bucket schemes, and pumps.

Description of Modern Process (29)

In its present form, the rotating biological contactor process consists of a series of closely spaced discs (10-12 feet in diameter) mounted on a horizontal shaft and rotated while about one half their surface area is immersed in wastewater. The media commonly used in Europe and originally introduced into the U.S. consists of a series of parallel, closely spaced flat discs. Now many U.S. manufacturers offer media with a lattice structure. This more complex structure offers more surface area per unit volume.

When the process is placed in operation, the microbes in the wastewater begin to adhere to the rotating surfaces and grow there until the entire surface area of the discs is covered with a 1/16 to 1/8 inch layer of biological slimes. As the discs rotate, they carry a film of wastewater into the air, where it trickles down the surface of the discs, absorbing oxygen. As the discs complete their rotation, this film mixes with the reservoir of wastewater, adding to the oxygen in the reservoir and mixing the treated and partially treated wastewater. As the attached microbes pass through the reservoir, they absorb other organics for breakdown. The excess growth of microbes is sheared from the discs as they move through the reservoir. These dislodged organisms are kept in suspension by the moving discs. Thus, the discs serve several purposes. They provide media for the buildup of attached microbial growth, bring the growth into contact with the wastewater, and aerate the wastewater and suspended microbial growth in the wastewater reservoir. The speed of rotation is adjustable.

The foregoing description was excerpted from a recent EPA technology transfer publication. It describes the operation of RBCs for organics removal and, with minor modifications, for nitrification. When RBCs are used for denitrification, the entire disc is submerged and rotation provides mixing but not oxygen exchange.

The rotating biological contactor process, like any other treatment technology, has inherent advantages and disadvantages of which prospective users should be aware.

Advantages

Rotating biological contactors have a number of characteristics which commend them to the design engineer. They can provide a very high degree of treatment. They require less area than most other comparable processes. They can be retrofit easily to existing plants.

RBCs show high efficiency in oxygen transfer. They handle organic overloading well due to the large biomass on the discs. Since they involve attached growth, they are much less likely to fail through washout when conditions adverse to the biological growth occur. There is no bulking, foaming, or floating of sludge to interfere with a plant's overall efficiency. Short circuiting in the biological reactor cannot occur.

In laying out a plant, RBCs offer advantages beyond their relatively low area requirements. Because most RBC units operate with a net increase in hydraulic head, pumping which would otherwise have been needed may be obviated. Less excavation is required for RBCs than for activated sludge aeration tanks, a characteristic of the process which is especially helpful in high water table areas. Finally, RBCs are versatile both in the functions which they perform and in the flexibility with which they can be configured. There is even a choice in the methods for rotating the discs. Mechanical drive units can be employed or an air drive mechanism can be used which has fewer moving parts and which uses less energy.

However the discs are rotated, RBC technology uses up to 50 percent less energy than activated sludge units. Over the lifetime of a plant, this can be a very important advantage. The low speed of the mechanical drive units reduces their maintenance requirements and prolongs their lives.

Rotating biological contactors are simple to operate. There are no sludge or effluent recycle streams, although recycle has been shown to be advantageous in some applications. The sloughed biomass settles well and can be more reliably removed than the solids from an activated sludge tank. Clarifier design and operation, which frequently limits the performance of plants relying on other processes, is far less critical in RBC installations.

Because Rotating Biological Contactor treatment is simple and stable with respect to most potentially upsetting fluctuations in influent flow and quality, it requires fewer process decisions by the operator than do activated sludge processes. Thus, satisfactory operation can be achieved with less highly skilled personnel than are needed for activated sludge. This factor could mean considerable savings in operating a treatment plant as well as making the operation of a plant more predictable. Since less than optimal operating procedures have been cited as a leading cause of plants failing to achieve the results for which they were designed, the advantages of the simple RBC process may be greater in practice than a comparison of design performances would indicate. The rotating disc process lends itself well to upgrading existing treatment facilities. Because of its modular construction, low head loss and shallow excavation, it can be installed to follow existing primary treatment plants, including Imhoff tanks and septic tanks.

Disadvantages

RBC technology is not without its share of problems. The oldest U.S. plants have been in operation for only 7 years. The structural integrity of RBC units is untested by time. In one instance, plastic media tore loose from its drive shaft. It has been a common experience for tie rods to loosen and cause uneven rotation and need for realignment. Oil leaks from drive units are common.

Although low maintenance costs are often cited as an RBC advantage, these costs are strictly proportional to plant capacity, exhibiting none of the economies of scale observed with other non-modular technologies. Similarly, area requirements for RBC installations are proportional to plant capacity so that an RBC advantage for small- and medium-scale plants becomes a liability in very large capacity applications. The use of air drives reduces the relationship between plant size and maintenance costs because each shaft does not require a separate electromechanical drive. Air drives avoid another disadvantage which has been cited for RBCs; rotational speed can be continuously adjusted by turning a few valves. Altering the rotational speed of electromechanical drives requires modifying each drive unit. A large plant may have dozens.

Enclosures are necessary where very low air and wastewater temperatures occur in order to achieve acceptable performance. Installations in warmer areas may also require enclosures for protection against wind, precipitation, and vandalism. Provision of enclosures increases an RBC installation's initial cost and is thus a disadvantage, although protected RBCs probably operate more stably, especially in winter.

When grit and primary solids removal is inadequate, suspended solids may accumulate in RBC reactors. Foul odors and falling process efficiency ensue. This is a potential RBC disadvantage which can be avoided by ensuring that the RBC's influent has had good primary treatment. When excess solids do pass through to an RBC unit, they can be periodically pumped out of the reactor.

While the RBC process is a relatively stable one, RBC operation can be disrupted by many of the same influent fluctuations which upset other processes. Organic and hydraulic shock loadings are handled comparatively well by RBCs, but some loss of process efficiency will occur. Toxic substances in the influent may cause a sometimes catastrophic loss of biomass from the discs. Process efficiency will fall. Recovery, however, is usually more rapid than that of trickling filters which have been similarly insulted. Extremes of wastewater pH have an adverse effect upon RBC system performance. This, of course, is a disadvantage common to all biological treatment processes.

It is common for organisms to develop on RBC media which are whitish in color. This white biomass, which is probably composed of Thiotrix or Beggiatoa, is of little concern when it appears in small patches. As these patches expand to cover a significant proportion of the discs, however, process efficiency falls. The white biomass phenomenon is associated with septic influents containing high concentrations of hydrogen sulfide. It can be prevented or cured by preaeration of the wastewater or by the addition of oxidizing materials such as hydrogen peroxide to the water.

Overloading of the first stage of an RBC installation can cause odors to develop and less than adequate removals to occur. Where this problem is observed or anticipated, extra surface area can be provided in the first stage, alleviating the overload conditions. When the overloads are episodic, equalization upstream from the RBC reactor can be as useful with this technology as with others.

Some disadvantages have been charged to the RBC process which will probably disappear as the technology matures. Extensive and intensive controversy exists regarding design criteria, optimum rotational speeds, matrix design, media configuration, recirculation requirements, surface-to-volume ratio for the reaction chambers, and appropriate scale-up procedures. Compared to many other modeling efforts, RBC modeling is in its infancy. Further operational experience, additional research, and symposia such as this one can be expected to remedy these shortcomings.

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EPA RESEARCH PROGRAM FOR RBC

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The EPA municipal wastewater research effort on Rotating Biological Contactors (RBC's) was initiated in 1967. A contract was awarded to Allis Chalmers to test a bench-scale unit in their laboratory. This contract was extended to conduct an evaluation using the laboratory RBC unit on domestic wastewater in Milwaukee, Wisconsin. This was then followed by a contract to pilot test the RBC's at Pewaukee, Wisconsin, in 1969, and it was during this period that Allis Chalmers sold their RBC interests to Autotrol Corporation. EPA supported these series of studies with approximately $360,000 in contract awards between 1967 and 1969. The Pewaukee pilot plant contract amounted to $33,000 and Autotrol continued the pilot study with corporate resources for an additional six months beyond the EPA project completion date.

The early research effort progressed from the laboratory scale in Milwaukee, through the pilot scale in Pewaukee to a full-scale demonstration in Pewaukee, between 1967 and 1971. These concentrated research and development initiatives accelerated the introduction of RBC's as an alternative secondary treatment process.

While EPA was supporting the research and development work in Wisconsin, a concurrent grant was awarded to Rutgers University to conduct a literature search. This was then followed with a research grant to study a bench-scale RBC unit.

These initiatives into laboratory and pilot plant scale evaluations were instrumental in establishing the feasibility of RBC's, and the next step was a demonstration with full-scale equipment at Pewaukee, Wisconsin in 1971. The Village of Pewaukee, along with Autotrol Corporation, conducted a comparative evaluation of the RBC process with an existing trickling filter plant and concluded that RBC's were an effective treatment process. Additional work was done on phosphorus removal by mineral addition. The EPA share of the project cost was approximately $400,000.

In 1972, EPA granted the West Virginia University an award of $16,000 to evaluate an RBC unit located at a summer camp, Camp Horseshoe in West Virginia.

Following these projects which were directed towards speeding the introduction of new technology, the EPA research program was decelerated. The private sector came to the forefront and took up the research slack and introduced novel equipment and process innovations to improve the cost effectiveness of the process. The number of manufacturers supplying RBC's grew and with this increase came further improvements spurred by competition for sales.

The EPA research program for RBC's became selective. In the mid 1970's grants were awarded for specific or unique situations. The University of Michigan studied nitrification with RBC's in a pilot facility at Saline, Michigan. The City of Edgewater, New Jersey, was awarded a grant to evaluate a novel application in which a primary clarifier was converted to a secondary system by installing RBC's above a false floor.

The entire RBC research effort was conducted via extramural grants, contracts, and cooperative agreements with universities and municipalities. This approach enabled EPA to handle a diverse program with a minimum of personnel and to perform these investigations within a minimal time frame.

The rapid growth in the number of treatment plants employing RBC's by the late 1970's has caused a re-evaluation of the research program. The present technology used for designing RBC facilities is being questioned. Peripheral speed as a scale-up factor is being questioned. DO sags in the initial stage require corrective action. The EPA research program had earlier taken the position that development efforts that lead to equipment modifications and result in performance advantages should remain in the domain of the suppliers. However, the technical questions that are being raised effect performance and capital costs. Both of these items impact the EPA Construction Grants Program and answers are required to improve the cost effectiveness of the process.

The RBC research program has again turned around to address these questions and a new program was developed to provide answers to these questions concerning RBC's.

The symposium on RBC's at Seven Springs is geared towards producing a state-of-the-art on the latest technology. There are 70 plus papers that will be presented during this symposium that will cover practically all aspects on RBC technology. In addition, there is a Research Needs Workshop on Tuesday night that is aimed at defining technical gaps in the present process so that solutions can be prescribed and evaluated for bridging these gaps and improve the overall effectiveness of the process.

The second phase of the EPA research program was an award of a contract to Roy F. Weston Consulting Engineers. They will assess the present design practices and evaluate the applicability of the various parameters to adequately predict the performance of RBC's. Secondly, they will study several RBC operating plants to determine if performance is within the design specifications. This phase of the program is well underway and Warren Chesner will bring us up-to-date on its progress later on this morning.

The third phase of the EPA research program is a cooperative agreement under evaluation with the City of Columbus, Indiana, to study the questions that are being raised on RBC's. The City of Columbus has 10 lines of RBC's. Each line consists of 8 shafts. At the present time only seven lines are used. It is our intention to modify two of these lines to evaluate various RBC design parameters with full-scale equipment. This would allow a direct comparison between the two lines. One of the major issues confronting the designers of RBC facilities concerns the application of hydraulic or organic loadings as the preferred basis for specifying the surface area requirements on RBC's. To evaluate this function, the following ground rules were established for conducting the test at Columbus.

1. A comparative evaluation between organic and hydraulic loadings would be performed.

2. The flows to both systems would be controlled.

3. The influent to both systems would be identical.

4. Diurnal variation would be incorporated into the flow control system with a maximum to average ratio of 1.5 and a minimum to average ratio of 0.7. The system will also be capable of operating between 50 and 200% of design flow to stress the RBC treatment trains.

5. The RBC's are plant scale facilities. That is, the use of pilot scale, or more specifically, less than 10 foot diameter disks were forbidden to avoid controversy over the peripheral speed scale-up parameter.

6. DO's would be continuously recorded at critical locations on each system.

7. Chemical characterization would be conducted on the influent, effluent and at the various stages.

Following this evaluation, the preferred mode of operation, organic or hydraulic, should be established. In addition, a secondary objective is to identify and improve the limiting factor governing RBC performance. If the data on DO identified this parameter as a contributing factor for limiting RBC performance, then provisions will be made to modify the process to improve DO levels. Several methods will be tried. Among them are:

1. Increase rotating speed at critical locations.

2. Force feed air at critical locations.

3. Evaluate the effect of lower immergence which increases the air contact time.

This project is scheduled to start during May, 1980, and its estimated completion date is October, 1982. The approach should provide definitive answers to many of today's questions and should advance the technology and performance of RBC's.

ASCE WATER POLLUTION MANAGEMENT TASK COMMITTEE REPORT ON "ROTATION BIOLOGICAL CONTACTOR FOR SECONDARY TREATMENT"

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Introduction

ASCE Environmental Engineering Division, Water Pollution Management Committee (WPMC) established a task committee in October 1977 to write a State-of-the-Art report on Rotating Biological Contactor for Secondary Wastewater Treatment. The task committee has completed a second draft of the report which is under review by the committee members. It is expected that after final reviews, this report will be published later this year. Selected portions of the second draft of the report are presented here.

Currently, the most common secondary biological treatment methods include the trickling filter process and its modifications, air or pure oxygen activated sludge process and its modifications, and rotating biological contactor (RBC) process. This paper will briefly summarize the present knowledge on the design, application and selection of the RBC process for municipal wastewater treatment.

Process Description

The (RBC) process is an aerobic, continuous flow, wastewater treatment system designed for municipal and many industrial wastewaters. The RBC process converts the influent soluble biodegradable organic wastewater constituents into biomass and off-gases. Biomass generated by RBC units is separated from the wastewater carrier stream in a sequential secondary clarifier. Settled wastewater is first introduced into a tank containing a series of high density polyethylene discs (media) attached to a horizontal shaft (Figure 1). In U.S. practice, the discs are mounted in the tank so that

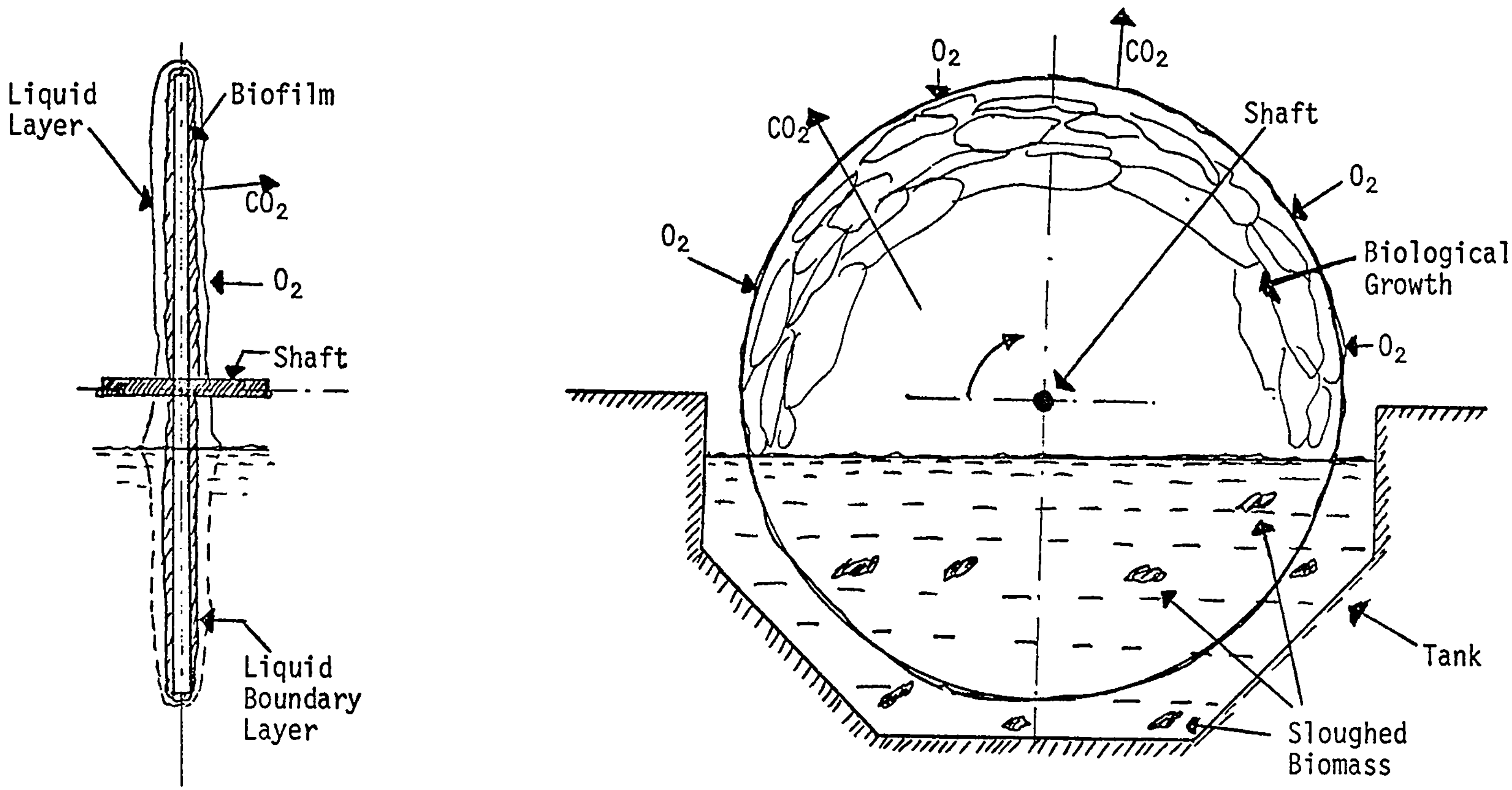

Figure 1. Rotating Biological Contactor Disc

about forty percent of the media area is submerged in the wastewater. The film of biomass growing on the media is responsible for the removal and conversion of the biodegradable organic wastewater constituents. The media is rotated continuously by mechanical or air drive systems so that the biomass film is alternately exposed to fresh wastewater in the tank and the air above the tank. For some specially designed RBC systems, additional treatment results from the development of unattached, suspended biomass culture in the mixed liquor of the tank.

To avoid short circuiting in the tank, groups of discs are segregated by baffling into stages. For small installations, the flow path is usually perpendicular to the disc faces. Larger installations are usually designed with a single shaft or a series of shafts constituting each stage with wastewater flow parallel to the disc faces. The microbial population in each stage can vary significantly depending on wastewater loading conditions. Heavy growth and substrate removal usually occur in the first stage followed by both decreasing growth and carbonaceous removal in succeeding stages. Where nitrification of wastewater is desired, the latter stages can be constructed with more media surface area per shaft length since biomass production is reduced and bridging between adjacent discs is less likely to occur.

As a result of continuous rotation, the media carries a film of wastewater into the air where oxygen is transferred through the liquid film surface. Both oxygen and organic substrate materials diffuse through the liquid film into the growing biomass film where they are consumed for growth and respiration purposes. Excess dissolved oxygen in the wastewater film is mixed with the contents of the bulk liquor in the tank and results in aeration of the wastewater carrier stream.

Shearing forces exerted on the growing biomass film result in excess biomass being periodically sloughed from the media into the wastewater carrier stream. This sloughing action prevents bridging and clogging between adjacent discs. The disc mixing action keeps sloughed biomass solids in suspension until they are removed from the RBC tank and separation occurs in the final clarifier. In essence, the rotating media is used to both provide a support surface for microorganism growth and to assure an opportunity for contact between the microorganisms, the substrate and oxygen.

Figure 2 shows a flow diagram for a typical RBC treatment plant. Floatable and settleable solids in the wastewater are first removed by primary treatment. The primary effluent then flows to the multi-stage RBC unit where biological removal of organic material occurs. Each RBC stage tends to operate as a completely mixed, fixed film, biological reactor. Treated wastewater and sloughed biomass flows from stage to stage with progressively increasing substrate removal occuring. Sloughed biomass is separated from the carrier stream in the final clarifier and the underflow solids are disposed of by conventional means. Operation of the process is on a once through basis with no need for effluent recycling.

The RBC process differs from the trickling filter by having a significantly longer retention time (8 to 10 times that of a trickling filter) and a dynamic, rather than stationary media; and from the activated sludge process, by having an attached rather than a suspended biomass and not dependent on suspended culture separation and recycle. In the former case, higher levels of treatment are achieved by the RBC process and in the latter, the RBC process has less susceptibility to upset from changes in hydraulic or organic loading in either the reactor or clarifier.

Process Applicability

The RBC process may be used to remove a major protion of the biochemical oxygen demand (BOD), and ammonia-nitrogen (NH_3-N) from any biodegradable wastewater. The process is applied to treat domestic sewage in plants ranging from small package facilities to large municipal sewage treatment plants. Also, wastewater from dairies, bakeries, meat and poultry processors, pulp and paper mills, animal feed lots, distilleries, canneries, refineries and other biodegradable industrial wastewaters can be treated by the process.

Process Hardware

The process hardware consists of closely packed circular plastic media mounted on a shaft. The shaft is supported on bearings and connected through a gear box to an electric motor. The plastic media consists of corrugated polyethylene material. In one instance, the media consists of a drum filled with 38 mm plastic balls (Bio-Drum Process). The shaft rotates at 2-4 rpm inside a concrete or steel tank. The shaft length varies from 5 to 20 ft. depending upon the size of the unit. The diameter of the packed media on the shaft varies from 4 ft to 12 ft. depending on the capacity of the unit. For higher degrees of treatment and larger flow capacities several modular units may be placed in parallel or in series depending upon the configuration desired as shown in Figure 2. The RBC media is about 40% submerged in a trapezoidal, semi-circular or rectangular tank, with intermediate partitions in some situations. To maintain performance under cold weather conditions, the modular RBC units are provided with fiberglas enclosure with access doors & ventilation. Alternatively, the RBC units can be housed in a conventional insulated structure that covers a whole battery of units.

In a recent development, air is introduced to aid in rotating the media in the tank. Figure 3 shows an air drive RBC system. In this process, plastic cups are welded onto the periphery of the media over the entire length of the contactor. A small air header placed in the tank underneath the media allows air to be released along the tank length. The released air is captured in the plastic cups causing buoyant forces to rotate the shaft. Radial passages in the media periphery cause a portion of the released air to flow upwards into the corrugated media sections. The supplementary aeration and increased turbulence achieved from this is sufficient to allow a reduction in rotational velocity of the media while still achieving the same degree of treatment. The air drive process requires about 25% less units for a given application compared to mechanical drive systems.

Historical Background

The RBC systems as presently used evolved from the research work of Pöpel and Hartman in West Germany in 1955 (1). However, earlier researchers in the USA had developed similar devices. Buswell, in 1929, developed a unit called "Biological Wheel", which was similar to the present RBC units and whose purification capacity was thought to be based on biological principles. Later in 1931, pursuing this line of thought, Maltby patented a process that was based

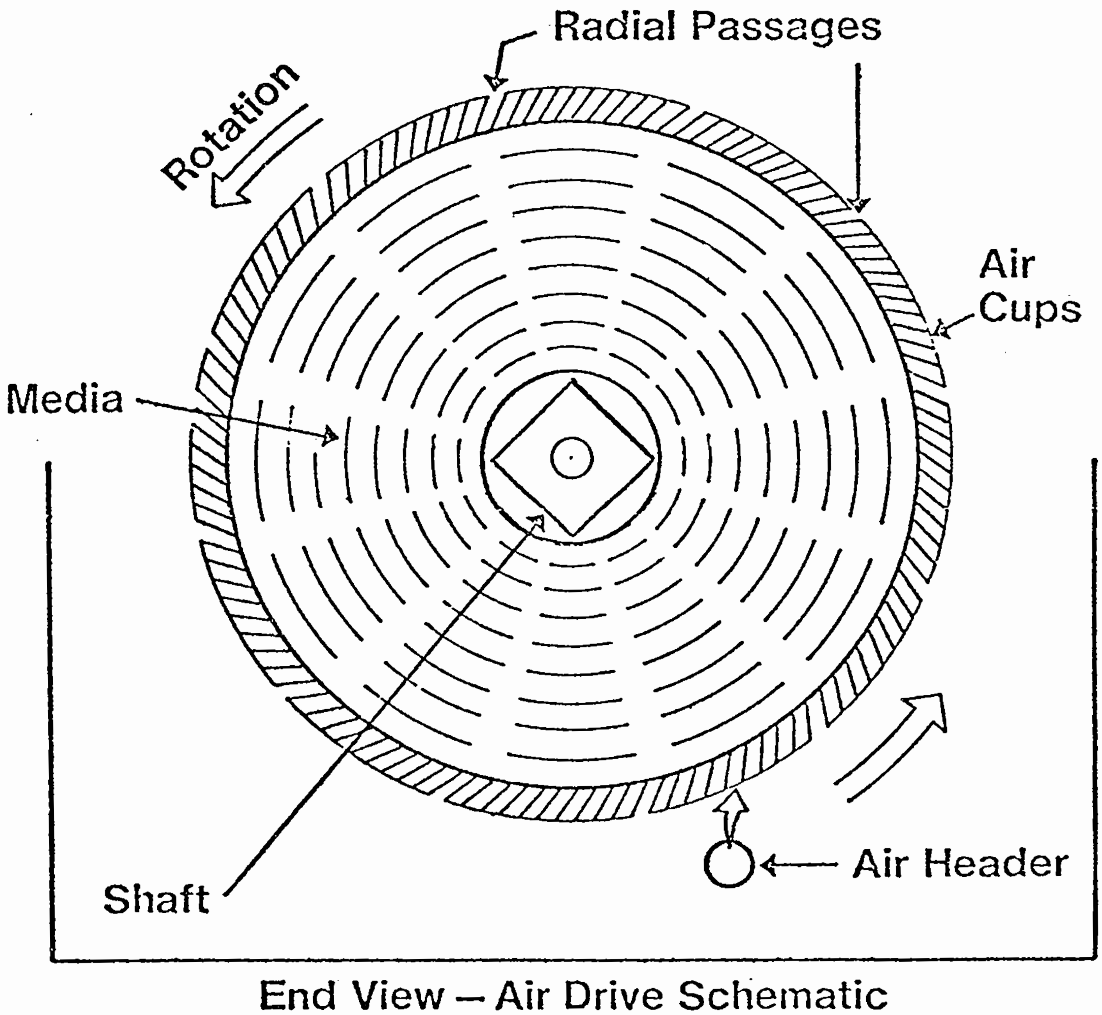

Figure 3: Air Drive RBC Unit

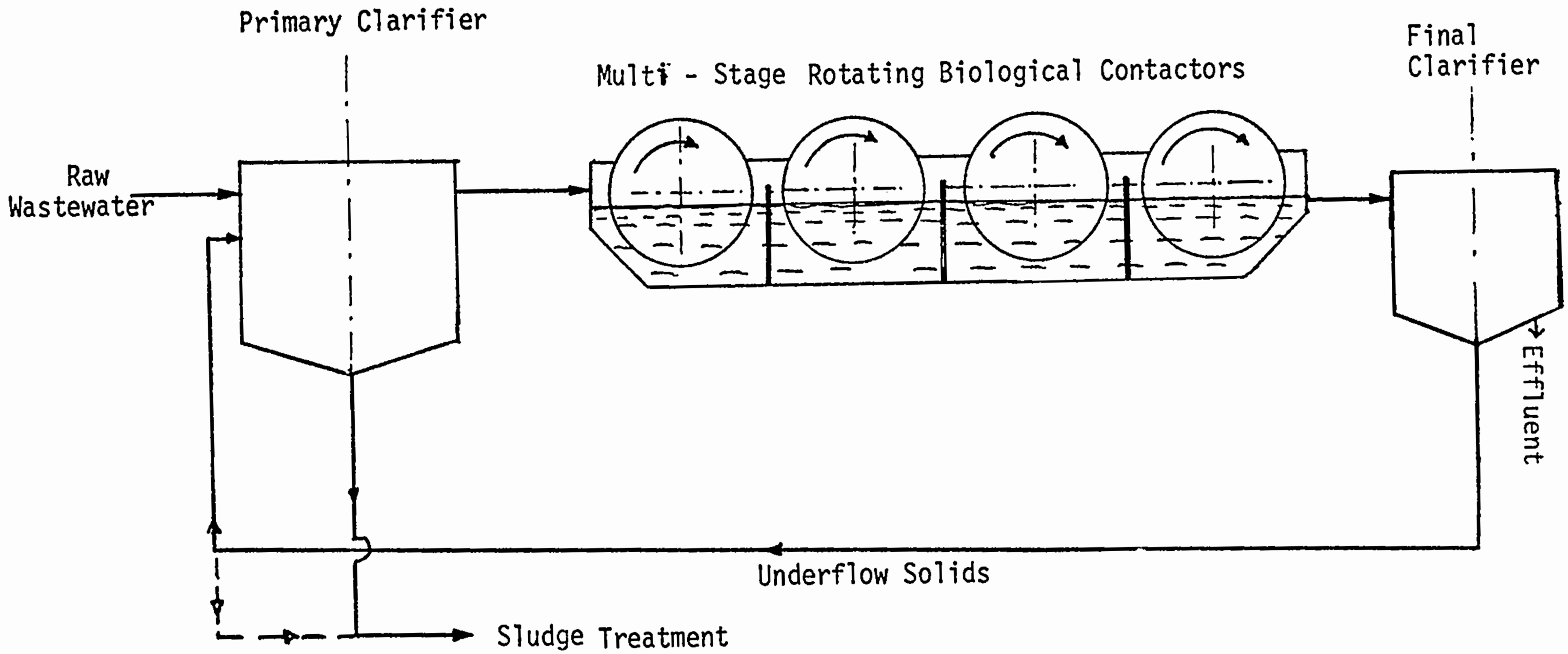

Figure 2. Flow Diagram for Rotating Biological Contactor Process

on the biological wheel principles (2). Hartmann (1) credits Travis for the idea behind the RBC process. Travis in 1901 installed wooden strips in "Hydrolytic Tanks", (settling tanks) that were to catch cloudy non-settling solids from the wastewater by adsorption. These strips accumulated solids on their surfaces & eventually these solids would slough off the strips to the settling tank hopper. The development of the contact aeration by Hays in USA (3) and others in Europe (4), was a logical improvement of Travis's idea. The application of air below the wooden slots was to retain sludge flocs in the aeration tank and improve effluent quality. However, these ideas were not integrated to produce the present RBC process until 1960 when Popel & Hartmann developed their immersion drip-filter (trauchtropfkörpern) (1). The first commerical RBC was installed in 1960 in West Germany & soon after it was widely applied throughout Europe (5). In U.S. Allis-Chalmers Company began development work in mid 1960's and presently there are several companies offering these systems for commerical applications.

PROCESS DEVELOPMENT

Operational Characteristics

The RBC systems employed for secondary waste treatment study usually involve 2-10 stages in small-scale laboratory units or 2-6 stage in full-scale pilot plants. Due to the change in physical and chemical properties of wastewater to be treated in each stage, the biochemical nature of micro-floral populations as well as metabolic end products varies significantly. Figure 4 shows the distribution of biochemical oxygen demand (BOD), chemical oxygen demand (COD), ammonia nitrogen (NH_3-N), nitrite (NO_2) and nitrate (NO_3) dissolved oxygen, suspended solids, dry weight of biomass, and pH in multi-stage rotating biological contactor systems (6,7,8).

Apparently, the organic carbon in both high-strength industrial waste and normal-strength domestic waste can be effectively removed by the RBC system, but the degree of BOD and COD removal is highly dependent upon the rate of hydraulic loading applied to the system and the number of stages employed. Observations from Figure 4 reveal that the majority of the removal of biodegradable organic matter is achieved within first six successive stages. It was also found, however, that the rapid uptake of carbonaceous matter occurred in the first three stages. It is believed that improved removal achieved by successive stages is due to improved residence time distribution obtained by staging and the development of a biomass population in each stage that has adapted to treat the specific waste characteristics found in that stage.

Study of nitrogen transformation as seen in Figure 4 indicates that the conversion of NH_3-N to nitrate and nitrate does not take place until stage 5 where the BOD concentration has been reduced to about 20 mg/l. Before Stage 5, the concentration of NH_3-N increases slightly over influent because of the hydrolysis of organic nitrogen in the biological growth. Figure 4 also shows that the minimum level of dissolved oxygen occurs at the first and second stages and, thereafter, the content of dissolved oxygen in wastewater increases in each successive stage. This result explains why the microbial activities as expressed by oxygen utilization, change in each stage of the RBC unit. Rapid oxygen consumption results in a low concentration of dissolved oxygen in Stages 1 and 2. This suggests that the most active cell growth apparently takes place within these stages, where the wastewater is initially contacted

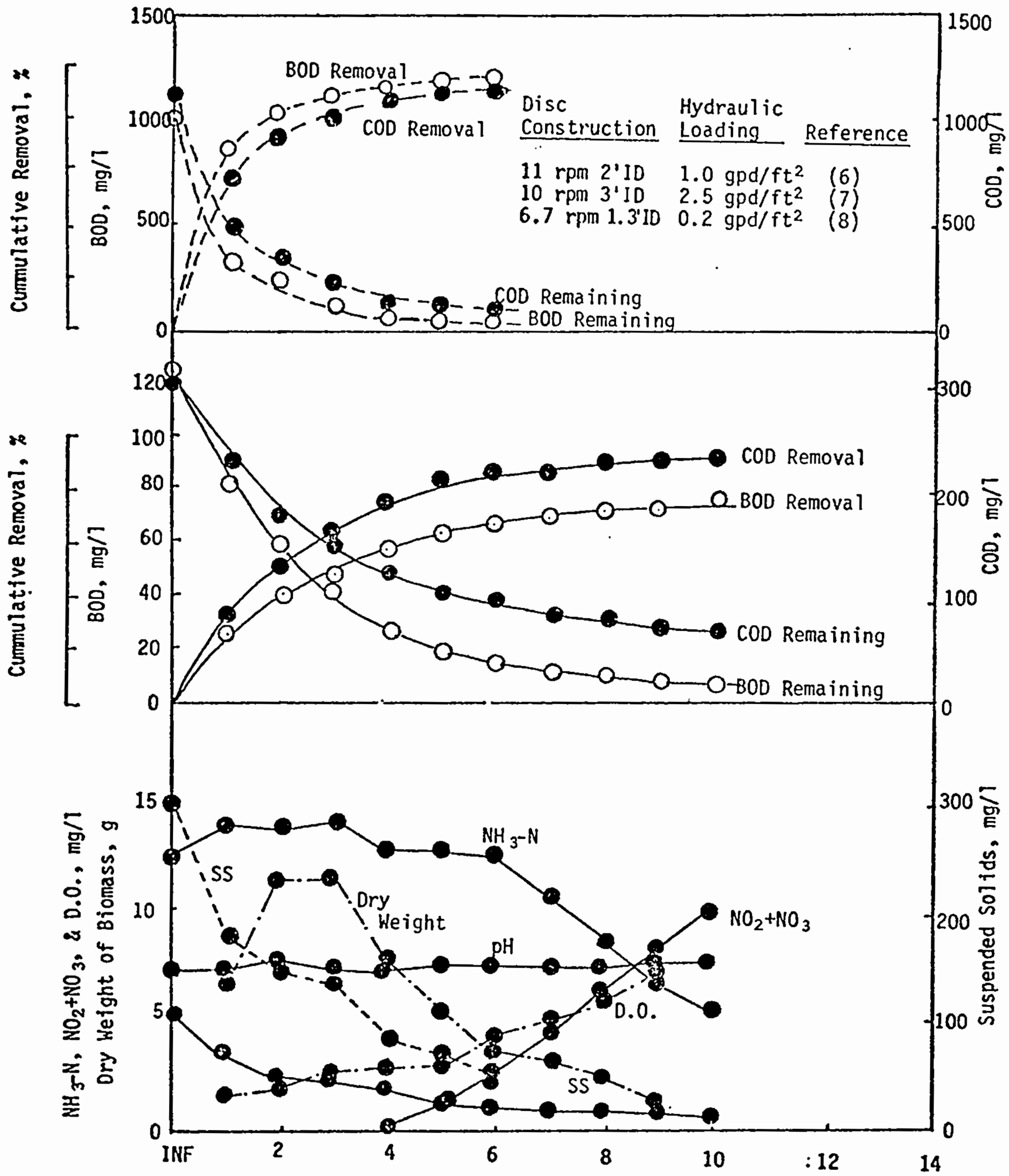

Figure 4. Effects of Stage Number on RBC System Performance

with biomass already developed on the disc surfaces. Studies of steady-state disc biofilm thickness at various stages by Pretorius (8), Pescod, et.al (9), and Sack, et.al (10) have confirmed the above statement.

The development of the types of microorganisms on the media is certainly dependent upon the nutrients in the wastewater entering each individual stage. Torpey, et. al (7,11), Pescod, et. al (9), Sack, et. al (10), and Antonie, et. al (12) conducted examinations to determine biological solids characteristics on the media under various operational conditions in treating domestic and industrial wastes. Sack, et. al (10) found that in a four-stage RBC system, the overall appearance of sludge organisms ranged from a black stringly growth with white gelatinous patches on first and second stages, to a greenish-brown slime on third and fourth stages. However, the general findings based upon these studies reported by the above investigators are:

> "The predominant organisims including Sphaerotilus and zoogleal bacteria are present on all discs. Besides these two important kinds, the diversity and abundance of free-swimming protozoa (Paramecium, Cyclidium, Ocomonas, Oxytrichia, and Euglena) are present in the first few stages. The growth of rotifers (Epiphanes and Proales), stalked ciliates (Vorticella), nematodes (Ethmolaimus), and a loop forming fungus (Athrobotrys) together with algae (Coelastrum, Chlorella, Fragilaria and Pinnularia) take place in the last few stages only when organic loading is low but high enough to support microbial growth. The quickly developed biofilm at the earlier stages of the RBC system is much thicker than bacterial slime produced on the later discs.
>
> "The mechanisms of attached growth in a RBC treatment system is described as the filamentous organisms (Sphaerotilus, Geotrichum, Bacillus) actually serving a sort of skeltal system on which other microorganisms are able to attach. The thickness of biofilm is substantially reduced in each stage as a result of significant reduction in filamentous populations, and that is caused by the marked change of carbonenergy level in wastewater after passing it through each stage. Both Pseudomonas denitrificans and Beggiatoa alba are also present in the RBC system indicating that there are the involvements of nitrogen and sulphur transfers inherent in the systems.

Operational Parameters

The major factors controlling the RBC system operation and performance are known to be:

1. Influent wastewater substrate concentration
2. Residence time of wastewater (or surface hydraulic loading)
3. Wastewater temperature
4. Media rotational speed.

In addition to the above control parameters, the effects of disc immersion depth and disc surface area configuration and density on the treatment efficiency of the RBC system may also be significant. However, these two parameters have been standardized for the purpose of optimizing the process design and operation. The current practice with regard to the immersion depth requirement is to ensure that 40% of the total disc surface area is submerged in the wastewater in the biological reactor. The total effective disc surface area for a fullscale treatment plant is determined for disc diameters commonly in the range of 10-12 ft., although the treatability study is often carried out by a relatively small pilot plant.

Effect of Influent Wastewater Concentration

The influence of the initial wastewater concentration on the removal of BOD and NH_3-N at various hydraulic loading rates is illustrated in Figures 5 and 6. It is apparent from Figure 5 that a linear relationship between % BOD removal and hydraulic loading is found when treating both industrial and domestic wastes. However, the rate of BOD removal is entirely dependent upon the initial concentration of BOD in the wastewater. At a specific hydraulic loading, the BOD removal for a domestic waste increases as the initial concentration of BOD increases. On the other hand, a decrease in the BOD removal is observed with increasing initial BOD concentration when treating low-biodegradable industrial wastes.

The removal of Ammonia nitrogen under different applied hydraulic loading rates is also affected by the initial wastewater characteristics. Figure 6 shows that in the range of 9.5-36.0 mg/l NH_3-N, the rate of ammonia nitrogen oxidation decreases as the initial concentration of ammonia nitrogen increases. In addition, under the high initial NH_3-N concentration, the fraction of ammonia nitrogen remaining increases significantly with an increase in hydraulic loading.

Effect of Residence Time of Wastewater (or Surface Hydraulic Loading)

The influence of varying residence time of wastewater on the efficiency of BOD and ammonia nitrogen removal is shown in Figure 7. It is evident that the system performance is closely associated with the liquid process retention time or the residence time of wastewater. At residence times less than 100 minutes, the removal of BOD and NH_3-N always decreases as the flow rate increases or the residence time decreases. The substrate removal does not increase significantly much beyond a residence time of 100 minutes.

As indicated earlier, the stability as well as the efficiency of a RBC waste treatment system is highly dependent upon the surface hydraulic loading rate. The parameter is normally expressed as flow per unit time per unit surface area (gpd/ft^2) covered by biological growth and is inversely related to residence time. Many researchers have reported that increasing disc surface area or decreasing surface hydraulic loading increases substrate utilization (24,25,27). The increase in substrate utilization is mainly attributed to the longer wastewater residence time and the fact that the amount of active biomass on the disc surface relative to the substrate loading has increased (lower F/M ratio).

The effect of hydraulic loading on the removal of BOD and ammonia nitrogen is shown in Figures 8 and 9. The studies of Antonie, et. al (15) and Tucker (21) indicate that the RBC process is approximately first order with respect to BOD and ammonia nitrogen removal, that is the rate of bio-oxidation reactior is proportional to the amount of oxizidable organic matter or inorganic nitrogen remaining. However, it is generally observed as in Figures 8 and 9, that the efficiency of substrate assimilation is reduced as the surface

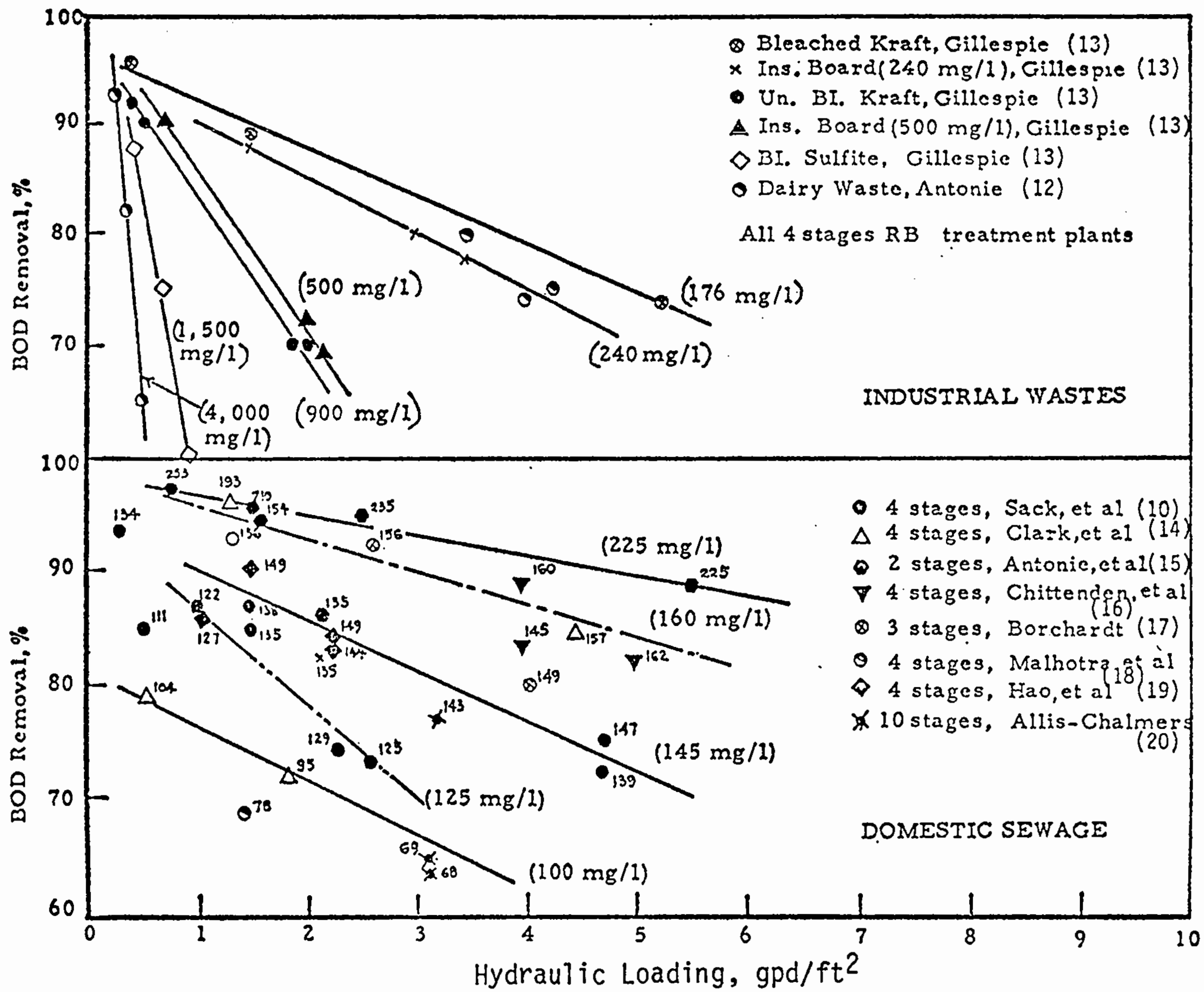

Figure 5. - Effect of Initial BOD Concentration on BOD Reduction at Various Hydarulic Loadings - (Initial BOD Concentration in Parentheis)

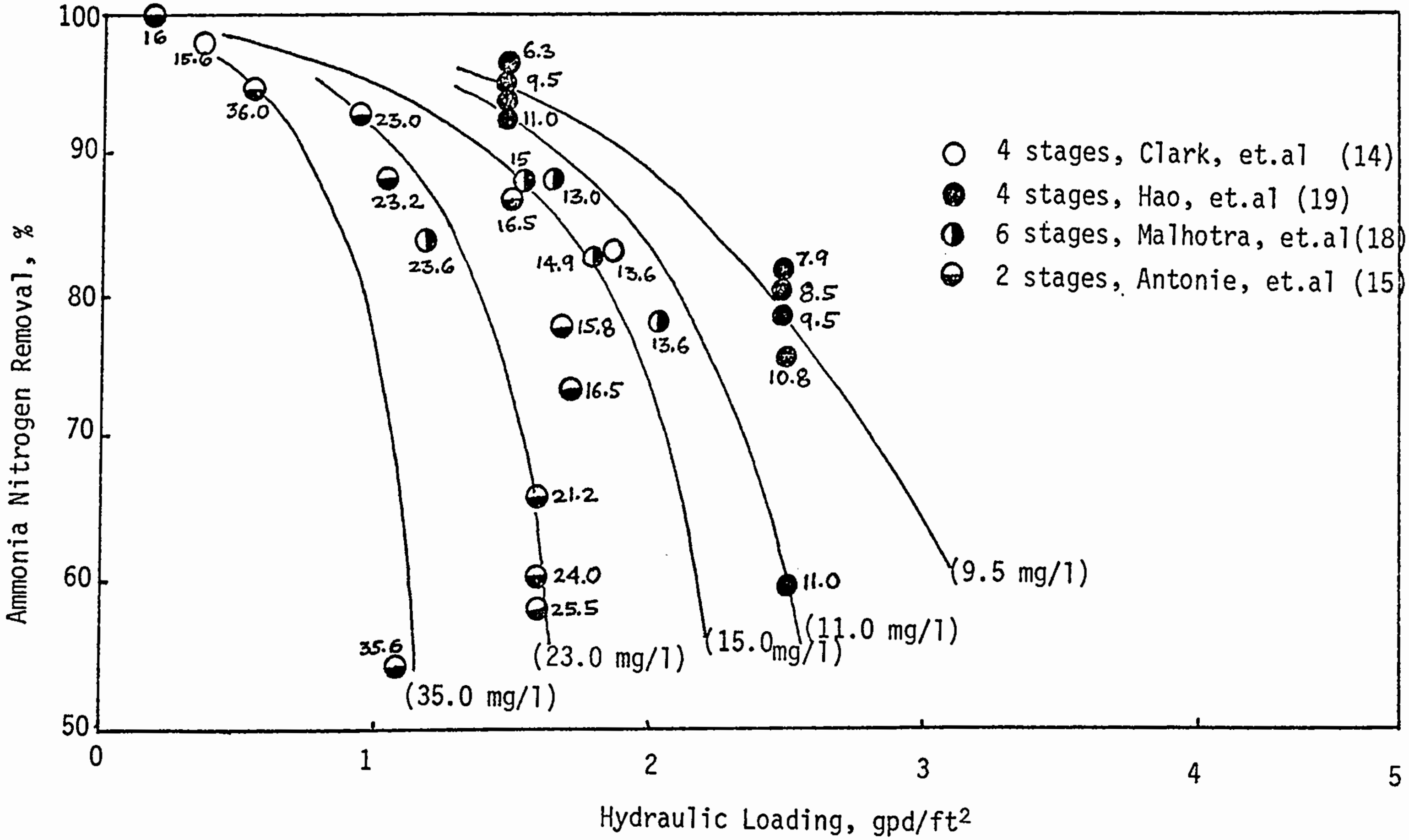

Figure 6. Effects Of Initial Ammonia Nitrogen Concentration on Ammonia Nitrogen Removal At Various Hydraulic Loadings (Initial NH_3-N Concentration in Parenthesis)

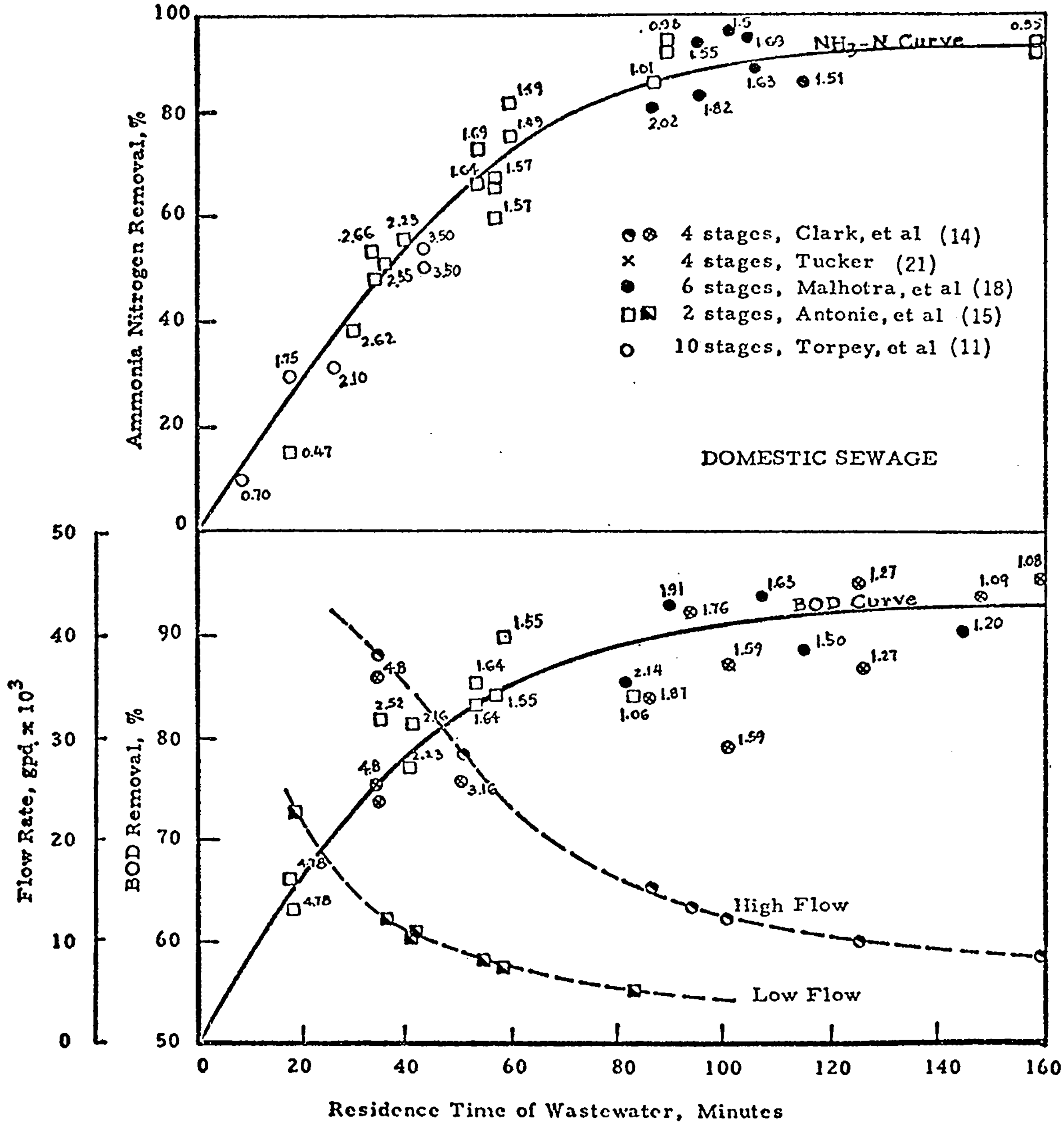

Figure 7. Effect of Flow Rate & Residence Time on BOD and Ammonia-N Removal

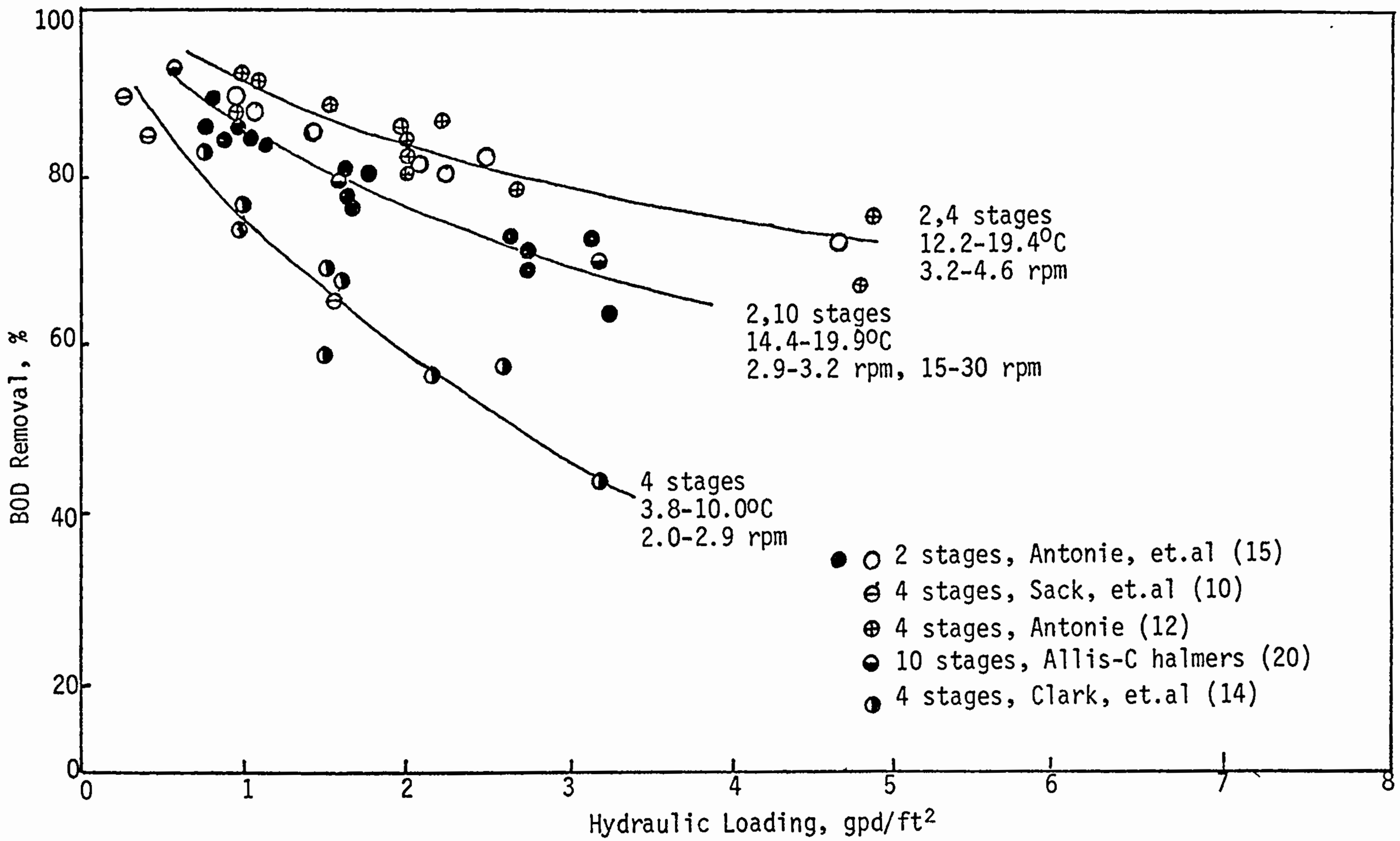

Figure 8. Effect of Hydraulic Loading on BOD Removal

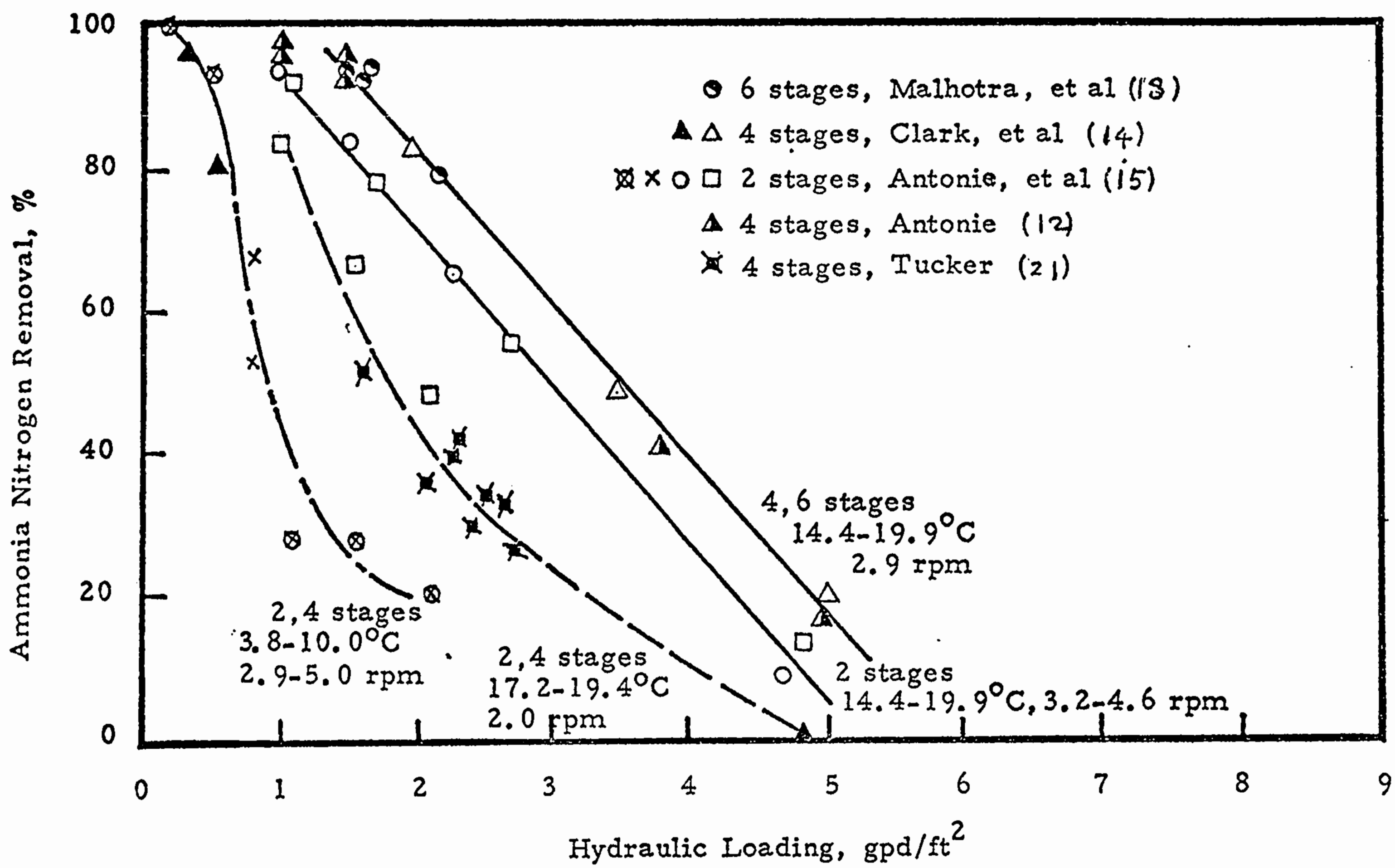

Figure 9. Effect of Hydraulic Loading on Ammonia Nitrogen Removal

hydraulic loading rate increases. Additionally, it is found that the influence of hydraulic loading on substrate removal is also dependent upon other parameters such as temperature, disc rotating speed, and number of stages.

Effect of Wastewater Temperature

The results of eight earlier investigations as shown in Figure 10 and 11 demonstrated that increasing temperature increases the rate of substrate utilization. However, the magnitude of the temperature effect on the removal of carbonaceous and nitrogen compounds is actually determined by the applied hydraulic loading rate as indicated earlier.

Figure 10 shows that at the wastewater temperatures of 55°F or above, the percentage BOD removal increases with temperature is not significant. However, at temperatures greater than 55°F, the hydraulic loading changes cause significant % BOD removal changes. As the hydraulic loading rate increases, the inhibition caused by the low wastewater temperature becomes more appreciable. A similar effect on ammonia removal efficiency by nitrification with temperature is shown in Figure 11. It seems that biological nitrification exhibits a greater sensitivity to temperature and hydraulic loading rate. In general, as the hydraulic loading rate is controlled at less than 1.0 gallon per day per square foot, the percentage of ammonia nitrogen removal is not greatly influenced by the wastewater temperature unless it reaches below 55°F. However, a significant decrease in the % ammonia nitrogen removal results from a temperature drop after the applied hydraulic loading rate is in excess of 1.0 gallon per day per square foot.

The temperature correction factor which is used for the conversion of treatment efficiency at any temperature to standard temperature of 20°C or 68°C has been studied by Antonie (22) and Weng (24) in both pilot-scale and full-scale operations. The results indicate that the temperature correction factor varies in each stage of a RBC system and also is related to the degree of temperature fluctuation. It is generally found that the correction factor θ (in the equation $K_T = K_{68}\ \theta^{T-68}$, where T is the temperature in °F) is approximately equal to 1.017 in the first few stages having a temperature higher than 55°F. The temperature correction factor is reduced from 1.017 to 0.645 as a result of temperature drop from 55°F to 40°F. According to Weng (24), a lower correction factor (<1.017) is commonly observed in the last few stages even at wastewater temperature higher than 55°F.

Effect of Media Rotational Speed

Disc rotation affects wastewater treatment in several ways. It provides contact between the biomass and the wastewater, it shears excess biomass, it aerates the wastewater, and it provides the necessary mixing velocity in each stage. Increasing the rotational velocity increases the effect of these factors. However, there is an optimum rotational velocity, above which treatment levels are no longer increased. This optimum velocity will vary with wastewater conditions, i.e., the optimum velocity is higher for more concentrated industrial wastes and lower for more dilute domestic wastes.

The effect of rotating disc speed on the performance of the RBC system is shown in Figure 12. It is apparent from Figure 12 that the effect of rotating disc speed on BOD or COD removal cannot be well defined unless the process operation is classified in accordance with disc size. For a 2-4 stage treat-

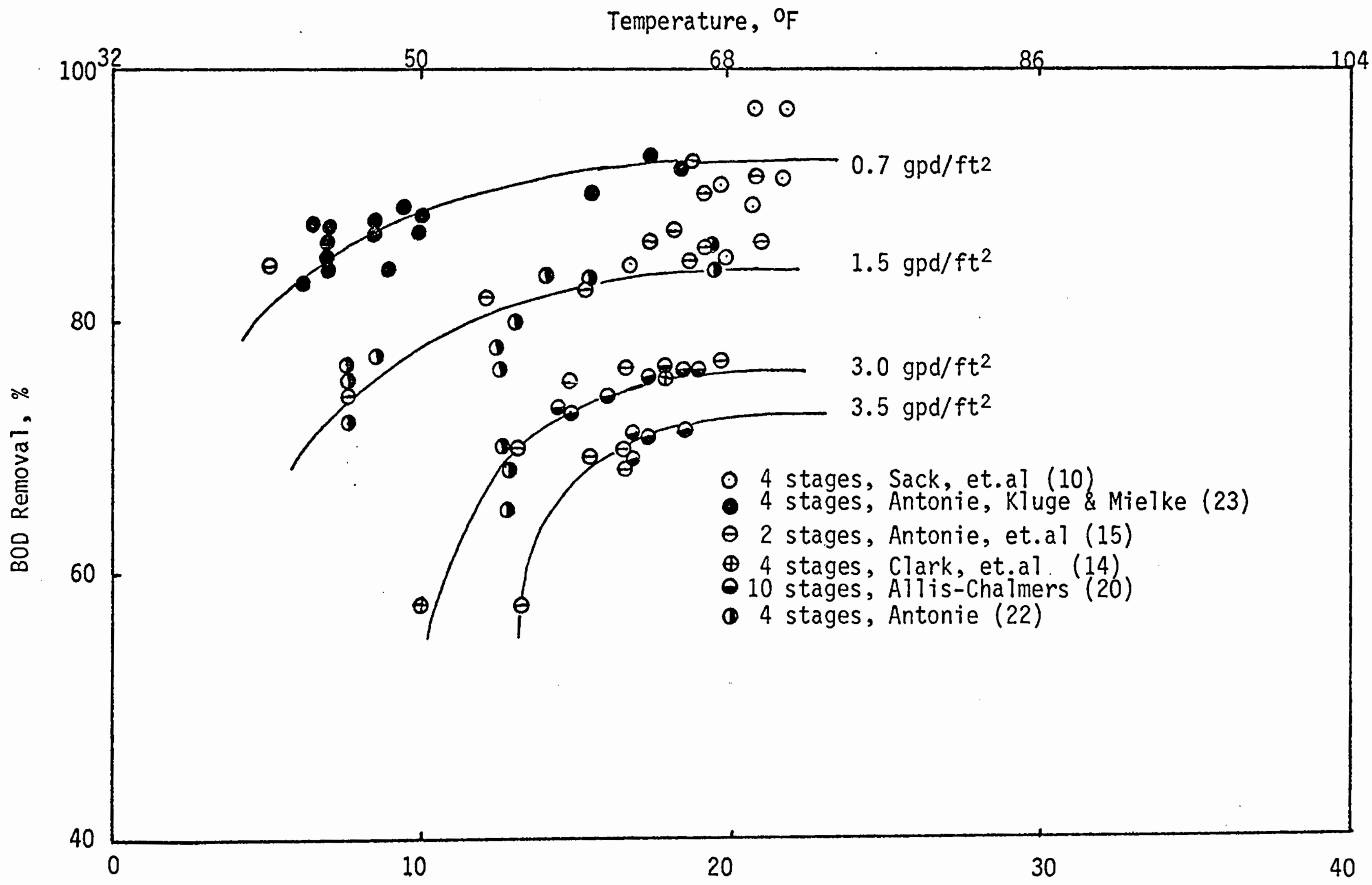

Figure 10. Effect of Temperature on BOD Removal

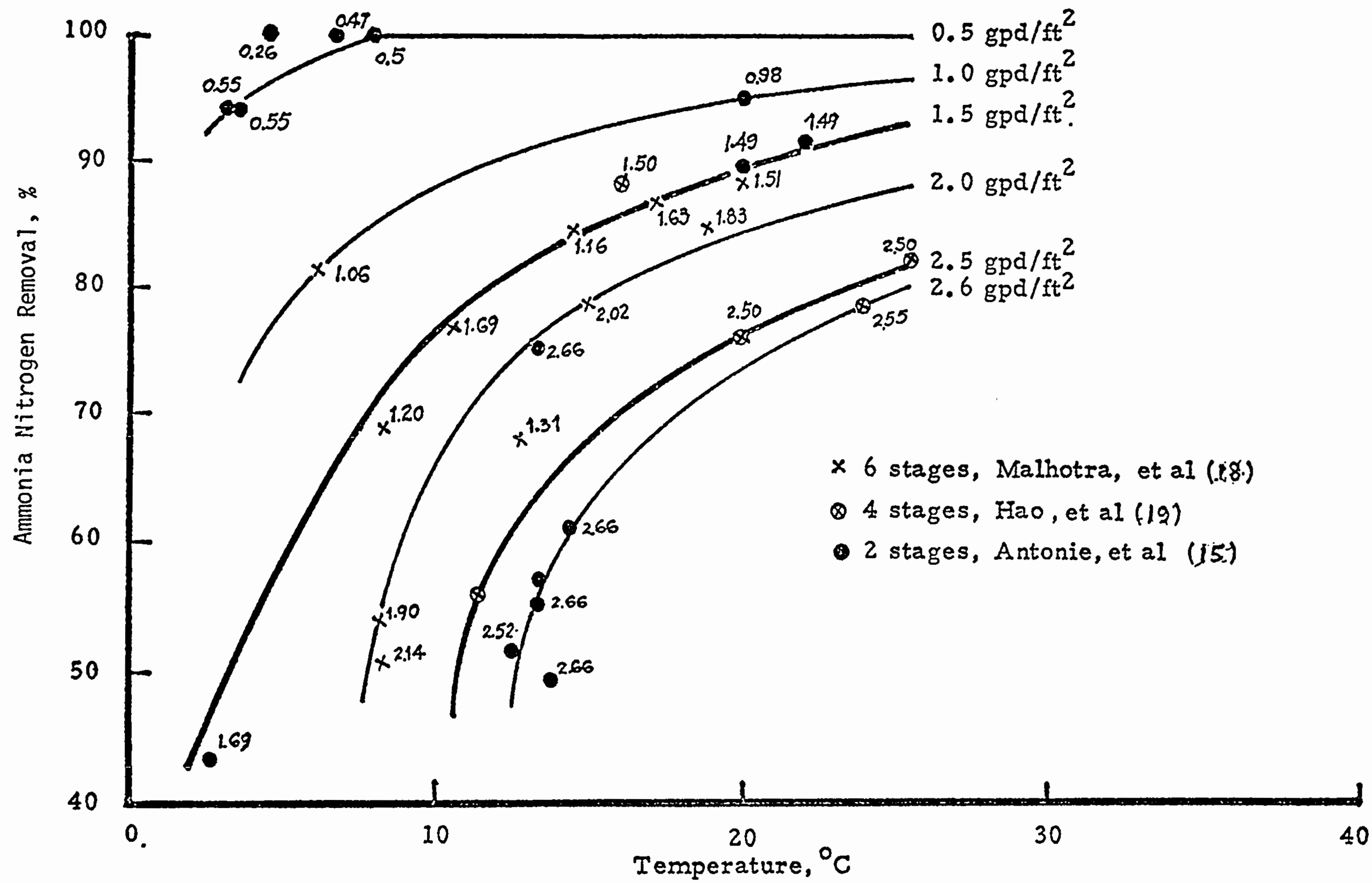

Figure 11. Effect of Temperature on Ammonia Nitrogen Removal

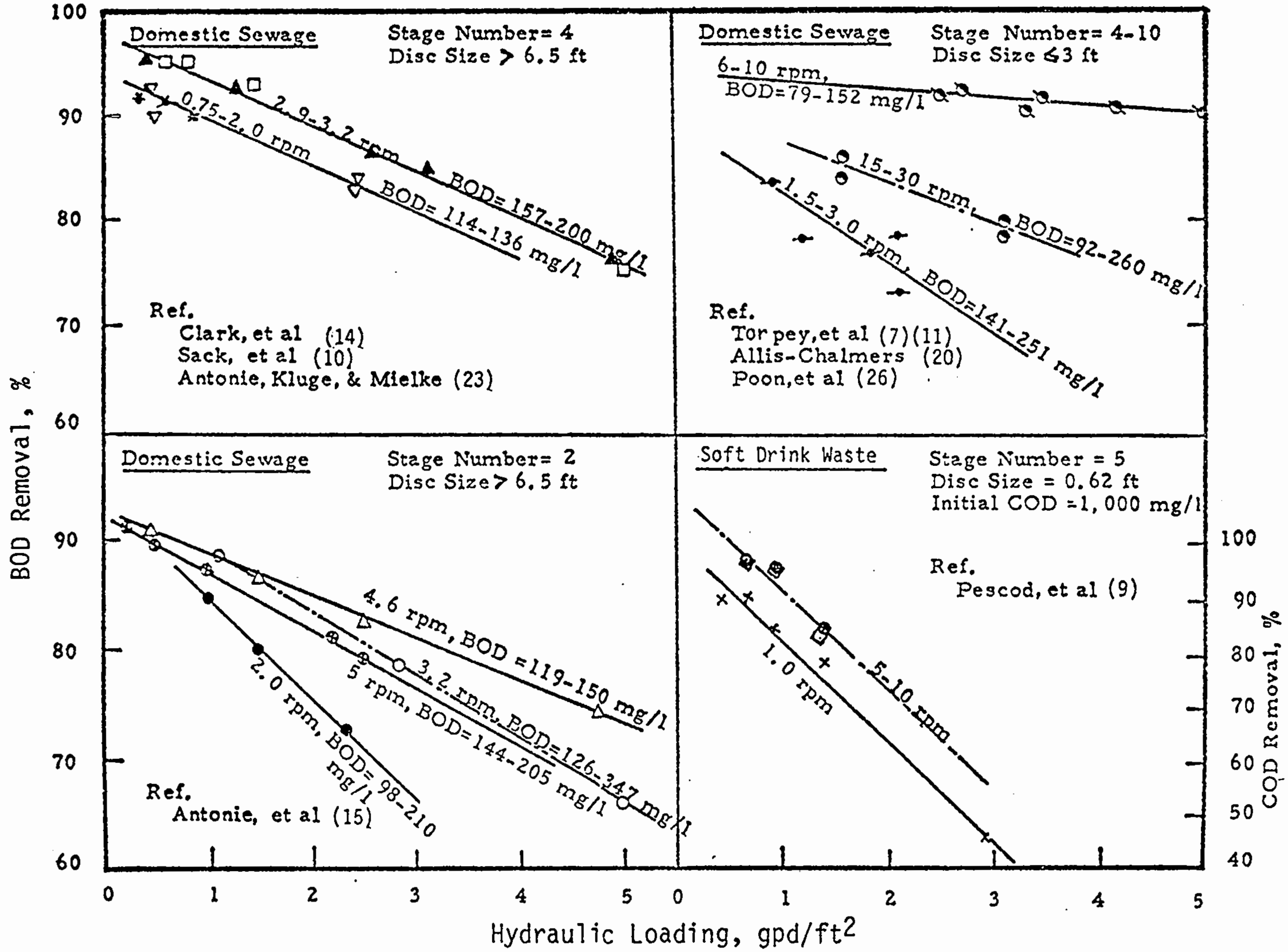

Figure 12. Effect of Rotating Disc Speed on BOD Removal

ment plant having the disc size greater than 6.5 ft., the optimum speed of disc rotation is approximately 4.6 rpm. Whereas in a small 4-10 stage pilot plant having a disc size of less than 3 ft., the most effective speed of rotation is somewhere between 6-10 rpm. The disc rotating speed in excess of the optimum velocity does not improve performance and wastes a significant amount of energy. Thus, rotational velocity of RBC media is an important design criterion. Testing of various diameter media has indicated that a fixed peripheral velocity can be used to determine the required RPM for any media diameter as long as there is no oxygen limitation on substrate removal. Under organic loading conditions where an oxygen supply limit exists, the angular velocity of RPM is the proper scale factor for various diameter media. The principle factor determining oxygen supply is the surface renewal rate per unit of wastewater flow and a proper scale-up for this factor is angular velocity. When an oxygen limit does not exist, the essential factors are hydraulic shear and mixing energy, both of which scale-up directly with peripheral velocity. Power requirements increase expenentially with increases in media velocity. For example, doubling the rotational velocity will typically increase the power consumption five-fold. Typicaly, full-scale RBC units consume from two to three KW per 100,000 sq. ft. of the media surface area when rotated at 1.6 RPM or a peripheral velocity of 60 ft. per minute. To significantly increase this rotational speed is not economically justifiable, particularly when considering that a 20-year present worth analysis of energy costs indicate that each horse-power of energy is worth $2,500 to $5,000. Therefore, a rotational speed of 1.6 RPM is considered a practical upper velocity limit to use, even when treating highly concentrated wastes (28).

Summary

The paper describes some of the factors that affect the performance of RBC process for wastewater treatment. The major factors controlling the performance of the RBC process are: Influent Wastewater Substrate Concentration, Residence Time of Wastewater (or surface hydraulic loading), Wastewater temperature and media rotational speed.

Acknowledgement

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CURRENT STATUS OF MUNICIPAL WASTEWATER TREATMENT WITH RBC TECHNOLOGY IN THE U.S.

By

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1.0 OVERVIEW

The U.S. Environmental Protection Agency's Municipal Environmental Research Laboratory (MERL) and Roy F. Weston, Inc. are currently engaged in the evaluation of the Current Status of Municipal Wastewater Treatment With Rotating Biological Contactor (RBC) Technology in the United States.

The study is outlined in six major categories:

1. Review of Available Process Equipment

2. Identification of Existing Facilities

3. Review of Existing Design Procedures

4. Evaluation of Field Process Performance Relative To Design

5. Evaluation of Field Operating Difficulties

6. Evaluation of O&M and Power Requirements

This paper summarizes the data collected and evaluated to date, briefly outlining equipment, existing facilities, design guides and power requirements, focusing on operating difficulties at surveyed facilities. A more detailed report is envisioned by early summer.

2.0 AVAILABLE PROCESS EQUIPMENT

RBC equipment can be placed into three basic categories: the media; the mechanical or drive components; and the tank or reactor. Each equipment manufacturer offers their own variation of media and drive components.

Table 2.1 highlights the nominal parameters associated with the media, mechanical components and tanks.

Some previous difficulties at RBC facilities have resulted from equipment problems, which affect both the mechanical and the process performance of the system. The media material, support, shaft strength, tank shape, baffling arrangement and clearance are some of the items which have adversely affected previous RBC performance. Further discussion of these problems are addressed in Section 6.0.

3.0 IDENTIFICATION OF EXISTING FACILITIES

There are approximately two hundred-sixty three RBC installations currently treating municipal wastewater, and fifty-eight installations treating industrial wastewater in the United States. Table 3.1 lists the distribution of municipal treatment facilities by flow range. Approximately twenty-five percent of the existing facilities are package plants provided as complete systems by RBC manufacturers. The largest operating RBC facility is an eighty shaft, fifty-four MGD facility at Alexandria, Virginia.

Figure 3.1 displays the regional distribution of RBC municipal facilities in the United States. The vast majority are located in the north, eastern and midwestern portion of the country. New York State has the largest number of facilities, but approximately eighty percent of the thirty-nine identified plants are package plants.

4.0 REVIEW OF EXISTING DESIGN PROCEDURES

Carbonaceous RBC removal rates per unit surface area are related to wastewater concentration, flow rate and temperature.

TABLE 2.1

BACIS RBC EQUIPMENT DIMENSIONS

MEDIA:	**Disc**	
	Shape	Circular
	Material	High Density Polyethylene
	Diameter	Standard: 12.0 Feet Range: 6.0 - 21.0 Feet
	Surface Area	Per 25-Foot Single Stage Shaft - Standard Media: 100,000 - 104,000 Square Feet High Density Media: 150,000 - 156,000 Square Feet
	Spacing	Standard Media: 1.25 Inches High Density Media: 0.75 Inches
	Construction	Segmented (8 Pieces): Steel Supported Unitized: Heat Welded Self-Supported
MECHANICAL:	**Shaft**	
	Shape	Cross-Section: Octagonal, Round Square
	Material	Steel
	Thickness	Municipal: 0.75 Inches Industrial: 1.50 Inches
	Length	Standard: 25 Feet Range: 6.0 - 25 Feet
	Motors	
	Horsepower Ratings	3.5, 5.0, 7.5
	Drive Units	Multi-V-Belts Chain and Sprocket Enclosed Cartridge
TANK:	Shape	Contoured, Filleted, Flat Bottom
	Material	Steel (0.25-inch Thickness) Concrete (1.0-Foot Thickness)
	Baffles	Overflow; Underflow

TABLE 3.1

TOTAL NUMBER OF OPERATING RBC INSTALLATIONS (1979)

Type	Flow Range (MGD)	Total
Municipal	0 - 0.1	74
	0.1 - 0.5	57
	0.5 - 1.0	41
	1.0 - 5.0	74
	5.0 - 10.0	8
	10.0 - 20.0	5
	> 20.0	4
	Sub-Total, Municipal:	263
Industrial	0 - 0.1	20
	0.1 - 0.5	13
	0.5 - 1.0	2
	1.0 - 5.0	5
	5.0 - 10.0	0
	(1)	18
	Sub-Total, Industrial:	58

(1) Size distribution not available

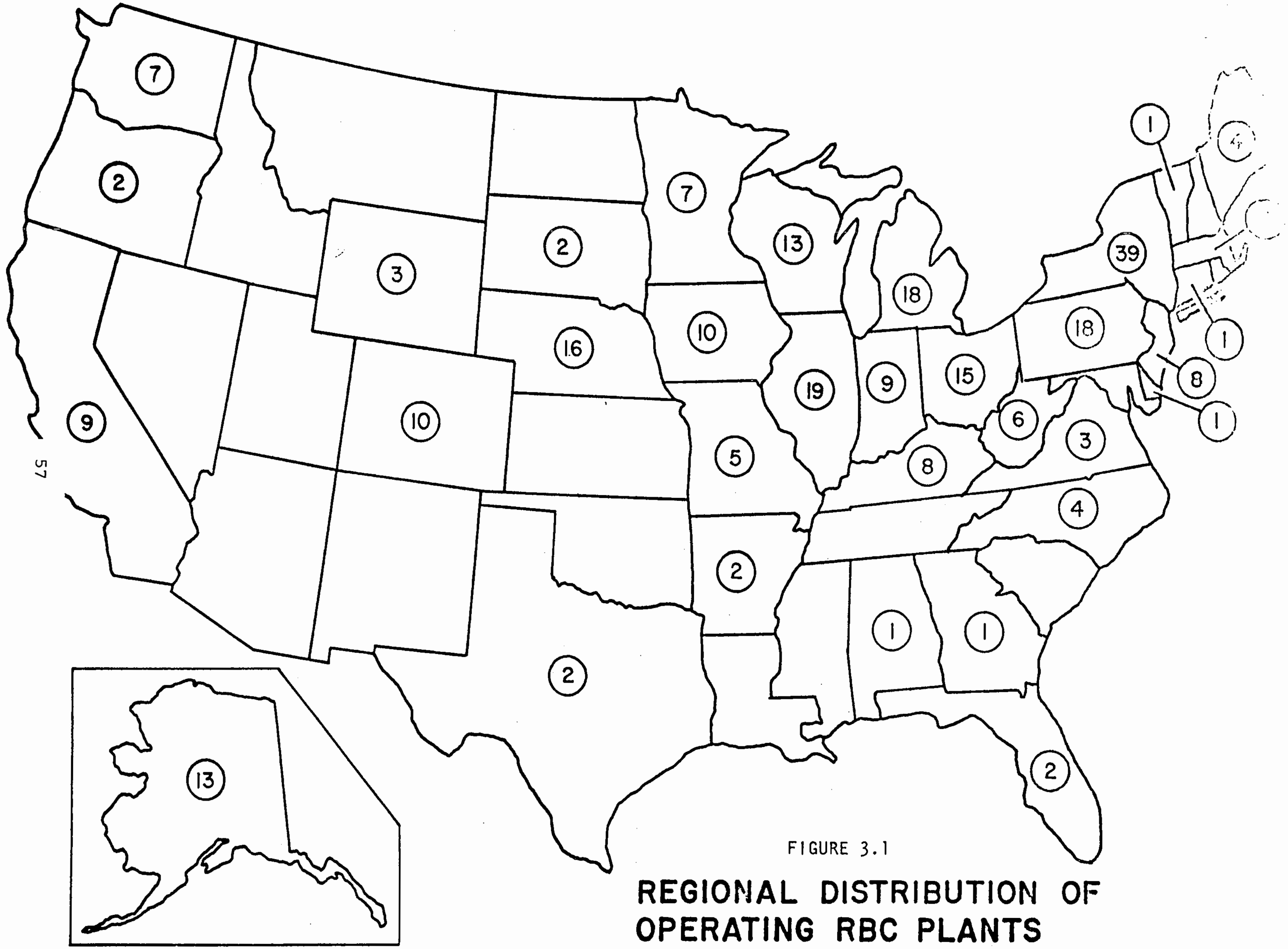


FIGURE 3.1

REGIONAL DISTRIBUTION OF OPERATING RBC PLANTS

In recent years, RBC systems have been mathematically simulated, attempts have been made to establish scale-up factors, and operating data has been fitted to design curves.

Current field design procedures are based upon manufacturers' established curves which define effluent concentrations (soluble or total BOD) in terms of applied hydraulic loading and influent concentration, as illustrated in Figure 4.1. Starting with a desired efficiency or effluent concentration and a given influent concentration, the designer simply selects the defined hydraulic loading which establishes the surface area requirement. These curves (Figure 4.1) are based upon field operation and have been established by developing relationships such as that presented in Figure 4.2. The shaded area shown in Figure 4.2 represents a range of disc performance as observed by the manufacturers.

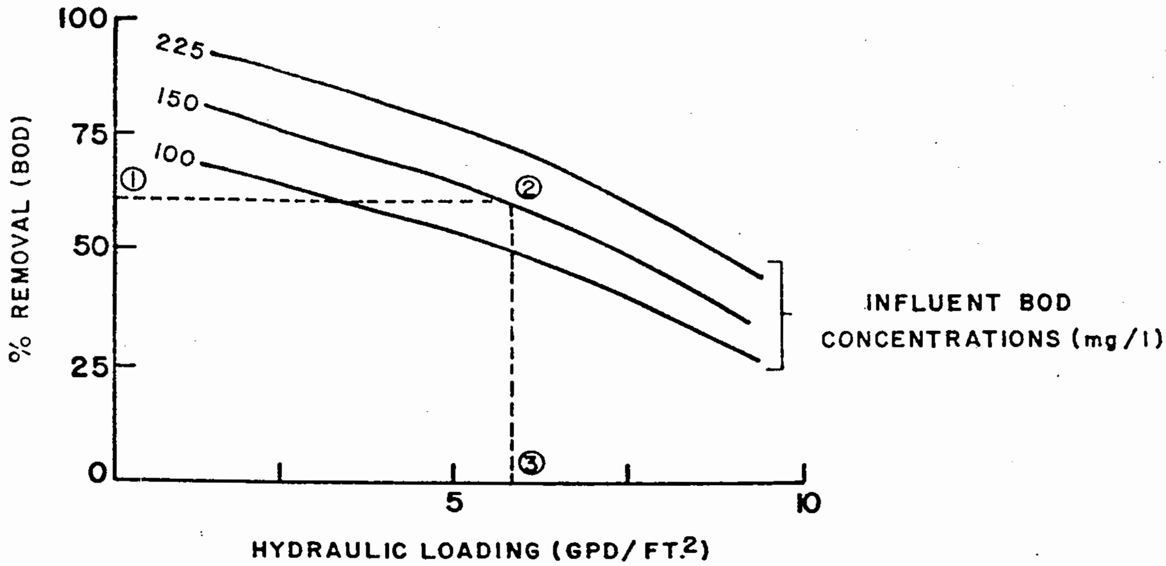
It is generally agreed that direct scale-up of RBC's, based upon unit area removal rates developed in pilot studies on small discs, cannot be extrapolated to full scale twelve-foot diameter installations. As a result, facility design is currently based upon previous operating experiences (design curves) or full scale pilot studies.

Staging RBC units should theoretically optimize the process by establishing desirable cultures in each stage and maximize removal rates. Current design guides recommend staging to achieve high removal rates, but have not definitively established the relationship between staging and existing design curves.

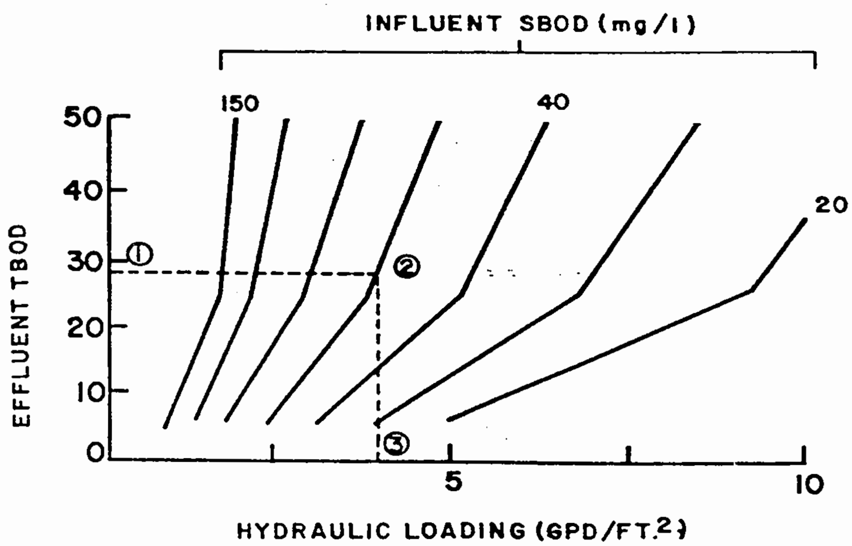
Surface corrections are recommended for wastewater temperatures below 55^{o}F. Figure 4.3 illustrates three different existing temperature correction curves currently recommended. The lower curve illustrates temperature correction based upon a decreasing reaction defined by $K_2/K_1 = 1.042^{(T-20)}$. The upper curves simulate greater reductions in reaction rates at these lower temperatures.

5.0 POWER REQUIREMENTS

Power drawn by operating RBC units is a function of disc surface area, rotational speed and the organic removal rate of the individual RBC shaft. Energy requirements increase as each of the above parameters increase. Variations in the media configuration of each manufacturer can also be expected to affect the drag and hence, the energy utilization of the system.



BASIS: TOTAL BOD



BASIS: SOLUBLE BOD

FIGURE 4.1

EFFLUENT DESIGN CURVES

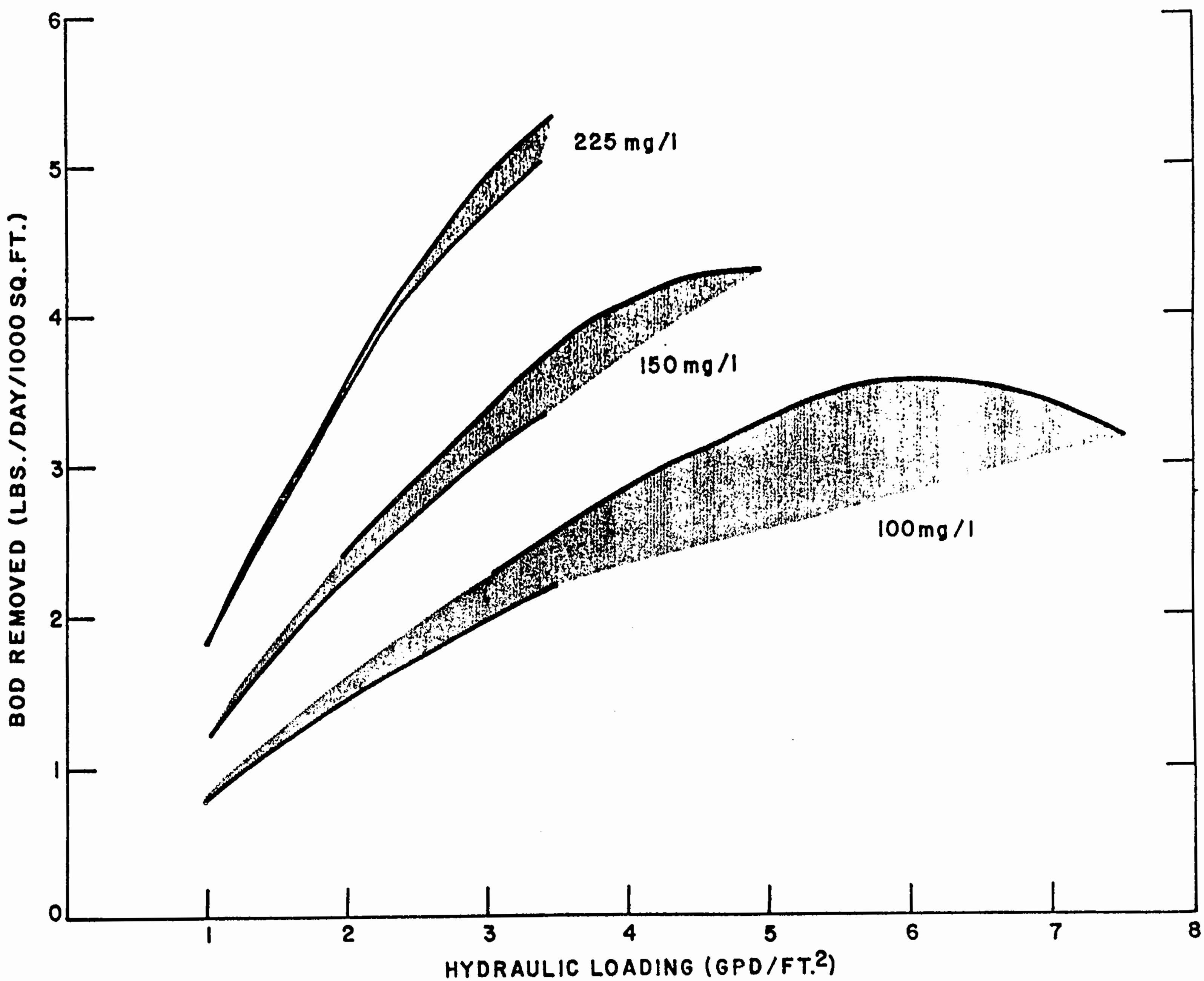


FIGURE 4.2
DESIGN LOADING CURVES

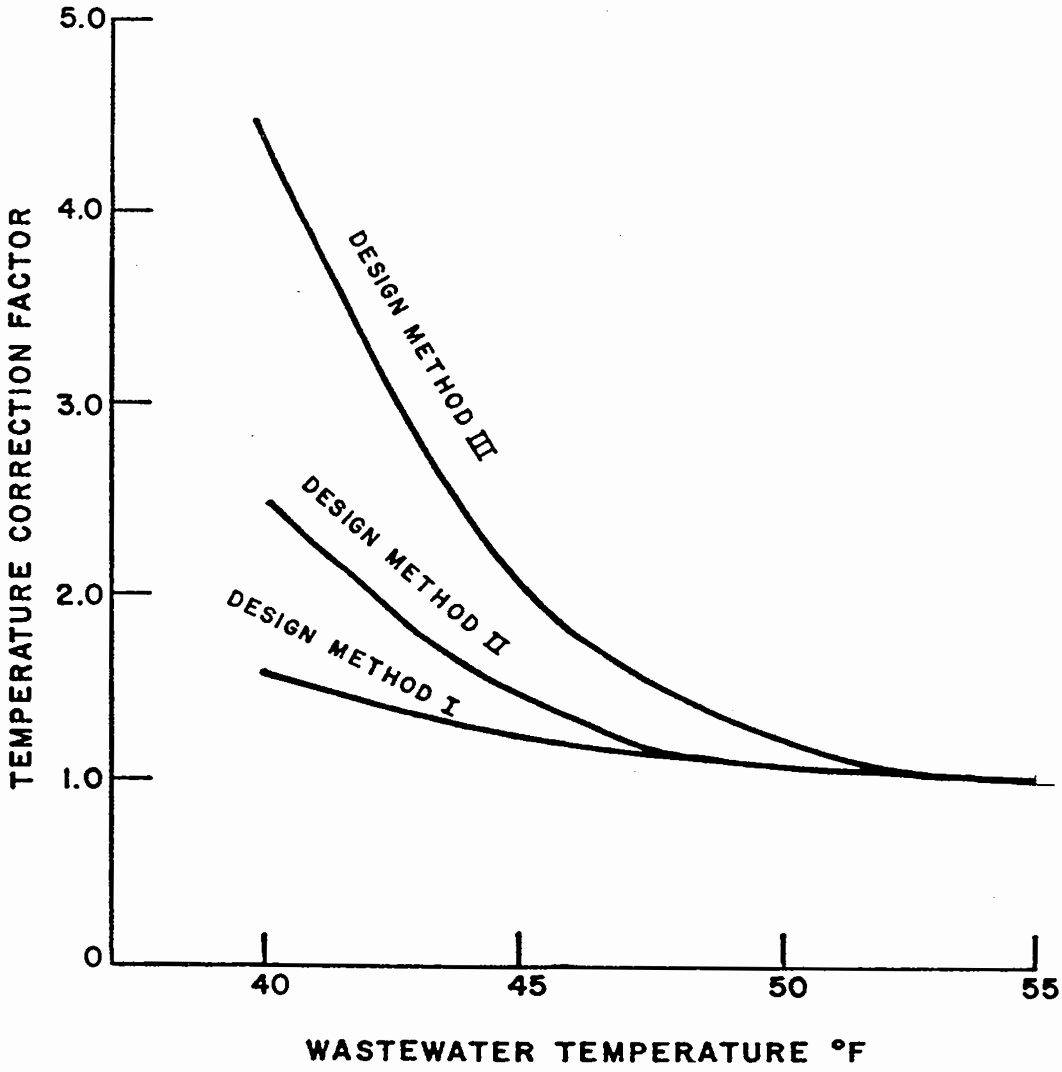


FIGURE 4.3

TEMPERATURE CORRECTION CURVE

Results of a literature survey of RBC power requirements with respect to trickling filters and conventional activated sludge is presented in Table 5.1, which categorizes RBC energy requirements per million gallons between trickling filters and activated sludge systems. Such an analysis is helpful in preliminary assessments of energy requirements, but must be tempered with organic removal rates and efficiences to fully assess the actual effective use of power supplied for treatment.

Preliminary results of field power studies are reported in Table 5.2, which depicts metered power drawn at five operating facilities. Power drawn per one-hundred-thousand square feet or shaft average 2.9 kw varying from 1.9 to 4.1 kw. Power tests (manufacturers) conducted on clean media at two locations, average approximately 1.4 kw per one-hundred-thousand square feet. Energy utilization per million gallons varied from five hundred-fifty five to three hundred-forty two kw-h/MG, averaging four hundred-ten kw-h/MG. This is consistent with the range observed in the literature survey (Table 5.1).

Energy utilization per million gallons is related to the hydraulic loading and the rotational speed. Higher hydraulic loadings result in more favorable energy/flow ratios; however, these loadings must be commensurate with effective treatment. Lower rotational speeds will also reduce metered kw, but must insure adequate mixing and aeration.

6.0 FIELD OPERATING DIFFICULTIES

Seventeen RBC facilities have been surveyed to date to aid in the review of operating and performance problems. These seventeen facilities were selected because they represent facilities which have equaled or exceeded eighty percent of their design flow and, as a result, were considered more likely to display process operating difficulties if they exist.

Table 6.1 outlines all reported difficulties. Some of these are minor and some are major. Some are directly attributed to RBC equipment and operation and some the result of design omissions and/or questionable operation practices. The analysis is designed to identify and review those reoccurring problems which limit RBC performance, and to identify problems which can be minimized in future facilities.

Table 6.1 tabulates the actual and design hydraulic loadings of each plant, highlighting structural difficulties reported along with design loading problems and their impacts.

6.1 STRUCTURAL PROBLEMS

Structural problems are divided into three categories: media; shaft; and bearing. Of the seventeen plants surveyed, four experienced some media difficulties; three shaft problems; and two bearing problems.

TABLE 5.1

LITERATURE SURVEY: ENERGY REQUIREMENTS OF SECONDARY TREATMENT PROCESSES

Process	Number Of Plants	Average KWH/MG	Range KWH/MG	Source
Trickling Filter	6	693	400 - 1100	(1) (2)
	1	(192)	----	(2)
RBC	2	803	720 - 887	(1) (3)
	5	(472)	(408 - 585)	(1) (4) (5) (6) (7)
Conventional Activated Sludge	5	2680	1300 - 4300	(1)
	2	(912)	(894 - 930)	(4) (8)

NOTE: Data on plants given in parentheses is for secondary portion of process only.

(1) Evaluation of Operation and Maintenance Factors Limiting Municipal Wastewater Treatment Plant Performance; EPA-600/2-79-034; June 1979

(2) Energy Use in Wastewater Treatment: Biological Waste Treatment Systems; Paper Presented At Second Mid-Atlantic Environmental Engineering Design Conference; May 1977

(3) Roy F. Weston Questionnaire; October 1979

(4) Design and Cost Considerations For A Biological Tower Treatment System; Paper Presented At First Mid-Atlantic Environmental Engineering Design Conference; May 1976

(5) Roy F. Weston Questionnaire; October 1979

(6) Roy F. Weston Questionnaire; October 1979

(7) Energy Conservation Dictates Innovative Treatment Plant Design; Public Works; 105.3.87 (1974)

(8) Energy Requirements For Wastewater Treatment; Water and Sewage Works; December 1976

TABLE 5.2

FIELD POWER TEST RESULTS

Plant No.	Flow (MGD)	Metered KW Draw/100,000 Ft.^{2} (1)	KW-Hr./MGD	Hydraulic Loading H.L.	Hydraulic Loading RPM
4	0.2	2.6	555	1.18	1.5
6	0.5	4.1	357	2.02	1.75
7	1.3	1.9	410	1.13	1.0
14	1.0	2.7	384	1.89	1.8
16	2.8	3.3	342	2.30	1.6
		Average: 2.9	Average: 410		
		Clean Media (2): 1.4			

(1) Standard Media

(2) Manufacturers Average Power Tests at Princeton, Illinois and Plainville, Connecticut

TABLE 6.1

REVIEW OF OPERATING RBC FACILITIES

Plant No.	Design Flow (MGD)	Percent Of Design	Design Hydraulic Loading GPD/Ft.^{2}	Structural Problems: Media	Structural Problems: Shaft	Structural Problems: Bearing	Design Loading Problems: Raw Wastewater	Design Loading Problems: Sidestreams	Impacts: D.O.	Impacts: Beggiotoa (5)	Impacts: SS Accumulations	Impacts: Hydraulic Overload: Washout and Diluted Waste	Effluent Requirement Problems: BOD	Effluent Requirement Problems: SS	Effluent Requirement Problems: NH_3-H
1	1.9	108	1.72		X		X					X		X	
2	1.13	83	2.17				X		X						
3	0.65	80	3.70												
4	0.20	85	1.18	X			X		X			X			
5	0.216	91	1.03												
6	0.5	100	2.02	X			X			X	X				
7	1.3	100	1.13	X		X	X		X	X	X				X
8	0.39	136	2.85				X					X			
9	1.0	80	2.5				X	X	X	X	X		X	X	X
10	0.5	100	1.56			X		X	X	X			X	X	
11	0.6	80	1.0				X								
12	0.65	1.00	2.6				X	X			X	X		X	
13	1.75	80	5.72 (1)		X		X				X	X			
14	1.0	80	1.89				X								
15	2.0	100 (2)	2.5		X		X					X			
16	2.8	103 (3)	2.3	X			X		X	X	X				
17	0.35	95	1.75				X	X	X	X		X	(4)	(4)	
			Totals:	4	3	2	14	4	7	6	6	7	2	4	2

(1) Follows trickling filter

(2) 1/2 of plant RBC's in use

(3) 2/3 of plant RBC's in use

(4) RBC;s followed by polishing lagoon

(5) Nuisance bacterial growth

Table 6.2 details the specific problems associated with the structural difficulties. Reported media problems include: movement of segmented, steel supported discs, which result in plastic shear, failure of the hub which links the media to the shaft in unitized systems, shifting media on support rods and ultraviolet degradation of the media. Reported shaft and bearing problems include a total of six shaft failures in three plants and two bearing problems; one associated with flooding, and the other with improper lubrication.

Corrective actions for structural failures in almost all cases require equipment replacements. Shaft failures and/or media movement or failure are major equipment problems, which must be corrected if RBC systems are to be considered viable treatment options. Media degradation from exposure to ultraviolet radiation, results in brittleness and can lead to ultimate structural failure.

Equipment manufacturers are aware of the importance associated with structural problems. Recent actions by major manufacturers have been aimed at improving the structural integrity of the media, their support and shaft strength. The addition of supplementary carbon to the media and insuring that the media is continually covered, should help reduce the affects of ultraviolet degradation.

6.2 DESIGN LOADINGS

Of the seventeen plants surveyed, Table 6.1 lists fourteen reported variations in raw wastewater loading and four reported variations resulting from sidestreams. Impacts associated with loading difficulties include: seven reporting low DO concentrations; six reporting undesirable bacterial growth; four reporting solids accumulation in undesirable locations; and seven reporting washouts or diluted wastewater resulting from hydraulic overloads.

Table 6.3 further defines the cause of the individual loading problems at each facility. Difficulties include excessive hydraulic loads resulting from inflow and infiltration (I/I); excessive flow resulting from water running in the winter to avoid freeze-up, and peaking. Unaccounted industrial contribution, sidestreams, and septage all increase organic loads above those anticipated. Sidestreams include aerobic digestor, anaerobic digestor, and sludge lagoon supernatant return. Poorly operated digestors can substantially increase the return solids as well as organic load to the RBC unit. Long collection system detention times were often associated with incoming septic waste. The absence of primary clarifiers increased the susceptibility of the RBC systems to potential solids and organic loading problems. Excessive solids detention time in the primary clarifier helped create septic conditions and poor settling. Solids deposition in the channel and the tank were deemed significant in creating septic environments. Solids deposition in tanks have been attributed to insufficient tank mixing and dead spots, resulting from poor designs and overflow baffles. Excessive equalization basin aeration in one facility increased the soluble organic loading to the RBC unit by hydrolizing some of the suspended solids.

TABLE 6.2

STRUCTURAL DIFFICULTIES

MEDIA DIFFICULTIES:

Plant No.	Description	Corrective Action
4	Movement Of Segmented Discs; Plastic Damage	Support Tightened
6	Plastic Hub Failure In Unitized Construction	Replaced
7	Exposure Of Media To Sunlight; Ultraviolet Degradation	Replace Damaged Cover (Carbon Black)
16	Shifting Media On Supports	Eliminate Play Between Media and Supports

SHAFT DIFFICULTIES:

Plant No.	Description	Corrective Action
1	Shaft Failure; Two Shafts	Replacement
13	Shaft Failure: Three Shafts	Replacement
15	Shaft Failure; One Shaft	Replacement

BEARING DIFFICULTIES:

Plant No.	Description	Corrective Action
7	Bearing Seizure; Pumps Flooded	Replacement
10	Uneven Bearing Wear; Lack Of Proper Lubrication	Repair Or Replace

Corrective actions for the itemized difficulties in Table 6.3 vary widely. Improper design loadings resulting from poor facility planning is difficult to correct after installation without increasing the plant capacity or reducing the loading. Sidestreams are loadings which are often omitted in the design of wastewater treatment facilities. Designers must account for these loads in the design of RBC systems. Since in many cases, sidestreams are dependent upon sludge thickening, dewatering, and treatment methods, current design curves (typical domestic wastewater) do not fully account for these loads. Given existing sidestream conditions, operators should avoid shock return loads and timewise distribute the recycled flow to the head of the facility. Channel, tank aeration, and the elimination of dead spots can help reduce solids accumulation in existing facilities. Where this is not feasible, solids must be withdrawn directly.

6.3 NUISANCE BACTERIAL GROWTHS

Microbes of many kinds are present in RBC films. Conditions which favor the growth of nuisance organisms that result in the colonization of initial stages of RBC media, have been reported. Resultant impacts include reduced organic removal rates and a shifting of loadings from the initial to the latter stages of a multistage system.

A sulfide oxidizing filamentous aerobe and microaerophile, from the Beggiotoa family commonly referred to as Beggiotoa, has in some cases been identified as the nuisance organism and has become associated with this particular problem in RBC systems. For purposes of this discussion, the nuisance organism will be referred to as Beggiotoa. This slow growing population has a milky white appearance when predominating RBC surfaces. The odor from the film in this condition is somewhat septic and unpleasant and markedly different from a healthy disc culture. The biomass is also thinner than would be expected for a first stage.

Beggiotoa problems have been associated with two reoccurring environmental conditions:

1. Low DO

2. Reducing environment capable of sulfide production

Since Beggiotoa are microaerophiles, they can exist at low DO concentrations. Low DO, coupled with sulfide availability, would appear to present optimum condition for Beggiotoa propogation at the expense of other aerobic microbes.

Six of the seventeen plants surveyed reported nuisance organism or Beggiotoa problems. Almost all of these plants reported low DO and sidestreams or suspended solids accumulation creating conditions for sulfide production. Two facilities surveyed did not report low RBC tank dissolved oxygen concentration, but did report suspended solids accumulation and localized septic conditions making sulfide production possible.

TABLE 6.3

LOADING DIFFICULTIES

Number Of Plants	Description	Predominant Effect
7	I/I	Hydraulic Overload; Diluted Waste
1	Winter Tap Flow (Avoid Freezing)	Hydraulic Overload; Diluted Waste
4	Peaking/Hydraulic Shock Loads	Solids Washout
4	Industrial Contributions	Organic Overload; DO Deficiency; Sulfide Production
4	Sidestreams	Organic Overload; DO Deficiency; Sulfide Production
1	Septage Contribution	Organic Overload; DO Deficiency; Sulfide Production
4	Long Collection System Detention Time	DO Deficiency, Sulfide Production
3	No Primary Clarification	High Organic and Solids Load; DO Deficiency; Sulfide Production
5	Solids Deposition In Channels	DO Deficiency; Sulfide Production
2	Solids Deposition In Reactor	Sulfide Production
2	Septic Primary Clarifier Sludge	DO Deficiency; Sulfide Production; Increased Soluble Organic Loading
1	Excessive Equalization Basin Aeration	Increased Soluble Organic Loading

It is currently felt that the condition which leads to sulfide production (incoming septic waste, suspended solids accumulation, long primary clarifier detention times) is the predominant factor associated with Beggiotoa problems. Beggiotoa are capable of surviving in highly aerobic or microaerobic conditions. Under microaerobic conditions, their competitive advantage is probably enhanced. Microaerobic conditions will exist within the RBC film at depths where oxygen penetration becomes limiting. If sulfides are available, this environment is probably capable of catalyzing a Beggiotoa take over.

Potential corrective actions include: the elimination of conditions which cause the sulfide production; oxidation of sulfide prior to the RBC systems.

7.0 CLOSING

Preliminary investigation of RBC systems have identified operating difficulties which are currently being addressed by manufacturers, design engineers and researchers. Some of these difficulties are common to biological systems, while some are unique to RBC systems. Major items needing resolution include:

- Solution to equipment problems
- Continued design procedure improvements
- Solutions to nuisance organism problems

Continued RBC research and data generation will continue to enhance our ability to utilize the treatment capabilities of these systems.

PART II. PROCESS VARIABLES AND BIOFILM PROPERTIES

HYDRAULIC CHARACTERISTICS OF THE RBC

By

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INTRODUCTION

In the course of pilot scale investigations with the rotating biological contactor (RBC) in the treatment of acid mine drainage (1), divergent results were obtained with different size treatment units when equivalent hydraulic loading and disc peripheral velocities were applied. It was the objective of this study to learn if the observation of lower ferrous iron removal efficiencies with a larger size RBC was related to hydraulic characteristics of the unit.

BACKGROUND

Full scale RBC systems, scaled according to pilot plant data, may not always meet design expectations. Godlove et al. (2) found that soluble COD removal from petroleum refinery wastewaters was about 14 percent less efficient in full scale RBC systems than in pilot plant units. Murphy and Wilson (3) reported about 16 percent lower COD removal efficiencies for a 2.0-m diameter RBC unit treating wastewater at a peripheral velocity equivalent to a parallel 0.5-m unit. Chesner and Molof (4) found that smaller discs operated at equivalent peripheral velocities to larger discs resulted in increased oxygen transfer with correspondingly higher COD removal efficiencies. Recently, Friedman, et al. (5) described experiments with a pilot scale RBC unit operating at different rotational speeds and it was concluded that heavily loaded or older plants approaching design load may not be able to meet design performance because differences in oxygen transfer exist between pilot and full scale systems operated at equivalent peripheral velocites.

In previous work performed by Olem and Unz (1) involving the treatment of acid mine drainage by the RBC, ferrous iron oxidation efficiency of the 2.0-m RBC was about 10 percent lower than that of the 0.5-m unit when operated in parallel and under equivalent conditions of peripheral velocity and hydraulic loading rate (Figures 1 and 2). The discrepancy was not believed due to an oxygen deficiency, however, since high reactor dissolved oxygen concentrations (greater than 10 mg/l) were observed for both units, presumably owing to the low temperatures (typically 10°C) and oxygen demand of the mine water relative to wastewaters rich in organic matter.

In this study, chemical tracers were employed to determine if significant hydraulic differences exist between the different size RBC units when operated at equivalent hydraulic loading rates and peripheral disc velocites.

METHODS

Rotating Biological Contactors

Field studies were conducted employing commercial RBC pilot units (Autotrol Corp. Milwaukee, Wisc.) equipped with 0.5- and 2.0-m diameter discs (Figure 3). Each unit consisted of four stages of closely spaced and corrugated high-density polyethylene discs suspended on a shaft with approximately 40 percent of the surface area immersed in the volume of a corrosion-proof trough. The individual stages were separated by baffles. Flow through the 0.5-m unit was facilitated by a 2.5-cm diameter hole in each baffle (Figure 4, left). A serpentine flow pattern was formed in the 2.0-m unit by an arrangement of piping on the outside of stage compartments (Figure 4, right). A rotating bucket mechanism controlled mine water feed from the influent chamber of each unit to stage compartments.

RBC units were enclosed in a building for protection from the natural environment. Mine waters were delivered from the source with the aid of two 0.25-kW centrifugal pumps in connection with a 7.6-cm diameter foot valve and appropriate lengths of flexible polyethylene piping. Mine water was pumped to the feed chambers of pilot units at slightly greater than the desired flow rate of 119 m^3/d. Excess flow from the feed chamber was discharged through an overflow pipe in order to maintain a constant liquid level of fresh mine water in the feed chamber. The static level of mine water in the feed chamber and the

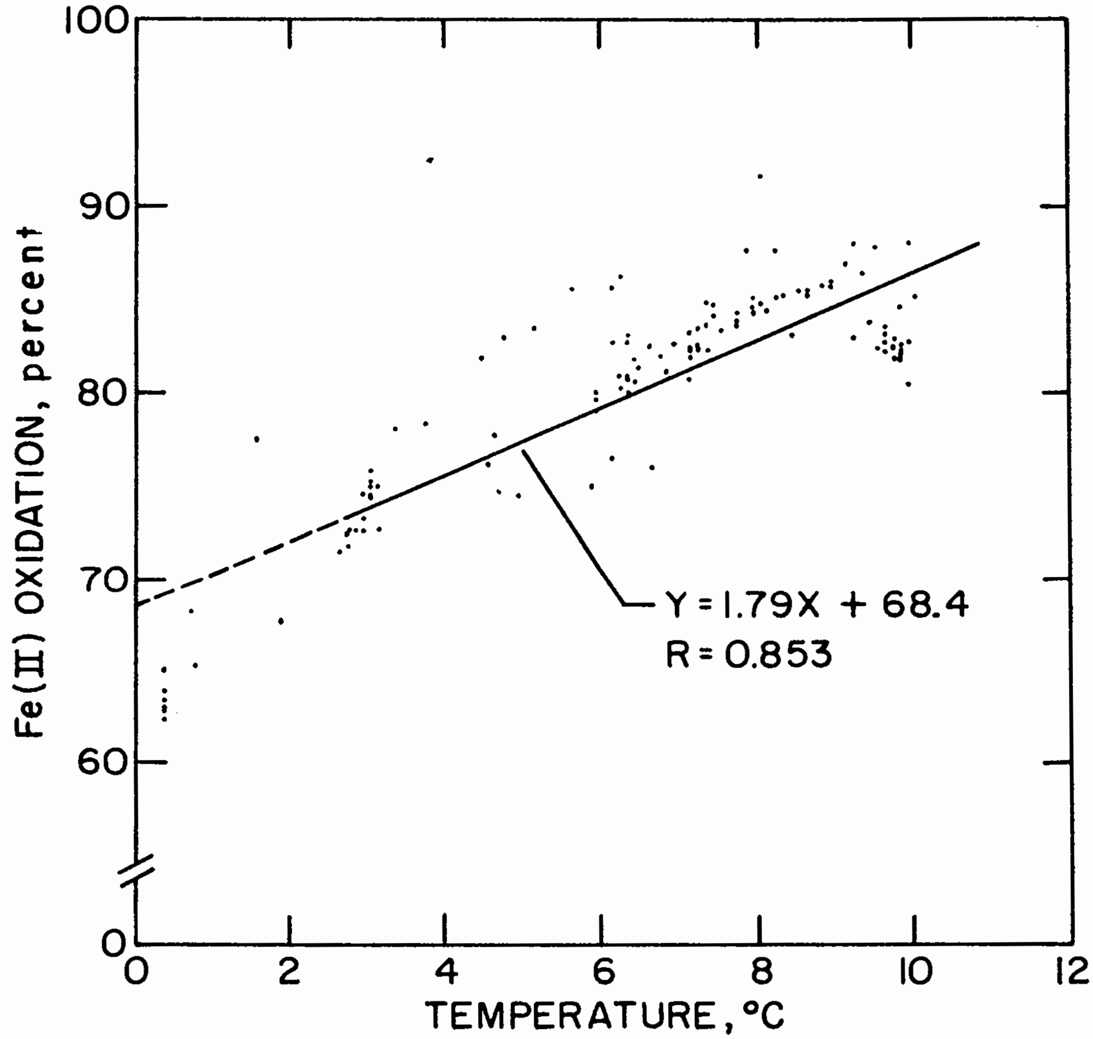

Figure 1. Ferrous iron oxidation efficiencies obtained at different acid mine drainage temperatures with the 0.5-m RBC.

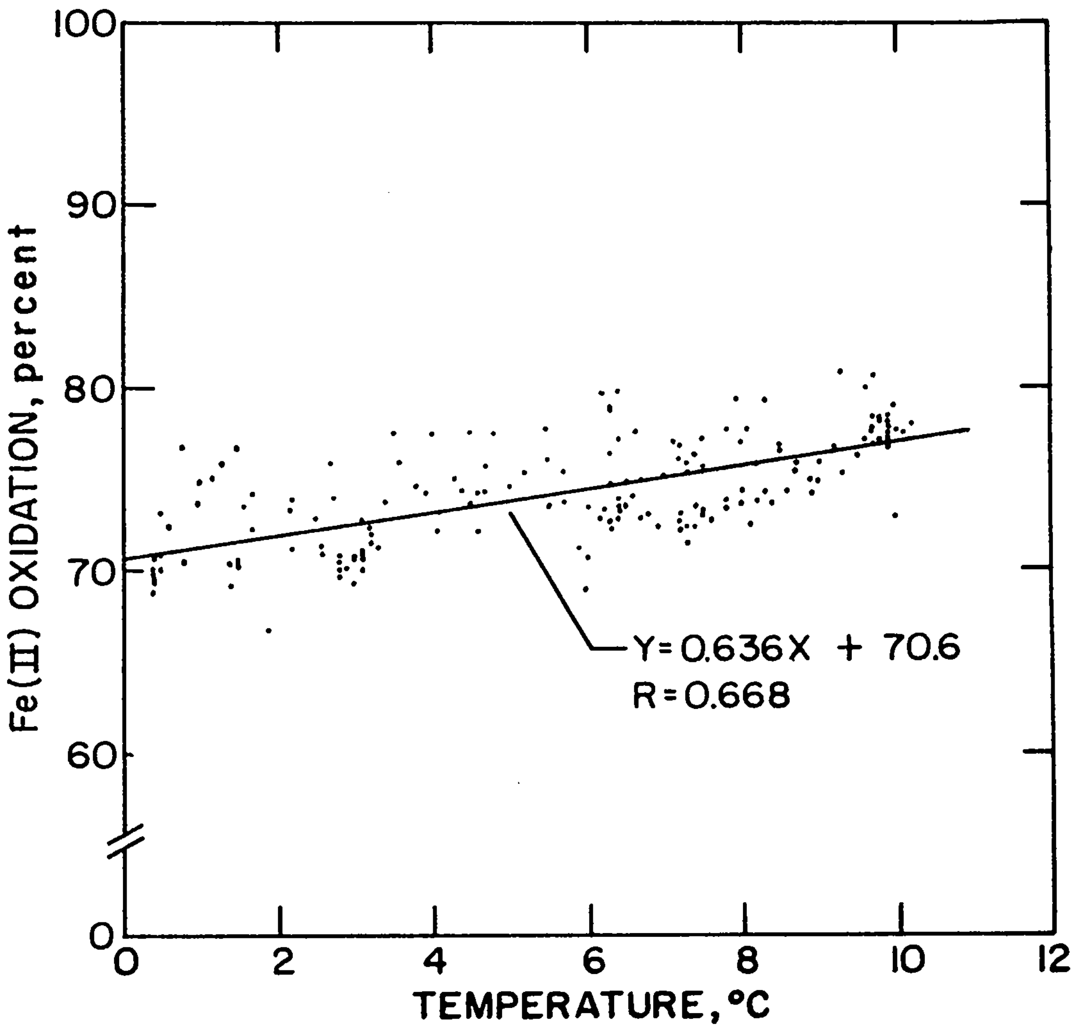

Figure 2. Ferrous iron oxidation efficiencies obtained at different acid mine drainage temperatures with the 2.0-m RBC.

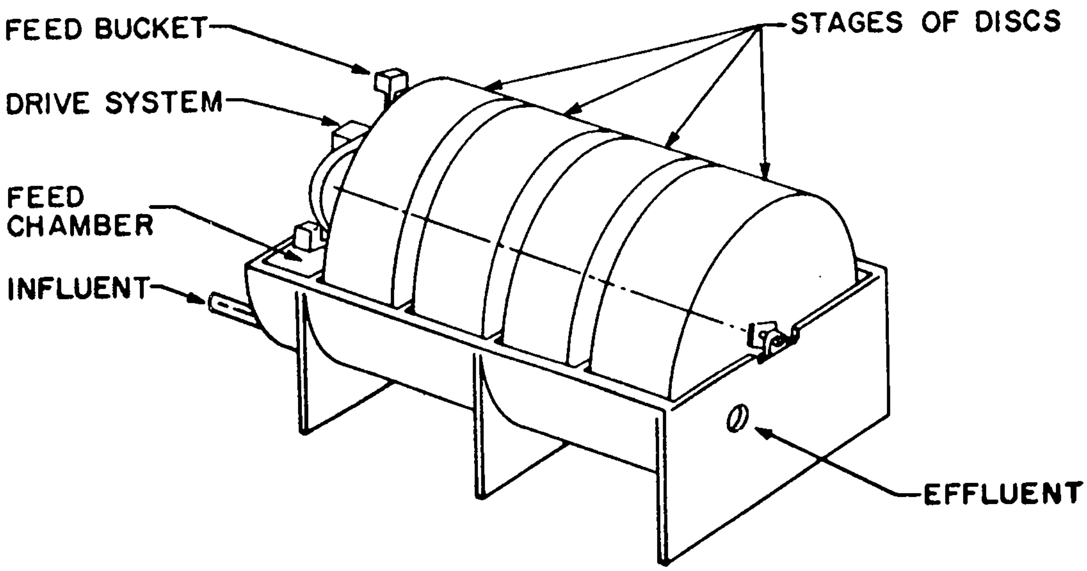

Figure 3. Schematic of rotating biological contactor pilot units.

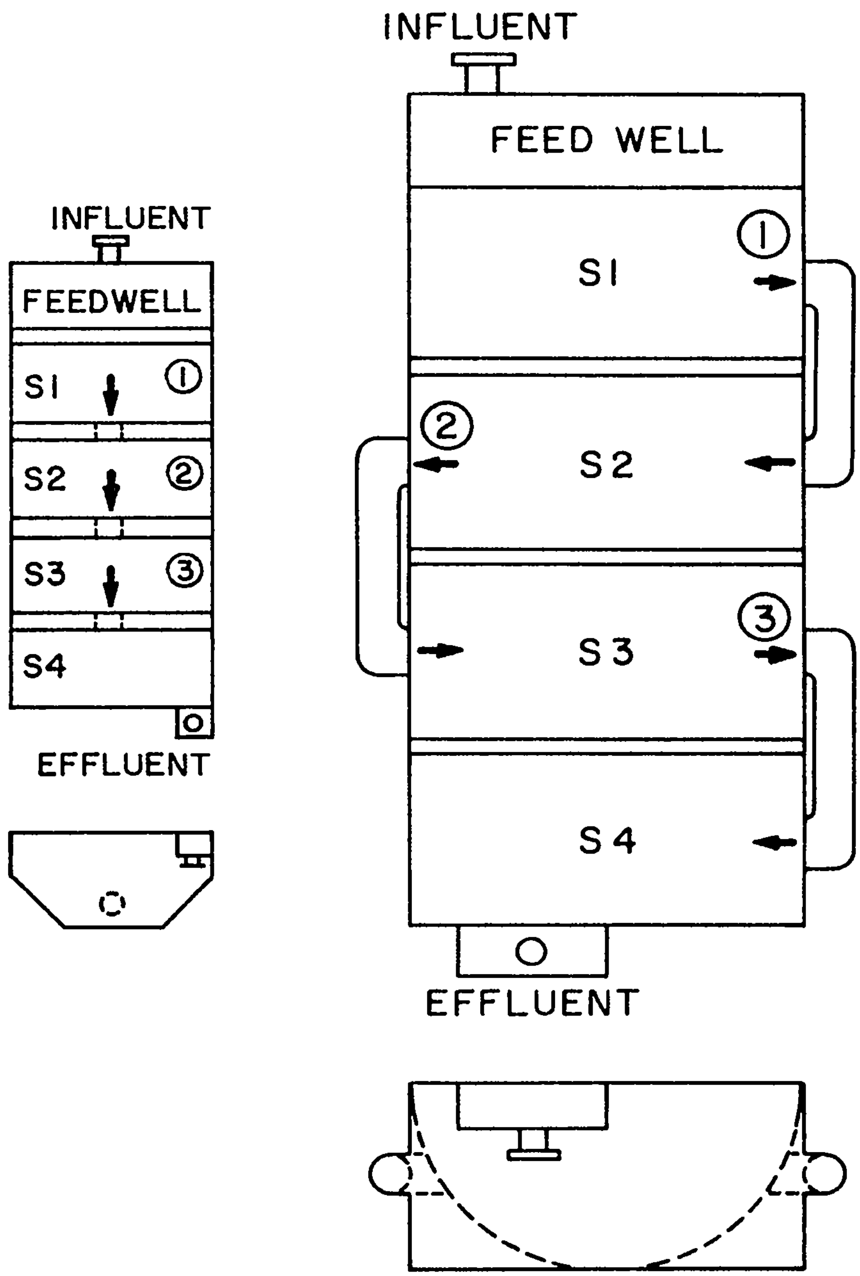

Figure 4. Plan and side view showing flow pattern and stage sampling locations for trough of 0.5-m RBC (left) and 2.0-m RBC (right). Drawing is not to scale.

number of feed buckets employed was in accordance with the desired flow rate. Flow rates were checked by collecting effluent samples for one minute in 3-l and 100-l containers, respectively, for the 0.5- and 2.0-m units. Both units were operated at an equivalent peripheral disc velocity of 19 m/min and a hydraulic loading of 0.16 $m^3/d\text{-}m^2$. Attainment of a peripheral velocity of 19 m/min required 13 and 2.9 rpm for the 0.5- and 2.0-m units, respectively.

Tracer Additions

During equilibrium operation of the RBC units, a concentrated solution of lithium chloride (1.0 g/l) was added instantaneously to one feed bucket of each RBC unit just prior to discharge into stage one. The volume of tracer solution added to the 2.0-m unit was proportionately larger on the basis of trough volume. Samples collected from each stage and effluent at specified time intervals were analyzed for lithium content by atomic absorption spectrophotometry. In addition, as a check on the validity of the lithium technique and for convenience of field analyses, sodium chloride was employed as an independent tracer and monitored by specific conductance. A concentrated solution of sodium chloride (230 g/l) was added simultaneously along with the lithium chloride solution.

RESULTS AND DISCUSSION

Recovery of lithium chloride and sodium chloride tracers in each stage of the RBC units is presented in Figures 5 and 6. There existed a background specific conductance in the mine water which ranged from 1020 to 1100 μmhos/cm. There was no detectable background level of lithium. The final set of samples for lithium analysis was collected 20 min following the attainment of baseline specific conductance (160 min).

Samples collected from stage one of both units revealed peak lithium content within 30 sec of tracer addition. Higher lithium concentrations were recovered from stage one of the 0.5-m RBC than from the same stage of the 2.0-m unit, which indicated more short-circuiting of flow in the larger size RBC to subsequent stages.

Pintenich and Bell (6) described a procedure for evaluation of tracer data to quantify the hydraulic characteristics of continuous flow treatment basins. The method is based on comparison of the shapes of standardized tracer recovery curves to those for an ideal completely mixed basin. Relative amounts of complete mix, plug flow, and dead space volumes may be obtained. Overall lithium recovery curves for the 0.5- and 2.0-m RBC units are presented in Figure 7. Percent recovery curves were developed by integration of the areas under a standardized recovery curve (C/C_o vs. t/T).

Both units were found to be approximately 80 percent completely mixed. Dead space in the trough of the 2.0-m RBC, although relatively low for continuous flow reactors, was 10 percent as compared to 2 percent

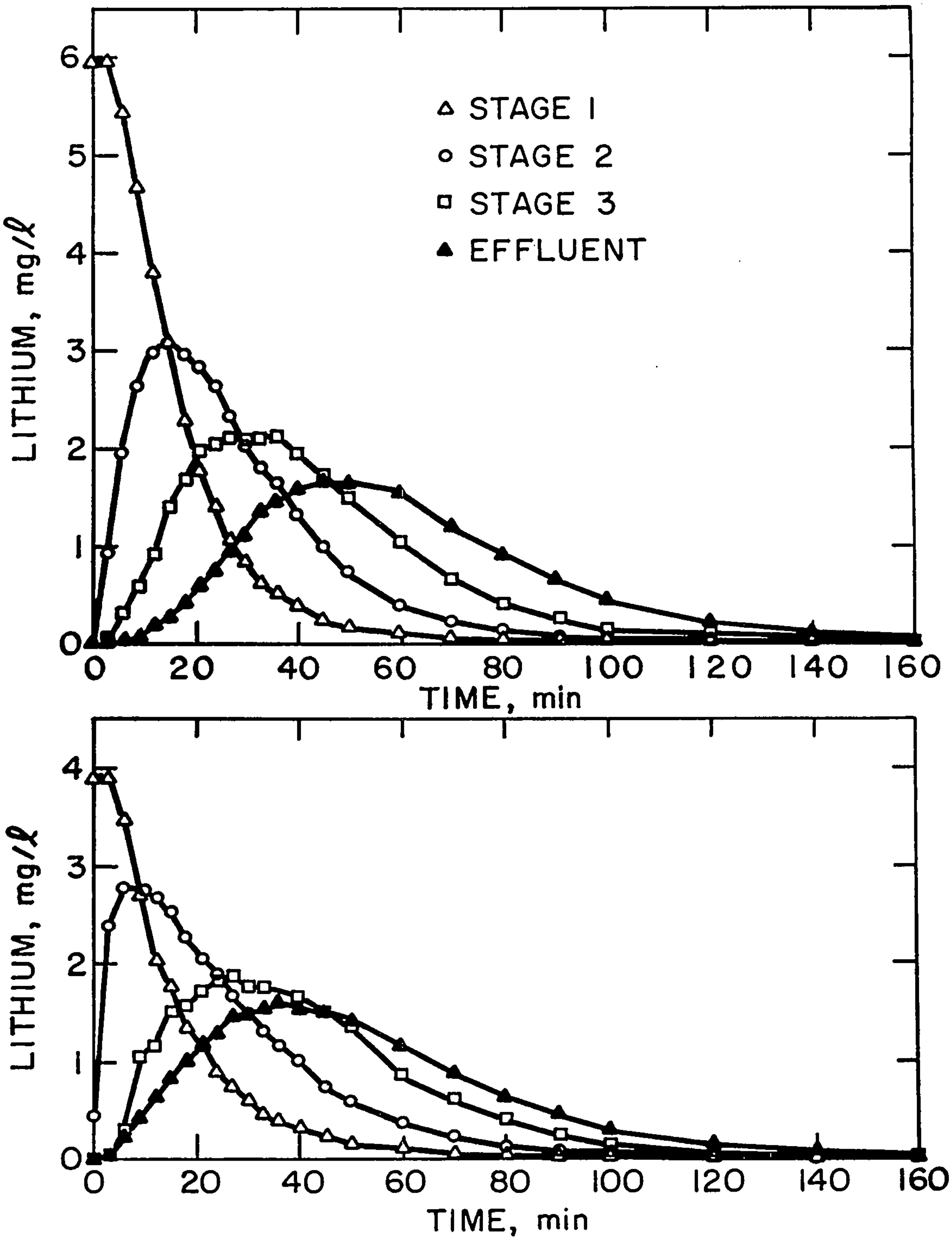

Figure 5. Recovery of lithium chloride tracer in each stage of 0.5-m (top) and 2.0-m (bottom) RBC units.

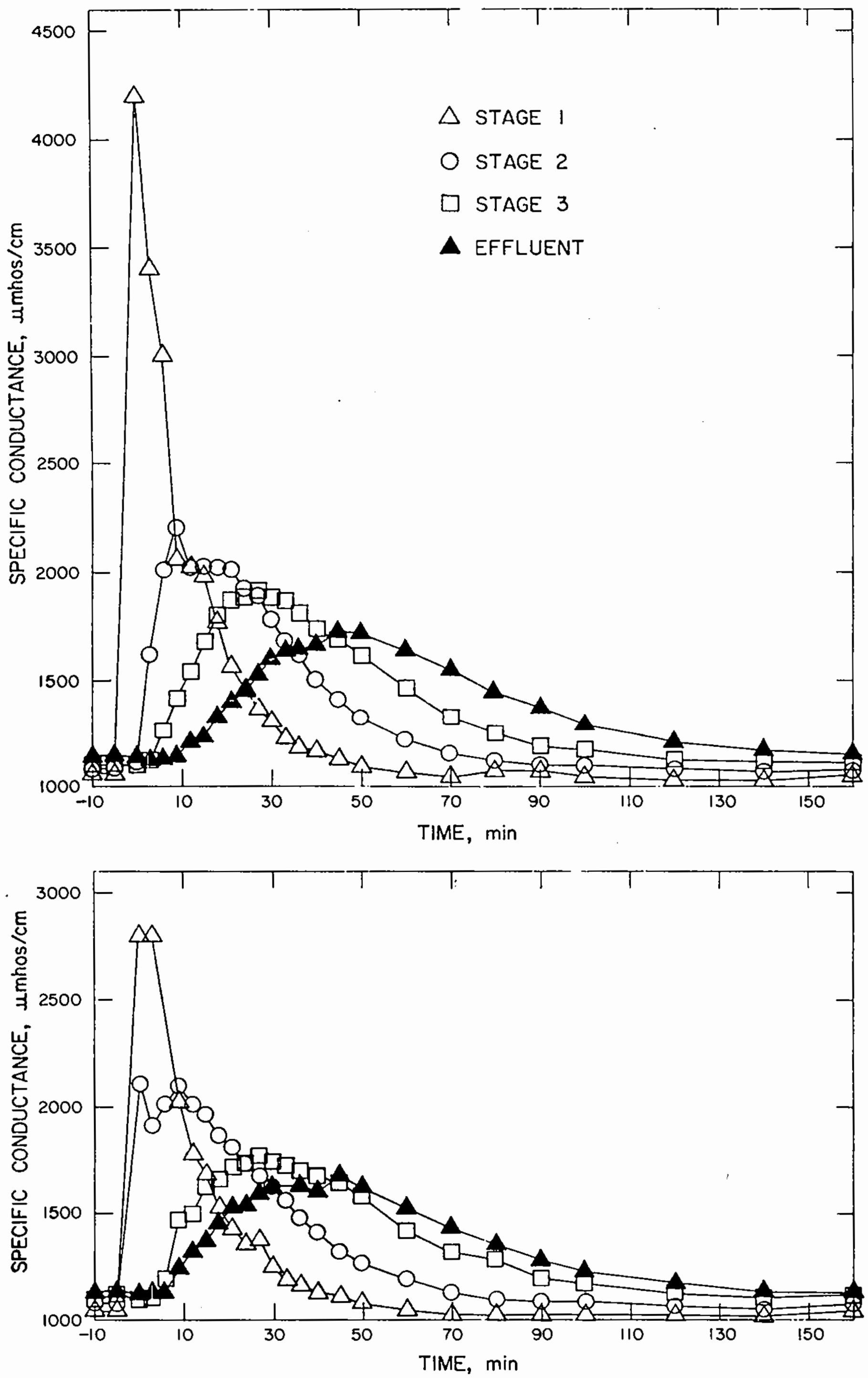

Figure 6. Recovery of sodium chloride tracer in each stage of 0.5-m (top) and 2.0-m (bottom) RBC units.

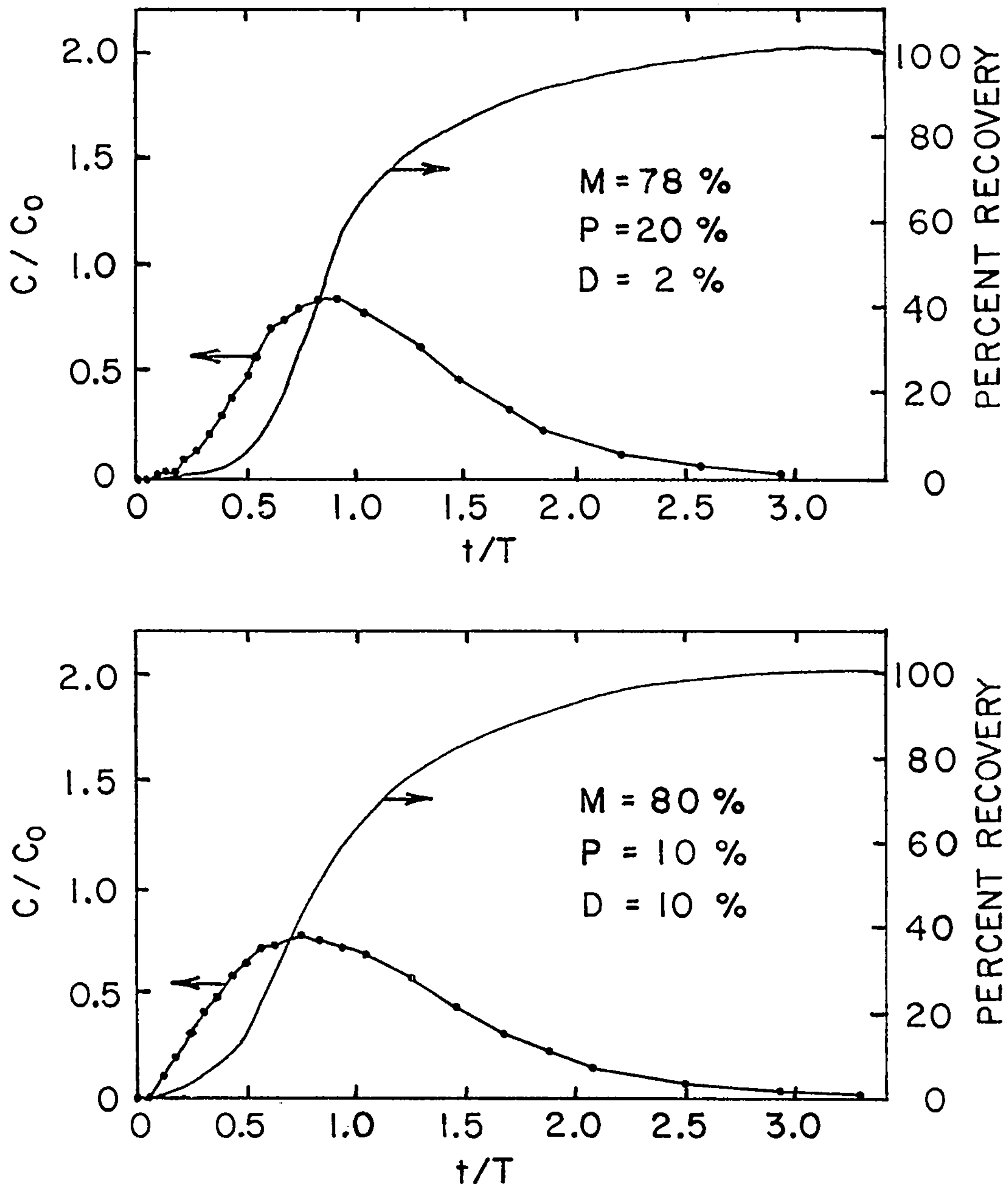

Figure 7. Lithium tracer recovery curves for effluent of 0.5-m (top) and 2.0-m (bottom) RBC units (M = mixed flow; P = plug flow; and D = dead space).

for 0.5-m unit. The larger RBC displayed only 10 percent plug flow characteristics while the 0.5-m unit had 20 percent plug flow volume. Pintenich and Bell (6) observed good performance for clarifiers with 15 to 20 percent plug flow characteristics. Baffled basins such as in the RBC should show similar plug flow characteristics. It was anticipated that the serpentine flow pattern of the 2.0-m RBC unit (Figure 4) would allow improved plug flow characteristics over the 0.5-m unit.

Villemonte and Rohlich (7) described certain dimensionless ratios which may be employed to evaluate the hydraulic efficiency of continuous flow reactors. These indices were applied to the results of lithium addition to RBC units (Tables 1 and 2). The ratio, t_p/T, is a measure of short-circuiting, dead spaces, and effective tank volume. It should be near zero for the first stage of the RBC (ideal mixing) and near unity for the final stage (ideal settling or plug flow). Comparison of this index for the two RBC units revealed more short-circuiting, dead spaces, and a lower effective tank volume in the 2.0-m unit. Another ratio, t_{90}/t_{10}, measures dispersion of reactor contents; a mixing function. The 2.0-m unit displayed near ideal dispersion of reactor contents in the first stage. The dispersion ratio decreased considerably for the latter stages of both units, presumably due to the baffled, multi-stage configuration.

Although RBC units were operated at equivalent hydraulic loading, different ratios of surface area-to-trough volume for the units resulted in different theoretical retention times (Table 1). The lower theoretical retention time for the 2.0-m RBC may also have been a factor in the observation of divergent treatment performance between the two units. In addition, there exists a sizeable difference in available surface area for the two RBC units (Table 3). Hydraulic loadings applied in this study were based solely on disc surface area. However, a greater proportion of the total surface area in the 0.5- than in the 2.0-m unit was attributed to the trough. Olem and Unz (8) observed similar viable iron-oxidizing bacterial densities in comparison of trough and disc surfaces of the same stage. Thus, calculation of the effective surface area of smaller diameter RBC units should include trough surfaces. It is likely that if the units had been sized on the basis of total available surface area, closer agreement would have been obtained in tracer evaluations and comparative treatment performance.

Friedman et al. (5) concluded that new, lightly loaded RBC systems should be able to meet design expectations when sized on the basis of pilot plant data. This would only be applicable to lightly loaded systems because differences in oxygen transfer between pilot and full scale would be less important. Similarly, differences in oxygen transfer for the two different size RBC units in treatment of acid mine drainage were likely very low owing to the high reactor dissolved oxygen levels present. The combination of a longer residence time for mine water in the trough of the smaller RBC and relative differences in total available disc surface areas probably accounted for much of the difference in performance observed between the two units. The exact contribution of observed hydraulic differences to the divergent treatment results is not known.

Table 1. Cumulative flow through times for lithium tracer in stages of 0.5- and 2.0-m RBC units

	Time, min							
	0.5-m RBC Stages				2.0-m RBC Stages			
Parameter	1	2	3	4	1	2	3	4
t_p[a]	3	15	36	47	3	6	27	37
t_{10}[b]	1.9	8.1	17.3	28.9	1.8	5.1	15.0	20.1
t[illegible][c]	33.9	54.3	77.0	98.7	36.1	53.4	76.3	96.5

Table 2. Comparison of dimensionless time ratios for 0.5- and 2.0-m RBC units.

Parameter	0.5-m RBC Stages 1	2	3	4	2.0-m RBC Stages 1	2	3	4
t_p/T [a]	0.22	0.55	0.88	0.86	0.25	0.25	0.75	0.77
t_{90}/t_{10} [b]	17.8	6.7	4.5	3.4	20.1	10.5	5.1	4.8

[a] Measures average short-circuiting, dead spaces, and effective tank volume. It is 1.0 for ideal settling and zero for ideal mixing.
[b] Measures dispersion. It is 1.0 for ideal settling and 21.9 for ideal mixing.

Table 3. Comparison of available surface area for different size RBC units.

RBC Unit	Disc Diameter	Surface Area, m^2 Disc	Trough[a]	Portion of Total Surface Area Due to Trough, Percent
Bench Scale	15 cm	0.438	0.10	19.1
Pilot Scale	0.5 m	21.8	2.14	8.9
Prototype	2.0 m	738.1	18.8	2.5
Full Scale	3.6 m	9,290	50	0.5

[a]Available surface area was calculated from dimensions of experimental RBC units and estimated for full-scale RBC by use of one 7.6 m shaft placed in a contoured basin.

SUMMARY AND CONCLUSIONS

Two conservative tracers were employed to hydraulically characterize two parallel RBC units with different disc diameters under equivalent conditions in order to determine possible explanations for the observation of lower ferrous iron removal efficiencies in the treatment of acid mine drainage. The observed decrease in treatment performance for scale-up to the 2.0-m RBC was due, in part, to the following factors:

1. The 0.5-m RBC more closely simulated the ideal hydraulic characteristics of a series of completely mixed treatment basins than did the 2.0-m unit.

2. The residence time of mine water in the 0.5-m RBC trough was longer than the larger unit when operated at equivalent hydraulic loading rates (54.4 min vs. 48.0 min) due to differences in surface area-to-trough volume ratios for the two units.

3. The proportion of total RBC surface area assumed by the trough floor was greater in the 0.5-m RBC than in the 2.0-m RBC. In the calculation of hydraulic loading, only the disc surface area is considered. Since the trough area, which is colonized by active iron oxidizing bacteria, does not enter into the determination of effective surface area, it is apparent that the 0.5-m RBC would demonstrate the greater treatment capacity of the two units.

ACKNOWLEDGEMENTS

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PHYSICAL FACTORS IN RBC OXYGEN TRANSFER

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Introduction

Although the Rotating Biological Contactors (RBC) process is classified as an aerobic system, oxygen transfer rates have not been sufficiently considered in present design. Rather, the emphasis has been centered on an increase in the RBC surface area.

When Hartman (1960) introduced the practical use of RBC, he claimed that the dissolved oxygen (DO) in the reactor did not have significance in treatment efficiency because adequate amount of oxygen could be supplied during the air exposed cycle. The importance of the DO in the mixed liquor has been cited by workers in defining a minimum oxygen balance. Welch (1968) presented data showing considerable decline of treatment efficiency when operational DO concentration dropped below 1.5 mg/l for his operating conditions of 500 mg/l COD and 30 minutes retention time. Weng and Molof (1974) found that nitrification took place only in the stages where DO was greater than about 2 mg/l in a six stage laboratory reactor for operating condition of 178 mg/l COD and 48 minutes per stage retention time. However there has not been a corresponding emphasis on the RBC as an aeration device to deliver the oxygen to maintain this DO level. Furthermore, treatment of high BOD industrial wastewater requires more study of the factors involved in RBC oxygen transfer.

The RBC biological film is fixed on the disk surface and is rotated alternately between the air and the liquid. One effect of rotation is to provide a means of better aeration by carrying a liquid film into the air after the disk completes its liquid immersion cycle.

Another effect is to provide turbulence in the mixed liquor surface and subsurface volume increasing mass transfer. Oxygen transfer takes place at the interfaces between the air-liquid film, liquid film-microbial fixed film, air-mixed liquor, microbial fixed film-mixed liquor and air-microbial fixed film.

This work involves a study of some of the physical factors affecting RBC oxygen transfer into the mixed liquor. Three different sized laboratory scale RBC units were used in non-steady state clean water tests. The volumetric oxygen transfer coefficient $(K_L a)_{20}$ was calculated from the laboratory data. The physical parameters studied included space between the disks, size of the disks, rotational velocity, peripheral velocity and number of disks per stage. It was the purpose of the study to correlate the volumetric oxygen transfer coefficient $(K_L a)_{20}$ with significant physical factors. As a result, the volume renewal number (N_V) is developed from the theory and data as a more practical efficient tool to predict physical oxygen transfer in the RBC process.

RBC Oxygen Transfer

Bintanja et. al (1975) used the Yamane and Yoshida solution to solve Fick's second law. The boundary conditions were:

$$t = 0, \quad 0 < x < \delta, \quad c = C_L$$

$$t > 0, \quad x = \delta, \quad c = C_S$$

$$t > 0, \quad x = 0, \quad \partial c/\partial x = 0$$

where δ was the liquid film thickness on the disk. They concluded that:

$$K_L = 2\left(\frac{D_m}{\pi t_R}\right)^{0.5} \text{ when } \frac{\delta}{(D_m t_R)^{0.5}} \geq 1.7 \qquad (1)$$

$$K_L = \frac{2\alpha}{(\pi)^{0.5}} \cdot \frac{\delta}{t_R} \approx \frac{\delta}{t_R} \text{ when } \frac{\delta}{(D_m t_R)^{0.5}} < 0.8 \qquad (2)$$

Experimental K_L value were 49% to 87% of the theoretical K_L value.

Chesner and Molof (1976, 1977) found that the smaller RBC with the higher RPM had better efficiency than the larger RBC with lower RPM. The peripheral velocity was set at the same level that is used in present plant design. They also reported that rotational velocity was the better DO scale-up function than the peripheral velocity.

Friedman et. al (1979) confirmed the significance of rotational velocity and presented an equation using Bintanja et. al (1975) data;

$$\ln K_L = 1.31 \ln \omega + 14.78$$

where the unit of K_L was 10^{-6} m/s and that of ω was RPM.

Zeevalkink et. al (1978) solved the Navier-Stokes equation for determining liquid film thickness on the RBC and also verified the equation by experiment. They concluded that:

$$\delta = 1.2 \cdot (v_c)^{0.5} \cdot (10^{-4} m) \tag{3}$$

where v_c is the vertical component of the peripheral velocity at the point where the disk emerges from the water.

Ouano (1978) correlated the overall liquid phase mass transfer coefficient (K_L) and Reynolds number by dimensional analysis. The result was:

$$K_L \cdot \frac{V_E/A_t}{D_m} = K\left(\frac{A_t}{A_p}\right)^a \left(\frac{D^2 \omega \rho}{\mu}\right)^b$$

where $\frac{D^2 \omega \rho}{\mu}$ was Reynolds number.

Zeevalkink et. al (1979) explained that the deviation of the Bintanja et. al (1975) model was due to incomplete mixing in the reactor and derived an experimental equation for

$$0.8 < \frac{\delta}{(D_m t_R)^{0.5}} < 1.7$$

as $$K_L = 2 \left(\frac{D_m}{\pi t_R}\right)^{0.5} \left(1 - 4.21 \exp \frac{\delta}{(D_m t_R)^{0.5}}\right)$$

Theoretical Considerations

In view of the present state of technology, there appears to be a need for development of a more practical method of predicting oxygen transfer in the RBC process. The volume renewal number method is an attempt to meet this condition.

To introduce the concept of the volume renewal number, the following assumptions were made.

1. Oxygen is only transferred via the liquid film on the disk during the air exposed cycle.

2. The contact time is sufficient so that the liquid film always has the same degree of saturation after completing the air exposed cycle.

3. The liquid film thickness is determined by the Zeevalkink et. al (1978) model.

The volume renewal number can be defined as the ratio between the liquid film flow rate and the effective reactor volume, or

$$N_V = Q_f/V_E \tag{4}$$

where, N_V: volume renewal number (T^{-1})

Q_f: liquid film flow rate, which is the total film flow volume per unit time (L^3T^{-1})

V_E: effective reactor volume (L^3)

Under the above assumptions, correlation of the volumetric oxygen transfer coefficient (K_La) and the volume renewal number (N_V) is expected as

$$K_La = a\ (N_V)^b \tag{5}$$

It should be noted that the use of a half circle submerged reactor is an ideal case in order to simplify the calculations. In case of any other physical conditions, the constants a and b will be changed accordingly. The conceptual scheme is shown in Fig. 1.

The liquid film flow rate can be derived by;

$$Q_f = K_1\ (\omega \cdot D^2) \cdot (\omega \cdot D)^{0.5} = K_1\ \omega^{1.5}\ D^{2.5} \tag{6}$$

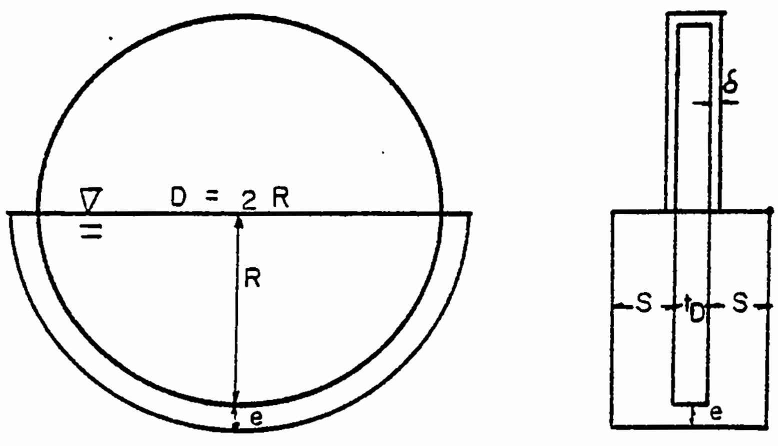

FIGURE I. CONCEPTUAL SCHEME

where $\pi/2 \cdot (\omega \cdot D^2)$ is a single disk surface area per unit time carrying the liquid film into the air and $(\omega \cdot D)^{0.5}$ is the liquid film thickness (δ) function from equation (3).

The effective volume is:

$$V_E = K_2 \cdot D^2 \cdot S. \qquad (7)$$

where the liquid volume underneath the disks, $\pi \cdot e \cdot (R + e/2) \cdot (2S + t_D)$, is neglected.

Therefore, the volume renewal number (N_V) is obtained by combining equation (6) and (7) without including constants K_1 and K_2.

$$N_V = \frac{Q_f}{V_E} = \omega^{1.5} \cdot D^{0.5} \cdot S^{-1} \qquad (8)$$

Substituting equation (8) into equation (5) results in;

$$K_L a = a(N_V)^b = a(\omega^{1.5}\ D^{0.5}\ S^{-1})^b \qquad (9)$$

Experimental

Three different size laboratory scale geometrically similar RBC units were designed to measure the dissolved oxygen concentrations with time under various physical conditions. The schematic drawing of the units and dimension data are shown in Figure 2 and Table 1 respectively. The disks were 6, 12, and 24 inch diameter flat circles made from 0.25 inch thickness (t_D) clear plastic.

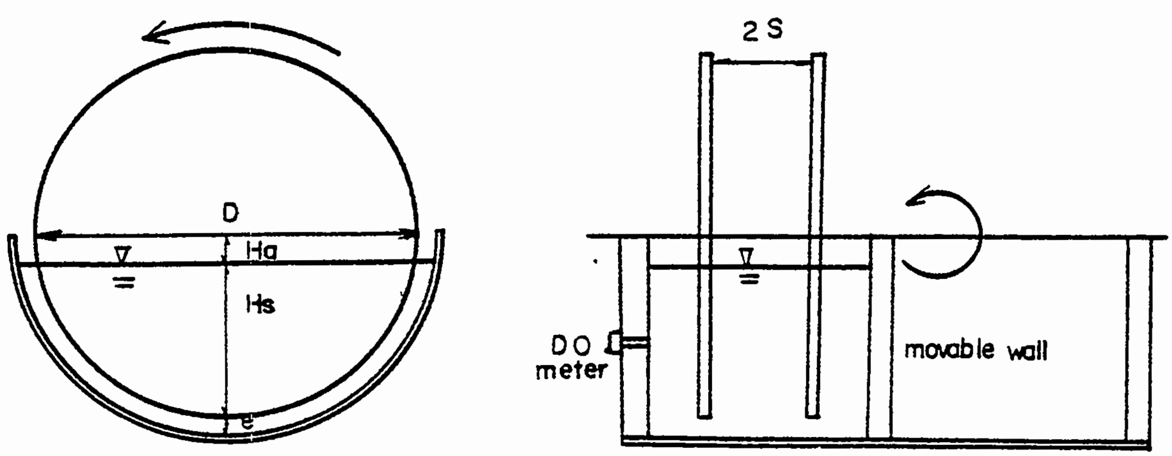

FIGURE 2. LABORATORY SCALE R B C UNITS.

Table 1. Reactor dimensions (unit: inch)

D	6.0	12.0	24.0
H_a	0.47	0.94	1.88
H_s	2.53	5.06	10.12
e	0.125	0.25	0.50
$R_T=H_a+H_s+e$	3.125	6.25	12.50
L	12.0	24.0	48.0

Movable walls and spacers were provided to vary the effective volume of liquid (V_E). Effective liquid volume in the case of a single disk at 85% submerged depth is shown in Table 2.

Table 2. Effective volume of reactor, unit-ml

S \ D	6 inch	12 inch	24 inch
(1/4) (1/4)	106	420	1700
(1/2) (1/2)	208	830	3330
(3/4) (3/4)	310	1240	4960
(1) (1)	412	1650	6590
(3/2) (3/2)			9840

The number of disks and the space between the disks were varied. The space between disks was twice the space between the disk and wall. Rotating velocity was varied from 5.2 to 164 RPM by a gear and sprocket combination connected to a 1/6 HP A.C. motor. The standard design peripheral velocity is about 1 ft/sec. The peripheral velocity in this study ranged from 0.5 to 4 ft/sec.

The physical variables are summerized in Table 3.

Table 3. Physical variables.

D (inch)	6.0	12.0	24.0
N (No. of Disks)	1-3	1-4	1-3
ω (RPM)	84-21	84-10.5	42-5.2
V_D (ft/sec.)	4.0-0.5	4.0-0.5	4.0-0.5
S (inch)	0.25-2.0	0.5-2.0	0.5-3.0

The oxygen transfer test was basically the same as the non-steady state clean water test procedure in Standard Methods (1975). New York City (Brooklyn, N.Y.) tap water was aerated about twelve hours to reach oxygen saturation. The water and room temperature was maintained at 20 ± 0.5°C. The DO meter (Beckman 777 oxygen analyzer) was calibrated with the oxygen saturated water. Winkler tests were performed to confirm the oxygen saturation. Reagent grade sodium sulfite and cobalt chloride were used to deoxygenate the water. The test run was over when the reaerated sample become about 90% to 95% saturated. Since accurate saturation values were known, the log deficit least square method was used to analyze the data. The temperature effect was compensated for by using

$$(K_La)_{20} = (K_La)_T \cdot (1.024)^{(20-T)}$$

Results

The volumetric oxygen transfer coefficients (K_La) were calculated from the experimental data under the various physical conditions.

The results of measured K_La are presented graphically on full log paper in Figure 3,4 and 5. Figure 3 depicts the disk size (D) effect on K_La. A linear relationship between the disk diameter (D) and K_La was shown. The averaged slope of 0.39 indicates that K_La is proportional to $D^{0.39}$. Figure 4 shows the effect of space (S) between disks on K_La. The average slope was - 0.86. K_La is proportional to $S^{-0.86}$.

Fig. 5 depicts the rotational velocity (ω) effect on K_La. The slope of the lines was not constantly linear in contrast to the results for D and S. The reason appeared to be a difference in hydraulic regime. A discontinuity in the liquid film flow occured somewhere on the dotted line in Figure 5 and K_La was drastically decreased under this hydraulic condition. The average slope of 1.07 indicates that K_La is proportional to $\omega^{1.07}$. The number of disks each having the same volume/area relationship was varied to detect the wall effect and turbulence characteristics. No appreciable effect on K_La was noticed.

The combination of effects of the rotational velocity, the disk size and the disk space resulted in:

$$K_La = K \cdot \omega^{1.07} \cdot D^{0.39} \cdot S^{-0.86} \qquad (10)$$

The equation coefficients strongly suggest that the experimental K_La correlates well with the proposed N_V. Figure 6 depicts the relationship between the calcualted N_V ($\omega^{1.5}$ $D^{0.5}$ S^{-1}) and K_La on a full log paper. The relationship can be expressed as

$$\log K_La = 0.732 \log (N_V) - 2.96$$

$$\text{or} \quad K_La = 0.0011 (N_V)^{0.732} \qquad (11)$$

with the correlation coefficient (r) of 0.991. This equation is valid for the clean flat disks, $e/R = 0.042$ and $H_a/R_T = 0.15$

Discussion

One of Bintanjal et. al (1975) solutions was equation (2);

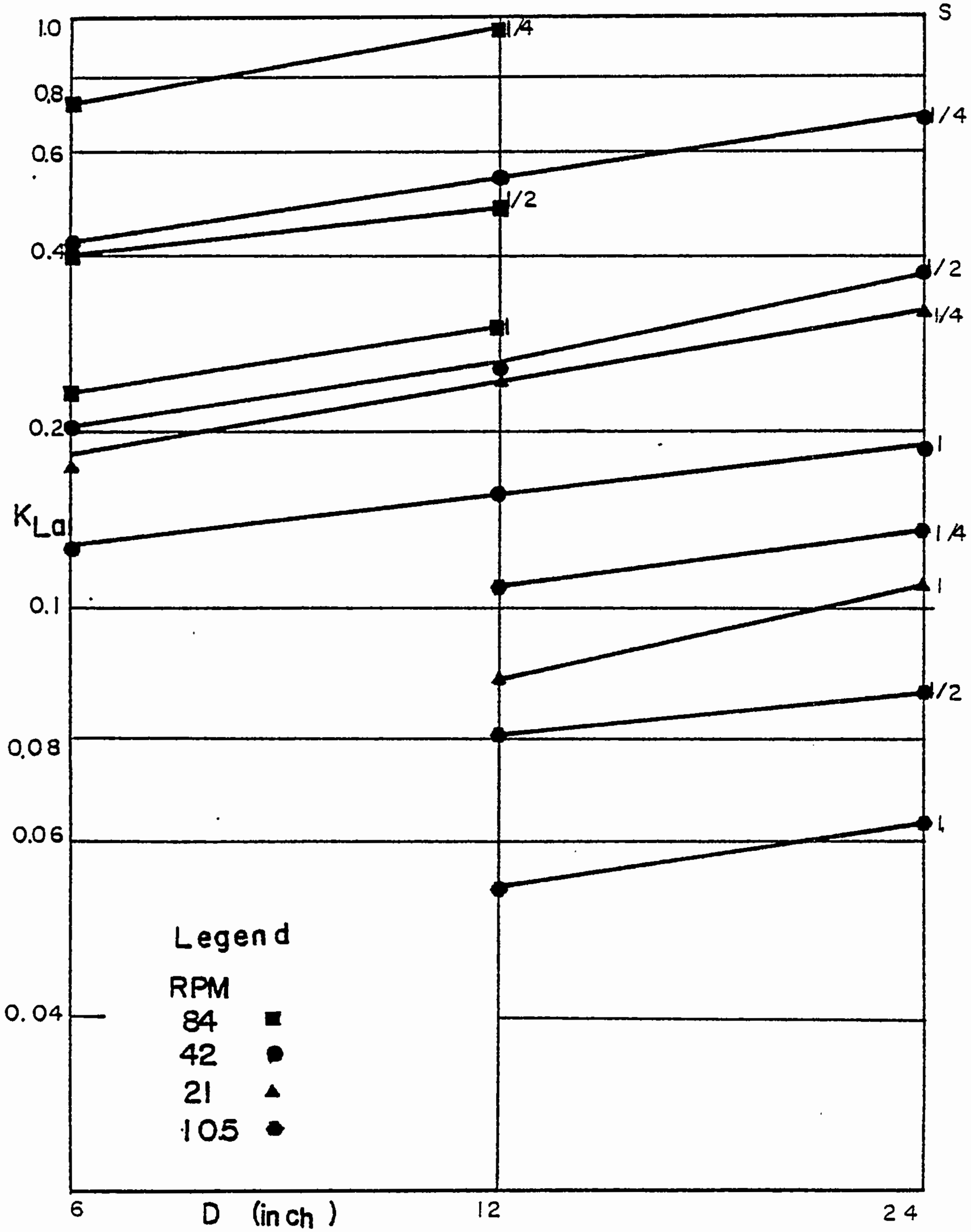

FIGURE 3. DISK SIZE EFFECT

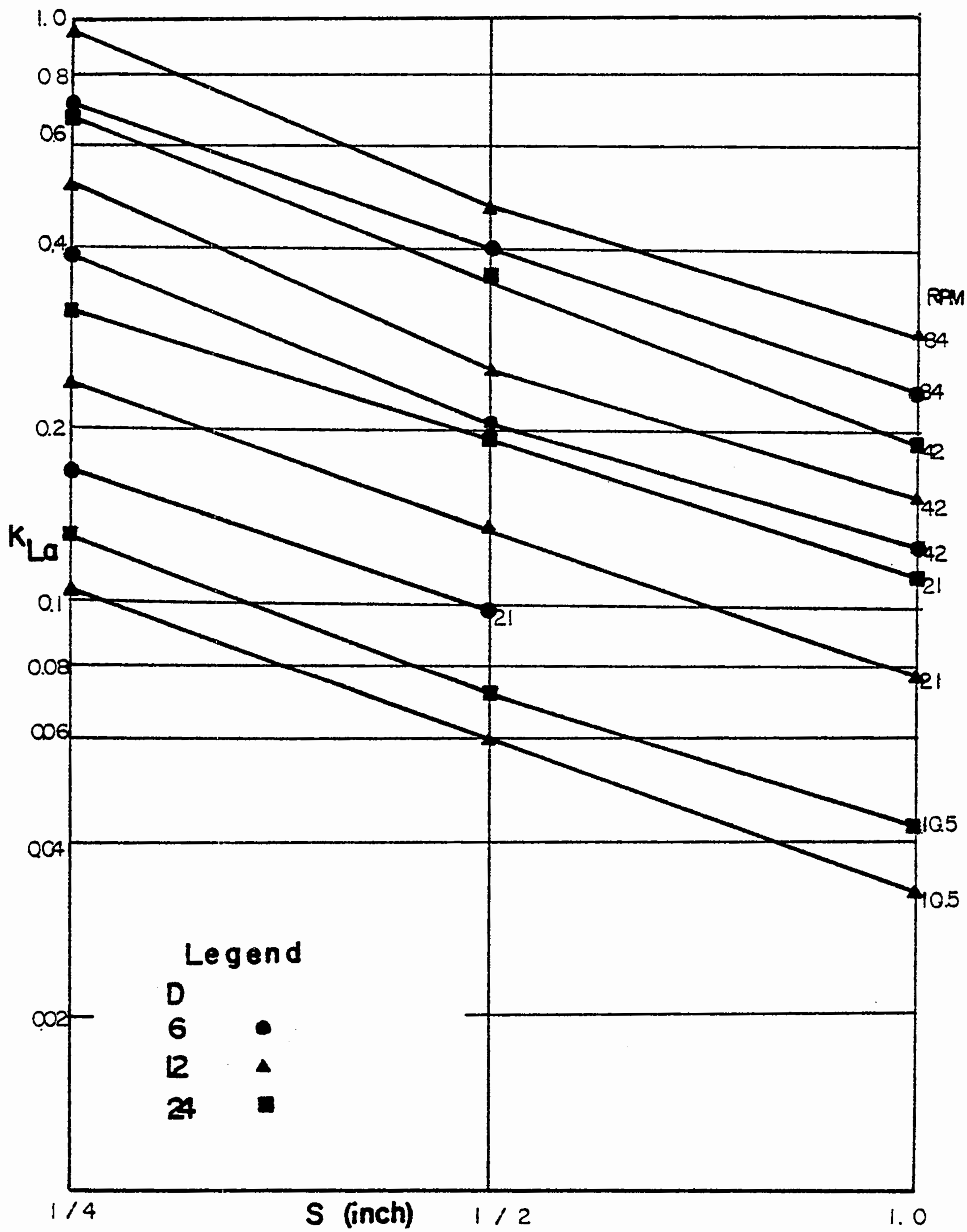

FIGURE 4. SPACE EFFECT

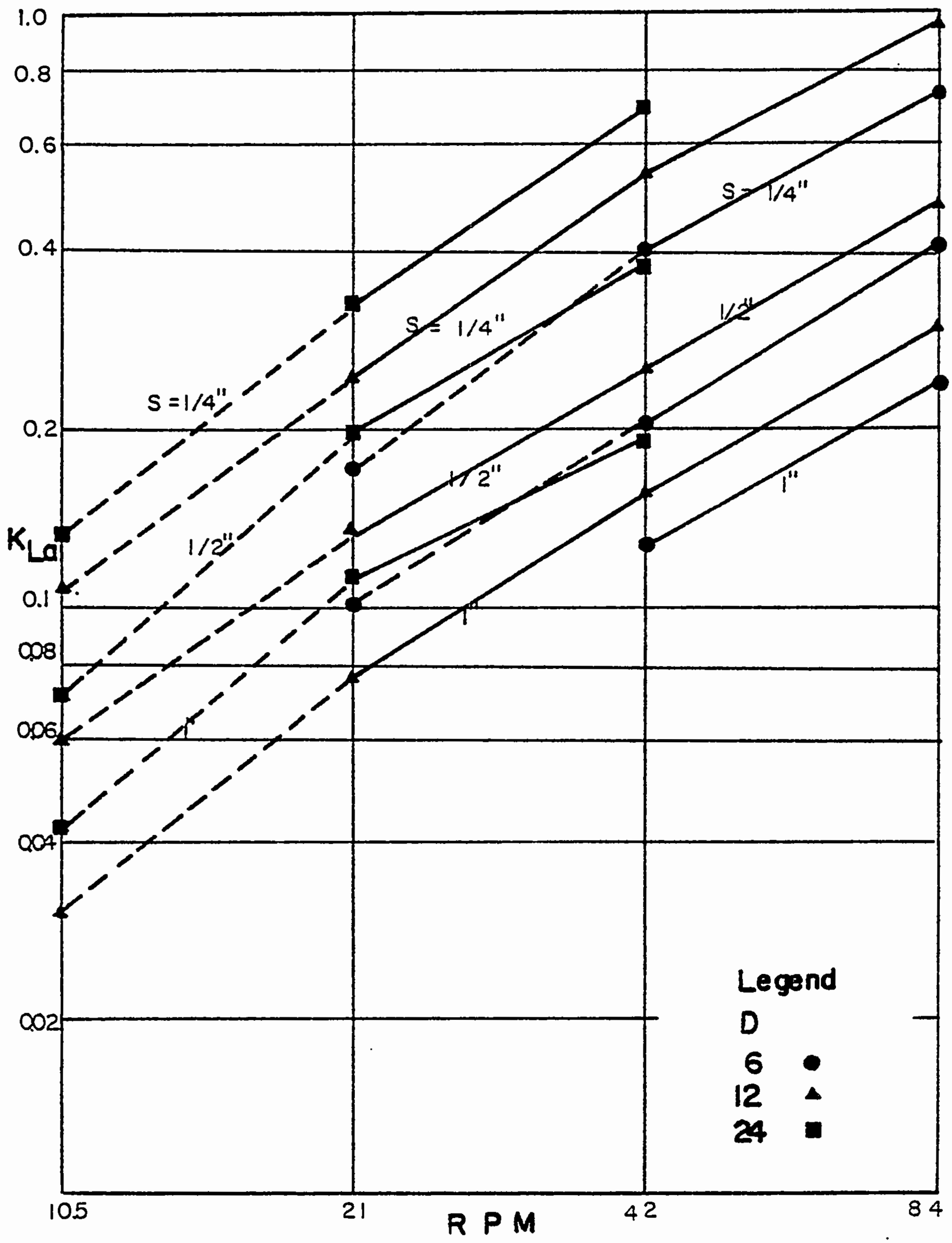

FIGURE 5. ROTATIONAL VELOCITY EFFECT.

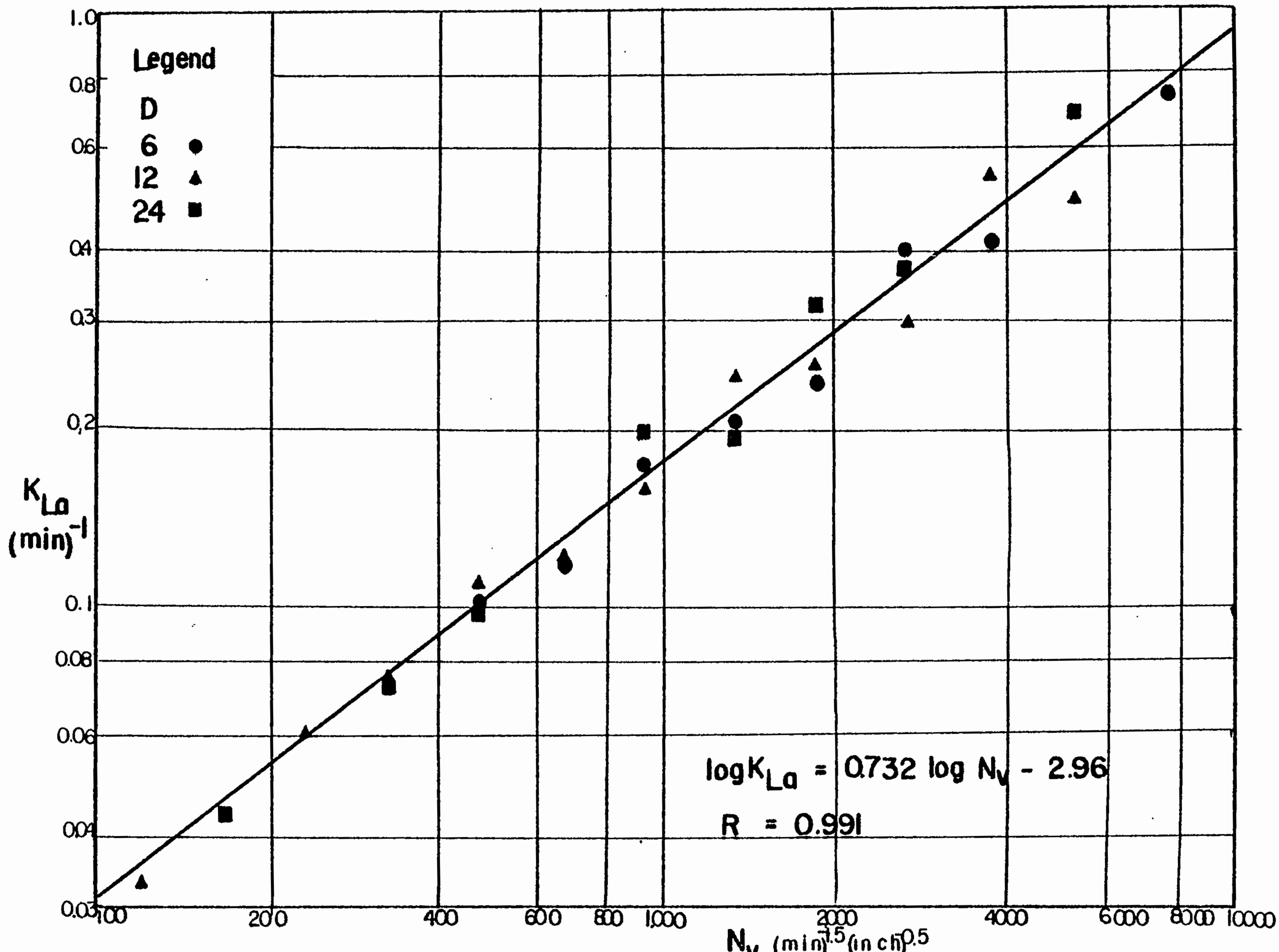

FIGURE 6. RELATIONSHIP BETWEEN K_{La} & N_v.

$$K_L = \frac{\delta}{t_R} \quad \text{when} \quad \frac{\delta}{(D_m\ t_R)^{0.5}} < 0.8 \qquad (2)$$

By using Zeevalkink et. al (1978) data for ω and δ, it can be calculated that at $\delta/(D_m\ t_R)^{0.5} = 0.8$ and 1.7, the corresponding peripheral velocity was about 1 ft/sec. and 2.2 ft/sec, respectively. As the disk size increases, t_R increases. Therefore, $\delta/(D_m\ t_R)^{0.5}$ will be less than 0.8 in most cases at plant scale . Thus, it appears that equation (2) would better represent full scale plant performance.

A comparison of the Bintanja et. al equation (2) with the volume renewal number (N_V) would be of interest. Since δ is proportional to $(\omega D)^{0.5}$ and t_R is inversely proportional to ω, K_L in equation (2) can be expressed as;

$$K_L = K\ \omega^{1.5}\ D^{0.5} \qquad (12)$$

The common functions are noticed in equation (8) and (12). This shows a possible application of the volume renewal number (N_V) concept to full scale plant operation.

Ouano (1978) obtained a straight line by plotting

$$\frac{K_L[V_E/A_t]}{D_m} \quad \text{and} \quad \frac{D^2\omega\rho}{\mu}$$

on full log paper. Since he did not vary the disk size, the plotting was simply the relationship of K_L and ω. The N_V concept shows that the ω term is more important than the D term. This is in contrast to the Ouano equation where D is more important than ω.

Conclusion

The volumetric oxygen transfer coefficient (K_La) in the RBC process was correlated with the volume renewal number (N_V) or $K_La=a(N_V)^b$. The volume renewal number represented the liquid film flow rate per unit effective reactor volume. The important physical factors in calculation of the volume renewal number included the rotational velocity (ω), the disk size (D) and the space between the disks(S). The volume renewal number (N_V) was defined as

$N_V = \omega^{1.5}\ D^{0.5}\ S^{-1}$.

The relationship of K_La and N_V under the conditions of $e/R = 0.042$, $H_a/R_T = 0.15$ and a clean flat disk is:

$$\log K_La = 0.732 \log (N_V) - 2.96, \text{ or}$$

$$K_La = 0.0011 (N_V)^{0.732}$$

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Nomenclature

A	:	Area where the mass transfer occurs (L^2)
A_p	:	Projected area of disk on water surface (L^2)
A_t	:	Surface area of reactor (L^2)
a,b,K	:	Constant.
D	:	Diameter of disk (D)
D_m	:	Molecular diffusion coefficient (L^2T^{-1})
e	:	Distance from the disk rim to inner lining of reactor (L)
H_a	:	Distance from the disk center to the liquid free surface (L)
H_s	:	Submerged disk depth (L)
K_L	:	Overall liquid phase mass transfer coefficient (LT^{-1})
K_La	:	Volumetric Oxygen transfer coefficient (T^{-1})
L	:	Reactor length (L)
N	:	Number of disks
N_V	:	Volume renewal number (T^{-1})
Q_f	:	Film flow rate (L^3T^{-1})
R	:	Radius of disks (L)
R_T	:	Radius of reactor (L)
S	:	Half space between disks (L)
t_R	:	Contact time per rotation (T)
v_c	:	Vertical component of the peripheral velocity at the point where the disk emerges from the water (LT^{-1})
V_E	:	Effective reactor volume (L^3)
v_p	:	Peripheral velocity of disk (LT^{-1})
δ	:	Liquid film thickness (L)
μ	:	Absolute viscosity of liquid (MLT^{-1})
ρ	:	Density of liquid (ML^{-3})
ω	:	Rotational velocity (T^{-1})

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EFFECT OF CARBON, AMMONIA NITROGEN AND HYDRAULIC LOADING RATES, RPM, AND EXPOSED SURFACE AREA VARIATIONS ON RBC PERFORMANCE

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INTRODUCTION

Sometimes a system being studied is well understood, and it is possible to postulate a plausible model from considerations of the governing physical laws. This type of model would be a theoretical or mechanistic model because it is based directly on the theory governing the process. However, the case often arises where the system is too complicated, or the important physical laws are not known, and a theoretical model cannot be postulated. In this case an emperical model might be of value. The responses of the system to changes in conditions often helps determine which physical laws are governing. Then an attempt can be made to postulate theoretical models. Even before an emperical model can be constructed, it may be necessary to determine which are the controlling factors, and what their effects are on the system. The experimental program reported on in this paper is such a screening study, designed to determine which of a number of variables are important ones, and what their effects are on process performance.

In such an experimental program it might appear necessary to study every factor over its entire range. This experimental design would examine many combinations of all factors at many levels. This is an inefficient and expensive way to experiment. In many cases a more realistic approach would be to design a sequence of more modest experiments so that variables can be dropped or added as information about the system is gained. A two level factorial design is often used, and this was the approach chosen by the authors.

Rotating Biological Contactor Study

Basic and independent factors which possibly effect RBC effluent quality and performance are hydraulic flow rate, carbon mass flow rate, ammonia nitrogen mass flow rate, percent of disc area exposed to the atmosphere at any given time, and disc revolutions per minute (rpm). These factors were studied in this experimental program. Influent concentrations were not chosen as variables for study because they are not basic factors, but are derived factors, equal to mass flow rates divided by hydraulic flow rates.

If experiments are run with each of the five variables set at only two levels, there are $2^5=32$ possible combinations of factors. From these 32 experiments the effects of changing variables one, two, three, four, and five at a time can be determined using standard experimental analyses. An excellent reference is Statistics for Experimenters, by Box, Hunter and Hunter (Wiley, 1978).

It is generally accepted that single factor and two factor interaction effects are greater than the higher order interaction effects. If this assumption is made, then sixteen or one-half 2^5 experiments can be run, using one half of all the possible combinations of five factors at two levels. From such a study, single and two factor interaction effects can be determined. However, the experimenter is gambling that the higher order effects are not significant. Before discussing levels of factors, it is necessary to discuss the experimental RBC because many of the currently accepted values for factor levels are based on RBC disc diameter and surface area.

Experimental RBC

It was felt that the results of the experimental program would be more applicable to full scale RBCs the larger the discs on the experimental RBC were. Three foot diameter discs were chosen after consideration of the logistics of operating various size systems. A three foot diameter disc has 1/16 the area of a full scale 12 ft. diameter disc, and 9 times the surface area of a typical laboratory unit.

The RBC used in this study has four stages, each stage with five discs, three feet in diameter. The total disc surface area is 284 square feet, or 71 square feet per stage. The discs were made of 1/8 inch thick plexiglass, and were spaced 1/2 inch apart. The tanks in which the discs rotate are 37 inches in diameter, and 4.5 inches wide per stage. Constant hydraulic flow rates were maintained through the use of a constant head tank. Constant feed rates were achieved through the use of multi-channel perfusion pumps. Flat plexiglass discs were used because the hydraulics of the flow on the discs would be easier to model later on, if necessary, and there is a much better definition of disc surface area. It was also felt that flat discs would give a baseline to which other types of disc surfaces could be compared.

Factor Levels

RBCs are often characterized on the basis of peripheral velocity (length-time $^{-1}$), hydraulic loading rates (volume-area $^{-1}$ - time $^{-1}$) and organic loading rates (mass-area $^{-1}$ - time $^{-1}$). Previous studies have suggested that peripheral velocity of 60 ft./min. is desirable. The experimental RBC was operated at peripheral velocities of 56 and 76 feet per minute, corresponding to 5.94 and 8.10 rpm. These rpms were dictated by gear ratios available.

Common hydraulic loading rates are in the order of 1 or 2 gallons per day per square foot of disc area. The experimental RBC had hydraulic loading rates of 0.96 and 1.63 gallons per square foot per day. These loading rates were obtained by maintaining flows at 1032 and 1750 liters per day.

Organic loading rates range from 0.25 to 10 pounds of BOD_5 per one thousand square feet per day. The loading rates used in this experiment were 258 and 500 grams of glucose per day, which are theoretically equivalent to 3.6 and 7 pounds of BOD_5 per thousand square feet per day.

Ammonia nitrogen mass loading rates are normally in the 0.1 to 0.2 pounds NH_3-N per thousand square feet per day range. By using NH_3-N mass flow rates of 41 and 65 grams per day, loading rates of 0.3 and 0.5 pounds NH_3-N per thousand square feet per day were achieved.

RBCs are usually constructed so that 50 to 60 percent of the disc surface area is exposed to the atmosphere at any time. The experimental RBC was built so that 60 or 74 percent of the disc area could be exposed by varying the water level in the tanks.

These five factors were varied systematically. The factor levels, condition numbers with corresponding coded factor levels and the order of experimentation are shown in Table 1.

EXPERIMENTAL PROCEDURES

RBC discs were rotated by a constant speed gear motor and a chain drive. RPMs were changed by changing sprockets. Hydraulic loading rates were varied by a valve in the water line from an aerated constant head tank. Tap water, which originated from shallow wells tapping an aquifer under a river, and receiving only chlorination was used as the feed water. Both the constant head tank and the water bath surrounding the RBC were heated to maintain a constant 20°C temperature. Percent of disc exposed was controlled by an adjustable overflow weir on the effluent side of the last stage. Carbon and nitrogen mass loading rates were controlled by varying concentrations of feed solutions which were pumped to the first stage by multi-channel perfusion pumps. The feed components were not all mixed together to prevent precipitation and to reduce the possibility of bacterial degradation. Table 2 shows feed solution compositions for both high and low level factor loading rates.

In addition to the carbon and nitrogen feed solutions, sodium bicarbonate and sodium hydroxide were added to the second stage to minimize pH and Alkalinity on nitrification. The pH of the second stage was maintained at pH 8.5 and effluent alkalinity was kept above 100 mg/l as $CaCO_3$.

TABLE 1

ONE-HALF 2^5 EXPERIMENTAL DESIGN

Factor Number	Factor	Level (-)	Level (+)
1	Revolutions per minute (RPM)	5.9	8.1
2	Hydraulic Flow Rate (liters/day)	1032	1750
3	Disc Surface Area Exposed (percent)	60	74
4	Nitrogen Mass Loading Rate (grams NH_3-N per day)	41.5	65.0
5	Carbon Mass Loading Rate (grams glucose per day)	258	500

EXPERIMENTAL CONDITION NUMBER	ORDER RUN	FACTOR LEVELS Factor 1	2	3	4	5
1	7	−	−	−	−	+
2	1	+	−	−	−	−
3	12	−	+	−	−	−
4	8	+	+	−	−	+
5	13	−	−	+	−	−
6	2	+	−	+	−	+
7	5	−	+	+	−	+
8	16	+	+	+	−	−
9	4	−	−	−	+	−
10	6	+	−	−	+	+
11	11	−	+	−	+	+
12	10	+	+	−	+	−
13	15	−	−	+	+	+
14	3	+	−	+	+	−
15	9	−	+	+	+	−
16	14	+	+	+	+	+

TABLE 2

FEED COMPOSITIONS

		Concentration (grams/Liter)	
		(+) Level	(-) Level
Carbon Feed Solutions			
Solution 1	$C_6H_{12}O_6 \cdot H_2O$	283.81	146.47
Solution 2	KH_2PO_4	119.325	61.577
	K_2HPO_4	242.520	125.15
Solution 3	$FeCl_3 \cdot 6H_2O$	0.1935	0.0999
	$CaCl_2$	2.064	1.065
Solution 4	$MnSO_4 \cdot H_2O$	3.339	1.664
	$MgSO4 \cdot 7H_2O$	33.3874	16.642
Nitrogen Feed Solution			
Solution 4	$(NH_4)_2SO_4$	194.112	100.162

The RBC was started up by pumping 18 liters of trickling filter effluent into the first stage over a 24 hour period. Constant carbon, ammonia and hydraulic flow rates were established and maintained. Effluent quality was monitored until steady state conditions were achieved.

Experimental runs all proceeded in the same manner. Factor levels were set as indicated by the experimental design. Because changes in nitrification will cause changes in pH and alkalinity, fourth stage pH and alkalinity, as well as nitrates, were monitored. When these values stabilized, it was assumed that the system was at steady state. During the initial phase of the experimental program a week was allowed to go by before this monitoring began. After three days of sampling we consistently found that the system had already reached steady state. During the later portion of the experimental program monitoring began immediately. We found that system steady state conditions were achieved in about four days.

Once steady state conditions had been reached the three to five day sampling program began. Samples were taken from the influent and first, second, third, and fourth stages. All samples were analyzed for filtrate COD (0.45 micron membrane filter), nitrates (electrode method), ammonia (electrode and distillation methods), total Kjeldahl nitrogen, pH, alkalinity, and total volatile solids. Influent and stage one, two, three, and four temperatures and dissolved oxygen concentrations were also determined. All analyses were done according to Standard Methods (14th Edition).

RESULTS

Experimental results which pertain to this report are presented in Table 3. A one-half 2^5 experimental design allows the experimenter to determine the effect of varying five different factors either one or two at a time. The effect reported is the average result of raising the factor or factors from the low (-) level to the high (+) level used in the experimental program while all remaining factors were held constant. In addition to analyzing the concentration data presented in Table 3, selected mass flow and mass removal (or generation) data were developed by multiplying concentrations or differences in concentrations by hydraulic flow rate. These generated data were also analyzed. All analyses are presented in Tables 4 through 7. "Factor" refers to the operational variable which, when raised from its (-) level to its (+) level caused, on the average, the effect listed in the table. "12", "13", etc., refers to raising 1 and 2 together or 1 and 3 together.

Caveats

The effects of changing the level of a factor, either by itself, or in combination with another factor, on a response whose value is dependent on the level of that factor should not be considered. For instance, the effects of changing flow rate on concentration should not be considered; since concentration is a function of the flow rate. Or, the effect of increasing carbon mass flow rate on COD mass removal or concentration should not be discussed.

If data is to be manipulated, the manipulations have to be done before the factorial analysis is performed. For example, efficiencies of removal must be calculated for each of the sixteen experimental runs and then analyzed to determine the effects of changing factors on efficiencies.

TABLE 3

EXPERIMENTAL RESULTS

CONDITION NUMBER	COD (mg/l)			NH_3-N (mg/l)			NO_3-N (mg/l)			D.O. (mg/l)	
	Influent	Stage 1	Stage 4	Influent	Stage 1	Stage 4	Influent	Stage 2	Stage 4	Stage 1	Stage 4
1	307.0	104.2	35.8	27.4	22.5	0.88	0.92	2.82	11.57	0.20	3.63
2	266.4	24.3	12.1	38.5	23.1	0.57	1.12	5.02	27.30	5.34	4.86
3	171.5	32.4	19.2	44.1	30.8	0.00	0.78	8.05	21.10	6.28	8.54
4	345.3	72.0	31.1	22.1	11.0	0.14	0.93	0.93	16.73	1.90	9.42
5	278.5	34.0	23.5	35.8	26.5	0.00	0.84	10.90	35.90	4.98	5.91
6	524.6	47.4	14.5	37.9	20.4	0.43	0.97	1.89	18.98	4.04	5.75
7	307.0	75.6	47.4	24.9	11.6	0.72	0.92	2.82	11.56	3.10	7.59
8	102.35	33.5	5.1	22.7	20.4	0.00	1.04	2.12	25.23	6.83	9.86
9	256.9	55.2	39.7	83.6	67.6	15.80	1.09	12.70	61.83	4.38	2.55
10	489.2	51.7	37.1	79.9	61.3	29.30	0.80	5.27	27.50	2.03	4.28
11	327.2	59.8	27.3	44.0	31.9	7.69	0.85	4.62	26.70	4.38	6.34
12	158.7	46.9	37.3	31.0	16.9	1.73	1.19	13.93	41.17	7.55	5.92
13	511.2	60.6	28.7	69.2	49.9	14.12	1.23	1.92	44.23	1.37	3.51
14	263.3	29.8	19.4	74.3	59.6	12.10	0.82	8.09	51.00	5.20	3.35
15	148.1	39.2	31.7	42.7	32.8	2.76	1.10	1.83	39.47	6.28	5.31
16	326.1	39.5	17.8	34.4	33.9	3.50	1.25	9.49	32.50	4.78	7.44

TABLE 4

FACTOR EFFECTS ON CHEMICAL OXYGEN DEMAND

	EFFECT			
	COD CONCENTRATION (mg/l)		COD MASS REMOVAL RATE (grams/day)	
FACTOR	Stage 1	Stage 4	Stage 1	Overall
AVERAGE	50.0	27.4	319	351
1	-14.1	-8.6	38	39
2	-1.0	-0.5	-6	7
3	-10.9	-5.2	25	24
4	-5.1	7.6	33	17
5	27.3	5.4	221	247
12	10.4	0.3	-36	-29
13	-0.7	-7.5	-30	-21
14	3.2	7.2	-20	-21
15	-7.9	-1.1	60	47
23	5.1	1.9	-53	-61
24	-1.9	-4.8	-11	-12
25	-3.6	2.4	55	56
34	-0.3	-3.2	-1	6
35	-5.7	-0.3	43	29
45	-16.4	-11.8	40	34

FACTOR NUMBER	FACTOR
1	RPM
2	FLOW RATE
3	SURFACE EXPOSED
4	NITROGEN MASS LOADING RATE
5	CARBON MASS LOADING RATE

TABLE 5

FACTOR EFFECTS ON AMMONIA NITROGEN

	EFFECT			
	NH_3-N Concentration (mg/1 NH_3-N)		NH_3-N Mass Removal Rate (grams/day)	
FACTOR	Stage 1	Stage 4	Stage 1	Overall
AVERAGE	32.51	5.61	14.1	51
1	-3.36	0.73	4.6	9
2	-17.71	-7.08	3.5	3
3	-1.26	-2.81	-6.3	-2
4	23.43	10.53	4.1	16
5	-4.40	2.98	-2.1	8
12	-2.83	-2.18	-4.1	-7
13	6.75	-1.12	-7.8	7
14	0.75	0.84	0.7	0
15	6.05	1.77	-2.8	5
23	3.29	2.17	-3.6	-2
24	-13.03	-6.83	-5.8	7
25	1.28	1.09	1.5	-1
34	0.87	-2.70	-0.1	6
35	-1.49	-2.00	1.6	3
45	4.40	2.58	-1.0	0

FACTOR NUMBER	FACTOR
1	RPM
2	FLOW RATE
3	SURFACE AREA EXPOSED
4	NITROGEN MASS LOADING RATE
5	CARBON MASS LOADING RATE

TABLE 6

FACTOR EFFECTS ON NITRIFICATION

FACTOR	EFFECT: NO_3-N Concentration (mg/l) Stage 2	NO_3-N Concentration (mg/l) Stage 4	NO_3-N Production Rate (grams/day) Stage 2	NO_3-N Production Rate (grams/day) Overall
AVERAGE	5.78	30.86	6.02	39.3
1	0.14	-1.37	0.68	0.1
2	0.60	-7.86	3.95	8.9
3	-1.79	3.00	1.55	3.9
4	2.88	19.63	3.76	25.6
5	-4.11	14.28	-2.53	17.4
12	2.15	5.82	3.09	7.4
13	0.90	0.51	0.10	0.4
14	3.79	-3.40	6.59	3.4
15	1.22	1.78	1.21	2.6
23	-1.02	-2.48	-2.32	-1.4
24	1.08	3.07	3.22	1.9
25	2.09	4.16	0.31	1.9
34	2.01	-0.75	2.18	0.1
35	2.41	3.20	3.29	2.9
45	0.30	-1.61	1.33	1.9

FACTOR NUMBER	FACTOR
1	RPM
2	FLOW RATE
3	SURFACE AREA EXPOSED
4	NITROGEN MASS LOADING RATE
5	CARBON MASS LOADING RATE

TABLE 7

FACTOR EFFECTS ON STAGE 1 DISSOLVED OXYGEN

FACTOR	EFFECT (mg/1 D.O.)
AVERAGE	1.29
1	0.84
2	1.70
3	0.57
4	0.41
5	-3.13
12	-0.58
13	0.44
14	-0.05
15	0.09
23	-0.35
24	0.81
25	-0.63
34	-0.74
35	0.63
45	0.42

FACTOR NUMBER	FACTOR
1	RPM
2	FLOW RATE
3	SURFACE AREA EXPOSED
4	NITROGEN MASS LOADING RATE
5	CARBON MASS LOADING RATE

The effects presented in a one-half 2^5 factorial analysis are the effects of changing the factors listed aliased with the effects of changing all the remaining factors not listed. For example, what appears to be the effect of changing factors 2 and 3 together could be the effect of changing 2 and 3, or the effect of changing 1, 4 and 5 together, or a combination of both effects. Remember, the experimenter is gambling that the single and two factor effects are greater than the three, four, and five factor interaction effects.

These experiments were not replicated, so there is no direct estimate of the variance. This means that there is no immediate way to determine which effects are statistically significant and which effects are merely due to random variations of system performance. This problem can be partially solved through the use of normal probability paper.

If the calculated effects were only due to random variations about a mean value, the effects would have been distributed approximately normal with mean zero. The mean would be zero because the effects are calculated from differences between pairs of responses, and, on the average, the difference between random numbers would be zero. Thus, the ordered effects, if not significant, would plot as a straight line on normal probability paper.

If the ordered effects do not plot as a straight line on normal probability paper, they are not distributed normal. If the removal of some effects allows the rest of the effects to plot as a straight line, the removed effects are not easily explained as chance occurrences, while the remaining effects are. The effects presented in Tables 4 through 6 were plotted on normal probability paper and analyzed. The results of these analyses are presented in Table 8, and are also discussed in the following section along with the results shown in Tables 4 through 7.

ANALYSES

When considering factor effects it is necessary to overlook the obvious effects and discuss that which may appear to be less important. One could guess that if flow rate was increased while mass loading rates remained the same, concentrations would decrease. Or, it could be anticipated that concentrations would increase if mass loading rates increase , flow being held constant. Similarly, mass removal rates would probably increase if mass loading rates were increased. Once the effects are calculated, it is very difficult, if not impossible, to rationalize apparent contradictions by multiplying effects of one factor by effects of another or subtracting interaction effects of several factors from effects of another. So, when considering the information in Tables 4 through 7 and in Table 8, a lot of the obvious will be ignored, In addition, even though analyses summarized in Table 8 indicated that only certain effects were significant, other effects revealed by analyses summarized in Tables 4 through 7 will be discussed. The authors think that the analyses summarized in Table 8, while helpful, are not absolute, and should be "taken with a grain of salt."

Factor Effects on Chemical Oxygen Demand

If the results of analyses summarized in Table 8 are believed, and the obvious effects ignored, there are no significant effects.

TABLE 8

RESULTS OF PROBABILITY ANALYSES

RESPONSE	SIGNIFICANT FACTOR OR FACTOR INTERACTION EFFECTS: INCREASE	DECREASE
Stage 1 COD Concentration	5	none
Stage 4 COD Concentration	none	none
Stage 1 Removal Rate	5	none
Overall COD Removal Rate	5	none
Stage 1 NH_3-N Concentration	4	24, 2
Stage 4 NH_3-N Concentration	4	24, 2
Stage 1 NH_3-N Removal Rate	none	none
Overall NH_3-N Removal Rate	4, 5, 2, 12	none
Stage 2 NO_3-N Concentration	none	23, 2, 5
Stage 4 NO_3-N Concentration	4, 5	none
Stage 2 NO_3-N Production Rate	14	23, 5
Overall NO_3-N Production Rate	4	none

FACTOR NUMBER	FACTOR
1	RPM
2	FLOW RATE
3	SURFACE AREA EXPOSED
4	NITROGEN MAS LOADING RATE
5	CARBON MASS LOADING RATE

One can choose to ignore Table 8 and the analyses for significance, and examine Table 4. It will be found that the factors which effect COD concentration the most, either decreasing or increasing it, are all trivial factors; i.e., flow rate, or flow rate in combination with an other factor, or carbon mass loading, singly or in combination. The same can be said for COD mass removal rates.

Factor Effects on Ammonia Nitrogen

Considering the results of the probability analyses, and ignoring the obvious effects, there are three significant effects. Overall ammonia removal rates were increased when carbon mass loading rate (5), flow rate (2), RPM and flow rate (12) were increased. The following non-trivial major effects were observed, but may or may not be significant, depending upon one's belief in the validity of the probability analyses.

Stage 1 Ammonia Concentration.

An increase in RPM and carbon mass loading rate (15) caused an increase in ammonia concentration. However, an increase in carbon mass loading (5) alone caused a decrease in ammonia concentration. It is easy to eliminate the contradiction by saying that neither effect is statistically significant.

Stage 4 Ammonia Concentration.

An increase in carbon mass loading rate (5) increased stage 4 ammonia concentration the most. An increase in surface area exposed (3) caused the greatest decrease.

Overall Ammonia Removal.

Increasing RPM (1) decreased removal the most. Increasing carbon mass loading rate (5) also increased removal. Increasing RPM and flow rate together (12) caused the greatest decrease in removal.

Factor Effects on Nitrification

There were four statistically significant, yet non-trivial effects. Increasing carbon mass loading rate (5) caused three of them. Stage 2 NO_3-N concentration decreased, stage 2 NO_3 production rate decreased, and stage 4 NO_3-N concentration increased. If one examines Table 6, ignoring Table 8, one can find no major non-trivial, but perhaps statistically non-significant effects.

CONCLUSIONS

The results of the analyses done to date are ambiguous. There were extreme variations in the performance of the RBC during the experimental program. One only needs to scan Table 3 to confirm that. Something caused those variations, and these experimenters feel it was variations in the factors being studied. Further analyses of the data are anticipated, and hopefully these analyses will give more insight into the system.

For the time being, some information has come out of the analyses which adds to the knowledge of the RBC system.

Consider first only information in Table 8.

RPM was a statistically significant factor, in combination with a non-trivial factors, once.

Flow rate was a statistically significant factor, in combination with other non-trivial factors, twice.

Surface Area Exposed was a statistically significant factor, in combination with a non-trivial factor, once.

Nitrogen Mass Loading Rate was never a non-trivial, statistically significant factor.

Carbon Mass Loading Rate was a non-trivial, statistically significant factor, by itself, four times.

If one includes the more subjective analyses presented in the "Analyses" section, Table 9 can be developed.

TABLE 9

FREQUENCY MENTIONED AS IMPORTANT FACTOR

FACTOR	FREQUENCY MENTIONED	
	ALONE	IN COMBINATION
RPM	1	3
FLOW RATE	1	2
SURFACE AREA EXPOSED	1	0
NITROGEN MASS LOADING RATE	0	0
CARBON MASS LOADING RATE	7	1

One must remember that the factors were identified as important or significant more times than indicated above. However, they were termed "trivial" because the factor was part of the response being mentioned. The data are still being analyzed, and attempts are being made to "normalize" the derived data so these factors become "non-trivial".

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HYDRAULIC AND ORGANIC FORCING OF A PILOT SCALE RBC UNIT

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The San Francisco Master Plan for water pollution control recommends that a new Southwest water pollution control plant (WPCP) be constructed. The Southwest WPCP would include secondary treatment for about 22 million gallons per day (Mgal/d) of dry-weather sewage to effluent quality levels required for ocean discharge. Since San Francisco has combined sewers, wet-weather sewage would also be treated in the secondary plant. A separate wet-weather plant would be constructed to handle wet-weather flows that exceed the hydraulic capacity of the secondary plant. Wet-weather flows as high as 500 Mgal/d could be possible depending on the design of the storage/transport system.

In support of the Southwest facilities planning effort, Metcalf & Eddy conducted a pilot plant program to study the serious technical problems associated with treating both wet- and dry-weather flows. During wet weather, the composition of the feed to the secondary plant would continuously change over an extremely wide range, and the effect of these changes on the quality of the secondary effluent was unknown. Thus, pilot testing of the secondary processes was necessary to determine (1) what percentage of the wet-weather treatment capacity could be achieved by hydraulically forcing the secondary plant, and (2) what design modifications would be necessary to achieve this maximum hydraulic capacity. Another objective of the pilot testing was to determine the stability of the secondary processes under extreme diurnal and wet-weather treatment load transitions.

This paper presents the results of the pilot plant tests designed to study the response of a rotating biological contactor (RBC) to rapid changes in hydraulic and organic loadings. These results should be valuable to others considering this process for locations experiencing similar wide variations in wastewater composition (for example, communities with combined sewers or unusually high infiltration/inflow rates).

CHARACTERISTICS OF THE TREATMENT PROBLEM

The proposed dry-weather treatment plant will be required to handle wide variations in both flow and pollutant loadings. The typical diurnal flow pattern includes a low of about 8 Mgal/d during the early morning hours and peak flows of about 35 to 40 Mgal/d during the late morning and early afternoon. The BOD concentrations vary in a similar fashion with concentrations as low as 50 milligrams per litre (mg/L) during the periods of low flow and peak concentrations as high as 250 mg/L coinciding with the peak flows. Typically, the time period between the early morning low and the later morning peak is 3 to 4 hours and, consequently, the BOD loading rate can increase by as much as 2,500% during this short time span.

During wet weather, the short-term variations in loading conditions could vary even more drastically depending on the peak hydraulic design capacity selected for the dry-weather plant. During a storm, the BOD and TSS concentrations may vary from as low as 15% of the dry-weather average to as high as 200 or 300%. Furthermore, the alkalinity and conductivity of the wastewater will vary more or less in direct proportion to the amount of stormwater dilution. These variations in influent characteristics may significantly affect the performance and stability of biological secondary processes and, consequently, must be carefully considered in the selection of the recommended process.

Since the RBC system is a fixed culture process, several advantages were anticipated over the competitive suspended biomass processes (air and pure oxygen activated sludge) for the proposed Southwest WPCP.

- Primary clarifiers could be operated at higher overflow rates during wet weather without seriously affecting the RBC process since the inert solids carried over in the primary effluent would tend to pass through the RBC units with little effect on BOD removal efficiency. In a suspended biomass system, these inert solids accumulate in the aeration basins and reduce the fraction of the total biomass, which is biologically active.

- Since the RBC biomass is attached to the rotating disks, high hydraulic loadings will not cause biomass "washout." Large increases in hydraulic loadings to a suspended biomass system will result in a temporary transfer (washout) of biomass from the aeration basins to the secondary clarifiers leading to lower BOD removal efficiency.

- Significant variations in loading conditions to an activated sludge process require careful operator attention to make proper adjustments to air or oxygen supplies, sludge recycle rates, and secondary clarifier operating conditions in order to maintain a viable biomass. Very little operator attention would be required during hydraulic forcing of an RBC system since the biomass is retained on the disks.

Pilot scale tests were performed to demonstrate the extent to which an RBC unit can be hydraulically forced relative to normal design loadings and to determine the response of an RBC system to rapid increases in BOD loadings over short-time periods. Additional tests were conducted to evaluate the effects of high feed solids from an overloaded primary clarifier; however, those tests will be described in a subsequent paper.

TEST FACILITY DESCRIPTION

The pilot scale RBC unit was a part of an extensive pilot plant facility designed to test a number of physical/chemical processes for wet-weather treatment in addition to the tests on biological secondary treatment processes. Photographs of the pilot plant facilities, located primarily in the courtyard of the Richmond-Sunset WPCP in Golden Gate Park, are presented in Figure 1. These photographs indicate the basic scale of the pilot units: e.g., the RBC unit was equipped with 2.0 metre diameter disks and the UNOX system was in the standard 8 ft by 40 ft trailer. All biological treatment units were elevated to provide gravity transfer to the secondary clarifiers.

The RBC pilot unit was an Autotrol 2.0 metre unit with four disks providing about 7,900 square feet (ft^2) of surface area. The nominal hydraulic capacity of the unit was 6 gallons per minute (gal/min) based on a hydraulic loading of 1.1 gallons per square foot per day ($gal/ft^2 \cdot d$). The pilot unit was operated at a rotational speed of about 3 revolutions per minute (rpm) during all tests. Feed to the RBC unit was typically primary effluent from a pilot-scale clarifier operated at a surface overflow rate of about 1,000 $gal/ft^2 \cdot d$. Effluent from the RBC unit flowed by gravity to a 5 foot diameter secondary clarifier. During the hydraulic forcing tests, a large portion of the RBC effluent was bypassed around the secondary clarifier to avoid excessive overflow rates.

A schematic process flow diagram of the RBC pilot facility is presented in Figure 2. Composite samples, taken at 8 hour intervals, were collected by the use of small metering pumps or timer-controlled solenoid valves at the flow points indicated in the figure.

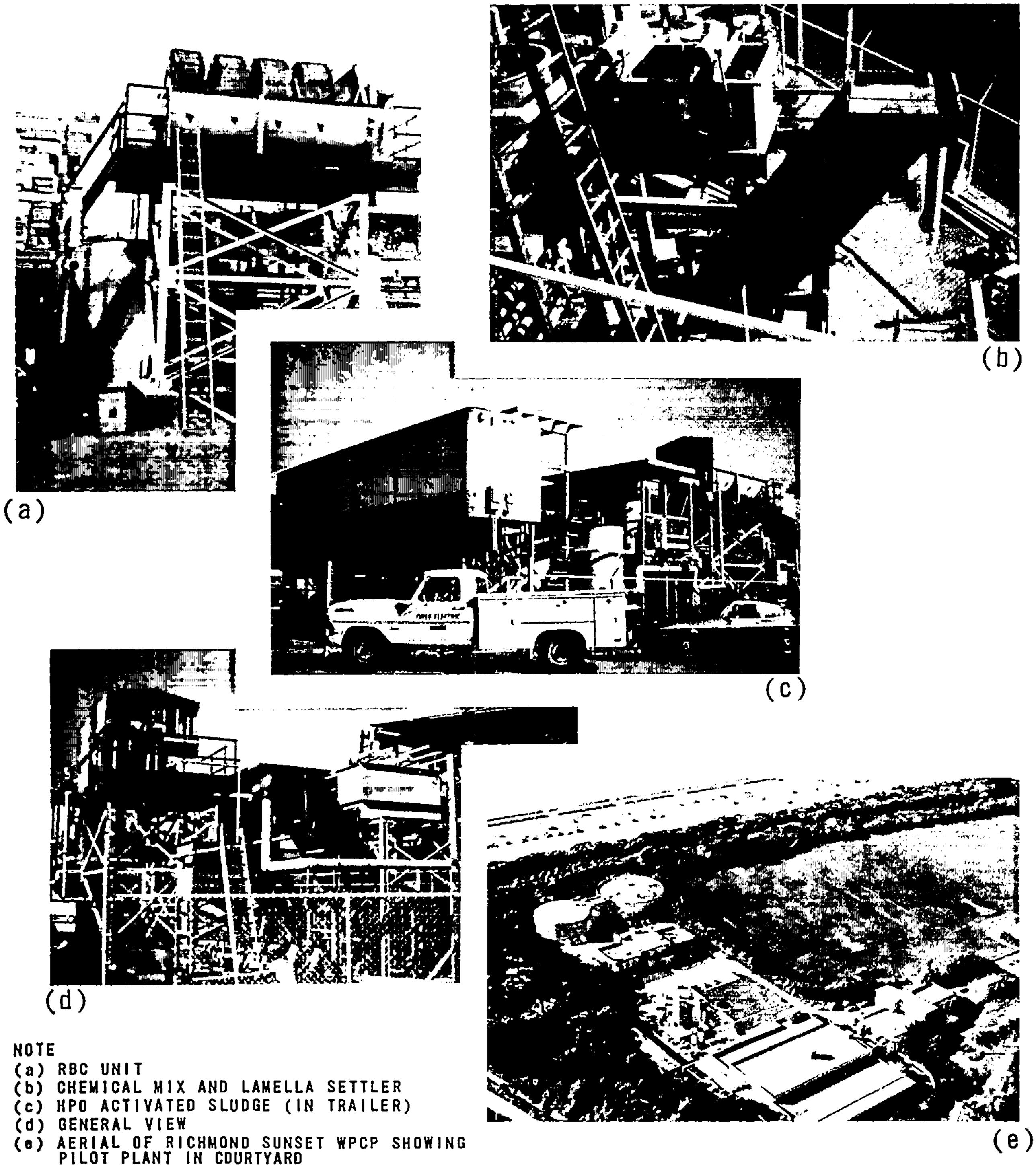

Figure 1. Southwest WPCP pilot plant.

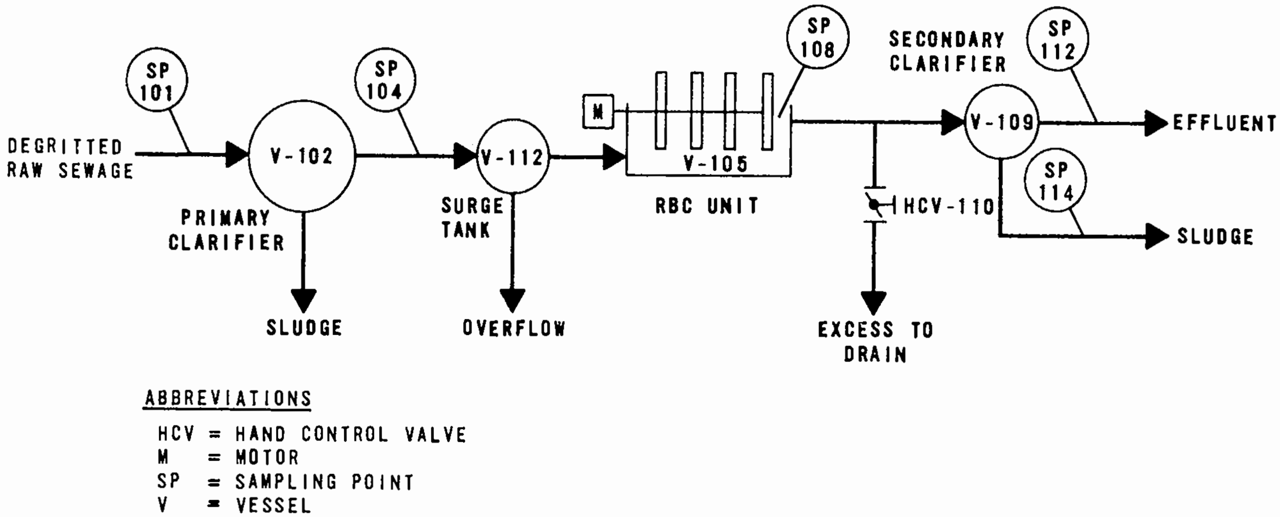

Figure 2. Process flow diagram, RBC unit.

HYDRAULIC FORCING TESTS

The experimental plan for the hydraulic forcing tests involved operating the RBC unit at progressively higher feedrates by applying the step increases shown in Figure 3. The unit received diurnal variations in wastewater composition but was operated at constant flow conditions after each step increase. Samples were collected for analysis according to the schedule shown in Table 1. The sampling points are shown schematically on the process flow diagram, Figure 2.

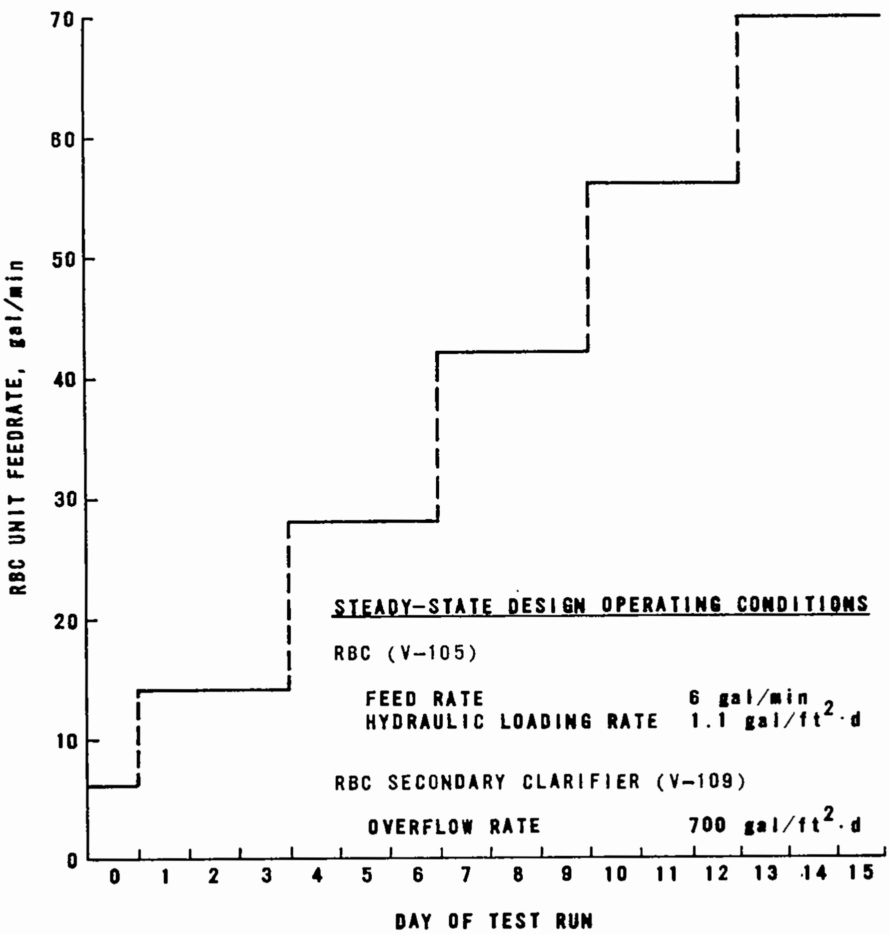

Figure 3. RBC flow increases, hydraulic forcing tests.

Table 1. SAMPLING PROGRAM, RBC HYDRAULIC FORCING TESTS

	Effluent primary clarifier V-102	RBC unit	Effluent secondary clarifier V-109
Sampling point No.	104	108	112
Sample type	Composite	Grab	Composite
pH	3	3	3
Temperature	3	3	3
Dissolved oxygen	3	3	3
Conductivity	3	--	3
Total suspended solids	3	3	3
Volatile suspended solids	3	3	3
Alkalinity	1	--	1
Settleable solids	--	3	--
Total organic carbon (TOC)	3	--	3
Soluble TOC	3	12[a]	3
Biochemical oxygen demand (BOD)	3	--	3
Soluble BOD	3	--	3
Chemical oxygen demand (COD)	3	--	3
Soluble COD	3	--	3
Ammonia nitrogen	1[b]	--	1[b]
Freon extractable material	1[b]	--	1[b]

a. One sample from each stage per shift.

b. 24 hour composite.

Results of the hydraulic forcing tests are summarized in Table 2 along with comparable data collected during another test run on this unit at essentially nominal design conditions. The RBC unit was operated at hydraulic loadings greater than 1,000% of design and organic loadings of up to 370% of design (based on a nominal design loading of 1.4 pounds of soluble BOD per 1,000 ft^2/d). The last line of Table 2 indicates whether the feed was all dry-weather flow or partially wet-weather flow.

Due to the rainfall that occurred during many of the test periods, the influent BOD concentrations varied widely from day to day. Thus, the average influent and effluent concentrations listed in Table 2 and the percent removals calculated from them must be interpreted only as trends with increasing hydraulic load.

Table 2. RBC HYDRAULIC FORCING TEST RESULTS

	Hydraulic loading rate, % of design rate[a]									
	97[b]	225	275	325	375	470	475	590	775	1,040
Test period	--	D	A	E	F	B	G	C	H	I
Duration, hours	216	96	104	40	80	72	88	40	168	256
Feedrate, gal/min	5.8	13.0	15.8	19.0	21.7	27.2	27.6	34.2	45.0	60.5
Hydraulic loading, $gal/ft^2 \cdot d$	1.06	2.38	2.88	3.45	3.96	4.96	5.02	6.23	8.20	11.03
Hydraulic detention, min.	172	77	63	53	46	37	36	29	22	16.5
Organic loading, $lb\ BOD_{sol}/1,000\ ft^2 \cdot d$	0.79	1.48	1.25	1.85	2.31	3.30	3.48	1.82	5.19	4.87
Organic removal, $lb\ BOD_{sol}/1,000\ ft^2 \cdot d$	0.75	1.24	1.13	1.21	1.88	2.72	2.60	1.09	3.69	2.48
Influent BOD, mg/L										
Total	160	133	102	123	108	145	140	78	115	101
Soluble	90	75	52	64	70	80	83	35	76	53
Effluent BOD, mg/L										
Total	24	29	23	43	44	46	49	52	55	55
Soluble	4.2	12	5	22	13	14	22	14	22	26
Effluent total suspended solids, mg/L	21	22	12	58	30	34	33	32	46	43
BOD removal, %										
Total	85.2	78.2	77.5	65.0	59.2	70.0	65.0	55.1	52.2	45.5
Soluble	95.3	84.0	90.4	65.6	81.4	82.5	74.7	60	71.1	50.9
COD removal, %										
Total	79.9	76.9	67.2	39.6	61.4	62.3	57.5	52.9	44.4	40.9
Soluble	70.2	71.5	47.4	36.3	46.6	63.5	56.5	59.9	32.2	36.0
TOC removal, %										
Total	74.2	64.5	53.5	41.0	59.3	54.7	57.6	30.7	--	--
Soluble	52.9	56.5	39.4	44.4	49.1	38.0	49.2	20.7	--	--
Feed type	Dry	Wet	Wet	Dry	Dry	Dry	Dry	Wet	Dry	Wet

a. Based on nominal design rating of 6 gal/min.
b. This column represents the performance at nominal design loading.

Scanning the data in Table 2 from left to right, it is clear that in general as hydraulic load increases, the effluent soluble BOD tends to rise from 4 mg/L up to 26 mg/L. Likewise, the soluble BOD removals tend to decrease from 95% down to 51%. However, to compensate for stormwater dilution of the feed, the data are best interpreted as a function of applied organic loading.

The organic removal achieved as a function of applied organic loading, both measured in units of pounds of soluble BOD per day per 1,000 ft^2 of disk area (lb BOD/1,000 $ft^2 \cdot d$), is shown in Figure 4. In Figure 4, the open circles represent dry-weather tests; the dark circles represent tests with some rain. Note that as the loading increases up to about 3 lb/1,000 $ft^2 \cdot d$, the removal increases linearly with a slope of 0.82, indicating a removal of about 82% on a mass basis.

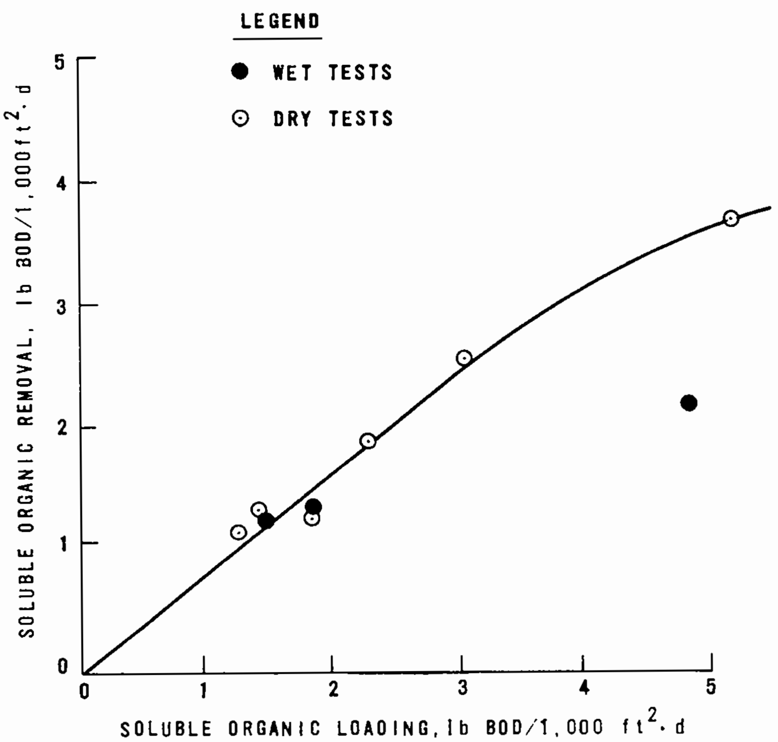

Figure 4. RBC performance, organic removal.

The data point at a loading of 5.19 lb/1,000 ft^2·d represents operation at 45 gal/min under wet-weather conditions. The organic removal was only 51%, a reduction due to a combination of short hydraulic detention time (16.5 minutes) and low influent soluble BOD concentrations (only 53 mg/L due to stormwater dilution, versus 70 to 90 mg/L for dry weather).

Another measure of performance is the effluent soluble BOD concentration. The effluent soluble BOD concentration as a function of organic loading is shown in Figure 5. Again, effluent BOD tends to increase more or less predictably as the organic loading increases. Only one data point is more than 5 mg/L from the correlating line, which is an excellent margin of error in BOD determinations at low values between 0 and 25 mg/L.

Similar conclusions hold for chemical oxygen demand (COD) removals, although the data are not presented in this paper.

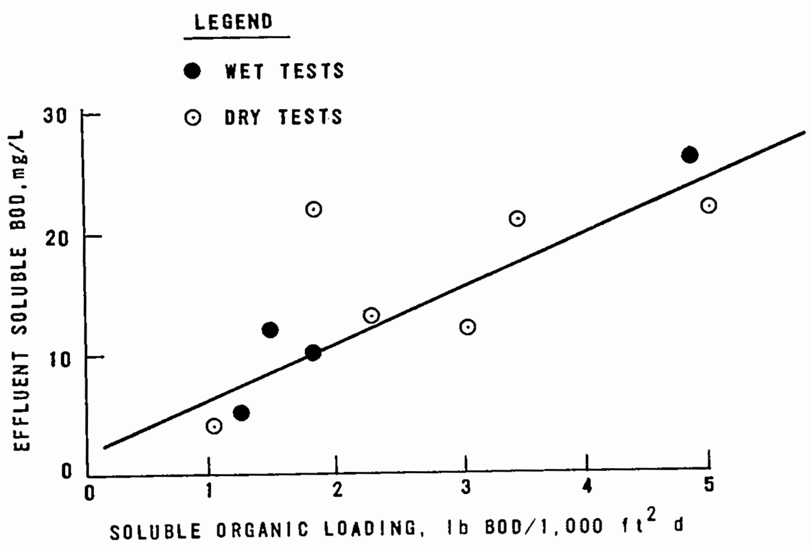

Figure 5. RBC performance, effluent soluble BOD.

The total organic carbon (TOC) reductions tend to decrease more rapidly than either BOD or COD because a portion of the TOC represents nonbiodegradable organics; thus, the maximum possible TOC reduction via biodegradation is less than 100%. Also, the effluent soluble TOC tends to increase more quickly as the organic load increases, indicating that some refractory metabolic byproducts may be produced under highly loaded, short residence time operations.

The overall performance data from Table 2 are superimposed on the vendor's RBC design curves in Figure 6. Only the data for test periods at feedrates up to 45 gal/min are shown. The hydraulic loading for the final test period at 60 gal/min is off scale.

Each operating point is represented in Figure 6 as two black dots connected by an arrow. In each case, the arrow runs from the measured performance to the theoretical performance. Thus, arrows pointing upward indicate that at the measured hydraulic and organic loading, the test resulted in a lower effluent BOD concentration than predicted by the design curves.

Note that in only one test period (E) does the arrow point downward, indicating that the actual effluent BOD was higher than that predicted by the design curves. Test E was a short test period that was terminated by very heavy rains. Thus, the data from Test E can probably be ignored in this analysis.

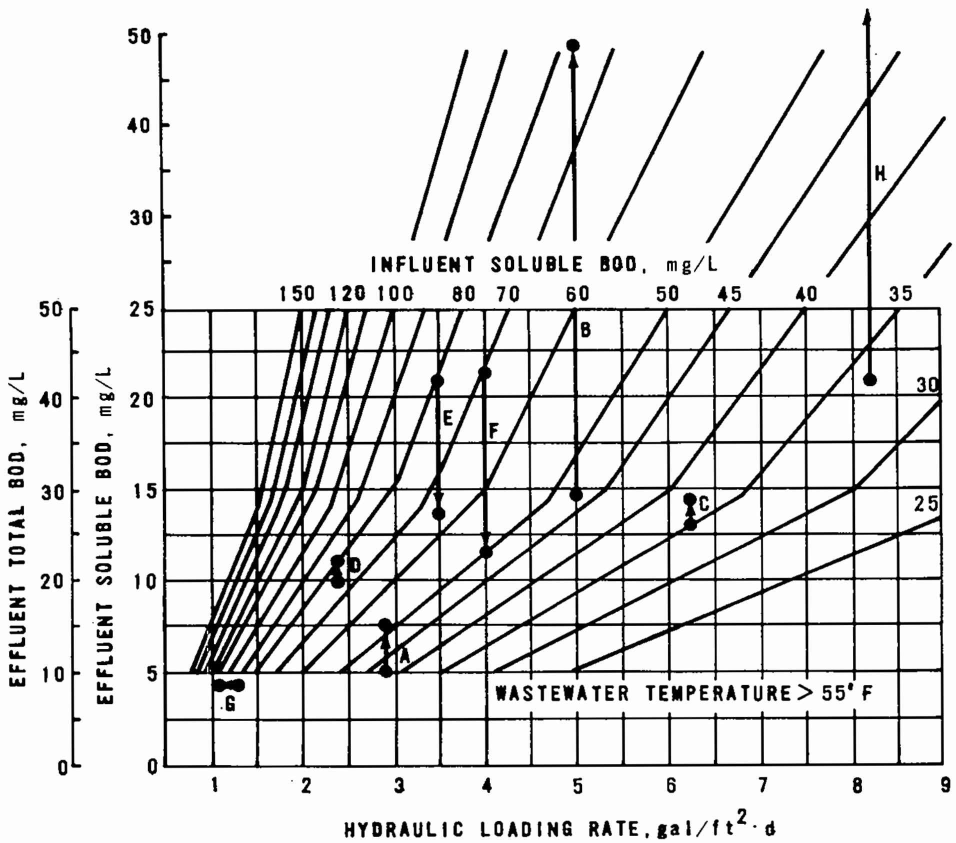

Figure 6. RBC design curves.

To consider Test B, it was necessary to arbitrarily extrapolate the design curves to much higher organic loads. The result is a predicted effluent BOD of 42 mg/L versus an actual value of 12 mg/L. The design curves could not be extrapolated in this manner to accommodate Tests H and I. In both cases, the predicted effluent concentrations would have been greater than the influent concentrations.

In general, it appears that the design curves are conservative and that the RBC can be much more heavily loaded for short periods of time than would be expected based solely on the design curves.

SIMULTANEOUS HYDRAULIC AND ORGANIC FORCING TESTS

A second series of tests was performed to evaluate the effects of more extreme forcing conditions over short periods of time. In these tests, the step increases in feedrate were made during the morning hours when the influent soluble TOC value changes from a low value of about 30 mg/L at 6:00-7:00 a.m. to a peak value of about 60 mg/L at 10:00-11:00 a.m. The tests were performed on 4 consecutive days with step increases in the RBC unit feedrate planned to follow the schedule shown in Table 3.

Table 3. PLANNED INCREASES IN RBC UNIT FEEDRATE

	RBC feedrate, gal/min			
Time	Day 1	Day 2	Day 3	Day 4
5:00	10	10	10	10
6:00	10	10	10	10
7:00	10	10	10	10
7:30	20	20	20	20
8:00	20	30	30	30
8:30	20	30	40	40
9:00	20	30	40	50
9:30	20	30	40	50
10:00	20	30	40	50
11:00	20	30	40	50
Noon	20	30	40	50
1:00	10	10	10	10

Grab samples of primary effluent (RBC feed), RBC 4th stage, and clarifier effluent were collected according to the following schedule:

	Sample frequency		
	5:00-7:00	7:00-10:00	10:00-1:00
Primary effluent	30 minutes	15 minutes	30 minutes
RBC 4th stage	30 minutes	15 minutes	30 minutes
RBC clarifier effluent	30 minutes	30 minutes	30 minutes

The RBC unit feedrates during the 4 days of testing are shown in Figure 7. The flowrate was adjusted according to the planned schedule except on April 18 when the operator had difficulty increasing the RBC feedrate to the desired rate of 20 gal/min because a bypass valve had been left open. Consequently, the RBC feedrate did not reach 20 gal/min until about 9:00 a.m. rather than at 7:30 a.m. as planned. The corresponding increases in soluble organic loading are shown in Figure 8. The response of the RBC unit in terms of soluble TOC in the feed, 4th stage, and secondary clarifier effluent for each day is shown graphically in Figure 9.

On April 18, 1978, the peak soluble organic loading was 2.1 lb TOC/1,000 $ft^2 \cdot d$ at about 10:30 a.m. The peak soluble TOC concentration in the RBC 4th stage was about 30 mg/L, and the clarifier effluent peak soluble TOC value was also 30 mg/L if the questionable data point at 10:30 a.m. is

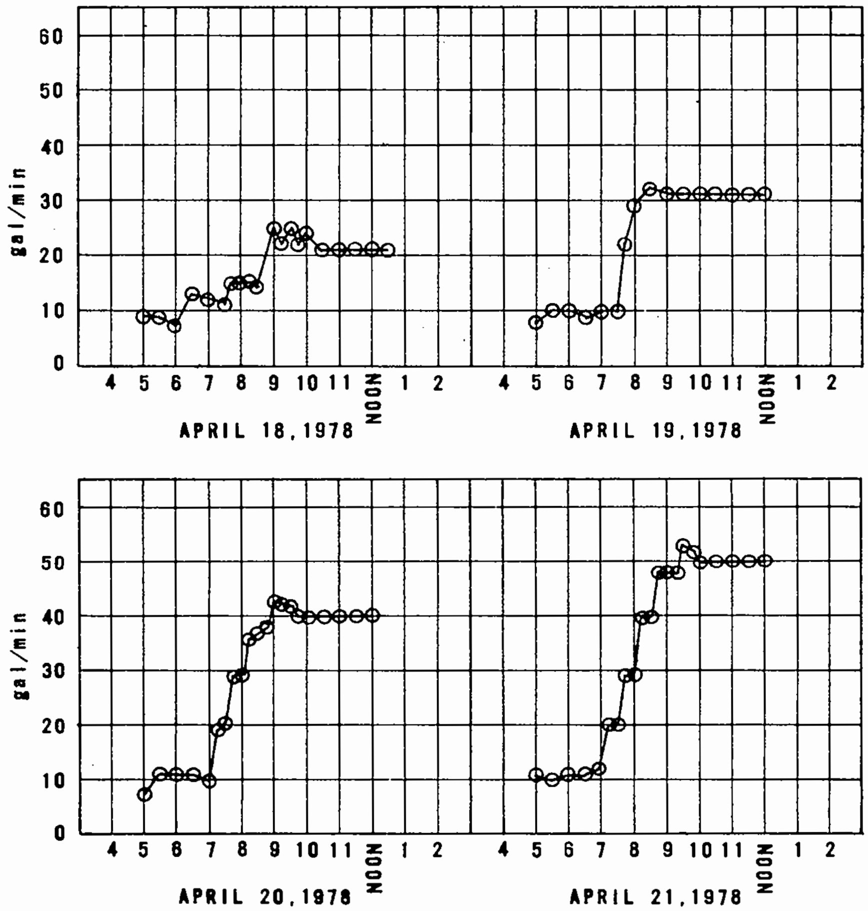

Figure 7. RBC feedrates.

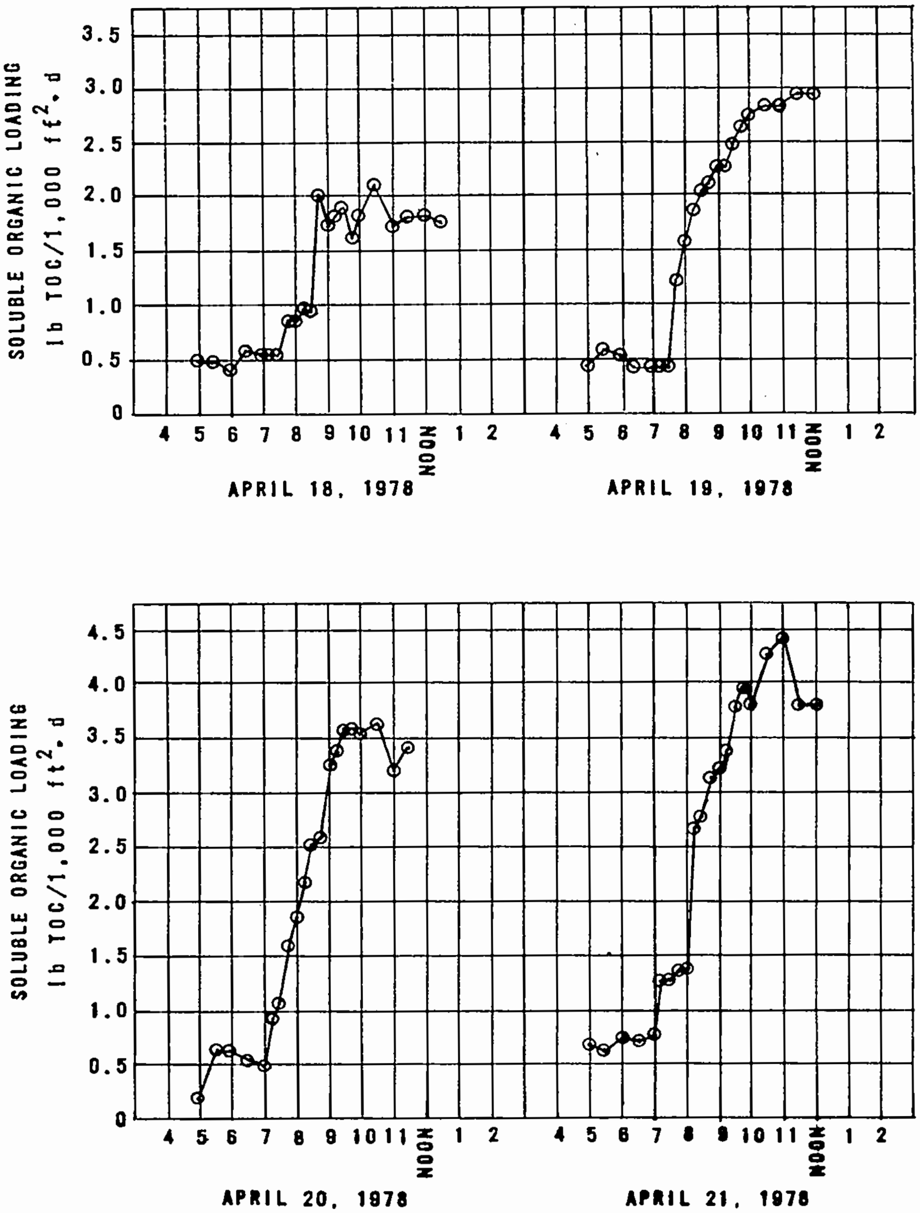

Figure 8. RBC organic loading rates.

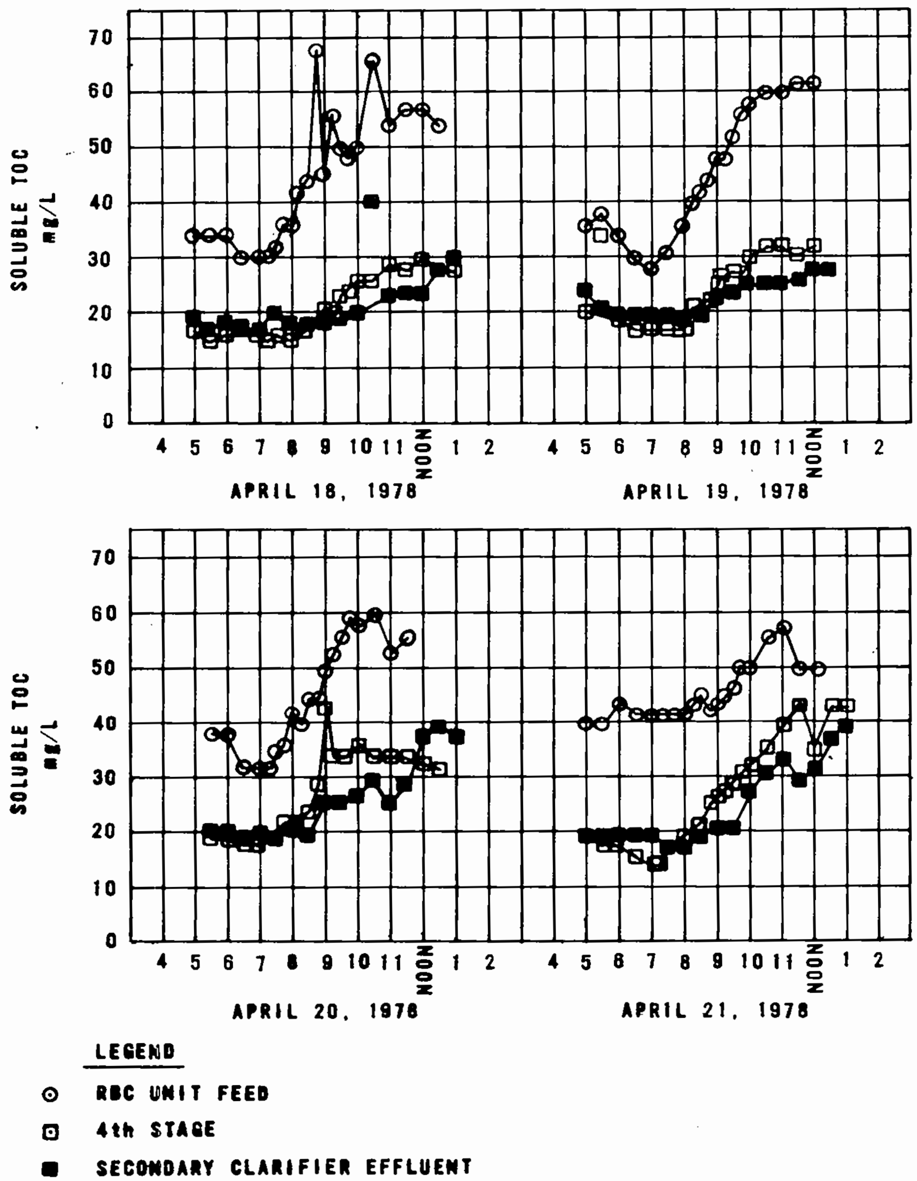

Figure 9. RBC soluble TOC response.

ignored. On the basis of hydraulic residence time alone, the response of the 4th stage should lag an increase in feed concentration by about 45 minutes during this test. Since the clarifier flowrate was maintained constant at about 15 gal/min for all tests, the response in the clarifier effluent should lag the 4th stage response by about 2.5 hours in each test.

On April 19, 1978, the deterioration in clarifier effluent quality was not significantly greater than in the first test even though the peak organic loading of about 2.8 lb/1,000 $ft^2 \cdot d$ was about 35% greater. The 4th stage soluble TOC did reach a peak of 34 mg/L, which was about 13% greater than the peak value in the first test.

On the third test day, April 20, 1978, the peak organic loading was about 3.6 lb TOC/1,000 $ft^2 \cdot d$, which was about 70% greater than in the first test. The clarifier effluent reached a peak value of 40 mg/L soluble TOC, which was about 33% higher than in the first test. The rapid increase in 4th stage TOC, which occurred between 8:30 and 9:00 a.m., was the initial response to the sudden increase in loading. Although a peak value of about 44 mg/L was apparently reached at 9:00 a.m., the subsequent TOC values leveled off at about 34 mg/L.

On the final day of the test, April 21, 1978, the RBC feedrate was increased to 50 gal/min, resulting in a peak organic loading of about 4.4 lb TOC/1,000 $ft^2 \cdot d$. This represents an increase of about 210% over that of the first test. The 4th stage TOC values during this test show a steady increase in response to the increased hydraulic and organic loading, peaking at about 44 mg/L. This is an increase of about 50% over the corresponding peak value in the first test. The clarifier effluent TOC showed a similar response.

Although the effluent TOC quality deteriorated as the organic loading increased during the series of tests, it is interesting to view the data in terms of the organic removal achieved as a function of applied organic loading. These data, along with corresponding data collected during the hydraulic forcing tests, are presented in Figure 10. It should be noted that the data points presented for the simultaneous forcing tests are average values corresponding to the final 2 to 3 hours of operation in each daily test, while the data from the hydraulic forcing tests represent relatively long-term, steady-state values.

Considering only the data collected during the simultaneous forcing tests (Curve I), for increases in soluble TOC loading of up to about 3.5 lb/1,000 $ft^2 \cdot d$, there is a net benefit in terms of total mass of TOC removed. At higher organic loadings, the organic removal apparently decreases, indicating that the process is stressed beyond its maximum capacity for soluble TOC removal. These results are consistent with the data from the hydraulic forcing tests, which are shown in Figure 10 for comparison.

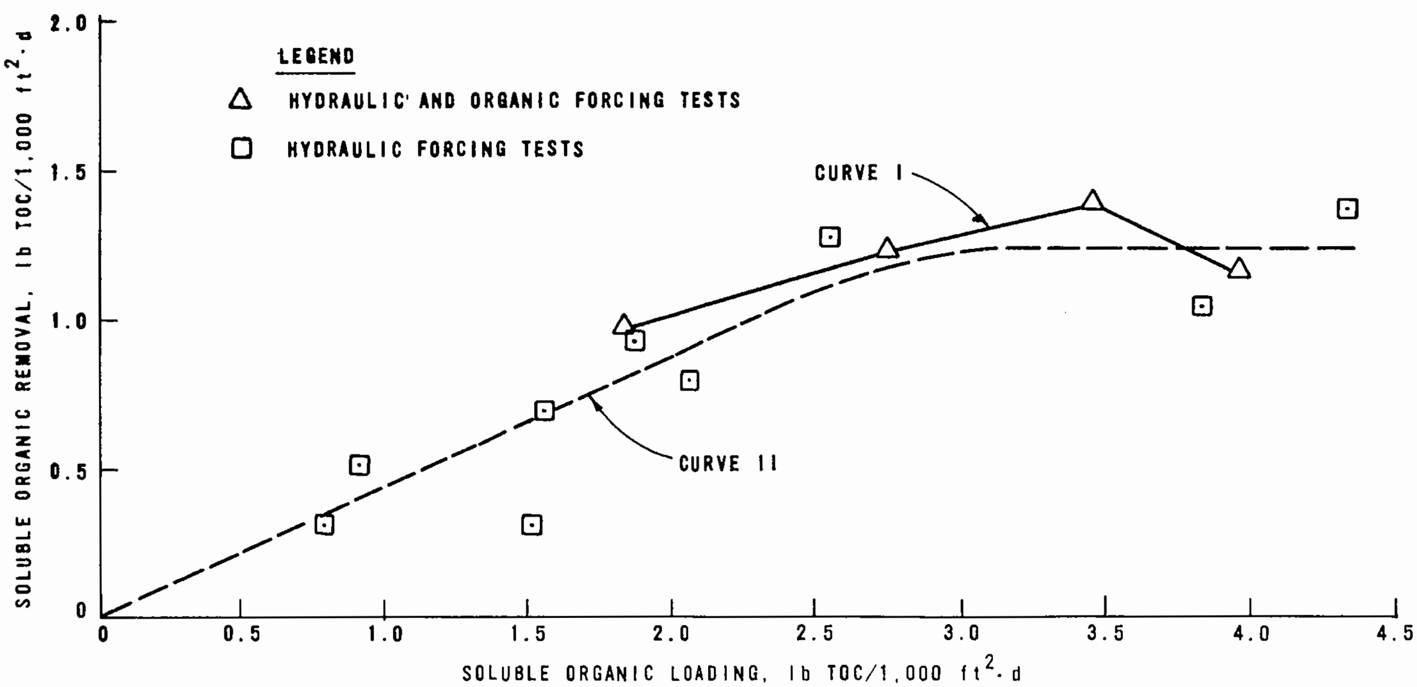

Figure 10. RBC performance, organic removal.

There were no apparent operational difficulties associated with these
severe shock loading tests. When the loading rates were returned to

EFFECT OF ORGANIC LOADING ON RBC PROCESS EFFICIENCY AND FIXED-FILM THICKNESS

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Secondary or biological treatment of wastewater has been provided principally by the activated sludge and trickling filter processes. The former can be described as a slurry or suspension process while the latter is a fixed film process. One substantial modification of the trickling filter process involves the use of biological films attached to rotating disks.

The rotating biological contactor (RBC) system, as now practiced, consists of a series of reactors each of which contains a number of closely spaced rotating vertical disks partially submerged in the wastewater. A biomass similar to trickling filter slime is established on the surface of the disks. As the disks rotate, they carry the biomass film saturated with wastewater into the air where it is aerated to provide the dissolved oxygen (DO) required for aerobic biological activity. In summary, the rotating disk is utilized as a supporting media for biological growth, as a mechanism for aeration, and as a means of contacting the microorganisms with the wastewater.

For better design and operation of the RBC process, it is important to understand the interaction of the loading and operating variables such as organic loading, flow rate, rotational disk speed, detention time, disk surface area, submerged disk depth, and wastewater temperature. It is also important to understand the effect of biological film age and thickness on the organic removal efficiency.

This work describes the effects of detention time, influent concentration, flow rate and biological film age and thickness on organic removal of RBC.

EXPERIMENTAL PROCEDURE

To study the above effects, a six-stage RBC system was constructed with each stage 6 in. (15.24 cm) deep, 4 in. (10.16 cm) wide, and 8 in. (20.32 cm) long. The system was partially submerged in a constant temperature water bath.

Each stage contained one or two 6 in. (15.24 cm) diam., 0.25 in. (0.64 cm) thick plastic disks. The available surface area from each disk was 0.442 sq. ft. or 63.6 sq. in. (410 sq. cm). When there were two disks in each stage, the two disks were spaced on 1.75 in. (4.45 cm) centers. These were mounted vertically and parallel to the flow on a 0.5 in. (1.27 cm) diam. horizontal stainless steel shaft. The shaft rotated at controlled speeds by a roller chain connected to a 1/6 hp (0.124 kw) ratiomotor.

In this type of laboratory reactor, the wall area effect could mask the effect of the rotating disk area. Therefore, it was necessary to prevent biological growth on the walls and bottoms of the reactor and restrict the biological film accumulation to the rotating disks. This was accomplished by using two sets of 0.25 in. (0.64 cm) thick reactor linings for each stage and changing the linings three times each day.

The substrate selected is shown in Table I. The protein was present in the nutrient broth representing 65 percent of the chemical oxygen demand (COD), the carbohydrate was present as glucose representing 25 percent of the COD, and the fatty acid was present as sodium oleate representing 10 percent of the COD.

The effect of film age on organic removal rate was studied under three different test conditions. These three conditions differed by flow, detention time, or organic loading. The first study was made at a flow rate of 21 l/day, a loading of 7.56 g COD/day, or 0.63 lb. COD/100 sq. ft. apparent disk area/day (30.8 g/sq. m/day), and a detention time of 96 min. per stage while the second study was done at a flow rate of 42 l/day, a loading of 15.12 g COD/day, or 1.26 lb. COD/100 sq. ft. apparent disk area/day (61.6 g/sq. m/day), and a detention time of 24 min. per stage. The third study was made at the same flow rate and organic loading as the second study, but at a detention time of 48 min. per stage. All three studies were made using a single disk per stage, an influent COD concentration of 360 mg/l, a rotational disk speed of 30 rpm, and a liquid temperature of 20°C. In the third study, the effect of film thickness on organic removal rate was also investigated.

The biological growths on the first three disks (stages) were removed successively starting from the third disk with the first disk being the last. This reverse procedure was used so that the feed to each stage would not be changed as would be the case if the disk cleaning went from stage one to stage three. At least four detention times of continuous slime removing was done before samples were withdrawn for COD testing in order to insure the purging of the original treated liquid in the reactor. Two successive runs were made on each disk.

The thickness of the biological film was determined with a Bausch and Lomb phase contrast microscope. The microscope was focused on the top, and then on the bottom of the film. The film thickness was determined from the number of divisions turned on the fine adjustment between the two focusings. The microscope was calibrated by focusing the top and bottom of one to five pieces of Corning cover glass which ranged in thickness from 174.78 microns to 873.92 microns. The actual thickness of the Corning cover glass was determined with a precision micrometer. By plotting the thickness against the number of divisions turned on the fine adjustment between the two focusings, it was found that each division on the fine adjustment of the phase contrast microscope used in this investigation represented 1.8863 microns of thickness.

Six and eight random film thickness measurements were made each time, respectively, on the four pieces of cover glass and the eight pieces of cloth tape which were attached symmetrically on the two sides of the disk.

All analytical determinations were made according to the recommendations of Standard Methods.[1]

RESULTS

Effect of Detention Time at Constant Organic Loading

The effect of detention time was studied by two different methods which are summarized in Table II. The detention time was first varied by keeping both reactor volume and organic loading constant, and varying both flow and influent concentratrion. This was studied at three different levels - 24, 48 and 96 min. per stage. Varying detention time was also achieved by keeping flow rate, influent concentration and organic loading constant, and varying only the reactor volume. This was tested at three levels - 17, 24 and 48 min. per stage.

Varying Both Flow and Influent Concentration

Data on the effect of detention time by varying both flow and influent concentration are summarized in Figure 1. It was shown that COD reduction for the first stage reactor increased as detention time was increased from 24 min. to 96 min. However, increasing detention time had little effect on the overall organic removal efficiency which was about 90% COD reduction.

Varying Reactor Volume

The results on the effect of detention time by varying reactor volume are shown in Figure 2 as COD reduction in each stage versus stage number. The COD reduction in the first stage increased only very slightly as detention time per stage was increased from 17 min. to 48 min. An increase in detention time had little effect on the overall organic removal efficiency which was about 91% COD reduction. The results show less fluctuation from decreasing detention time by varying reactor volume than from varying both flow and influent concentration.

From both Figure 1 and Figure 2 it is readily seen that stage 4 removed more substrate than stage 3 under all detention time conditions studied. This was contradictory to the first order reaction theory if all conditions except organic concentration were the same in both stages. It was found that mixed liquor dissolved oxygen was absent in the first three stages, but was present in the last three under all detention time conditions investigated. Therefore, it appears that the absence of the mixed liquor dissolved oxygen was the cause for this reverse condition.

Further insight into controlling mechanisms can be gained by assuming the reaction was first order. If the biochemical reactions taking place in the RBC were "first order" in character, or the rate of the reaction was proportional to the amount of oxidizable organic matter remaining at any time, a straight line should be obtained when the logarithms of the concentrations remaining were plotted against the linear scale of time. Since the percentage COD remaining is proportional to the concentration remaining and the detention time of the liquid in the RBC is proportional to the number of stages, data were plotted as the logarithm of the percentage COD remaining versus stage number.

The results showed that two or three straight lines in series instead of one were obtained depending on the organic loading. Typical results exhibited three straight lines and are shown in Figure 3. The first straight line extended over the first three stages where mixed liquor dissolved oxygen was absent. The second line covered the next two stages where mixed liquor dissolved oxygen was present. The last line represented the last stage where the influent organic concentration was low. The slope of the second straight line is greater than those of the other two. Therefore it can be stated that organic utilization in the first three stages was dissolved oxygen limited, the fourth and fifth stages was diffusion limited, and the last stage was organic concentration limited.

Effect of Influent Concentration at Constant Flow

The effect of influent concentration was studied under two different slime area conditions - single disk and double disks per stage.

Single Disk Reactor

The effect of influent concentration was first tested using a single disk per stage at four feed levels - 180, 360, 540 and 720 mg/l COD at a flow rate of 42 l per day and a detention time of 48 min. per stage. The results are shown as stage number versus COD reduction rate in Figure 4 and as percentage COD reduction in Figure 5.

As expected, organic utilization in each stage increased as influent COD concentration was increased to 730 mg/l except in the fourth stage where more organics were removed at an influent COD of 347.5 mg/l than at influent COD concentrations of 533 and 730 mg/l. This was a result of the mixed liquor dissolved oxygen being absent at influent COD concentrations of 533 and 730 mg/l and being present at an influent COD of 347.5 mg/l.

For a six-stage reactor, the total COD reduction rate was increased from 6.70 grams per day to 20.70 grams per day as the influent COD concentration was increased from 178 mg/l to 730 mg/l. The two highest feed levels yielded almost equal COD reductions in the last two stages even though the COD concentrations in the reactors were different. At an influent COD of 533 mg/l, the effluent COD's from the last two stages were 151.7 and 98.7 mg/l. At an influent COD of 730 mg/l, the effluent COD's were 288 and 237 mg/l. This indicates that the RBC system was overloaded at these two high feed levels.

At low influent concentration, the last few stages are at low COD levels and do not need to utilize their organic removal capacity. As a result, nitrification takes place. As can be seen from Figure 4, the last three stages at an influent COD of 178 mg/l and the last two stages at an influent COD of 347.5 mg/l did not need to utilize their organic removal capacity and as expected, nitrification was found in these stages.

Since the mixed liquor dissolved oxygen was absent in the first three stages at an influent COD of 347.5 mg/l and the first four stages at influent COD concentrations of 533 and 730 mg/l, stage 4 removed more organics than stage 3 when the influent COD was 347.5 mg/l, and stage 5 removed more organics than stage 4 when the influent COD concentrations were 533 and 730 mg/l. Stage 4 also removed more organics at an influent COD of 347.5 mg/l than at influent COD concentrations of 533 and 730 mg/l. Again, the depletion of mixed liquor dissolved oxygen was found to have a negative effect on the organic removal efficiency of the RBC.

Figure 5 shows that the percentage COD reduction increased markedly as influent COD concentration decreased from 730 mg/l to 178 mg/l for the RBC up to three stages. For systems of more than three stages, this was also true as the influent COD concentration decreased down to about 350 mg/l, and then leveled out and became independent of influent concentration at low feed concentrations.

Double Disk Reactor

The effect of influent concentration was also studied using double disks per stage at three levels - 360, 540 and 720 mg/l COD at a flow rate of 42 liters per day and a detention time of 48 min. per stage. The results are summarized in Figures 6 and 7.

It is of interest to note the similarity in COD removal rates in stage 2 even though the effluent concentrations from stage 2 were different, namely, 103.1, 272.5, and 432.1 mg/l COD for feeds of 345.4, 531.5, and 717.7 mg/l COD respectively. As the number of stages increased, the COD utilization decreased more rapidly with decreasing influent concentration. As can be seen from Figure 6, the last three stages at an influent COD of 345 mg/l, the last two stages at an influent COD of 532 mg/l, and the last stage at an influent COD of 718 mg/l did not need to utilize their organic removal capacity and as expected, NH_3-N removal took place in these stages.

Again, the absence of the mixed liquor dissolved oxygen was found to have a negative effect on the organic removal efficiency. The mixed liquor dissolved oxygen was absent in the first stage at an influent COD of 345 mg/l, the first two stages at an influent COD of 532 mg/l, and the first three stages at an influent COD of 718 mg/l. Stage 3 was found to remove more organics than stage 2 when the influent COD was 532 mg/l and stage 4 removed more COD than stage 3 when the influent COD was 718 mg/l. Stage 3 also removed more organics at an influent COD of 532 mg/l than at an influent COD of 718 mg/l. Stage 2 utilized more organics at an influent COD of 345 mg/l than at an influent COD of 532 mg/l.

Figure 7 shows that the percentage COD reduction decreased markedly as influent COD was increased from 345 mg/l to 718 mg/l for the RBC up to four stages. For systems of more than four stages, the percentage COD reduction became independent of influent concentration.

Effect of Flow Rate at Constant Influent Concentration

Flow rate (hydraulic loading) determines the rate of organic addition and the detention time of the liquid in the RBC system. The effect of flow rate was studied at four levels - 14, 21, 42 and 63 l per day at an influent COD concentration of 360 mg/l.

Figure 8 shows the effect of flow rate on the COD reduction rate. The COD reductions in the first two stages increased only very slightly when the flow rate was increased from 42 l/day to 63 l/day. This indicates that the first two stages were overloaded at a flow rate of 63 l/day.

Since the mixed liquor dissolved oxygen was absent in the first one, three, and four stages at flow rates of 21, 42 and 63 l/day respectively, and present in all stages at a flow rate of 14 l/day, stage 1 removed more COD at a flow rate of 14 l/day than at a flow rate of 21 l/day, and stage 4 had a higher COD reduction at a flow rate of 42 l/day than at a flow rate of 63 l/day. Stage 2 also removed more COD than stage 1 when the flow rate was 21 l/day, and stage 4 had a higher COD reduction than stage 3 when the flow rate was 42 l/day. Stage 5 also removed more COD than stage 4 at a flow rate of 63 l/day. However, in general, increasing flow rate increases the COD reduction rate.

Figure 9 shows that the percentage COD reduction increased markedly as flow rate was decreased from 63 l/day to 14 l/day for RBC up to three stages. For reactors of more than three stages, this was also true as flow rate was decreased down to about 42 l/day and then leveled out and became independent of flow rate at low feed rates.

Effect of Biological Film Age

The results of the biological film study were calculated in terms of COD removal rate (mg/hr) and correlated with film age as shown in Figures 10, 11 and 12 for the first, second and third tests respectively. There was a rapid organic utilization by films in the early periods of their growth. Utilization rates were then increasing more slowly and relative stability was eventually achieved. The slight COD reduction at time zero was most likely achieved by the suspended microorganisms. Figure 10 shows that it took the first, second and third disks 25, 34.5 and 38 hours respectively to recover the efficiencies to their former levels. Figure 11 indicates that 33, 50 and 42 hours were required, respectively, by the first, second and third stage disks to reestablish the normal efficiencies of treatment. It was also found in a third test that 34, 44 and 33 hours were required, respectively, by the first, second and third disks to recover their normal efficiencies as shown in Figure 12. In these three figures, the two straight line portions of each curve were drawn by the method of least squares.

While Figures 10, 11 and 12 delineate average rates, they do not demonstrate trends that might have occured during the individual runs. However, the scatter of data shown in these figures suggests that there may be significant trends. Therefore, data are presented separately for runs 1 and 2 of the second stage disk of the first study in Figure 13. The trends indicated by the data were thought to be a reflection of the dynamic nature of the mixed-culture microbial films. Sharp decreases were observed in the COD removal rates between 40 and 60 hours. However, a relative stability in substrate utilization rates would be apparent when films were older than approximately 70 hours as shown in Figure 13, and as evidenced by the results obtained from the equilibrium period[2]. These results of an increase, a decrease and an increase in substrate utilization support only those of Hoehn[3], although he used film thickness instead of film age as the independent variable.

Effect of Biological Film Thickness

Figure 14 shows the effect of film thickness on the COD removal rate. The two straight line portions of each curve were drawn by the method of lease squares.

It is evident that there was a rapid uptake of organics by films in the early periods of their growth. Utilization rates were then increasing more slowly and relative stability was finally achieved. Similar trends of rate changes have been reported by other investigators[3,4,5]. Figure 14 shows that reestablishments of the normal efficiencies of treatment were achieved when the films on the first, second and third disks were, respectively, 135, 220 and 265 microns thick.

While Figure 14 depicts average rates, it does not manifest trends that might have occurred during the individual runs. Therefore, data are presented separately for runs 1 and 2 of the third stage disk in Figure 15. Sharp decreases were observed in the COD removal rates between 350 and 600 microns. A relative stability in organic utilization rates was apparent when films were thicker than approximately 600 microns or more than 70 hours old as shown in Figures 15 and 13 respectively, and as evidenced by the constancy of the data from studies during the equilibrium period[2]. These results support only those of Hoehn[3].

DISCUSSION

Effect of Detention Time at Constant Organic Loading

A decrease in detention time from 96 min. per stage to 24 min. per stage by varying flow and influent concentration at a constant organic loading was found to have no appreciable effect on the organic removal efficiency of the RBC. It was also found that decreasing detention time from 48 min. per stage to 17 min. per stage by varying reactor volume at a constant organic loading had no significant effect on the organic removal. Due to the restrictions in the physical conditons, lower detention time levels were not obtained. The results indicate that the critical detention time was equal to or less than 17 min. per stage, and the system could be operated at a detention time of 17 min. per stage without losing efficiency.

Effect of Influent Concentration at Constant Flow

It was found that increasing influent concentration at a constant flow resulted in an increase of the COD removal rate while the percentage COD reduction decreased. These results are not unexpected. Organic removal is also mass transfer limited even if the mixed liquor dissolved oxygen content is a controlling factor. Mass transfer is directly proportional to the concentration gradient. Increasing influent concentration increases the concentration gradient which, in turn, increases mass transfer and thereby increases the organic removal rate. However, above a certain loading the biomass and the resulting organic removal rate is not sufficient to prevent the overall percentage reduction from dropping to a lower level.

Effect of Flow Rate at Constant Influent Concentration

Increasing flow rate at a constant influent concentration resulted in an increase in both effluent COD concentration and rate of COD reduction with a decrease in the percentage and mg/l COD reduction. The most plausible explanation for these results is that organic removal is mass transfer limited because organics in the layer of liquid immediately adjacent to the slime layer is depleted rapidly. Mass transfer is directly proportional to the concentration gradient. At low flow rates the concentration gradient penetrates into the bulk liquid film thus reducing the magnitude of the concentration gradient. Increasing feed rate reduces penetration of the concentration gradient into the bulk of the film, until at high feed rates depletion of organics is limited to the area immediately adjacent to the slime; as a result, the concentration gradient extends over a shorter length and is numerically larger[3]. At high flow rates, the concentration gradient is limited to a thin liquid film adjacent to the slime layer, and further increases in liquid feed rate have no effect. Therefore, organic removal becomes independent of flow rate at high feed rates, as seen in Figure 8.

Effect of Film Age and Thickness

Each run made in this investigation to assess the organic removal by biological films indicated clearly that at film ages between 40 and 60 hours or thicknesses between 350 and 600 microns the rates of organic removal changed markedly. The composite data of Figures 10, 11, 12 and 14 would suggest that the removal rates stabilized at some constant values when films were between 25 and 50 hours old or between 135 and 265 microns thick.

Torpey et al.[6] found that only 18 hours after cleaning were required to restore a biological growth on the disk surfaces of the first stage with the reestablishment of normal efficiency of treatment. They did not report the procedures used for this evaluation. However, it seems that no continuous slime removing was made by them before samples were withdrawn in order to purge the original treated liquid in the reactor. This is a possible explanation for the difference between the limiting film ages obtained by Torpey et al.[6] and this investigation. Another possible explanation for this difference is that different cultural and operational conditions existed in the two studies.

Perhaps the most significant trait of the composite removal rate curves (Figures 10, 11, 12 and 14) is that they tend to support the theory that the assimilation capacity of the biological films remains constant beyond some limiting film age or thickness. It should be emphasized, however, that this is based on composite data. It is useful, therefore, only in the determination of the average results.

The limiting film ages found for disks one and two in the first study were much less than those found in the second and third studies. This is most likely due to the fact that the organic loading in the first study was only one-half of that in the other two studies. Therefore, the organic removal rate in the first study was less than those in the other two studies. This, in turn, enabled the films in the first study to reestablish their normal efficiencies in shorter times. However, the limiting film age for the third disk in the first study, 38 hours, was more than that in the third study. This event along with the fact that each disk in each study required different periods of time to recover their former efficiency may possibly be explained on the basis of different cultural conditions on each disk.

Figure 14 shows that reestablishments of the normal efficiencies of treatment were obtained when the films on the first, second and third disks were, respectively, 135, 220 and 265 microns thick. Greater film thicknesses resulted in no increase in the organic removal rates. These values were taken as the thicknesses of the active microbial films. The above three active film thicknesses differ from each other. This again may possibly be explained on the basis of different cultural conditions on each disk.

Data presented by Kornegay and Andrews[7] and Tomlinson and Snaddon[5] show that the organic removal stabilized at the maximum rates observed when the limiting thicknesses were reached. However, both groups of data are of composite data.

Even in studies conducted with strictly controlled laboratory systems, the investigator cannot duplicate exactly the cultural conditions each time. Differences in microbial metabolism are brought about by slight shifts in predominance of organism-type. Therefore, if the results of metabolic studies of several film cultures are considered together, the fluctuations in data from the individual studies will most likely be masked. An average result will therefore be defined.

Cultures were not reproducible in every detail in this investigation. The individual organic removal curves reflect differences that were masked by the composite curves.

The individual utilization rate curves (Figures 13 and 15) were interpreted as indicating that some factor or factors caused organic uptake by the microbial films to decrease sharply when they were between 40 and 60 hours old or 350 and 600 microns thick. Relative stability of organic removal rates and recovery to their former levels were achieved by the time ages of approximately 70 hours or thicknesses of about 600 microns were reached. Thereafter, a quasi-steady state with respect to organic removal was apparent.

Sanders[8] showed that organic removal rates for microbial films decreased after the limiting film thickness for oxygen diffusion to the lower microbial layer had been exceeded. However, there is evidence, based on Sanders' own data, that the rates were starting to increase at the time the experiments were terminated. Had the growth not dropped off, recovery of the films might have been manifested by further increases in the organic removal and a reestablishment of the former rates.

Maier[4] did not study films less than 480 microns thick. It is possible that changes in organic removal rate would occur before this thickness was reached.

The observed decreases in the individual organic removal rates might be caused by some film instability brought about by the limitation of some necessary growth factor. This growth factor may have been oxygen, as proposed by Sanders[8], Kornegay and Andrews[7], and Maier[4], or it may have been nutrient as proposed by Tomlinson and Snaddon[5]. The limiting of either oxygen or nutrient would result in population changes within the film, and a period of readjustment would be required. It is postulated that during this period the organic removal rate would decrease particularly if endogenous respiration were increasing. The latter would result in utilization of nutrient either stored in the film or supplied by the microorganisms themselves and, in effect, reduce the removal of organics from the incoming wastes. Organisms within the film would be dying during this period which would provide an additional source of nutrient to the living microorganisms.

The findings of Hoehn and Ray[3] for the reversal effects of film thickness on nutrient utilization rates are in good agreement with the findings of this study, although the absolute magnitude of the maximum removal rates and the film thickness at which they occurred differ from those of this investigation due to different cultural conditions in the two studies. Their finding that composite results mask the reversal effect was also noted in this work.

SUMMARY AND CONCLUSIONS

1. Organic removal in the RBC can be limited by the dissolved oxygen content, diffusion or organic concentration.

2. Increasing influent concentration at a constant flow resulted in a decrease of both the percentage COD reduction and the dissolved oxygen content of the mixed liquor, while the rate of COD removal increased.

3. Increasing flow rate at a constant influent concentration resulted in an increase in both effluent COD concentration and rate of COD reduction, with a decrease in the percentage and mg/l COD reduction and dissolved oxygen content of the mixed liquor.

4. At low organic loadings, most of the organics is removed in the first few stages of the RBC with very little being utilized in the following stages. This allows NH_3-N removal to take place in these later stages when the COD and BOD are reduced to 50 mg/l and 14 mg/l respectively.

5. Organic removal by mixed-culture biological films initially decreases when some limiting film age or thickness is reached. However, the organic removal will increase again to recover its former level as the films get older or grow thicker. A quasi-steady state (with regard to organic removal) will then be established.

ACKNOWLEDGEMENTS

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TABLE I

Synthetic Wastewater Substrate

Material	Amount*
Trace salt solution, **ml	16.66
$(NH_4)_2SO_4$,g	0.264
NH_4Cl,g	2.537
$CaCl_2.2H_2O$,g	0.294
$MgCl_2.6H_2O$,g	0.407
$CaSO_4.2H_2O$,g	1.350
$MgSO_4.7H_2O$,g	3.650
$Na_3PO_4.12H_2O$,g	3.650
KH_2PO_4,g	6.434
K_2HPO_4,g	16.430
$Na_2HPO_4.7H_2O$,g	1.500
Nutrient broth, ***g	9.000
Glucose,g	3.798
Sodium oleate,g	0.603
COD,mg/l	360

*Diluted to 45 l with deionized tap water.

**Dilute 5.0 g of $FeCl_3.6H_2O$, 0.672 g of $AlCl_3.6H_2O$, 0.342 g of $CoCl_2.6H_2O$, 0.15 g of $MnSO_4.H_2O$, 0.06 g of $(NH_4)_6Mo_7O_{24}.4H_2O$, and 0.01 g of $ZnCl_2$ to 1 l with distilled water.

***Bio Cert™ Nutrient Broth, Dehydrated, J-1089-C, Fisher Scientific Company, Pittsburgh, Pa.

Table II

Detention Time Study Conditions

Study	Organic Loading as COD g/day	Organic Loading as COD lb/1000 ft^3/day	Organic Loading as COD lb/100 ft^2/day	Flow Rate (1/day)	Influent COD Conc. (mg/1)	Reactor Volume (1/stage)	Detention Time/Stage (minutes)	Total Detention Time (minutes)
Varying Flow	15.37	114.2	1.28	84	183	1.40	24	144
and Influent	14.59	108.4	1.22	42	347.5	1.40	48	288
Concentration	15.25	113.3	1.27	21	724	1.40	96	576
Varying Only	14.64	108.8	1.22	42	349	0.50	17	102
Reactor	15.02	111.6	1.25	42	358	0.70	24	144
Volume	14.59	108.4	1.22	42	347.5	1.40	48	288

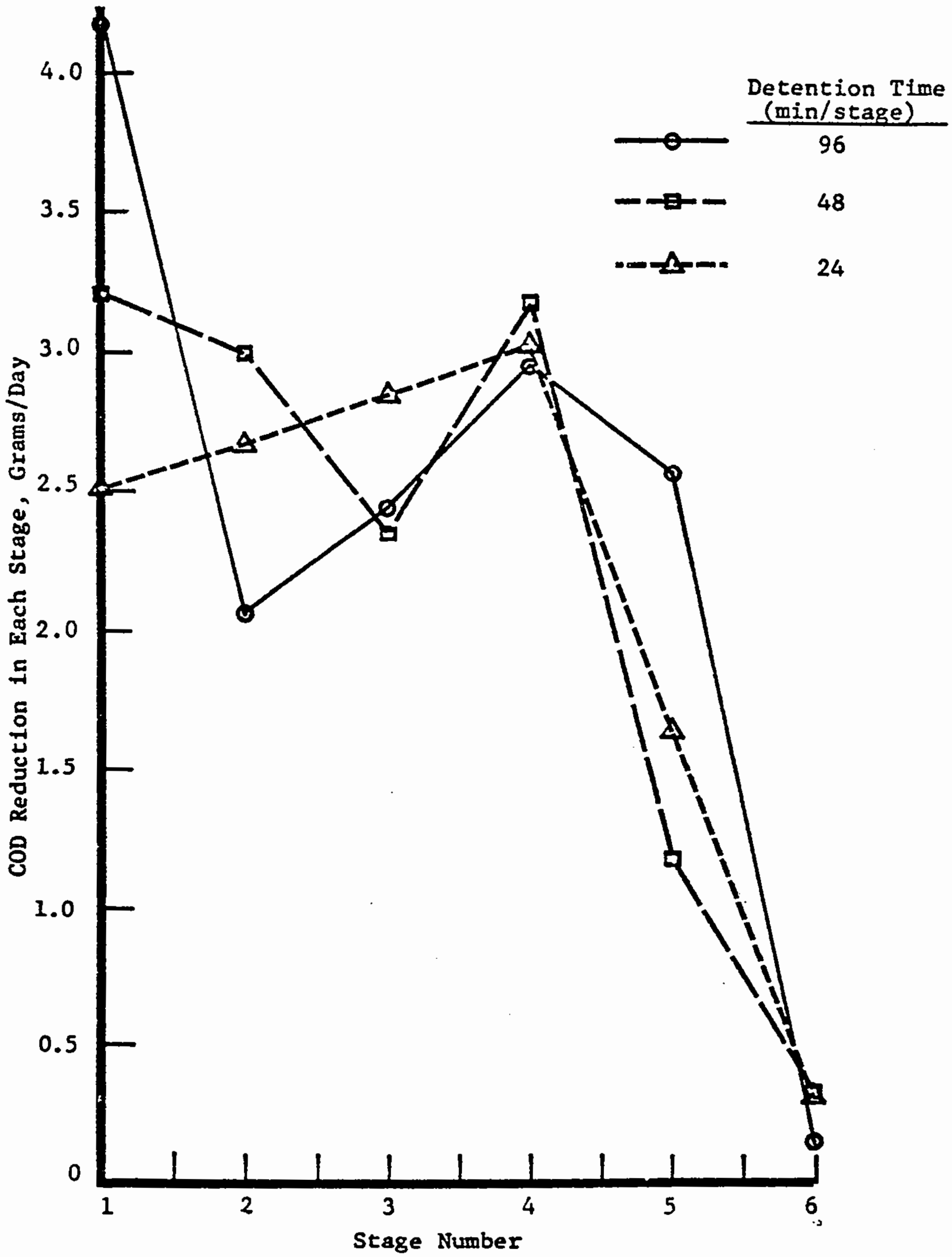

Figure 1 Effect of Detention Time by Varying Flow and Influent Concentration at Constant Organic Loading

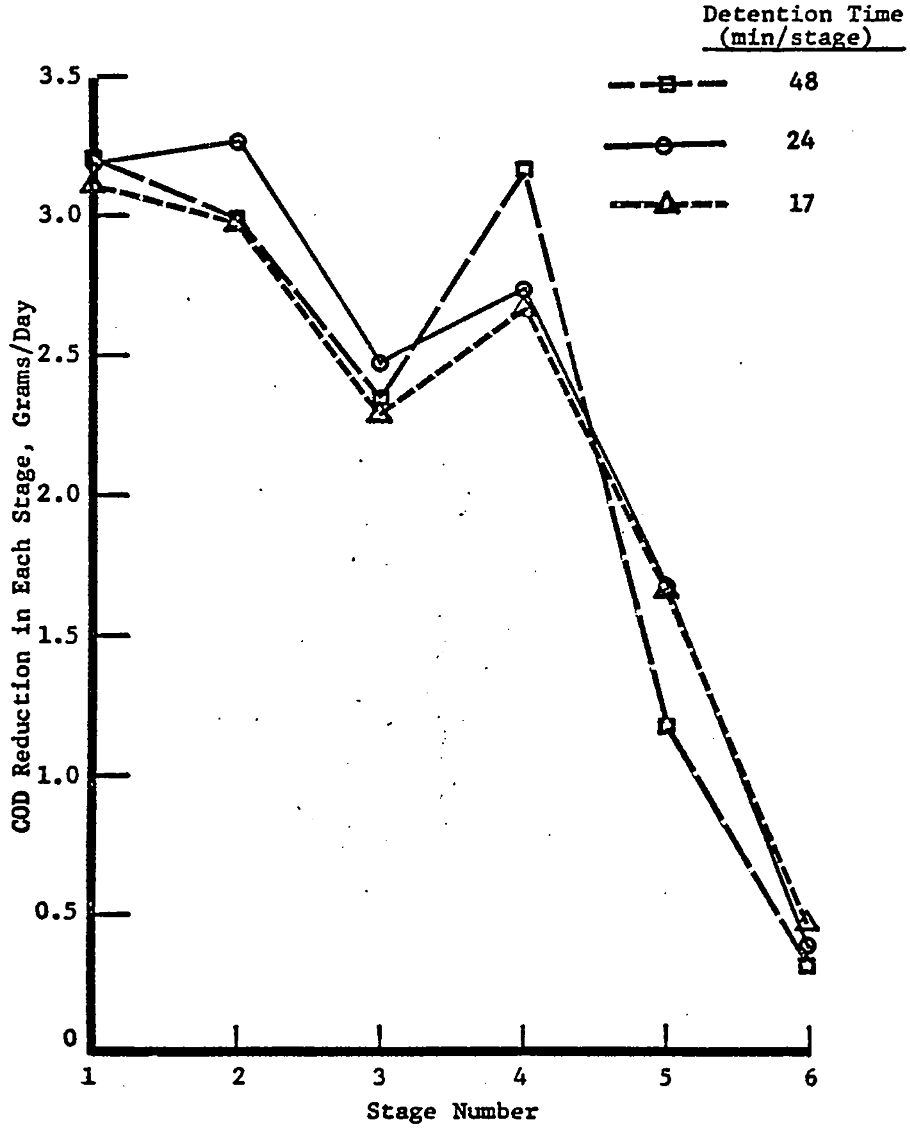

Figure 2 Effect of Detention Time by Varying Reactor Volume at Constant Organic Loading

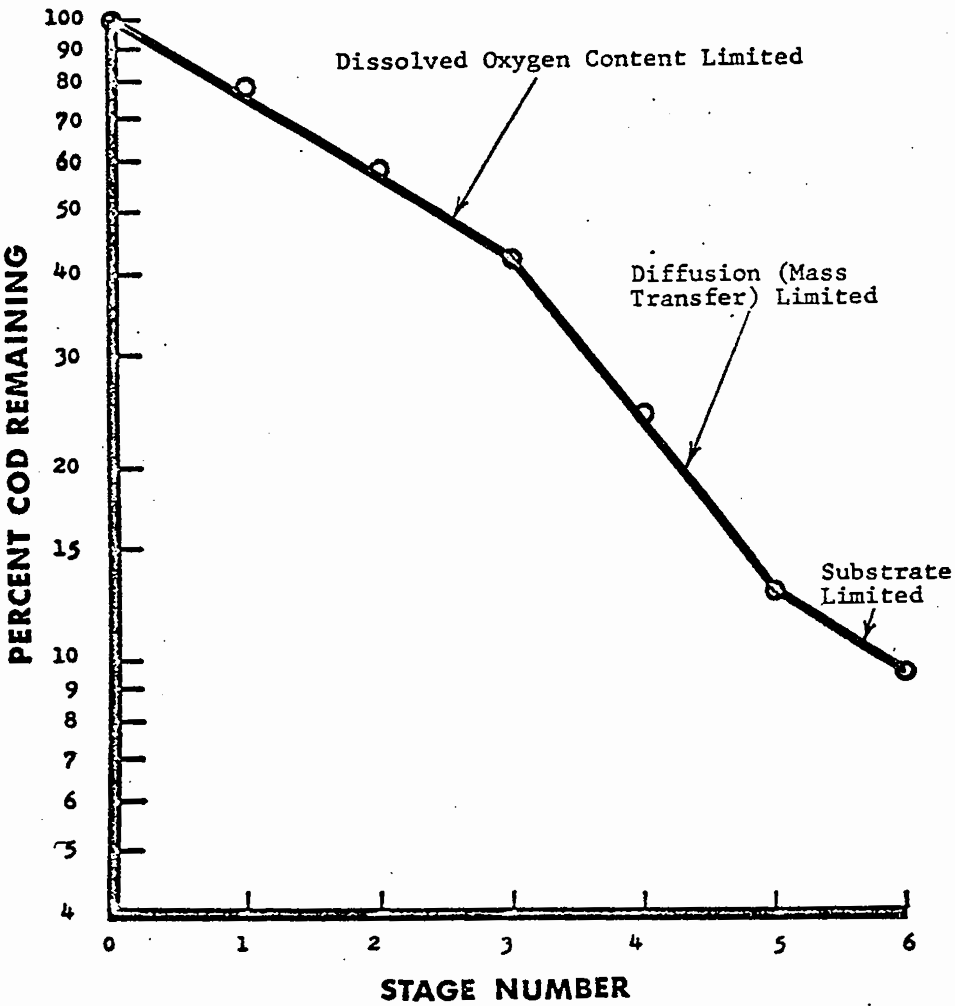

Figure 3 Effect of Dissolved Oxygen and Substrate Concentration

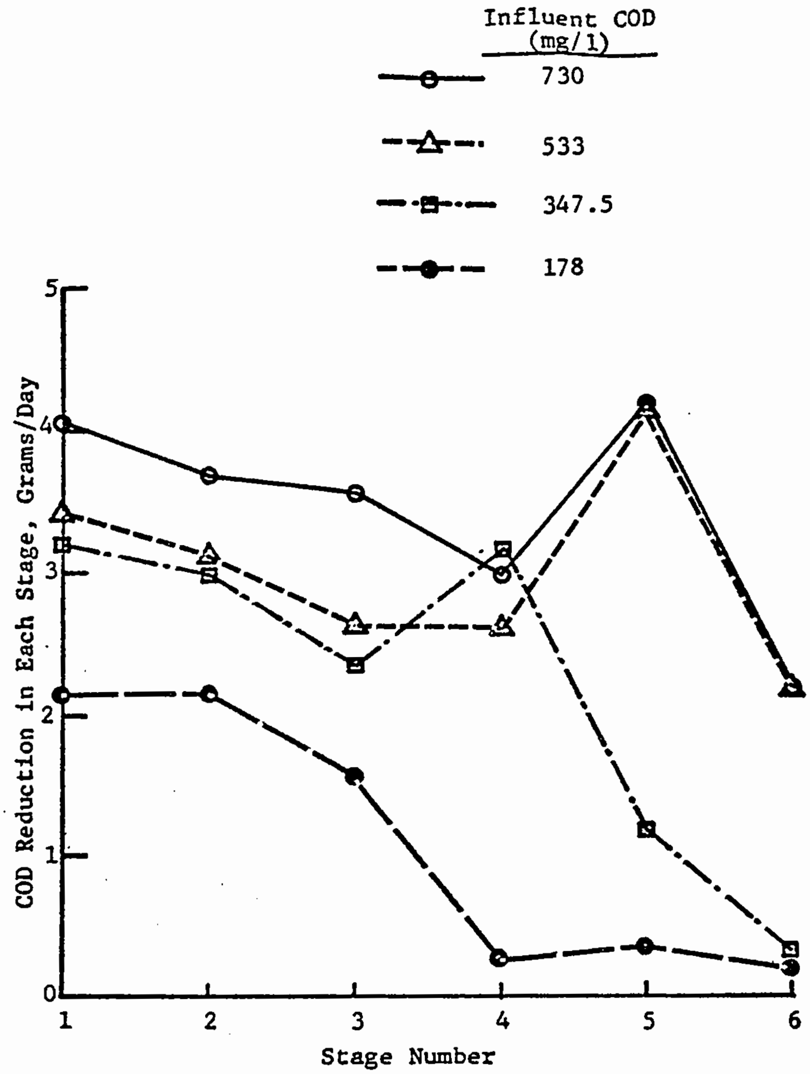

Figure 4 Effect of Influent Concentration on COD Reduction at Constant Flow with Single Disk

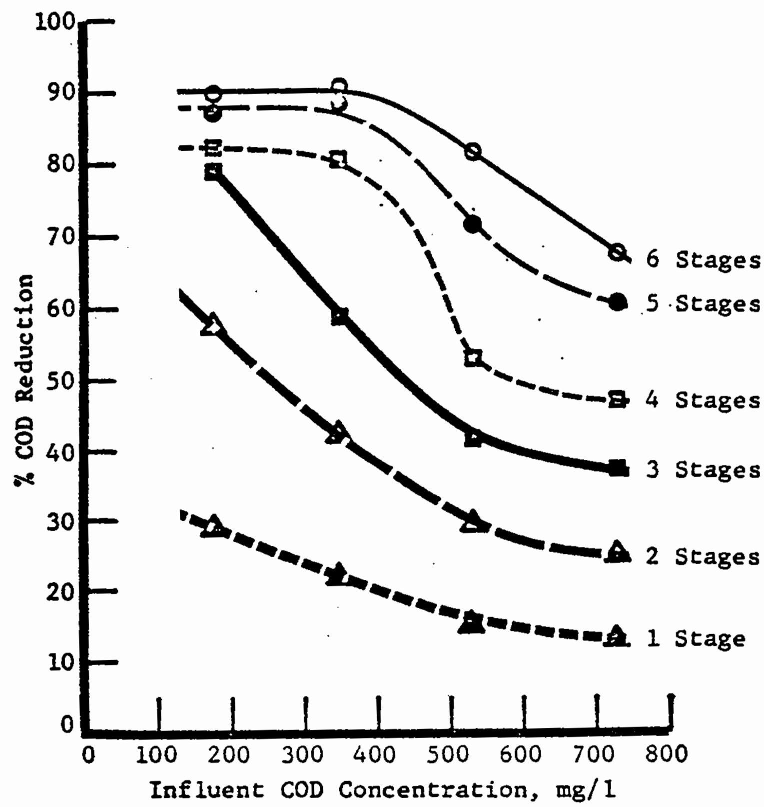

Figure 5 Effect of Influent Concentration on % COD Reduction at Constant Flow with Single Disk

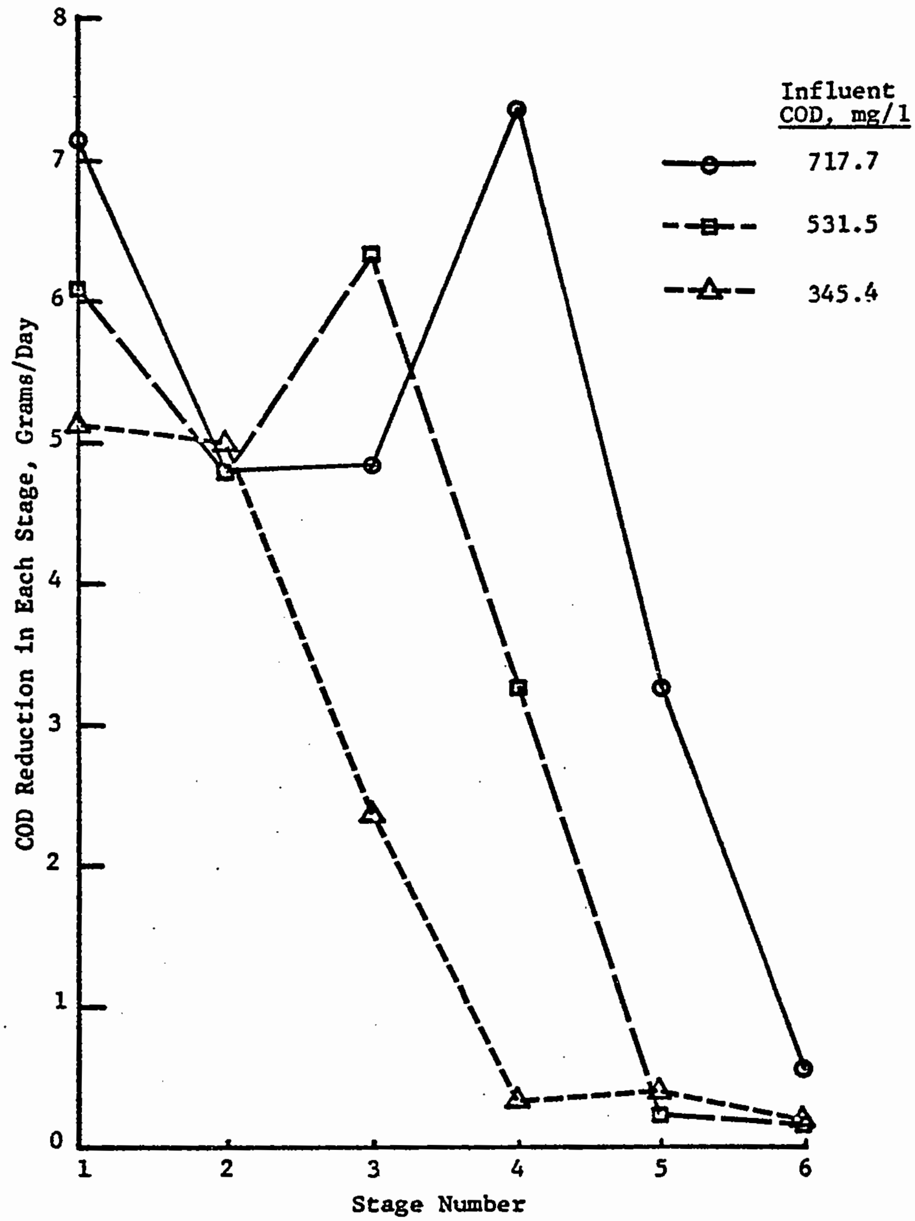

Figure 6 Effect of Influent Concentration on COD Reduction at Constant Flow with Double Disk

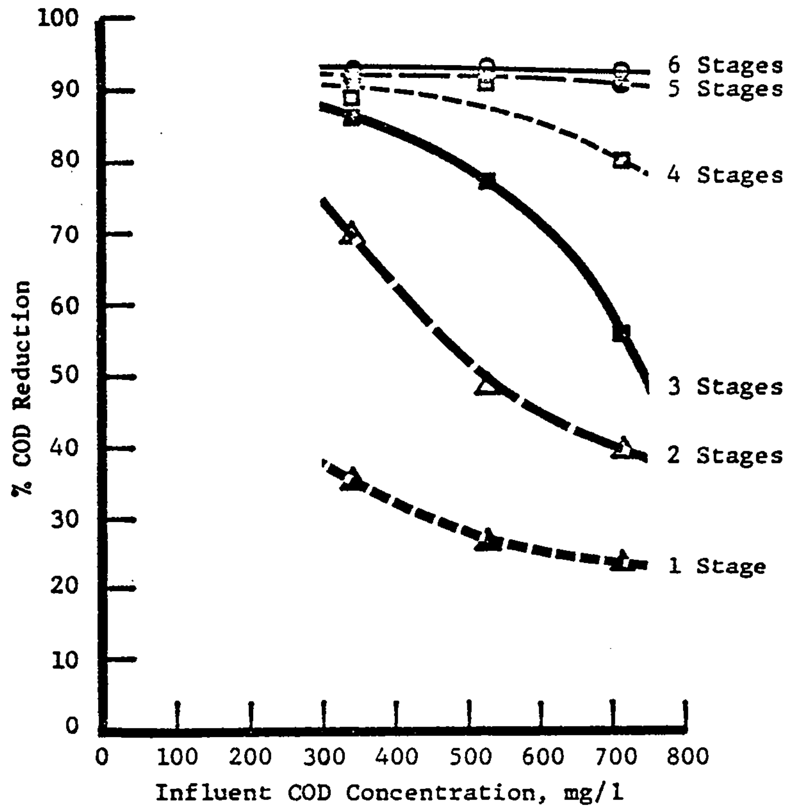

Figure 7 Effect of Influent Concentration on % COD Reduction at Constant Flow with Double Disk

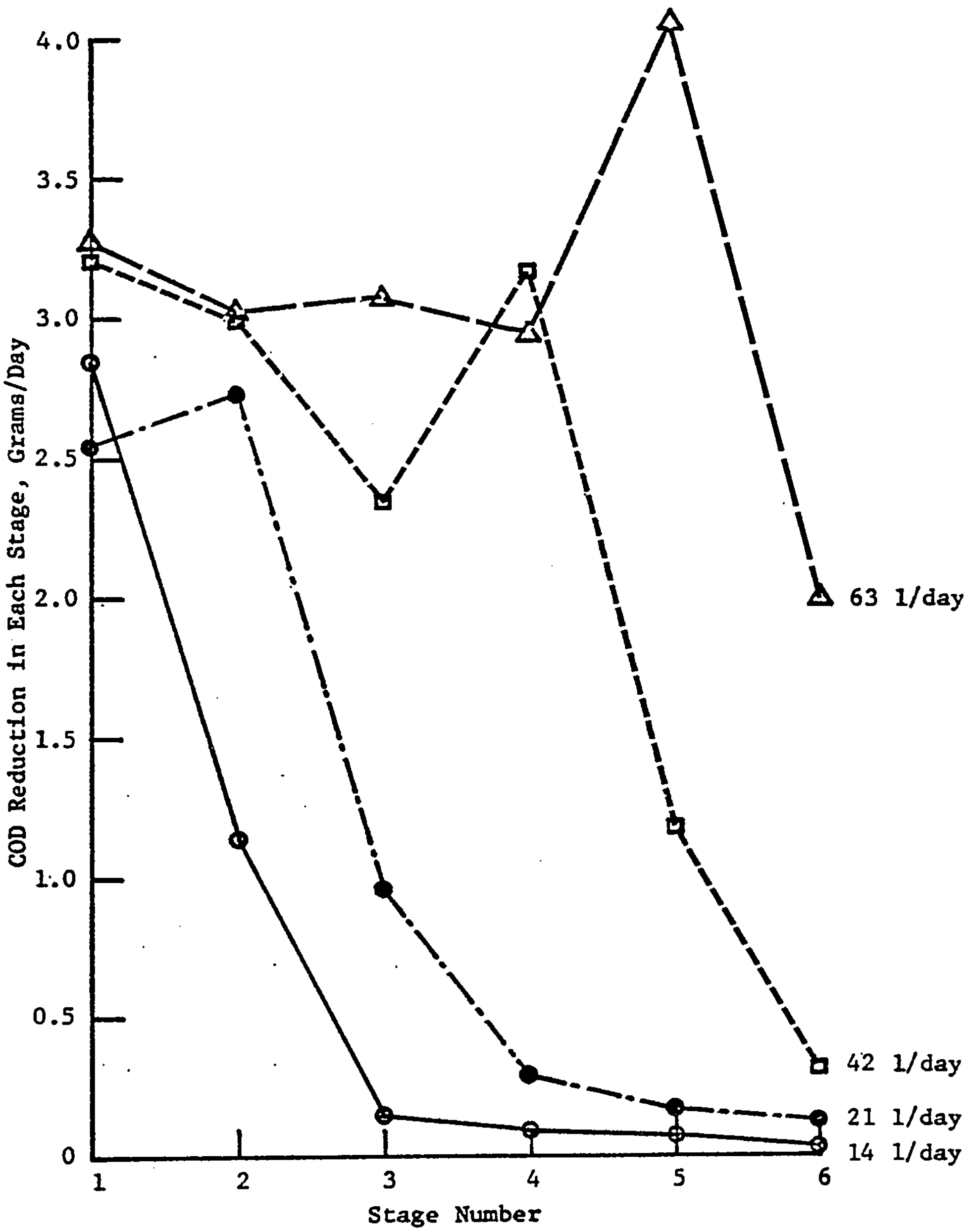

Figure 8 Effect of Flow Rate on COD Reduction at Constant Influent Concentration

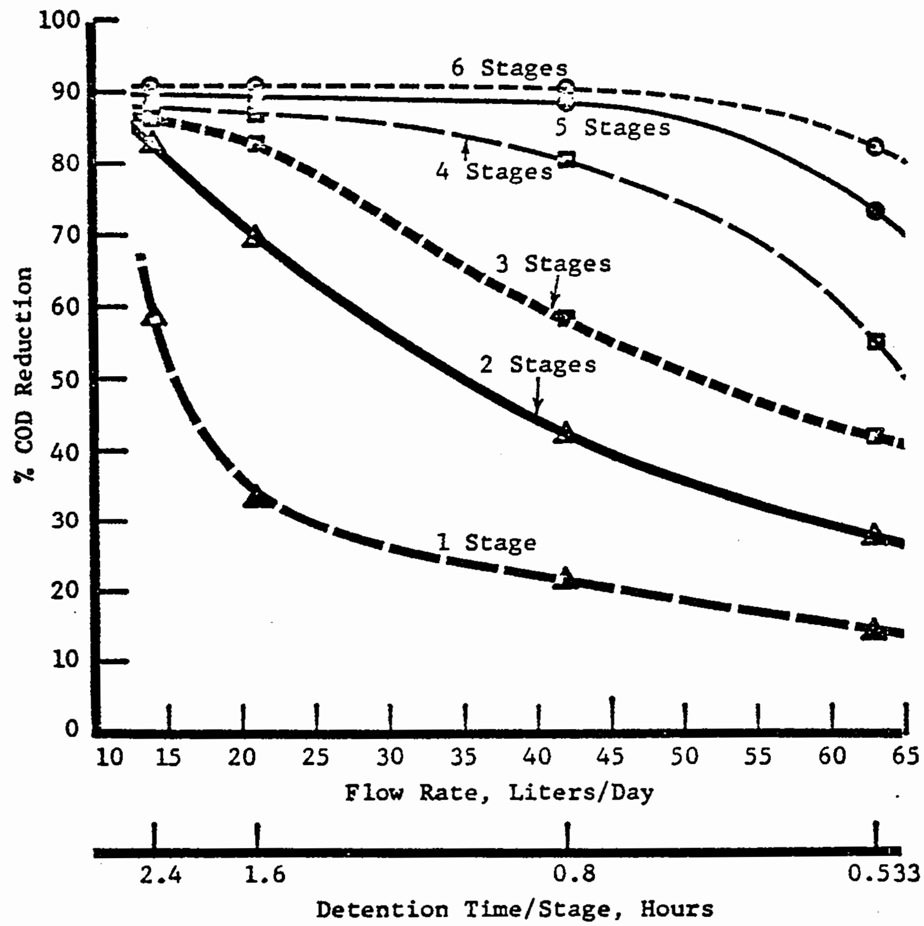

Figure 9 **Effect of Flow Rate on % COD Reduction at Constant Influent Concentration**

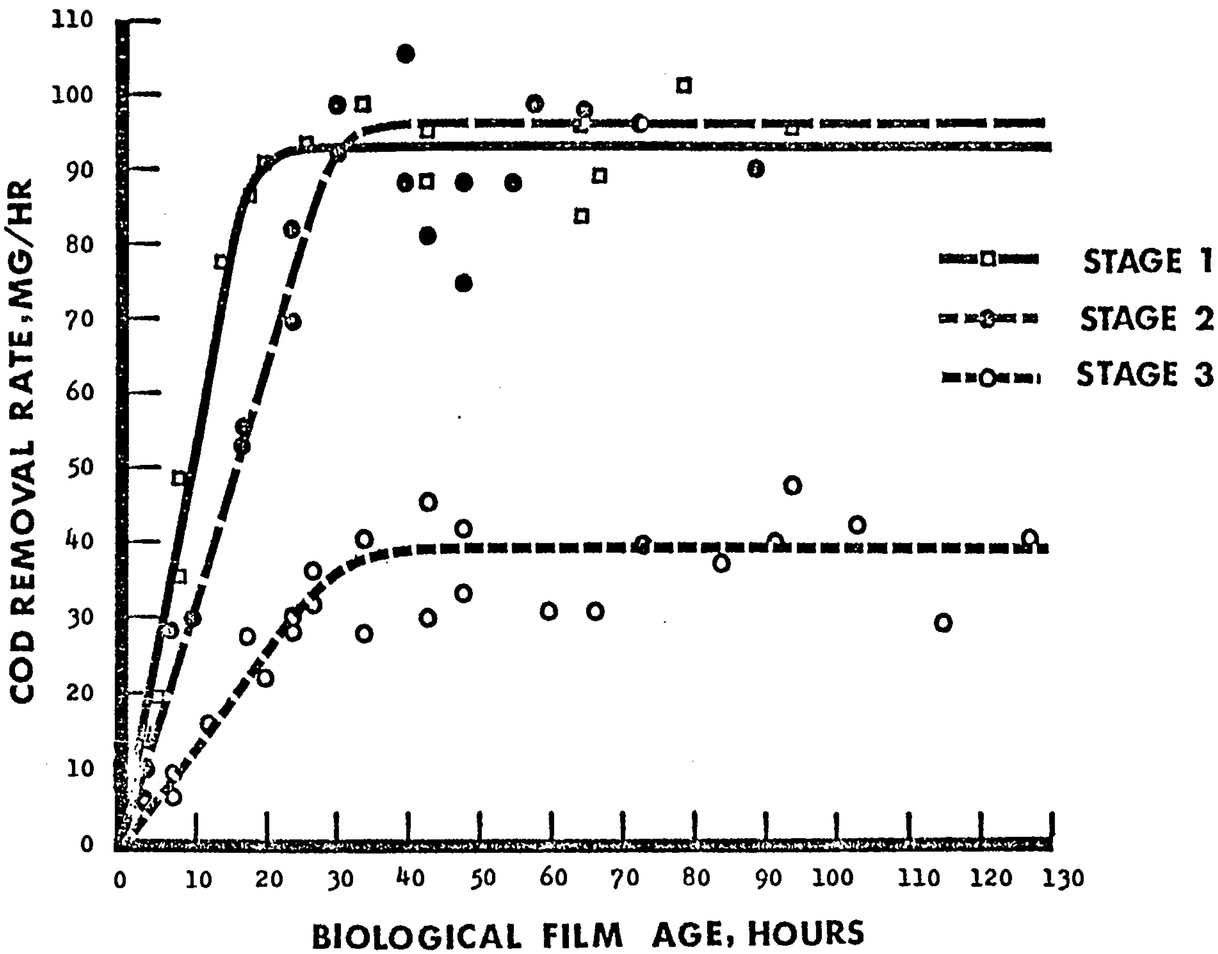

Figure 10 Effect of Slime Age on COD Removal Rate - First Study

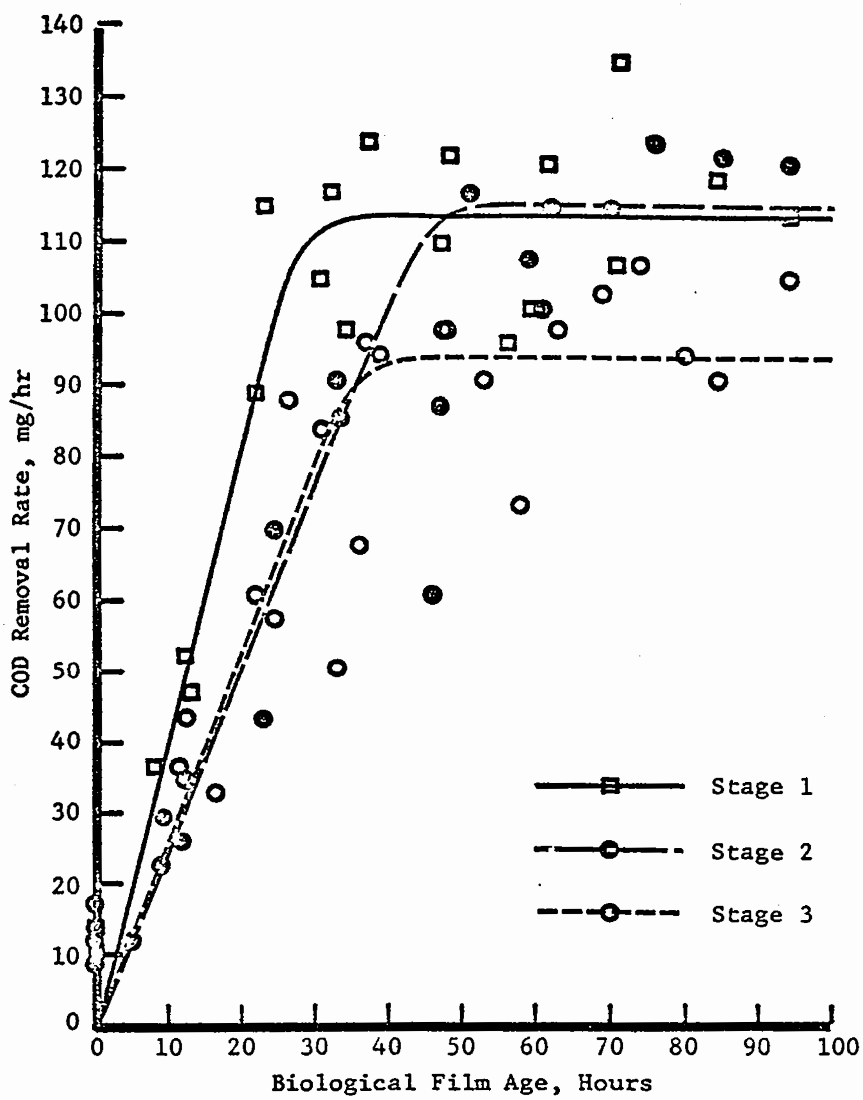

Figure 11 Effect of Slime Age on COD Removal Rate - Second Study

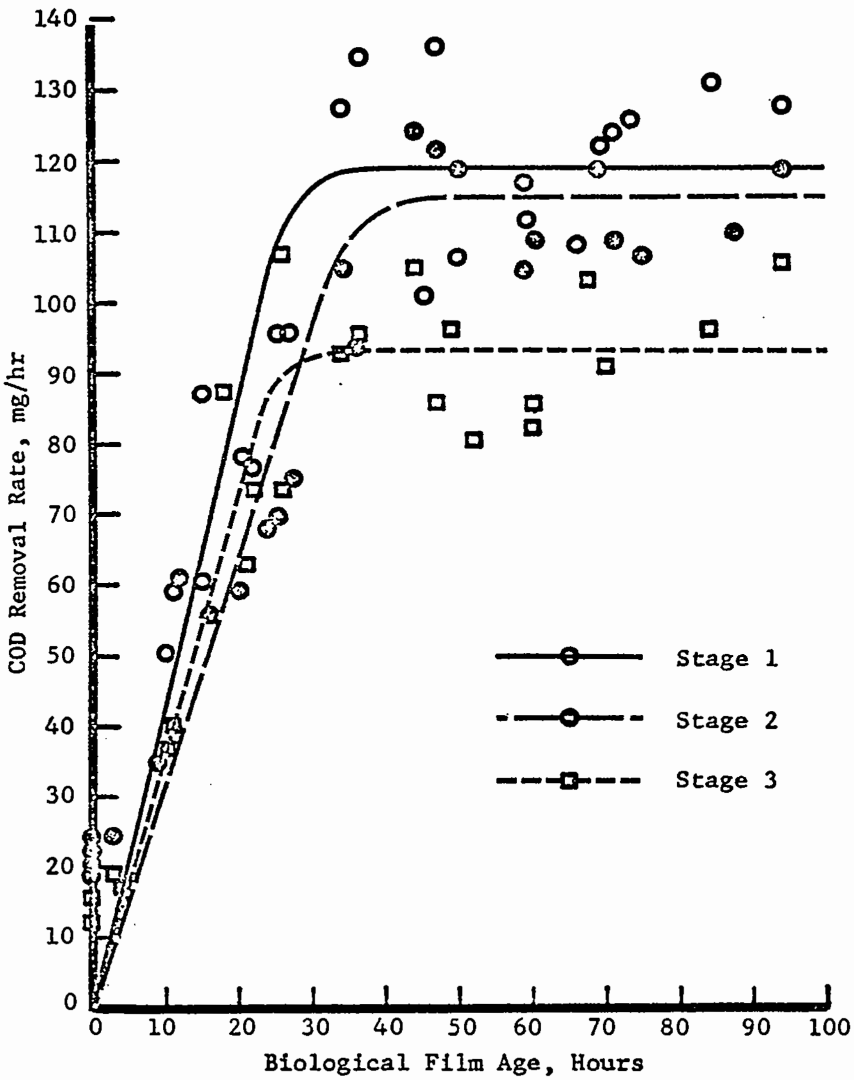

Figure 12 Effect of Slime Age on COD Removal Rate - Third Study

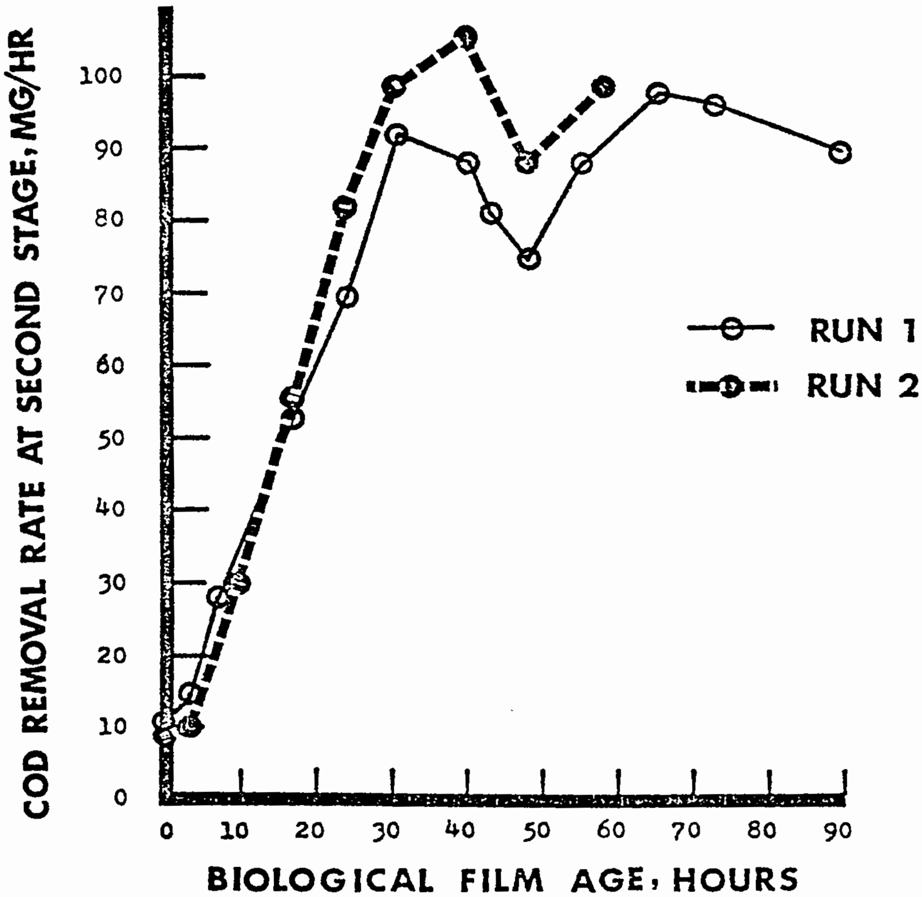

Figure 13 COD Removal Rate at Second Stage

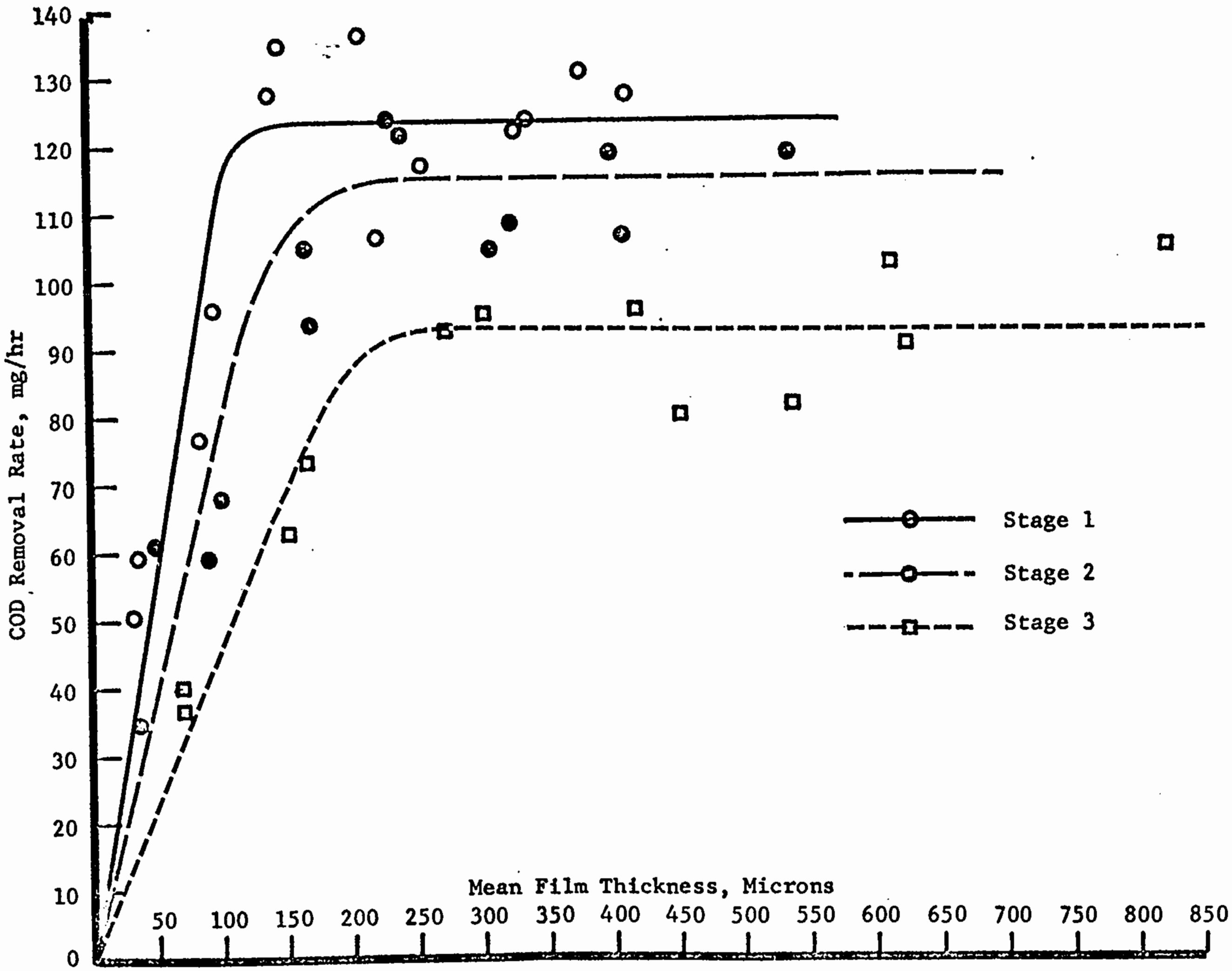

Figure 14. Effect of Slime Thickness on COD Removal Rate

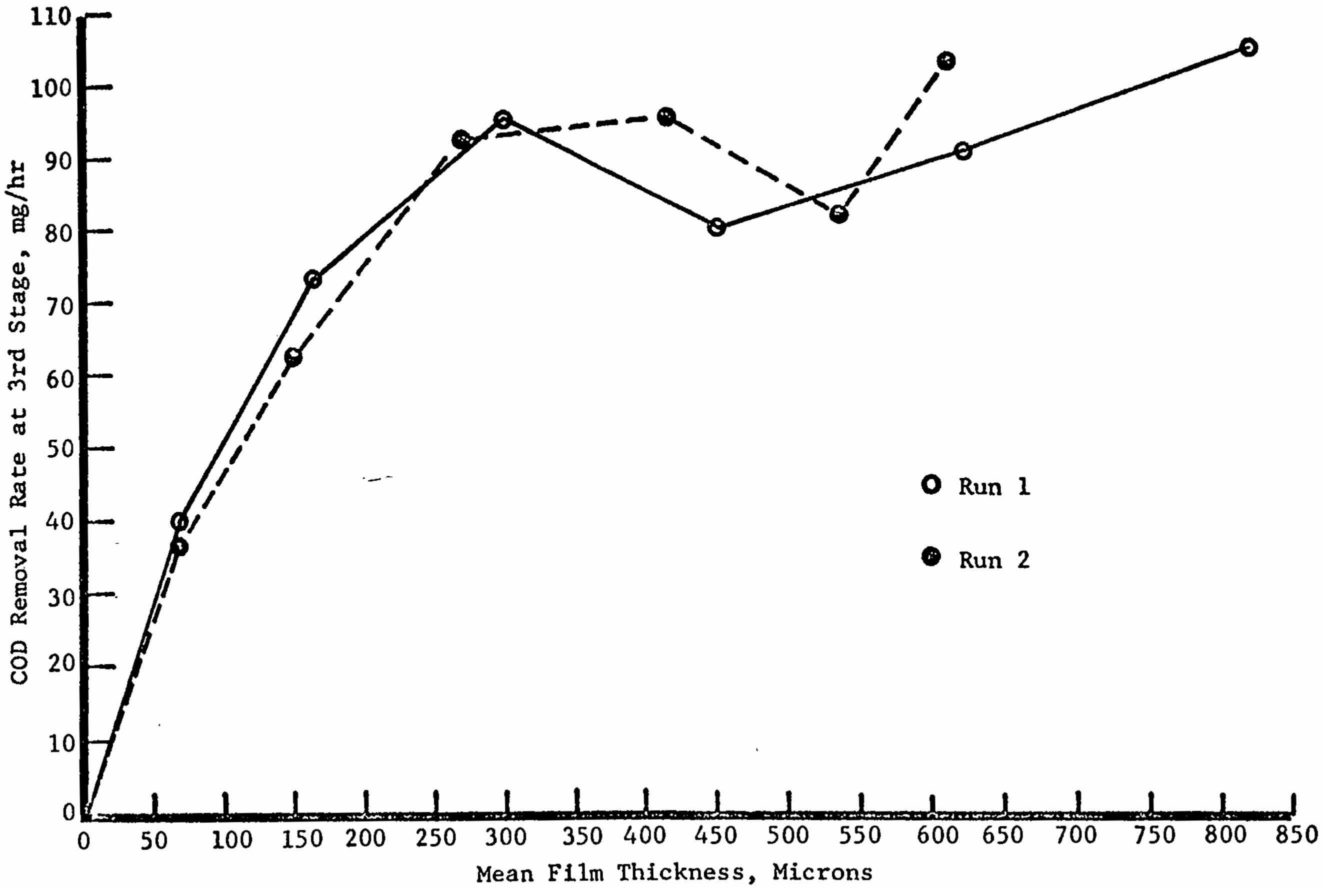

Figure 15 COD Removal Rate at Third Stage

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MICROFAUNA AND RBC PERFORMANCE: LABORATORY AND FULL-SCALE SYSTEMS

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Introduction

The transformation of soluble substrate to cells, metabolic products and excreta is the basis of the rotating biological contactor process. Metabolic activities associated with the rotating biological contactor occur in fixed-film, or as they are more commonly termed, biofilm cultures. Bacterial populations exist as a response to combinations of various environmental factors. The sum of all these factors constitute, in an ecological sense, a niche, or a specialized environment suitable for microbial occupation. These saprophytic bacteria are in turn predated by holozoic organisms such as the Phyla: Protozoa, Rotifera and Nematomorpha. Some members of these Phyla may prey upon either bacteria, algae or other members of the above-mentioned Phyla. Thus, the niches for these organisms are defined by the presence of other organisms as well as environmental factors.

Factors that affect the growth of microbial populations are generally: I. Physical, II. Chemical, and III. Biological. These factors individually and synergistically alter suitability and as a consequence, abundance of particular organisms. The abundance of an organism in an environment can be used as a measure of its importance in that ecological structure.

Biological treatment processes select for particular organisms by creating very defined niches. However, while these processes are fundamentally dependent on the activity of living organisms, their design does not account for the specific characteristics of the organisms.

Recently, problems associated with activated-sludge bulking have stimulated morphological studies of sludge-floc particles.[1,2] Control of activated sludge filamentous bulking by increased hydraulic mixing was examined by Chudoba, et al. (1973).[3] They found that increased staging caused more distinct biological cultures, and a decrease of filamentous organisms in pre-clarified effluent.

Hawkes (1963) integrated fundamental relationships of biology and ecology to trickling filter and activated sludge operations.[4] He recognized that there was a succession of microfaunal organisms affecting the efficiency of biological treatment. An informative representation of microfaunal food chains was developed for both processes.

McKinney (1962) briefly discusses predominance of microfauna in wastewater treatment processes and presents an interpretation of microfaunal succession with respect to degree of treatment.[5] He indicates that protozoa are particularly valuable indicators of the performance and stability of treatment processes.

Various interactions between aquatic organisms have been developed by Bungay and Bungay (1968).[6] A list and simple definitions are found in Table 1.

TABLE 1

MICROBIAL INTERACTION

Interaction	Definition
Neutralism	No interaction
Commensalism	One benefits, other unaffected
Mutualism	Each member benefits from the other
Competition	A race for nutrients and space
Amensalism	One adversely changes the environment for the other
Parasitism	One organism steals from the other
Predation	One organism ingests the other

Although Table 1 contains a convenient list of interactions, and all are possible in the rotating biological contactor. However, clear cut boundaries cannot always be defined.

Recently, Commensalism, Mutualism, Amensalism and Predation have been regarded as important factors mediating activities of aquatic organisms.[7]

Competition and predation, in these authors' view, are of great importance in the biofilms associated with rotating biological contactors. Changes in process parameters such as hydraulic retention time, cellular retention time, and specific growth rates of bacteria and protozoa may have an effect on the predator-prey relationships in the biofilm.

The importance of predator-prey interactions was examined in continuous flow systems by Pirt and Bazin (1972).[8.] The necessity of maintaining predator and prey growth rates either greater or less than the dilution rate was discussed. They claimed that more efficient substrate conversion and/or microfaunal predation can be attained if these dilution rates are properly controlled.

A reduction in effluent biomass due to protozoa predation of bacteria is reported to be significant (Sudo and Aiba, 1972).[9.] They state that the average yield of protozoan mass from bacterial mass is 0.5.

Most investigations of microfaunal interactions as a response to changes in operating parameters are conducted in either pure or mixed suspended growth systems. The list of environmental factors affecting organisms in the RBC process may be different from the list of factors that affect organisms in other processes. Where the lists have factors in common, the biological responses to the common factors may be different.

Full-scale activated sludge and trickling filter culture studies have revealed that particular species and taxa occur with greater frequencies in these cultures. Curds and Cockburn (1970a, 1970b) examined a variety of trickling filters and activated sludge units. They determined the frequency that individual species were present in these processes.[10., 11.] They found that ciliated protozoa were generally the most abundant microfauna in effluents from both processes. Also, the microfauna of both processes were similar in that the organisms were from the same orders. Individual species within these orders did vary. Activated sludge units that did not contain ciliated protozoa contained significant populations of flagellates. Effluent from these plants was turbid and of low quality.

To date, there is a lack of published investigations addressing the types, abundance and ecology of microfauna in rotating biological contactors. In view of the importance of microfauna in biological wastewater treatment processes, we decided to examine the organisms associated with these biofilms. The following were objectives of this study:

I. To examine what types of bacteria are responsible for soluble substrate conversion.

II. To identify predatory microfauna that ingest substrate removing bacteria.

III. To investigate successional patterns exhibited by the microfauna in a rotating biological contactor.

IV. To evaluate the possibility of using microfauna as indicators of effluent quality and stability of rotating biological contactor biofilms.

Methods and Materials

Results from the examination of biofilms from two rotating biological contactors are presented in this paper. A plexiglass laboratory scale unit was constructed and an experimental program initiated to analyze various components of the system. Results from a $\frac{1}{2}$-2^5 factorial design experiment have been reported elsewhere.[12] A full-scale Clow-Envirex rotating biological contactor used to treat an overloaded trickling filter effluent was examined during the course of study. It was hoped that during the course of study, the full-scale RBC would replace the trickling filter. However, this did not occur. Consequently, the sampling program on the full-scale unit was limited. These results will be discussed in this paper.

The laboratory unit was a continuous flow, four stage reactor. Specific design of the unit and the experimental program has been described by Hoag and Hovey (1980). Some basic characteristics at the laboratory and full-scale units are found in Tables 2 and 3.

TABLE 2

CHARACTERISTICS OF ROTATING BIOLOGICAL CONTACTOR UNITS STUDIED

	Laboratory Unit	Full-Scale Unit
Number of Stages	4	6
Discs per Stage	3	--
Disc Diameter (ft)	5	12
Surface Area per Stage (ft^2)	71	100,000
Total Surface Area (ft^2)	284	2.4×10^6
BOD Mass Loading Rate (lbs/1000 ft^2 - day)	3.6 or 7	0.12
NH_3-N Mass Loading Rate (lbs/1000 ft^2 - day)	0.3 or 0.5	0.17
Peripheral Velocity (ft/min)	56 or 76	57

TABLE 3

ONE-HALF 2^5 EXPERIMENTAL DESIGN

Factor Number	Factor	Level (-)	Level (+)
1	Revolutions per minute (RPM)	5.9	8.1
2	Hydraulic Flow Rate (liters/day)	1032	1750
3	Disc Surface Area Exposed (percent)	60	74
4	Nitrogen Mass Loading Rate (grams NH_3-N per day)	41.5	65.0
5	Carbon Mass·Loading Rate (grams glucose per day)	258	500

EXPERIMENTAL CONDITION NUMBER	ORDER RUN	FACTOR LEVELS Factor 1	2	3	4	5
1	7	-	-	-	-	+
2	1	+	-	-	-	-
3	12	-	+	-	-	-
4	8	+	+	-	-	+
5	13	-	-	+	-	-
6	2	+	-	+	-	+
7	5	-	+	+	-	+
8	16	+	+	+	-	-
9	4	-	-	-	+	-
10	6	+	-	-	+	+
11	11	-	+	-	+	+
12	10	+	+	-	+	-
13	15	-	-	+	+	+
14	3	+	-	+	+	-
15	9	-	+	+	+	-
16	14	+	+	+	+	+

In brief, the full-scale treatment plant was an overloaded trickling filter plant upgraded through the use of rotating biological contactors. At the time of the experiment, the rotating biological contactor was treating trickling filter effluent. Characteristics of the full-scale RBC given in Table 2. This treatment plant did not control pH or alkalinity. Low alkalinities, low pH and low nitrogen mass loading rates resulted in a significant underloading of the process.

EXPERIMENTAL PROCEDURE

During each of the 16 laboratory experimental runs, biofilm samples were removed from the discs by means of a glass slide. The culture was microscopically examined immediately after it was removed from the discs. Microfauna present in the cultures were identified using taxonomic keys by Kudo (1966)[13.], Martin (1968)[14.] and Gibbons (1974).[15.] Due to the difficulty of removing a fixed mass and volume of biofilm and then making representative serial dilutions, a quantitative analysis of population number was not possible. Relative numbers of individuals of a particular species were determined in a manner similar to a method used by Curds and Cockburn (1970).[10.] Population levels were recorded by estimating relative numbers of a species using the following levels: (+) few, (++) many, (x) very many, and (xx) extremely high level. While the numbers of organisms are not numerically estimated, the levels presented indicate proportionate sizes of the populations.

Biofilm was removed from the laboratory discs approximately six inches from the outer edge of the discs. Biofilm samples from the full scale unit were taken at a distance of 2.0 feet (0.6 m) from the outer edge of the discs. Samples used for the determination of wastewater chemical characteristic were taken from the bulk liquid in all cases.

In addition to the microscopic examination of the biofilms, many visual changes of the texture, color and density were observed during the experimental period.

Results

The characteristics of the biofilm cultures varied from stage to stage and often shifted between experimental runs. Biofilm thickness, in the first stage, while not measured, was observed to become very rough textured and thick with increases in carbon mass loading. Similarly, the latter stages increased thickness and roughness with increases in nitrogen mass loadings.

Microfauna of the biofilm samples also varied between stages and runs. Four classes, represented by 16 orders of protozoa, were present at some time during the laboratory study. Frequencies of occurrence of the various microfauna taxa, separated by stages, are found in Table 4. While Protozoa were classified to species, the Phyla Rotifera and Nematomorpha were classified to order.

As indicated in Table 4, members of the class Sarcodina occurred 69% of the time in the 3rd and 4th stage laboratory biofilm cultures. Amoeba

TABLE 4

CLASSIFICATION OF MICROFAUNA FROM

LABORATORY RBC BIOFILMS

TAXONOMIC CLASSIFICATION	FREQUENCY			
	Percentage of Biofilms containing Species			
Phylum Class Subclass Order Genus Species	STAGE			
	1	2	3	4
Protozoa				
Class Ciliata				
<u>Order Gymnostomatida</u>				
Chilodonella cucculus	-	6	6	-
Traecophylum pusillum	-	6	6	6
Didinium nasutum	-	-	-	6
Orthodonella guttula	6	6	-	6
Coleps hirtus	-	-	-	6
Spathidium spathula	-	6	-	6
Prorodon griseus	-	-	6	-
Prorodon teres	-	-	6	-
Litonotus lamellae	19	50	63	56
Litonotus fasciola	13	44	44	44
<u>Order Trichostomatida</u>				
Colpoda inflata	6	6	6	-
<u>Order Suctorida</u>				
Podophyphra elongata	-	6	-	-
Podophyphra fixa	-	13	6	13
<u>Order Hymenostomatida</u>				
Saprophilus muscarum	31	19	25	38
Uronema marina				6
Colpodium colpada	6			
Paramecium caudatum	31	13	38	13
Paramecium bursaria				6
Glaucoma scintillans	44	56	50	38

TABLE 4 (cont.)

	1	2	3	4
<u>Order Peritrichida</u>				
Vorticella campanula		19	31	44
Vorticella microstoma		6	6	13
Vorticella acquilata		38	38	19
Vorticella nebuifera			19	
Opercularia coarctata		25	13	13
Epistyllis plicatis		6	6	
Telotrochidum henneguyi		6	19	13
<u>Order Hypotrichida</u>				
Euplotes patella			6	6
Euplotes harpa		6	6	13
Stylonchia mytilus	6			
Stylonchia pulsata			6	
Osytricha				6
Class Sarcodina				
<u>Order Amoebina</u>				
Amoeba guttula	6	-	19	25
Amoeba proteus	-	19	69	56
Amoeba verucosa	-	-	13	-
Amoeba gorgonia	-	6	38	38
Amoeba limax	-	-	-	6
Amoeba striata	-	-	13	13
<u>Order Testacea</u>				
Diffuglia oblongata	-	-	25	69
<u>Order Proteomyxida</u>				
Nuclearia simplex	-	6	44	31
<u>Order Heliozoida</u>				
Heterophrys myriopoda	-	6	6	25
Class Mastigophosa				
<u>Order Chrysomoadida</u>				
Monas amoebina	6	19	6	6
Monas obliqua	6		6	
Monas vulgaris	6	6	13	19

TABLE 4 (cont)

	1	2	3	4
Order Phytomonadida				
Chlamydomonas sp.				6
Carteria globosa			6	
Order Euglenida				
Astasia dangeardi				6
Peranema trichophorum	6		6	31
Class-Zoomastigophosea				
Order - Rhizomastigina				
Mastigamoeba longifillum	6			
Mastigamoeba reptans	13		13	19
Order Protomastigida				
Bodo caudatus			6	13
Bodo globosus		13		
Bodo lens			6	6
Bodo mutabilis				6
Cercobodo radiatus				6
Pleuromanas jaculans				6
Order Polymastigida				
Tetramitus pyroformis			6	
Phylum Rotifera				
Class Digonata				
Order Bdelloidea				
Philodina		31	63	38
Rotatia sp.			6	
Class Monogononta				
Subordis Ploime				
Encentrum sp.				6
Phylum Nematomorpha	25	88	88	38

guttula, A. proteus, A. gorgonia, Diffuglia oblongata, Nuclearia simplex and Heterophrys myriopoda were the species most commonly found of this Class. Class Ciliata was represented by 31 species and occurred in 44%, 56%, 63%, and 56% of the cultures in Stages 1 through 4, respectively. Both Free-Swimming Ciliates and Attached Ciliates were numerous during the course of experimentation. Class Mastigophora species (Flagellates) occurred infrequently. The only flagellates which were present in 20% or more of the biofilms per stage was Peranema trichophorum (31% of 4th Stage Cultures). Phylum Rotifera was absent in all 1st stage cultures but occured in 31%, 63%, and 38% of the 2nd, 3rd, and 4th stage cultures, respectively. Phylum Nematomorpha occurred with the greatest frequency of any group (88% in both 2nd and 3rd stages).

Filamentous Bacteria were always dominant in the first stage cultures. A very feathery like series of filaments were quite evenly distributed over the entire discs. Although some of the experimental runs had lower depths of submergence, the cultures tended to cover the entire disc. At times, the bacterial growth near the center of the disc was not as thick and/or rough as was found on the submerged (wetted) area. The filiamentous bacteria were not confined to the first stage cultures in all cases. Two types of filamentous bacteria were identified during the study. A Sphaerotilus sp. (probably S. natans) and a Norcardia sp. were the major filamentous species. A Zooglea sp. (probably Z. ramigera) was present most of the time in the 2nd, 3rd, and 4th stages cultures. These bacteria formed a biological matrix in which the other microfauna lived. The amount of surface area these bacteria matrices provided to the predatory microfaunal was quite astounding to the microscope viewer.

The roughness and thickness of these bacterial matrices varied from species to species. The roughest cultures were Sphauotilus and Norcardia. Cultures associated with the Zoogleal colonies were the smoothest. Roughness and texture of the individual species did change with changes the process factors. Table 5 indicates which bacteria were dominant in the four stages for the 16 experimental conditions.

Analysis

Reduction of the number of individual taxa was conducted for the following reasons: First, while individual species may not be present in all RBC biofilms, members of the same Order may perform a function similar to that of the absent species. Second, examination of all species may not be necessary because of similarities in types food ingested and in means of motility. Third, this methodology would have a much greater application for treatment plant operators. Fourth, the necessity of tedious taxanomic identification to the level of species is alleviated. Accordingly, species were grouped together by motility. The groups, and orders contained in the groups are shown in Table 6.

Another method of reducing the burden of species identification is counting the number of species per stage without identifying the species. A list of the number of microfauna species per stage (excluding bacteria) for the 16 experimental runs is shown in Table 7. The average number of microfauna species per stage is also shown in Table 7. Although the number of species per stage increases from stage to stage, the rate of increase is much lower going from Stage 3 to Stage 4.

TABLE 5

BACTERIA IDENTIFIED IN BIOFILMS

Expr. Run	STAGE 1 S*	1 N	1 Z	2 S	2 N	2 Z	3 S	3 N	3 Z	4 S	4 N	4 Z
1	x	x	x	+		xx			xx			xx
2												
3	+	xx			x	x	++	++	xx			xx
4	xx			x	xx	++	+	+	xx	x	x	x
5	++	xx			x	x	+		xx	+		xx
6	x	x		++	++	xx	x	x	x	++		xx
7	x	x			x	x	+	+	xx			xx
8	+	xx		++	x	x	++	++	x	++	++	xx
9	xx			++	++	x			xx			xx
10	xx	++			x	++	+	x	x	x	x	x
11	+	xx			xx	++	+	xx	x			xx
12	+	xx		+	xx		x	xx	x	+	+	xx
13	x	x		+	+	xx	x		xx			xx
14		xx		+	x	x			xx			
15		xx		xx					xx			xx
16	+	xx		x	x	x	x	x	x	+	+	xx

\+ = few
++ = many
x = very many
xx = extremely high level

*S = Sphaerotilus sp.
N = Norcardia sp.
Z = Zooglea sp.

Another response was that, in some of the runs, the number of microfaunal species decreased from stage 3 to 4. This condition was observed in experiments 1, 4, 6, 10, 13 and 16.

Table 6

Grouping of Microfauna

Group	Orders or Phyla included in groups
Sarcodina	Amoebina Testacea Proteomyxida Heliozoida
Flagellates	Chrysomoadidg Phytomonadida Euglenida Zoomastigophorea Protomastigida Polymastigida
Free swimming Ciliates	Crymnostomuatida Trichostomatida Hymenostomatida Hypotrichida
Attached Ciliates	Suctorida Peritrichida
Rotifers	Bdelloidea Monogononta
Nematodes	Phylum Nenatomorpha

These 6 runs represented 6 of the 8 runs with high glucose mass loading rates (500 g/day).

Table 7

Numbers of Microfaunal Species per stage

Exp.		STAGE		
Run	1	2	3	4
1	1	6	11	9
2	-	-	-	-
3	4	11	10	11
4	1	7	7	5
5	5	10	10	13
6	1	3	4	4
7	3	6	11	12
8	5	6	9	12

Table 7 (cont.)

Numbers of Microfaunal Species per Stage

Exp Run	STAGE 1	2	3	4
9	4	2	5	6
10	1	4	8	7
11	3	6	8	9
12	3	7	15	17
13	3	4	15	12
14	-	-	-	-
15	4	12	22	15
$\bar{x}$	$\overline{2.9}$	$\overline{6.3}$	$\overline{9.9}$	$\overline{10.3}$

Microfauna in the 4 stage in all of the laboratory experiments runs were classified according to the groups listed in Table 6. Relative populations of each group were determined. An organism was identified as the dominant (most numerous) organism in each of the biofilm cultures. The group that this organism was a representative of was then identified as the dominant (most numerous) group. The dominant group in each stage was recorded. In some instances more than one organism was dominant. To investigate the possibility that a particular group was repeatedly dominant in a stage, Figure 1 was developed. The number of times a group was dominant was recorded for each stage, then plotted in Figure 1.

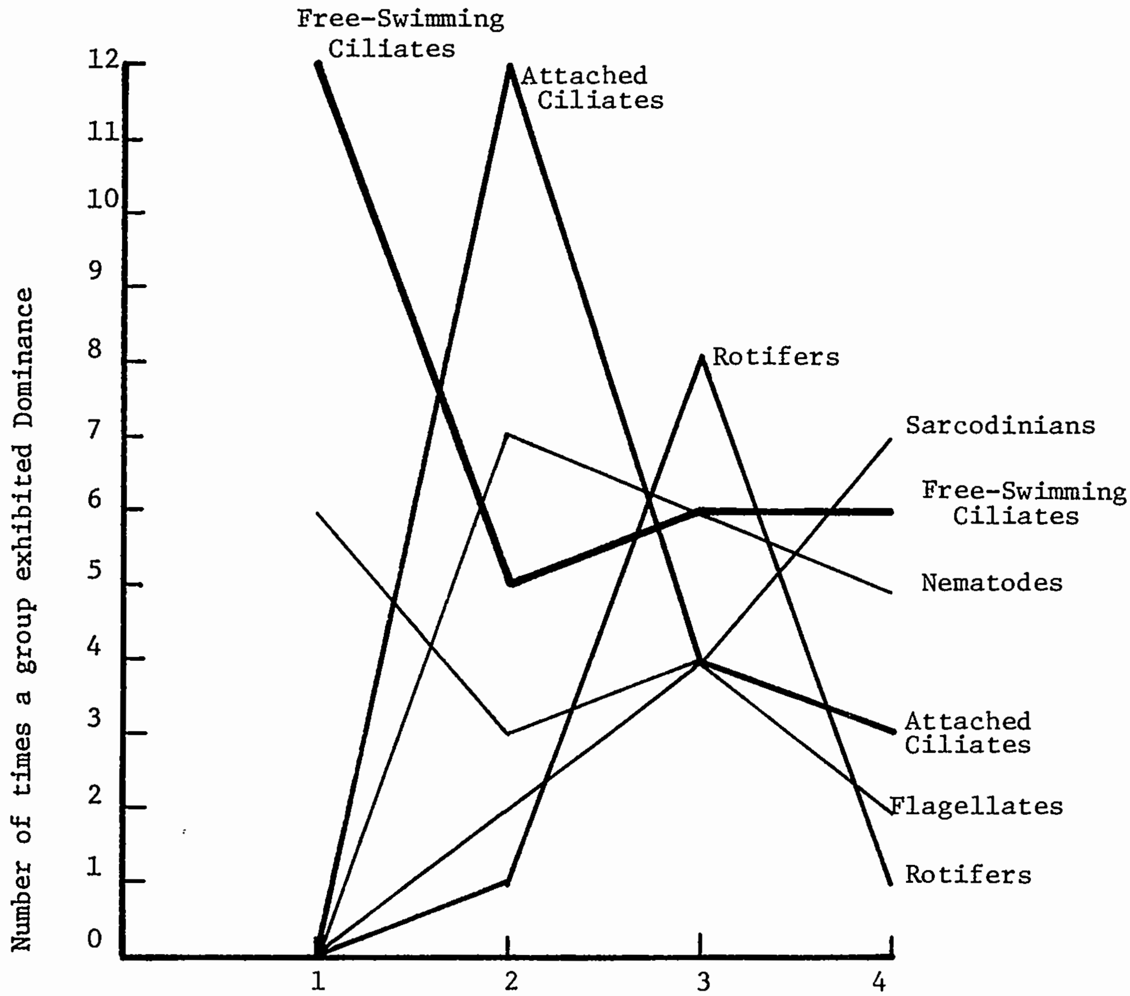
While only two groups were ever dominant in Stage 1 (Free-swimming Ciliates and Flagellates), all groups exhibited dominance in the latter stages in at least one of the experimental runs. This indicates that the niche provided in stage 1 is more specialized and/or restrictive than niches found in later stages.

The second stage was most frequently dominated by populations of attached ciliates. In 12 of the 14 runs examined they were either the dominant organisms or shared dominance with another organism while the attached ciliates were a dominant organism in stage 3 four times and in stage 4 three times their dominance in the laboratory rotating biological contactors was most pronounced in stage 2. Nematodes were a dominant group in stage 2 in 7 runs.

Rotifers were the group most often predominant in the 3rd stage (8 of 14 runs). They showed large increase in the number of times they were a predominant organism from stage 2 to 3 (from 1 to 8 times). A decrease in dominance from 8 to 1 times was also observed from stage 3 to 4 cultures.

Attached Ciliates exhibited a similar large decline in dominance from stage 2 to 3. Because they exhibit a large increase, followed by a rapid decrease in dominance, these two groups of organisms could be used as indicators of RBC process kinetics. The environment in stage 3 varied more than that of any other stage. All groups were dominant in at least 4 of the 16 runs.

Dominance in the 4th stage biofilms was most frequently shown by the Sarcodinians. This group was the only one that steadily increased the number of times it was a dominant group going from stage 1 to stage 4. Free-swimming



RBC STAGE

Figure 1

Dominance of Microfauna in Laboratory RBC Cultures

Ciliates and Nematodes were also predominant in some of the 4th stage cultures.

After the dominant groups in a biofilm were determined, COD and NH_3-N concentrations in the bulk liquid were noted for that particular stage and run. Population of a group in the biofilm was plotted against concentration of either COD or NH_3-N (Figures 2 and 3).

Concentrations of various chemical species are descriptive of the environmental condition the biofilms are exposed to. The development of a relationship between concentration of a chemical species and microfauna population would provide increased understanding of the species succession found in RBC biofilms. Figures 2 and 3 are graphs of concentration versus population levels of Free-swimming Ciliates, attached Ciliates, Rotifers and Sarcodinians.

It can be seen from examination of Figure 2, that these are two basic differences in the types of areas enclosed for each of the four groups.

First, consider the range in COD values that the groups occurs in. Free swimming ciliates occur over the broadest range of COD concentrations, followed by attached ciliates and then Sarcodinians. Rotifers exist over the narrowest range. Secondly, it appears that free-swimming ciliates occurred in high COD concentrations than attached ciliates. Rotifers occurred in still a lower COD concentration. Sarcodinians occurred at the lowest concentration of COD while there is overlap of the areas in Figure 2, there are also areas unique to some groups.

Ammonia Nitrogen (NH_3-N) concentration and population levels are related for the four groups in Figure 3. Similar effects in the concentration ranges of NH_3-N and COD were observed. Both free-swimming ciliates and the attached ciliates occur in wide ranges of NH_3-N concentrations. Rotifers and Sarcodinians were restricted to more narrow conditions, especially at higher population levels.

A similar methodology to the one used in this study, relating COD and NH_3-N concentrations to group population levels, may also be used for individual species. While this has not been done by these authors, there are indications that the area enclosed for an individaul organism would be less than that for a group.

<u>Results From Full Scale Analyses</u>

Microscopic examinations of the Plainville, Connecticut, RBC treatment plant indicate that the number of species per stage decreased from 17 in the 1st stage to 12 in the 6th stage. Chemical and biological characteristics of the full scale operations are found in Table 8. The pH and alkalinity concentrations are noticeably less than those maintained in the laboratory study. 66% of the nitrification occurred in the 1st stage.

The succession of dominant groups from stage to stage is shown in Figure 4. The peak of attached ciliates followed by a large decline in dominance from stage 1 to 2 is very similar to that observed in the laboratory unit.

A peak in the dominance of flagellates, similar to that observed in the laboratory unit was more pronounced in the Full-scale process. The flagellate

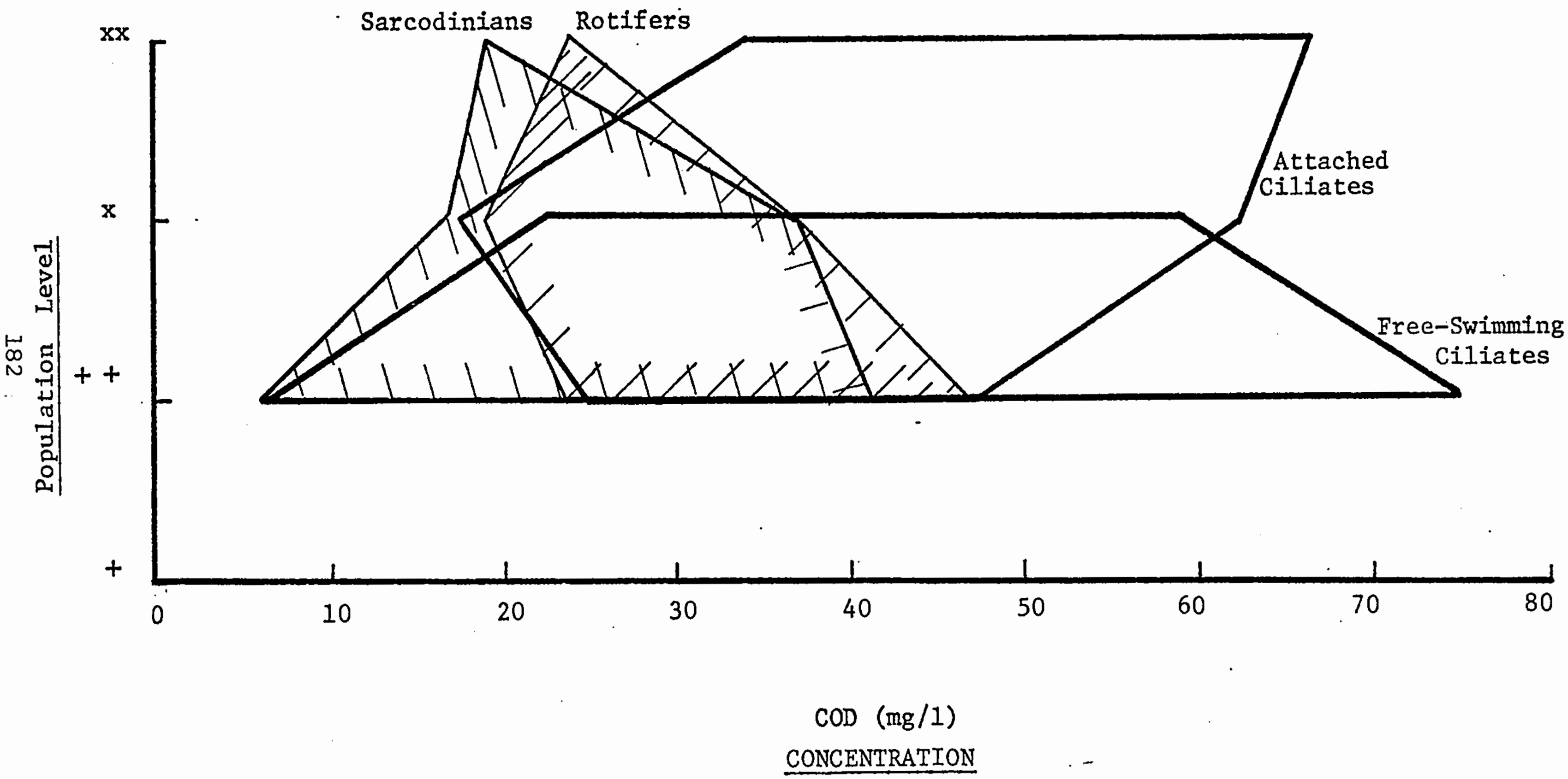


Figure 2

Effects of COD concentration on Microfaunal Populations

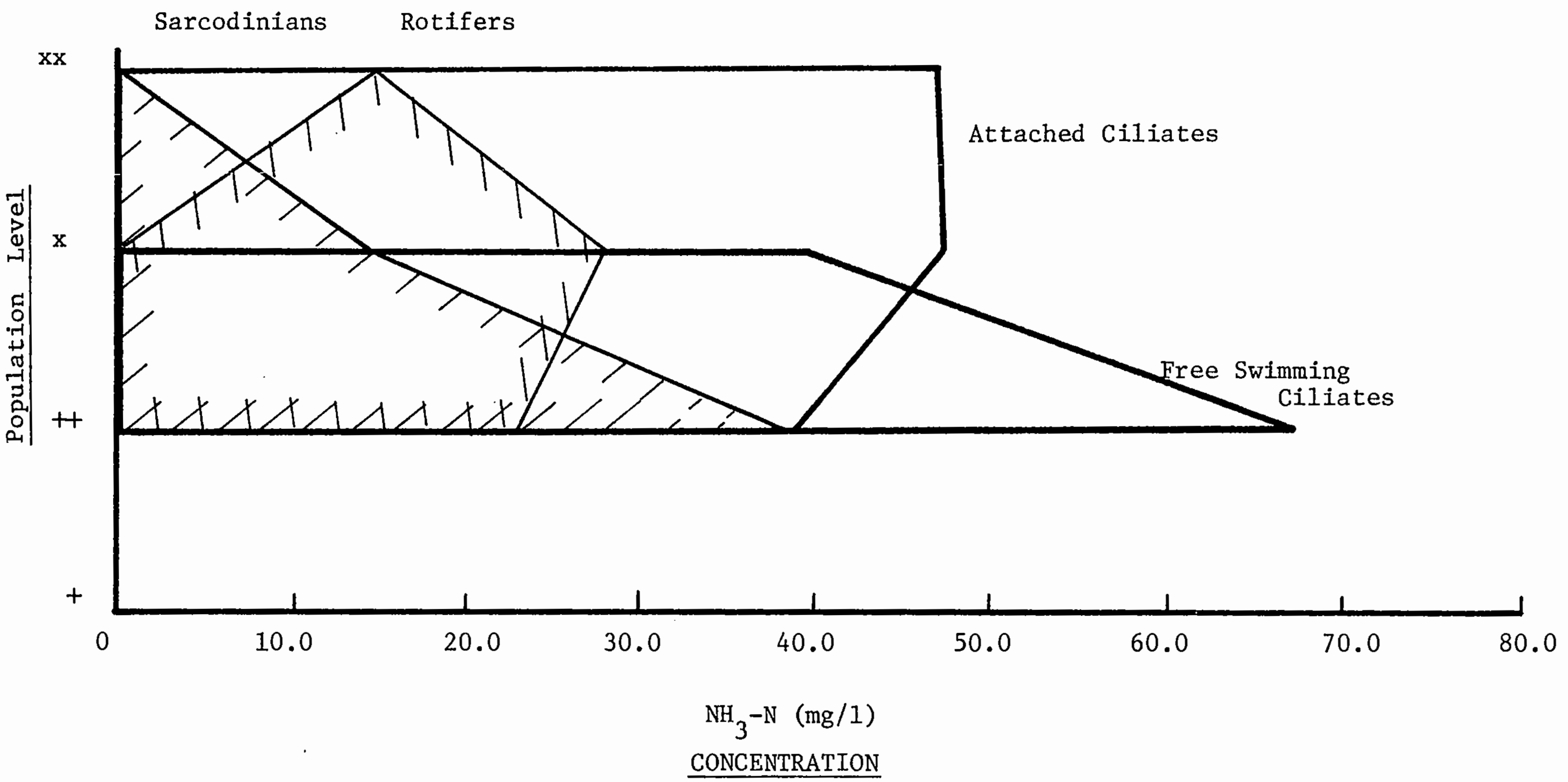


Figure 3

Effects of NH_3-N Concentration on Microfaunal Populations

Table 8

Chemical and Biological Characteristics
of the Full Scale RBC

STAGE	CONCENTRATION (mg/l)				NUMBER OF MICROFAUNAL SPECIES
	NH_3-N	NO-N	pH	Alkalinity*	
INFLUENT	36.7	0.86	7.27	232.0	
1st	9.4	24.4	6.83	112.0	17
2nd	3.86	28.8	6.63	70.5	17
3rd	2.26	30.4	6.56	62.0	13
4th	1.30	32.6	6.51	54.2	13
5th	1.67	32.0	6.57	57.0	12
6th	1.38	36.6	6.63	57.0	12

* Alkalinity expressed as
mg/l as $CACO_3$

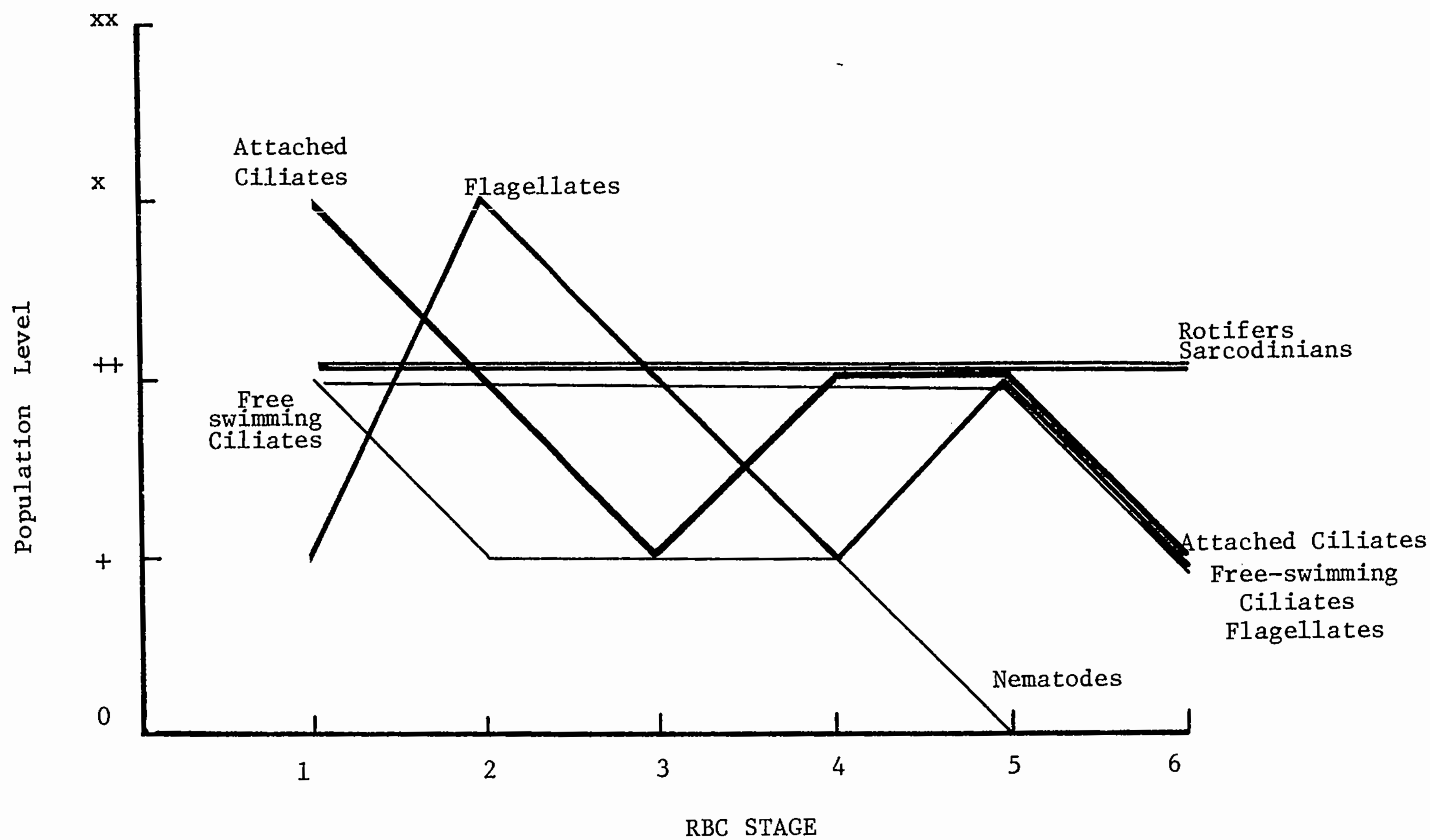


Figure 4

POPULATIONS OF MICROFAUNA IN FULL SCALE RBC species responsible for this peak was Monas obliqua and Mastigamoeba reptans. Dominance of Rotifers and Sarcodinians in the full scale biofilms was similar to that of the laboratory study. However, the slow increases of the Sarcodinians and the rapid increases and decreases of Rotifers was not observed in the laboratory.

Although there are similarities between the laboratory and Full-scale studies, no major comparisons should be made. The lack of pH and Alkalinity control creates a very different niche for organisms and their abundance is very likely an indication of this. The biofilm cultures nearly had identical zooglea matrices when examined with a microscope. The full scale biofilms were much thinner and in some instances did not uniformly cover the entire discs.

Conclusions

There is a succession of microfauna from stage to stage in both laboratory and full scale biofilms. In the first stage filamentous bacteria dominate biofilms and are responsible for the removal of BOD. Free-swimming ciliates were the primary predators of the filiamentors bacteria in the first stage of the laboratory unit. Two genera of filamentous bacteria were either singly or in combination identified in the first stages of all 16 laboratory runs. A Zooglea sp. (probably Zooglea ramigera) played an increasely important role in latter stage biofilms associated with nitrification.

Attached Ciliates were the most frequently dominant microfaunal group (excluding bacteria) in the 2nd stage-laboratory and 1st stage-full scale units. Rotifers and Sarcodinians were most frequently the dominant groups in the 3rd and 4th stage laboratory biofilms and in the 3rd through 6th stages of the full scale unit.

While there were similarities in the full-scale and laboratory scale results, nitrogen-mass loading rates, pH and alkalinity regimes were quite different. The decrease of total microfaunal species per stage (excluding bacteria) in the full-scale unit may have been the result of lower Nitrification reaction rates.

Microfauna was classified to the level of species whenever possible, but results in this paper indicate that this may not be necessary in order to assess the overall microbiological condition of the biofilm. Until there is an adequate data base, classification of microfauna to the level of species should be attempted whenever possible.

The average number of microfauna per stage (excluding bacteria) increase with increased degree of treatment in the laboratory study. Some individual runs did exhibit decreases in the 4th stage cultures.

Population levels of four microfaunal groups were related to concentration of COD and NH_3-N. There were also shifts of microfaunal groups related to changes in COD and NH_3-N concentrations. Ranges for many of the microfaunal groups were quite large. Future investigators should address the relationships between microfaunal groups or individual species and chemical concentrations in effluents from RBC reactors. Rotating Biological Contactor researchers should incorporate whenever possible, microbiological examinations of biofilm cultures, to expand a minute data base. Hopefully this will provide further insights into the biological activities of the organisms responsible for the efficient operations of the process.

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THE CHARACTERISTICS OF ROTATING BIOLOGICAL CONTACTOR (RBC) SLUDGE

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Introduction

The concept of using natural aerobic microbe in municipal sewage treatment has become an important design factor. Activated sludge process has been popularly adopted since it useless space, treats large quantity of sewage quickly and obtained better quality effluent. However, requirement of high degree management skill and high power consumption are disadvantage.

The RBC process which applies aerobic microbe treatment is become more important and popular. This process has a buffer character against variation of load. It generates less sludge and eliminates the bulking and foaming problems which usually cause great trouble in activated sludge process, and also has no odor and clogging problem which trickling filter process has. It also requires less power and operation as well.(1)

The article discloses data from a pilot experiment of sludge characteristics using RBC process as reference in design RBC sludge treatment facilities.

EXPERIMENT FACILITIES AND METHOD

Experiments were performed with 2 RBC pilot plants, one with 2 stages and the other 4 stages. The inlet water was municipal sewage which has been primarily treated and then introduced by pump to the RBC for treatment, or had been pretreated with coarse screen, grit removal and fine screen and then introduced to RBC for direct treatment. Pilot plants are described as follows:

(1) 2-stages Pilot Plant: (Unit: mm)

Overall Size: 1550L x 440W x 210H
Inlet: 50L x 440W x 75H
Reaction Tank: 950L x 440W
Gross Capacity: 68.9 liters. Net Capacity: 61.3 liters.
Liquid Volume/Disc Surface Area: 3.64 $1/m^2$
RBC Body: 405 Ø x 440W x 2 stages, acrylic board: 2 mm thick. (30 pieces in 1st stage at 13 mm interval, 23 pieces in 2nd stage at 15.3 mm interval, total surface area is 16.86 m^2.)
Settling Tank: 500L x 440W x 200H (Net Capacity: 0.044 m^3)
Settling Tank Outlet: 50L x 440W x 60H
RBC Rotation Speed: 2-20 rpm
Driving Motor: 0.25 KW

(2) 4-stages Pilot Plant: (Unit: mm)

Overall Size: 2000L x 720W x 560H
Reaction Tank: 1550L x 616W x 265H
RBC Body: 473 Ø x 300L x 4 stages (foamed plastic board, 9 pieces per stage, 36 pieces in total.)
Total RBC Surface Area: 23 m^2
Reaction Tank Capacity: 151 liters
Rotation Speed: 13 rpm
Driving Motor: 0.4 KW

BOD REMOVAL

RBC process is a multi-stage continuous treatment, therefore, the number of stage is an important factor for BOD removal. Fig 1 gives the experimental results of the rate of BOD removal by a 2-stages and a 4-stages RBC process respectively under the same load. It shows that the more the stages are, the better BOD removal efficiency is. Fig 2 indicates the process of the 1st stage can remove about 65% of BOD and the sequential stages have lower efficiency. The BOD removal curve demonstrates that while the number of stages is up to 3 or more, the removal efficiency is negligible. However, an appropriate number of stages will have buffer effect on the shock loading of flow and water quality change. Generally 2-4 stages may be recommended.

BOD removal speed has a close relation with the property of raw wastewater. Fig 3 shows its reaction. It is proceeded in two stages and its reaction is a first order reaction.

$$\ln \frac{C_t}{C_o} = K_1 t$$

Where C_o: Influent BOD (%)

C_e: Remaining BOD (%) after t time

K_1: BOD removal speed (hr^{-1})

t: reaction time (hrs)

The 1st stage reaction has a BOD reduction ranging from 100% to 35% and K_1 is 5.0 (hr^{-1}). During the 2nd stage reaction, the BOD reduction is from 35% to 10%, K_1' is 1.75 (hr^{-1}). These are much greater than those

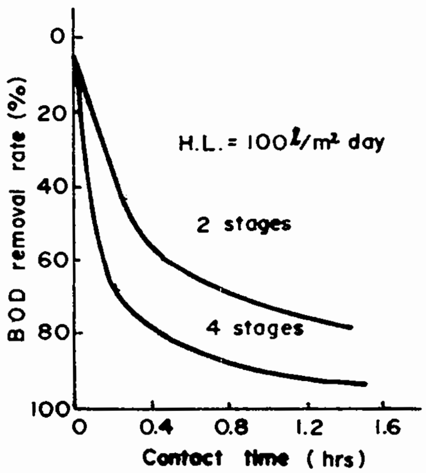

Fig. 1. No. of RBC stages and BOD removal rate

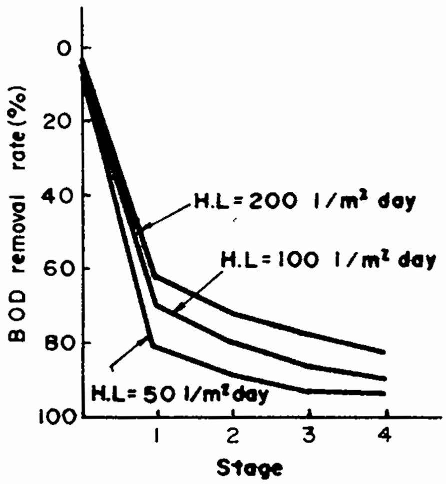

Fig. 2. BOD removal rate under each stage

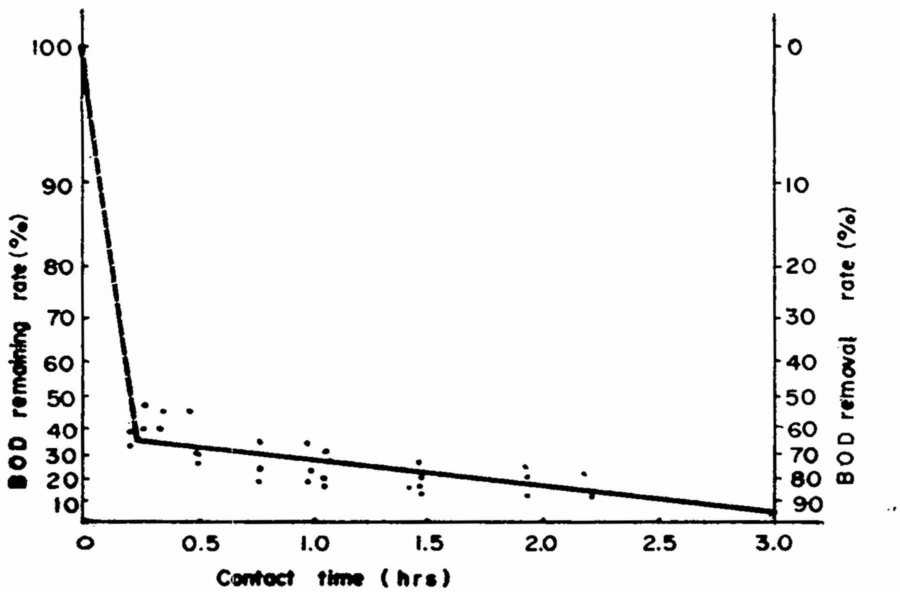

Fig. 3 BOD removal speed

in actived sludge process. The time of reaction treatment will be completed within 2 hours.

According to our long term experiment, the following two assumptions are made:

(1) Organic load and flow load of RBC are factors which affect BOD removal rate.

(2) Rotation speed of RBC and G value (Liquid Volume/Disc surface area) should be of the optimum data selected according to experience.

With reference to the data from continuous operation, relations of influent wastewater concentration and removal rate under different load are shown in Fig 4. The result can be written as:

$$y = 100 - (BL \times \alpha) \qquad (1)$$

where y = BOD removal rate (%)

α = the slope of BOD loading and BOD removal rate

Then, common slope of all influent wastewater concentrations are analyzed:

$$\alpha = \frac{126.1}{Cin + 10.07} \qquad (2)$$

where Cin = Influent wastewater BOD concentration (mg/l)

Institute (2) into (1), then,

$$B.L. = \frac{Cin + 10.07}{126.1} (100-y) \qquad (3)$$

$$H.L. = \frac{Cin + 10.07}{126.1 \quad Cin} (100-y) \times 1{,}000 \qquad (4)$$

Remaining BOD

$$X (\%) = \frac{126.1 \; B.L.}{Cin + 10.07} \qquad (5)$$

$$X (\%) = \frac{126.1 \; Cin \; H.L.}{(Cin \quad + 10.07) \times 1{,}000} \qquad (6)$$

where B.L. = BOD loading (g BOD/m^2 day)

H.L. = Flow rate (l/m^2 day)

CHARACTERISTICS OF BIO-FILM

RBC treatment exposes the disc to air under an appropriate rotation speed and provides oxygen for reproduction requirement for various microbes, then, bio-film is grown. Microbes, by means of Exo-enzymes, absorb basic elements in sewage and perform metabolism. Part of it becomes thermal energy and the other becomes the film.

Under normal operation, compositions of bio-film in different stages are as follows: (SS composition in Table 1 and VS composition in Table 2.)

Table 1: Composition of Bio-film

Stage	Qty. (g/m^2)	Conc. (%)	Chemical Composition (%)					Thermal Value (Cal/g.vss)	VS/TS (%)
			C	H	N	O	Ash		
1	32.6	4.3	35.2	5.4	5.8	21.4	32.2	5,790	67.8
2	28.7	4.4	39.1	6.7	6.9	23.4	23.7	4,230	76.3
3	28.8	5.8	37.1	5.7	6.8	25.3	24.5	4,000	75.5
4	22.4	5.9	37.2	6.0	6.8	25.5	24.5	3,010	75.5

Table 2: Chemical Composition of VS of Bio-film

Stage	Organic Composition (%)				Thermal Value (Cal/g.vss)
	C	H	N	O	
1	51.9	8.0	8.6	31.5	8,540
2	51.2	8.8	9.0	31.0	5,540
3	49.1	7.5	9.0	33.4	5,300
4	49.3	7.9	9.0	33.8	3,990

The molecular formula of this composition is $C_{4.2}H_{8.1}N_{0.6}O_{2.1}$.

Content of volatile substance is 67.8% in the 1st stage and 77-78% in the 2nd and sequential stages.(2) Ash content in the 1st stage is higher than other stages. It might be due to the absorption of inorganic matter in the influent wastewater. Sludge concentration is 4.3% in the 1st or 2nd stage and 5.8% in the 3rd or 4th stage. The latter film is harder due to the bio-film formed from the growth of nitrobacteria. No significant difference is occurred in the CHNO ratio among these stages.

Regarding the appearance of bio-film, during the 1st stage, because of the high load, the sludge is in light dark color and the micro-organism appears are merely Sphaerotilus, Zooglea and other filamentous bacteria. During the 2nd stage, the film is in tan or brown color with slurry; the protozoa therein is Rotaria, Diplogaster, Zoothonmium etc. During the 3rd stage, the film is harder then previous one and it is in tan color with Rotaria etc. Microbes therein are less than that in the 2nd stage. The film in the 4th stage is similar to that in the 3rd stage. There are only a few microbes, such as Podophrya etc. Since microbes appear at the RBC film is much more than that in activated sludge process, the ecosystem is more stable, and is more resistant to the variation of influent wastewater

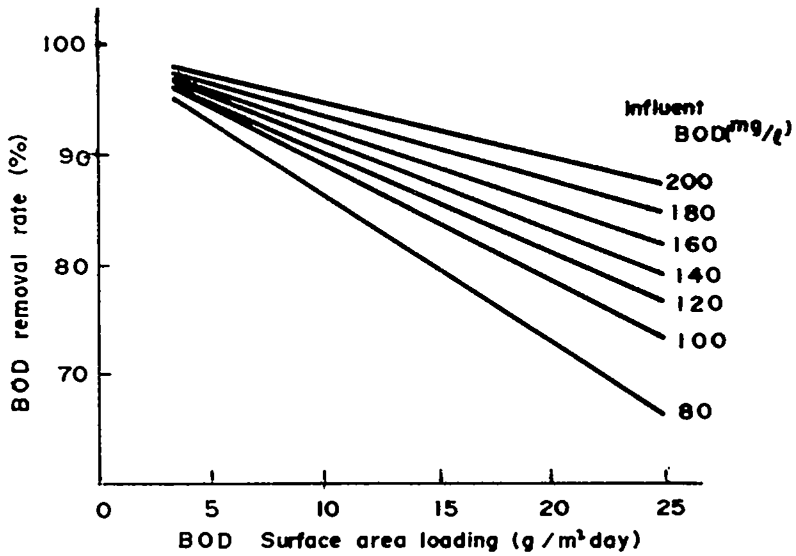

Fig. 4. relation of BOD removal rate with influent BOD and surface area loading

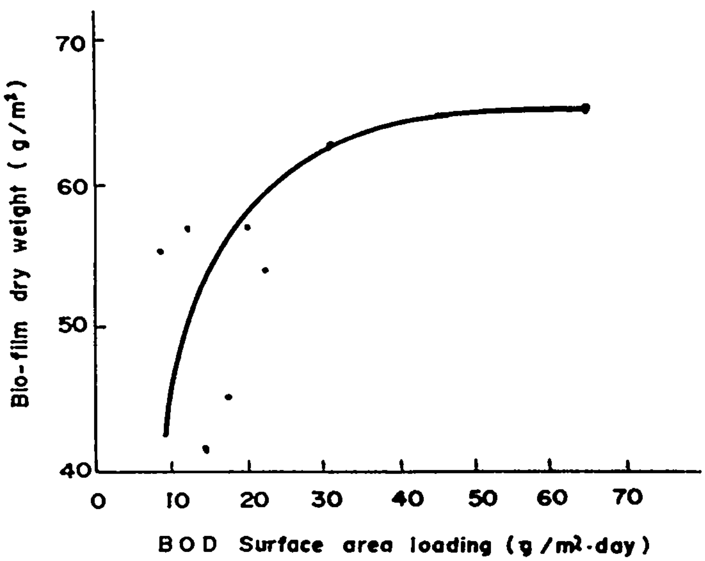

Fig. 5. relation of BOD surface area loading and bio-film dry weight (4 stages)

quality. This is one of the reason why this method is easier to operate.

In the experiment of pilot plant, the thickness and the dry weight of the bio-film due to the BOD surface loading (see Fig 5,) vary at each stage, due to the rotating speed. Fig 6 shows the relationship between the thickness of the bio-film and rotating speed.

RBC is relying on the bio-film to treat wastewater. The MLSS in the activated sludge aeration tank is generally considered to be equivalent to the sum of the microorganism and SS in the RBC divided by the net capacity of the contact tank. This is defined as equivalent SS (ESS), which is a function of removing the organic matter in the wastewater. Thus it can be written as[4]

$$F/M = B.L./DS$$

Where B.L.: BOD surface area loading ($g.BOD/m^2.day$)

DS: Dry sludge weight ($g.SS/m^2$)

F : BOD (mg/l)

M : MLSS (mg/l)

In 4-stages pilot plant experiment, the first stage ESS is 10,000 - 13,500 mg/l, and decreases stage by stage. The 4th stage ESS is about 4,000 - 8,500 mg/l. The average ESS is 7,000 - 11,500 mg/l for the whole process. The average ESS for the two stage pilot plant under different rotating speed is shown in table 3. It is found that the average ESS for the 2-stages experiment is approximately the same as that for 4 stages experiment, specially under the same speed where the average ESS is 9,000 mg/l for 2-stages experiment.

Table 3: Average SS and ESS of 2-stages experiment

Rotating speed (RPM)	Peripheral velocity (m/min)	Average SS in tank (mg/l)		Average bio-film dry weight (g/m^2)		ESS (mg/l)
		Stage 1	Stage 2	Stage 1	Stage 2	
4	6.5	466	263	66.3	30.9	13,940
7	9.9	370	100	60.7	23.9	12,230
10	14.1	327	78	54.8	22.6	11,170
15	21.2	294	76	28.4	21.6	6,960

ESS and BOD relation under the aerobic conditions vary, depending on the loading. From the experiment, no definite relation between F/M loading and removed rate at each stage can be evaluated. The BOD removal in the experiment can be up to 50% at the 1st stage and it differs substantial at each stage. This may be due to the reason that very high adhesion of the SS on the disc at the 1st stage experiment result on the reduction of non-organic SS in the waste. Fig 7 show in the relationship between average F/M and BOD removal rate.

Table 4 described the relation between the F/M loading and BOD removal for the 4-stages experiment. Because the aeration time for the activated sludge process is about 4 times of the RBC process, the F/M ratio of 0.2 -

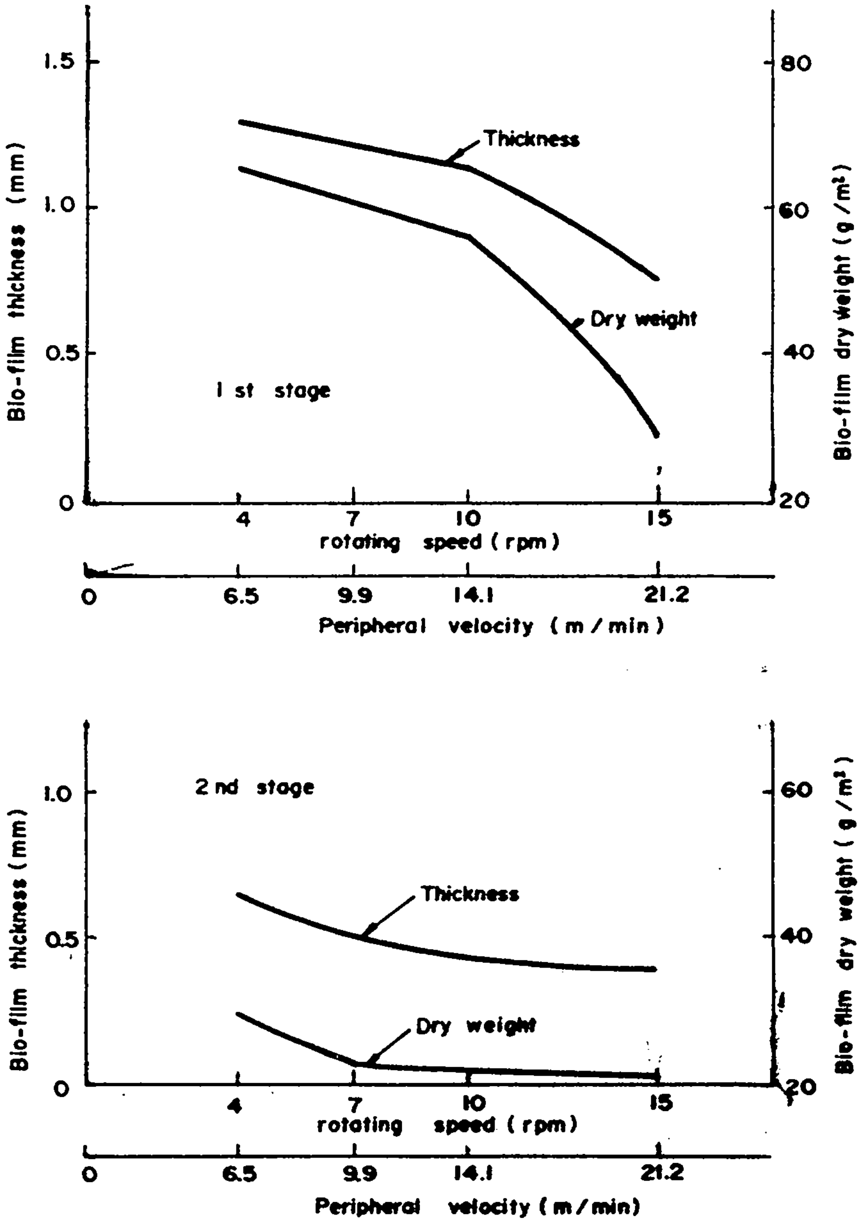

Fig. 6. relation of peripheral velocity with Bio-film thickness and dry weight

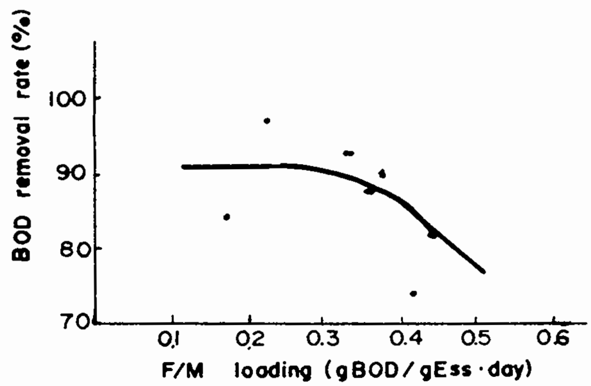

Fig. 7 average F/M loading and BOD remoal rate

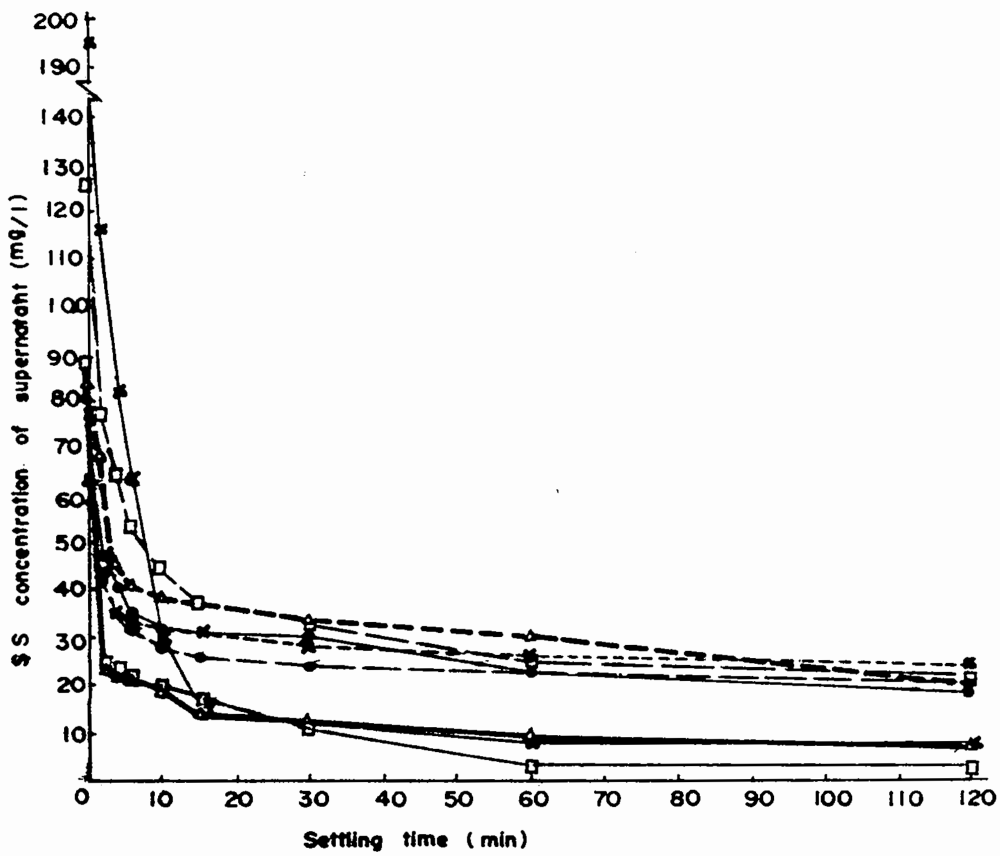

Fig. 8 Settling speed of the RBC effluent water

0.4 g.BOD/g.ESS.day for the RBC process can be considered as the same as that for the activated sludge process.

Table 4: Aerobic digestion of RBC sludge

days	T.S. (mg/l)	V.S. (mg/l)	V.S. reduce rate (%)	BOD (mg/l)	SV_{30} (%)	Supernatant BOD (mg/l)
raw sludge	12,630	9,230		11,400	98.0	
5	11,000	7,500	18.6	6,000	95.0	255
10	6,300	4,010	56.4	2,900	73.0	140
15	5,800	3,330	63.8	1,200	40.0	110

SETTLING AND THICKENING CHARACTERISTICS OF SLUDGE

Suspended solid in RBC reaction tank is merely down-scaled bio-film and a few SS of influent wastewater. Concentration of SS in the tank varies upon the concentration of influent wastewater and rotating speed. It increases while the load is high and thickness of bio-film is increasing. When the load is lowering, the sloughing of bio-film will be also increased.

Removed sludge radiates outwards with the Zoogloea sludge as a center a great amount of sponge-like bacteria filament. Under microscope they are branch-like suspension with lower specific gravity but larger diameter. The diameter found is about 0.1 mm - 1.2 mm.

Generally, SS in the final reaction tank is less than 200 mg/l, the solid loading in final settling tank is about one-tenth of that in activated sludge process. In the measurement at the overflow weir of reaction tank, SS is about 100-150 mg/l, the SV_{30} is about 1-1.5%, hence, SVI is about 100. The settling character is excellent.

Settling speed varies upon the difference of raw wastewater concentration However, large SS sinks quickly and freely within the first 10 minutes. The curve in Fig 8 shows the initial sinking speed. If the outlet port has an average SS of 150 mg/l and SV_{30} is 1.5%, then, after a settling for 30 minutes, the sludge concentration is about 10,000 mg/l.

The lowest settling speed of slough off bio-film as mentioned is about 6-15 cm/min, i.e. 80-216 m/day. If the surface loading of a clarifier is under 30 m^3/m^2.day, it is enough to settle the flow and separate it from water.

Sludge from final sedimentation tank is withdrawn to a 1,000 ml cylinder. The interface is measured at definite time intervals. The interface change with various sludge concentrations are shown in Fig 9. Thickening speed lowers when concentration increases. After keeping still for 6 hours, the thickening nearly terminated.

Since SS is high, the interface of the liquid is obvious. In the beginning, the zone falls with a constant speed. However, the lower concentrated layer is thickened and the sinking speed at the middle is lowered. Thereafter, by the accumulated weight, the inter water being

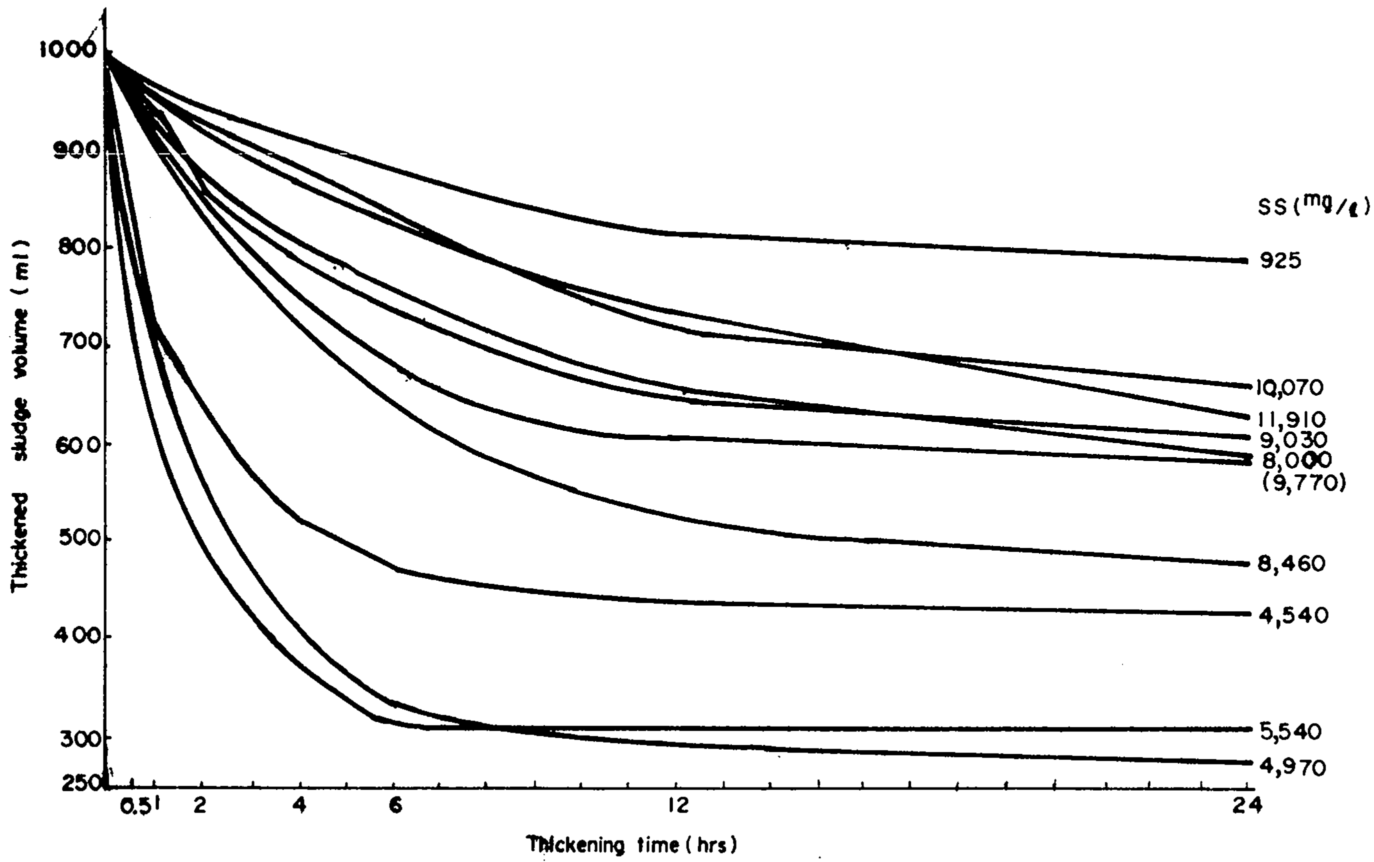

Fig 9. Volume occupied by the thickend sludge with thickening time

pressed out. Settling speed becomes very slow (compression zone). The thickening character is identical to that of activated sludge process and in a layer deposit.

PRODUCTION OF SLUDGE

Treatment of sludge is the most difficult problem in future sewage treatment. We have researched the sludge production by directly treat the sewage which is merely through grit removal and fine screen and the sludge production from presettled sewage. The actual sludge production from direct treatment and pretreatment is respectively shown in Fig 10. In direct treatment has a wider range of 0.37 - 1.15 Kg.SS/Kg.BOD.removal.day, average is 0.67 KgSS/KgBOD removed.day. Thus indicates that the conversion rate between BOD removal and sludge is very low. For secondary treatment, depending on the flow rate, the sludge production rate is 0.34 - 0.55 Kg.SS /Kg.BOD removal-day is about one third of that from activated sludge process, which is considered as the major advantage of this process.

SLUDGE DIGESTION AND DEWATERING

The sludge generated from the direct treatment by the RBC process, has a water content of about 98.8% with the average quality of SS about 9,200 mg/l. The VSS has been 18% for the duration of 5 days and 56% for 10 days respectively, if the aerobic digestion is applier under the room temperature of 15 - $20^{\circ}C$. This result indicates that the digestion due to the RBC process has been more efficient than that of the primary plus activated sludges.[3] Fig 11 shows the relationship between the VSS reduction rate and the aerobic digestion duration. The digestion characteristics due to the RBC sludge with the aerobic digestion are also given in Table 4.

Further, the raw sludge obtained from the RBC process went through the thickening process. It was found that the sludge has a water content of 97.5%. The sludge then was proceeded by anaerobic digestion with a temperature of about $36^{\circ}C$ and under the room temperature of 18 - $20^{\circ}C$. Table 5 shows the characteristics of anaerobic digestion. After 30 day the VS reduction rate is approximately 38% and the sludge alkality is about 1,500-2,400 mg/l.

Table 5: Anaerobic digestion of RBC sludge

Sludge		S.S. (mg/l)	BOD (mg/l)	T.S. (mg/l)	V.S. (mg/l)	Alk. (mg/l)	Water cont.(%)
Raw sludge		23,720	25,000	25,300	14,500		97.40
$36^{\circ}C$ Digestion sludge	15days	18,460	9,200	17,750	11,200	1,580	98.20
	30days	16,600	4,800	18,500	8,900	2,430	98.26
$19^{\circ}C$ Digestion sludge	15days	19,270	11,600	20,610	13,150	1,195	97.93
	30days	18,230	5,400	17,460	9,290	1,440	98.35

The raw sludge produceed by the RBC process,then, is dewatered after the anaerobic digestion for the duration of 15 and 30 days respectively. The dewatering characteristics of the obtained sludge cake can

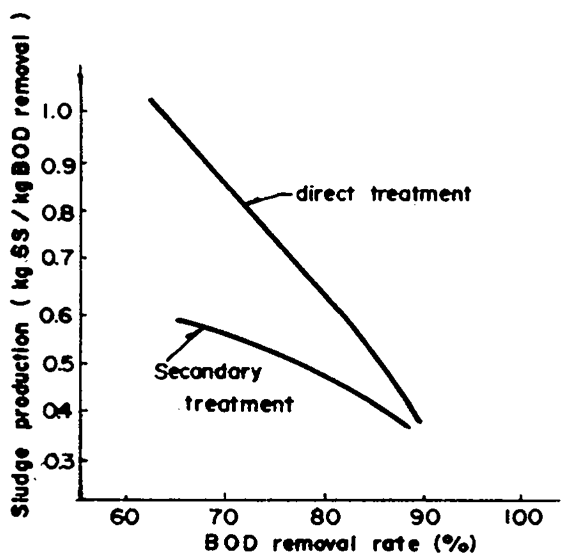

Fig. 10. BOD removal rate and sludge production

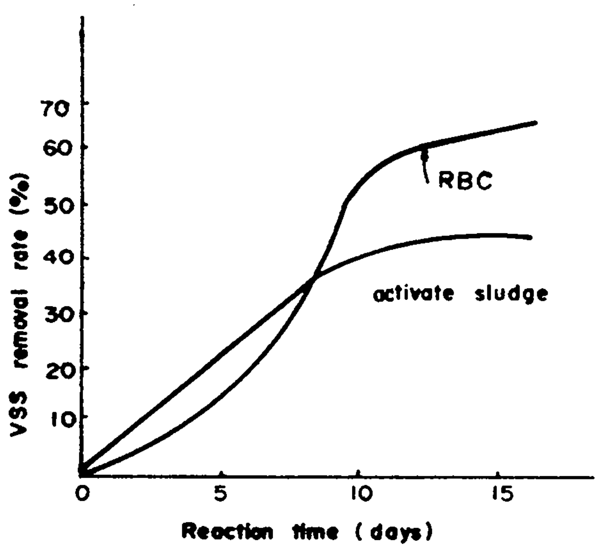

Fig. 11 relationship of VSS removal rate and aerobic digestion reacton time be compared with other of the sludge as shown in Table 6.[2] It appears that the specific resistance of the dewatered sludge is lower than those of the primary plus activated sludge as well as those of the activated sludge. The water content of the obtained sludge cake is also lower.

Table 6: The characteristics of RBC sludge dewatering and compared with others

Sludge		Feeding $FeCl_3$(%)	Water cont. of Sludge cake	Specific resistance $R \times 10^7 (sec^2/g)$
RBC direct treatment sludge	raw sludge	5 10	84.9 78.8	18.4 12.8
	15 days digestion	5 10	83.1 68.3	19.6 12.6
	30 days digestion	5 10	80.5 65.0	25.8 10.8
Primary plus activated sludge		5 10	79.8 83.1	27.0 27.0
activated sludge		5 10	82.3 86.1	16.8 15.2

CONCLUSIONS

1. The speed of BOD removal in the RBC treatment is a 1st order reaction. The process can be proceeded into two stages. The BOD remaining rate ranged from 100% to 35%, $K_1 = 5.0$ (hr^{-1}) and from 35% to 10% $K_1' = 1.75$ (hr^{-1}) in the 1st and 2nd stage respectively.

2. The thickness of the Bio-film was grown depending on the sewage loading and the rotating speed. Under the two stages process and the peripheral velocity under 14.1 m/min, the dry weight of bio-film are 54.79 g/m^2 and 22.56 g/m^2 for the 1st and 2nd stages respectively. When the peripheral velocity is high than 21.2 m/min, the dry weight of the bio-film is only 28.41 g/m^2 for the first stage and 21.56 g/m^2 for the 2nd stage.

3. The ESS in the contact tank varies depending on the load. Under the 4-stages process of the RBC treatment, the ESS is 6.970 - 11,460 mg/l with the average of 9,000 mg/l, if the BOD loading is under 10 - 641 $g.BOD/m^2$. day. For the 2-stages process, the ESS has an average of 11,000 mg/l and has a concentration about 3-4 times of that of MLSS from the convential activated sludge process. Hence, the contact time by the RBC process in only below ¼ of that of the activated sludge process. This may me the reason where the RBC is more efficient.

4. Generally, the larger the F/M loading is, the lower the BOD removal. The F/M loading of the RBC process is about 0.2 - 0.4 g.BOD/g.ESS.day, which is approximately equal to that of the conventional activated-sludge process.

5. The bio-film from the RBC process has an average VSS of about 74% and a water content of 95%. Its chemical composition is $C_{4.2}H_8 N_{0.6}O_2$. The

thermal value is about 8,500 cal/g.SS in the 1st stage, which is comparetively higher than that in the 2nd-4th stage (about 5,200 cal/g. SS).

6. The sludge has settled freely and relatively fast. After 10 minutes, the settling almost terminated, because, only limit amount of the settled sludge was found. The early settling of the sludge has a velocity between 6 - 15 cm/min, which is faster than that due to the activated sludge process. A surface loading of 30 m^3/m^2.day for the clarifier design is recommended of a rate of less than 20 mg/l for the effluent SS is required.

7. The sludge production under the RBC secondary treatment process varies depending on the influent loading and its BOD. The average sludge production rate is approximately equal to 0.54 Kg.SS/Kg.BOD.removal. The sludge production due to the direct treatment process has a wider range of 0.37 - 1.15 Kg.SS/Kg.BOD.removal (average is 0.67 Kg.SS/Kg.BOD. removal).

8. The VSS reduction rate can reach up to 56% for a duration of 10 days if the RBC direct treatment sludge using aerobic digestion applied. Consequently, it seems to have better digestion than conventional activated sludge.

9. After the thickening of the sludge from the RBC direct treatment, the raw sludge and anaerobic digested sludge is dewatered by feeding $FeCl_3$. Its specific resistance is less than those of primary plus activated sludge, and activated sludge.

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PART III: MUNICIPAL WASTEWATER TREATMENT

DATA EVALUATION OF A MUNICIPAL RBC INSTALLATION, KIRKSVILLE, MISSOURI

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INTRODUCTION

The Rotating Biological Contactor System, RBC, is becoming more and more competitive with the conventional fixed film and suspended growth systems because of its simplicity, ease of operation, and low operation and maintenance costs. Design of RBC systems has been based primarily on surface hydraulic loading, HLA, as gal/d/sq ft of media surface, and only recently have the manufacturers suggested BOD influent concentration as having some effect on system performance. Independant researchers have considered both organic and hydraulic loading in analyzing significant parameters and results have been somewhat contradictory. Some researchers have suggested system performance was strictly a function of HLA and some investigators have said surface organic loading, OLA, as lb BOD/d/1000 sq ft of media surface was the controlling parameter. These initial analyses were conducted using pilot plant results and suffered from a severe lack of long-term operational data from full scale RBC installations. The research conducted at the University of Kansas and summarized in this paper was aimed at clarifing some of the inconsistencies reported in the literature. Full scale operational data from a 5 MGD RBC plant at Kirksville, Missouri was analyzed to determine the parameters that had a significant effect on system performance, to check field results against design theory, and to make recommendations on parameters that should be used for future design and analysis of the RBC system.

EFINITION OF TERMS

Many different terms have been used in the analysis of RBC process fficiency to describe basically the same parameters. Shown in Figure 1 urface hydraulic loading, HLA, as gal/d/sq ft is determined simply from he flowrate in MGD divided by total media surface area in million sq ft. he HLA was found to have common units by all authors reporting hydraulic oading data. The second parameter of interest is organic surface loading, LA, as lb BOD or COD/d/1000 sq ft media surface area. As seen in Figure 1, rganic concentration and flowrate or hydraulic loading have an effect on LA. The significant characteristic of OLA is that it not only reflects he driving force for the diffusion of substrate into the biofilm through he influent BOD concentration term but also the effect of reaction time ı the system as indicated by the hydraulic loading term.

FIGURE 1 VARIABLES USED IN DATA ANALYSIS

/DRAULIC SURFACE LOADING--- HLA

$$HLA = \text{GAL/DAY/SQ FT OF MEDIA SURFACE AREA.}$$

$$= \frac{\text{Q in MGD}}{\text{SA in million sq ft}}$$

ˎGANIC SURFACE LOADING--- OLA

$$OLA = \text{LB BOD or COD/DAY/1000 SQ FT OF MEDIA SURFACE AREA}$$

$$= \frac{(\text{Q in MGD})*(\text{BOD influent concentration in mg/1})*8.34}{\text{SA in 1000 sq ft}}$$

Correspondingly:

$$OLA = \frac{(\text{BOD influent concentration in mg/1})*(\text{HLA})*8.34}{1000}$$

Many investigators have presented OLA as lb BOD/d/1000 cu ft media ·lume. The RBC system is a fixed film process and as such, microbial tivity is concentrated at the media surfaces. It seems more theoretilly sound to be concerned with loading on a surface basis rather than on volumetric basis and all data in this analysis is presented in this manner. tal organic loading, OL, and BOD removal both as lb BOD/d, were used for oss comparisions between specific periods within the overall study period. rcent BOD and SS removal values were also used in the analysis for comrison of Kirksville data with data in the literature.

EVIOUS STUDIES

Initial use of rotating media for treatment of wastewater was by Doman) in 1925. His experimental "contact filter" consisted of concentrically unted 20 gage steel discs, 1 inch on centers, that was rotated ½ rpm in e effluent from a septic tank. He used the discs to accumulate "active robic material" for removal of colloids from the wastewater. The biological ture of the process was not appreciated by Doman and the system was doomed failure. Anaerobic conditions plagued the system's operation and poor results caused abandonment of the concept for almost 30 years. The next use of rotating media for wastewater treatment came in the late 1950's with the investigations of Hartman and Pöpel (2) at the University of Stuttgart. One meter diameter plastic discs were used as support media and tests were run to gather information for process development of what they called "Immersion Drip Filter." These investigations, along with the development of expanded polystrene as an inexpensive construction material, led to commercial use of the "Immersion Drip Filter" in Europe as early as 1960.

Allis-Chalmers developed the RBC in the United States independently from European efforts in the late 1960's and upon learning of European activities, came to a licensing agreement with J. Conrad Steneglin Company of Tuttlinger, West Germany for sales and manufacturing rights for the RBC process in the United States under the trade name BIO-DISC (2). The process was subsequently bought by Autotrol Corporation in 1970 and has been marketed since then under the trade name BIO-SURF.

Most early process performance analyses were carried out by manufacturers and it has been only since about 1970 that results of independent investigation have been readily available. Table 1 is a summary of important RBC performance analysis results and design recommendations found in the literature. Hartman's study in 1965 (3) was the first presentation of modern RBC design criteria. He recognized the effect of influent BOD concentration on BOD removal through "decomposition curves" he presented. These "decomposition curves" represented percent BOD removal versus 1/HLA for varying wastewater strengths ranging from less than 100/mg/1 to over 600 mg/1 influent BOD. He also recommended a maximum OLA of 20.5 lb BOD/d/1000 sq ft to the first stage of the RBC for efficient process performance.

Welch and Antonie presented manufacturer's pilot plant results in their studies of synthetic waste from 1968 to 1970. Welch (4) investigated both hydraulic and organic loading variables in his early studies but design and operational analyses, based on organic loading, were abandoned by manufacturers in the later studies in lieu of HLA or detention time in the reactor. All of the studies showed an initial rapid increase in percent removal with an increase in detention time, leveling off to a constant percent removal between 30 and 90 minutes depending on the waste type used in the study. Welch found a constant slope for BOD removal versus OLA, both as lb/d/1000 cu ft, to an OLA of 650 lb/d/1000 cu ft, beyond which a constant BOD removal as lb/d/1000 cu ft was achieved.

After 1971 studies began to look into the relationship between loading and removal on a pound basis. Pescod and Naïr (5), Cochran, Burn, and Dostal (6), and Stover and Kincannon (7) all showed constant rates of removal of organics on a pound basis for loadings up to 250 lb COD/d/1000 cu ft for bottling waste, 10.6 lb BOD/d/1000 sq ft for cannery waste, and 3.48 lb COD/d/1000 sq ft for synthetic sucrose waste, respectively.

The study by Malhorta, Williams, and Morley (13) in 1975 is the only presentation of full scale, long-term RBC plant data found in the literature. A large amount of average monthly data was provided, but no attempt was made to quantify performance in terms of graphs or correlations between significant parameters. In analyzing the data it was observed that once the plant established its microbial population, organic removal was more a function of

TABLE 1 REVIEW OF RBC EFFICIENCY ANALYSIS STUDIES

RELATIONSHIPS	WASTE	RESULTS	RECOMMENDATIONS	AUTHOR(S)
%BOD removal vs 1/HLA	variable	decomposition curves to yield SA requirements for different WW with different influent BOD's	use of decomposition curves for design, max OLA to first stage of 20.5#/d/KSF	Hartman (1965)
BOD removal vs OLA #/d/KCF	synthetic	constant slope to 650 lb/d/KCF then pounds removed/d/KCF is constant		Welch (1968)
%COD removal vs t_D	synthetic	decreasing COD influent yield higher constant rate of removal. all level off at 30 min		Welch (1968)
%COD removal vs t_D	synthetic 500 mg/l	logrithmically approach 90% removal at 60 min. t_D	t_D 60 min for 90+% removal	Antonie (1970)
%BOD removal vs HLA	primary eff 170 mg/l	approach 90% removal at approx. 90 min	t_D=50 min for 20 mg/l SS and BOD eff, 80% NH_3-N removal	Antonie (1970)
HLA vs %BOD removal	anaerobic packinghse waste	decreasing BOD inf 206-145 mg/l through the study. 65% removal HLA=8gal/d/SF, 83% removal HLA= 4 gal/d/SF. 95% of total removal by first stage	single stage unit at an HLA of 4gal/d/KSF	Chittenden Wells (1971)
%COD removal vs OLA #/d/KCF	bottling 1000 mg/l municiple 400 mg/l	95% removal@250 #/d/KSF 85 % removal @300#/d/KSF	OLA=250 #/d/KSF for bottling waste	Pescod & Nair (1971)
%BOD removal vs HLA	winery 1003 mg/l	95% removal@HLA=0.75 gal/d/SF (OLA=6.31 lb/d/KSF)	HLA=0.75 gal/d/SF design	Labella et al (1972)
BOD removal vs OLA #/d/KSF	cannery varying Q, BOD, SS	constant rate of removal for OLA=3-10.6 #/d/KSF		Cathran, Burn Dostal (1973)
BOD effluent vs OLA #/d/KSF	primary eff varying BOD, Q, SS	increased BOD effluent for increased OLA	OLA=1 #/d/KSF for BOD eff quality of 15 mg/l	Ahlberg & Kwong (1974)
%COD removal vs COD inf	synthetic sucrose= 236 mg/l	constant HLA=0.5 gal/d/SF, constant rate of COD removal= 90%@ 1.9x and 3.6x initial COD inf. peak OLA of 3.48 #/d/KSF		Stover & Kincannon (1975)

Full scale plant data	municiple 1 MGD	OLA=1.11-2.41 #BOD /d/KSF
		HLA=0.97-2.14 gal/d/SF
		removal a function of temperature
		T= 45-68°F 86-93% removal T <50°F
		94-97% removal T >50°F
				Malhorta, Williams, Marley (1975)

%BOD removal vs no stages	synthetic carbohydrate	95% removal through the unit with same OLA from 3 HLA and BOD inf combinations BOD inf / HLA / OLA 550 / 0.75 / 3.13 850 / 0.50 / 3.54 1870 / 0.25 / 3.48		Stover & Kincannon (1976)
COD removal vs OLA #/d/KSF	slaughterhse	70% COD removal for OLA=3-7 #/d/KSF 4.7 #/d/KSF removed at OLA 7#/d/KSF		Stover & Kincannon (1976)
%COD removal vs OLA	carbohydrate	92% COD removal for OLA=6.7-26.7#/d/KSF logrithmic decrease of % removal to 40% at 47 #/d/KSF		Stover & Kincannon (1976)

temperature than organic or hydraulic loading. Resultant removal efficiencies were 86 to 93 percent BOD removal for temperatures below 50°F and 94 to 97 percent BOD removal for temperatures above 50°F. The overall temperature range was 45 to 68°F while OLA varied from 1.11 to 2.41 lb BOD/d/1000 sq ft and HLA varied from 0.97 to 2.14 gal/d/sq ft.

Stover and Kincannon's 1976 study (14) of carbohydrate and slaughterhouse waste treatment with RBC's is significant in that it showed definitively that percent COD removal is not strictly a function of HLA or influent COD concentration, but is dependent on the combination of the two factors, namely the total organic loading to the system. Not only was the driving force for diffusion into the microbial film, as given by COD concentration, an important variable affecting the performance of the RBC unit, but also the ability for the biomass to metabolize the waste in a given detention time, as reflected by HLA, was important. Influent COD concentrations ranged from 550 to 1670 mg/l and HLA values ranged from 0.75 to 0.25 gal/d/ sq ft to produce OLA values ranging from 3.4 to 3.5 lb COD/1000 sq ft. Resultant percent COD removals varied only between 94 and 96 percent and indicated a definite relation between OLA and COD removal efficiency as shown in Figure 2. Decreased COD removal efficiency resulting from organic overloading was also shown in their results. In Figure 3 their plot of COD removal versus COD applied as lb/d/1000 sq ft shows a constant rate of removal of COD up to an OLA of 7 lb/d/1000 sq ft, beyond this a constant mass of COD of 4.7 lb/d/1000 sq ft was removed, causing a decrease in removal efficiency expressed as a percent as shown in Figure 4. The effect of varying wastewater characteristics is also evident from Figure 4 from the relative removal efficiencies for the carbohydrate and the slaughterhouse wastes used in the study. The soluble, readily degradable synthetic carbohydrate waste showed 90 percent COD removal up to 26.7 lb/d/1000 sq ft before the constant COD removal situation developed, resulting in decreased percent COD removal. The slaughterhouse waste, on the other hand, showed a maximum 70 percent removal with decreasing efficiency beginning at only 7 lb/d/1000 sq ft because of the large amount of fats, oils, and slowly degradable organic materials in this wastewater.

Several other studies were important in developing a conceptual background for the understanding of the microbiological relationships within the RBC biofilm. The discussion of substrate utilization in relation to oxygen and substrate diffusion in the RBC biofilm by Hoehn and Ray (15) in 1973 was based on work first presented by Sanders (16) in 1964 and by Kornegay and Andrews (17) in 1968 and with RBC pilot plant results described above serve as the basis for Figure 5. Three theoretical limitation regions and the limiting parameter in each are identified for BOD removal as percent and lb/d/1000 sq ft versus influent BOD or OLA. Initially substrate is limiting because of diffusion limitation into the biofilm. Once a minimum load or concentration is reached the driving force for diffusion into the biofilm is such that metabolism and corresponding percent substrate removal becomes independent of substrate concentration and is limited by the mass of substrate reaching the organisms on the RBC discs. More substrate would be removed as more is applied as shown in Figure 5b. In this region the mass of organisms on the discs increases proportionally to the substrate applied as neither oxygen nor substrate are diffusion limited within the biofilm.

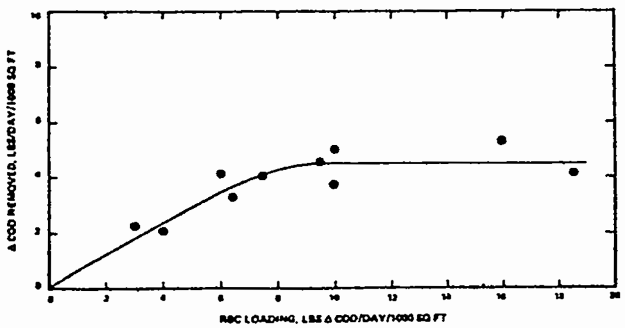


Figure 2. Relationship of ▲ COD applied (lbs./day/1000 sq.ft.) with △ COD removed (lbs./day/1000 sq.ft.) for slaughterhouse wastewater at various applied organic loadings.

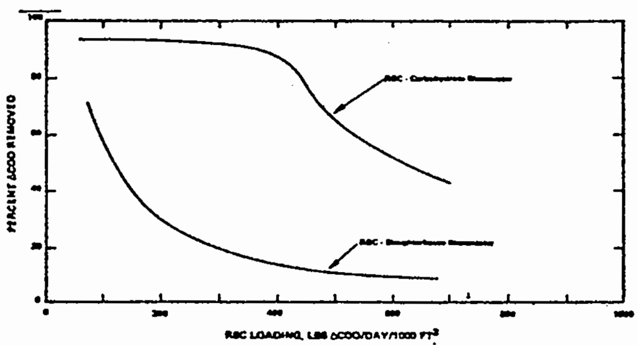


Figure 3. Comparison of percent ▲ COD removal versus applied organic loading to the RBC for carbohydrate and slaughterhouse wastewaters.

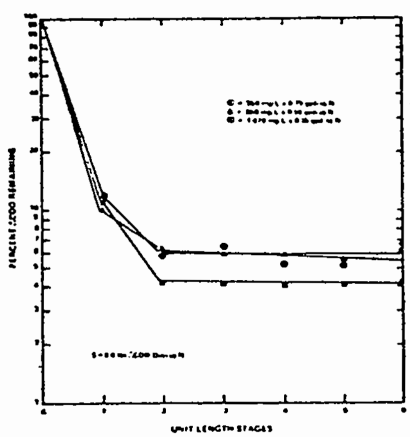
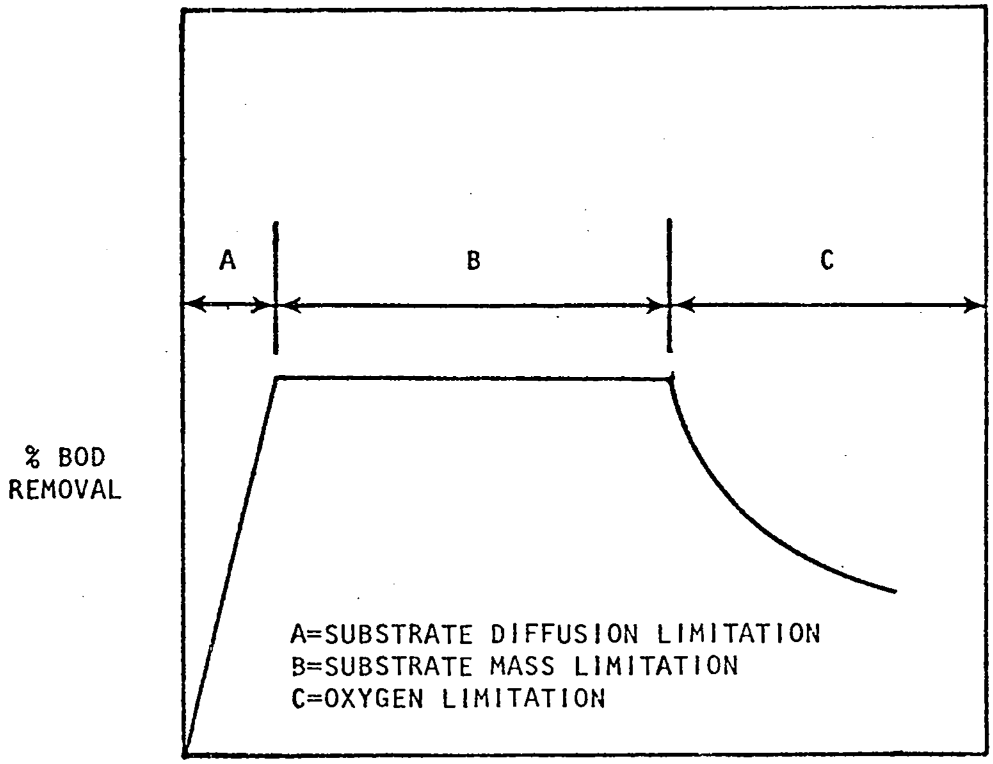
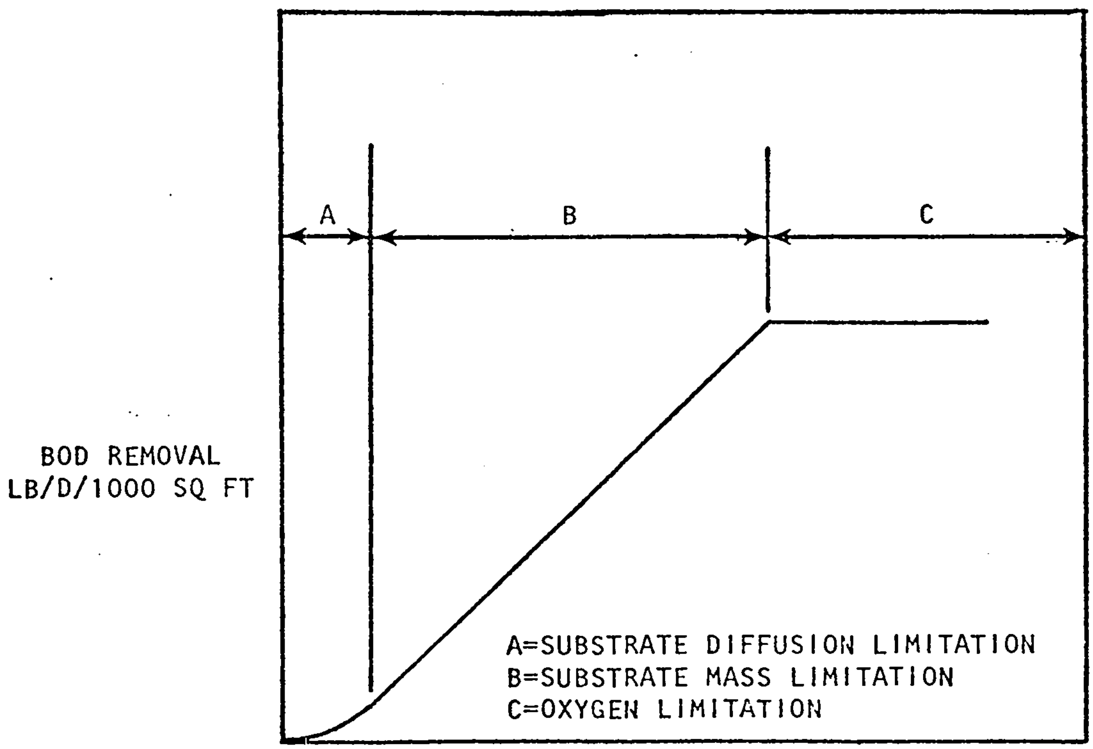


Figure 4. Percent △ COD remaining per stage with carbohydrate wastewater for various organic concentrations and various flow rates (all resulting in the same total applied organic loading).

(From Stover and Kincannon, 1976 (14), Water and Sewage Works)



5a. BOD INFLUENT or OLA



5b. BOD INFLUENT or OLA

FIGURE 5 THEORETICAL LIMITING REGIONS WITHIN THE RBC BIOFILM

Higher organic loadings cause the biofilm to grow to a depth where oxygen diffusion within it is limited due to oxygen depletion from microbial metabolism of substrate within the biofilm. An active aerobic layer at the biofilm surface is established having a constant depth independent of total biofilm depth, and removes a constant mass of substrate independent of substrate loading. The removal of a constant mass of substrate produces the logrithmic decrease in percent removal of BOD in this oxygen limiting region as shown in Figure 5a.

From this brief review of the literature it becomes obvious that analyses and results from past RBC work have been all but uniform and results have not been generally comparable nor compatible. They do reflect, however, the importance of influent BOD concentration and HLA, used together as OLA, to describe RBC performance.

PLANT DESCRIPTION

The Kirksville,Missouri wastewater treatment plant was the first full-scale municipal RBC plant to be built in Missouri and serves a population of roughly 20,000. It went into operation in June, 1976, as part of a $7 million sewerage system improvement program. Two trickling filter plants and 12 single-celled lagoons were replaced by the RBC system. A schematic flow diagram of the Kirksville plant is shown in Figure 6.

Flow entered the plant through a 36 inch influent line and was detained in a wet well where two, 66 inch diameter, screw pumps lifted the sewage to a level where it could flow by gravity through the rest of the plant. An emergency holding basin was provided with a capacity of 2.1 MG to prevent bypassing of excess flows. From the screw pumps the flow passed through a grit chamber and comminutor/bar rack structure before being split and entering the two primary clarifiers. Effluent from the primary tanks was split again before it entered the four flow paths of the RBC media. Once through the RBC tanks, the flow was combined and split to flow into two secondary clarifiers. Clarified wastewater flowed through the chlorine contact chamber, through a Parshall flume for measurement, and into an adjacent stream.

Sludge from the secondary clarifier was recycled to the front of the plant via a 10 inch return line to the wet well. It was settled in the primary tank along with primary solids and the combined sludge was pumped from the primary clarifier to a two stage digester. Digested sludge was then disposed of primarily as liquid for application to farm lands via a tank truck. A storage basin for digested sludge was also provided for winter months when land application was not possible. Vacuum filters were provided as well for dewatering of raw or digested sludge for final disposal in the city sanitary landfill.

The rotating biological media and accompanying equipment was provided by the Autotrol Corporation. Four flowpaths of five shafts each were used for this 5 MGD plant and each shaft of media consisted of 11 foot diameter rings of corrugated polyethylene media, concentrically mounted on a square steel shaft 24 inches on a side. Shafts were 25 feet in overall length and were

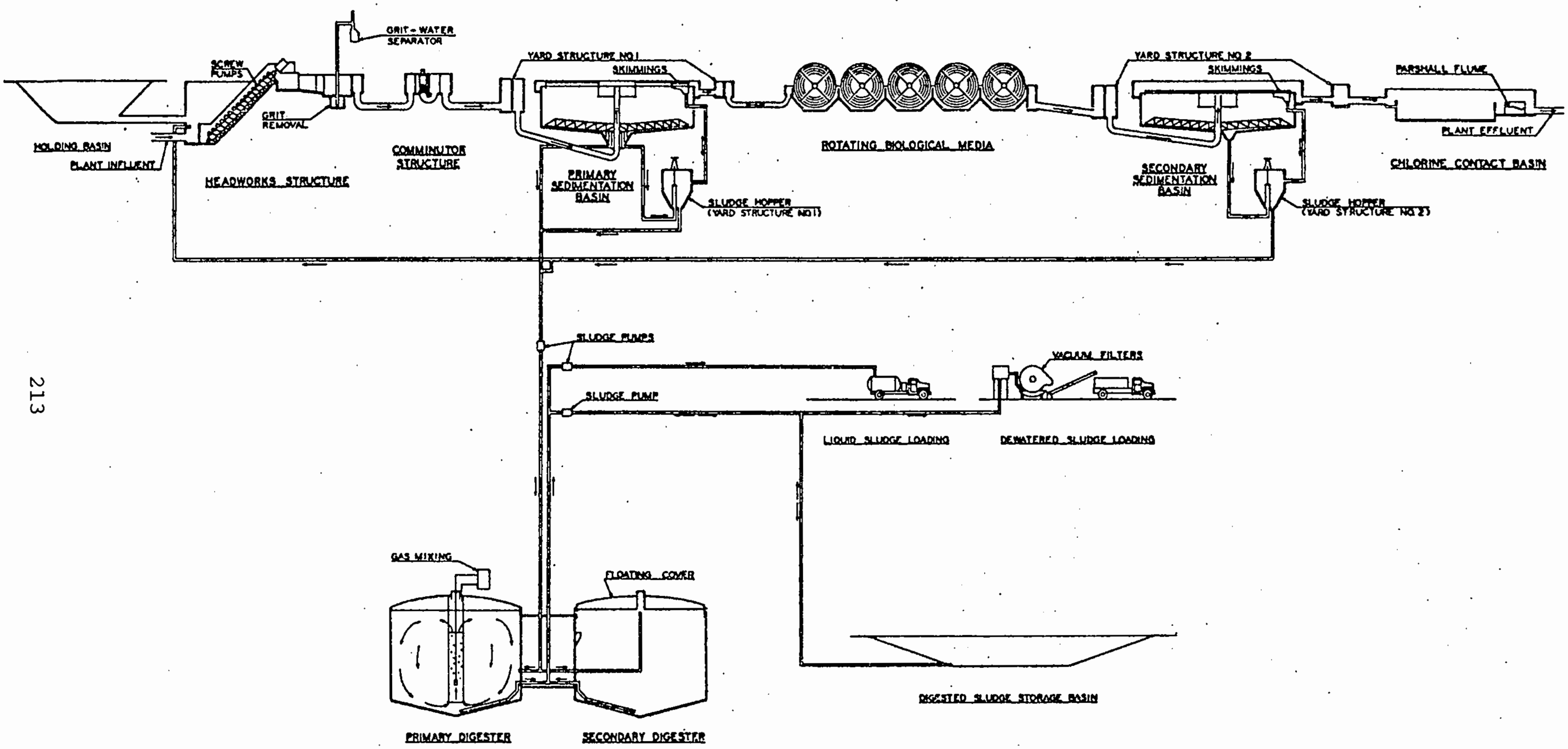


FIGURE 6 KIRKSVILLE WASTEWATER TREATMENT PLANT---- SCHEMATIC FLOW DIAGRAM - SECTION

supported at their ends and rotated with approximately 40 percent of their area submerged in wastewater. Flow entered the trapezoidal concrete reactor chambers perpendicular to the shafts and an overflow weir separated each compartment so short circuiting was prevented and each isolated tank approximated a completely mixed reactor. Each of the 20 shafts was rotated by its own drive unit rated at a maximum 7½ HP. Table 2 shows pertinent equipment and design specifications for the Kirksville RBC plant. No information was readily available for design values for influent BOD and SS so average values of 200 mg/l were assumed for each in the table.

TABLE 2 RBC SYSTEM DESIGN SPECIFICATIONS

CLARIFIERS:		
MFG	Smith & Loveless	
Type	2 primary, 2 secondary, 70' Diameter	
SWD	10'	
Feed	Center	
Overflow	Peripheral Weir	
Sludge Withdrawl	Mechanical Scraper, Full Surface Skimming	
RBC UNITS:		
MFG	AUTOTROL, Inc.	
Type	BIOSURF	
Configuration	20 Shafts, 4 Paths @ 5 Shafts	
Rotator Motor HP	20 @ 7.5	
Surface Area		
Shaft	95,400 sq ft	
Total	1,900,000 sq ft	
Tank Construction	Trapezoidal, One Interstage Baffle/Shaft	
Cover	Building	
Mode of Operation	Flow Perpendicular to Shafts	
INFLUENT:		
	Design	Peak
Q, MGD	5.0	12.3
BOD_5. mg/l	(200)	(200)
SS. mg/l	(200)	(200)
Min Temp. °F	45	
R.T., Hrs.	1.25	0.51
HLA, gal/d/sq ft	2.6	6.4
Equiv MLVSS, mg/l	15,000	
Break HP/Shaft	3	
Break HP Total	60	
Clairifier SOR gal/d/sq ft	650	1600
EFFLUENT:		
BOD_5, mg/l	30	30
SS, mg/l	30	30

DATA PRESENTATION

Data used in this evaluation were from the monthly reports of the Kirksville, Missouri WWTP for the period extending from January 1977 to November 1978. Two monthly reports were not available, June and July 1977, and were omitted from the data analysis. The data included flow, influent and effluent BOD, influent and effluent SS, and water temperature. Flow and temperature data were generally measured on a daily basis, while BOD and SS were initially determined on a daily basis until mid-March 1978 when they were measured on an average of three times a week.

Derived parameters from the available data included HLA as gal/d/sq ft, OLA as lb BOD/d/1000 sq ft, BOD removal as lb BOD/d/1000 sq ft, percent BOD removal and percent SS removal. Unfortunately no primary, secondary or individual stage measurements of these parameters were made and analysis of the treatment efficiency of the RBC in isolation was not possible. System evaluation then was based on overall plant performance in the removal of SS and total BOD as loading conditions changed. Average weekly values were used to indicate overall system performance while extreme daily values were used to indicate performance under extreme loading conditions.

In looking at the data several periods of data within the study period become important. August to September 1977 and December 1977 to May 1978 were periods of reduced surface area in the system because of structural failure of the first shaft of two separate flows paths. The surface area reduction from 1.9 to 1.4 million sq ft was taken into account in the data and allows analysis of the RBC operations under high surface loading conditions. Spring 1977 was also an important time at the plant because a discharge of highly concentrated metal plating wastes from an industry in the city stopped all biological activity in the digesters. Subsequently solids were allowed to build up in the primary clarifiers before being vacuum filtered for disposal rather then being digested. This period allows for comparison of RBC removal efficiency under adverse high solids conditions to operations under more normal conditions assumed to occur during the balance of the study period.

Three other specific periods become important in the analysis. First, September to December 1977 because it was a time of highly variable flow with relatively constant influent BOD and SS. Second, August to December 1978 because it was a time of relatively constant flow with highly variable influent BOD and SS. These two periods are of interest in determining if changes in flow and concentration effect removal efficiency for the same organic loading. The third period of interest was February 1978 because of the extremely high BOD and relatively low SS received at the plant during that time.

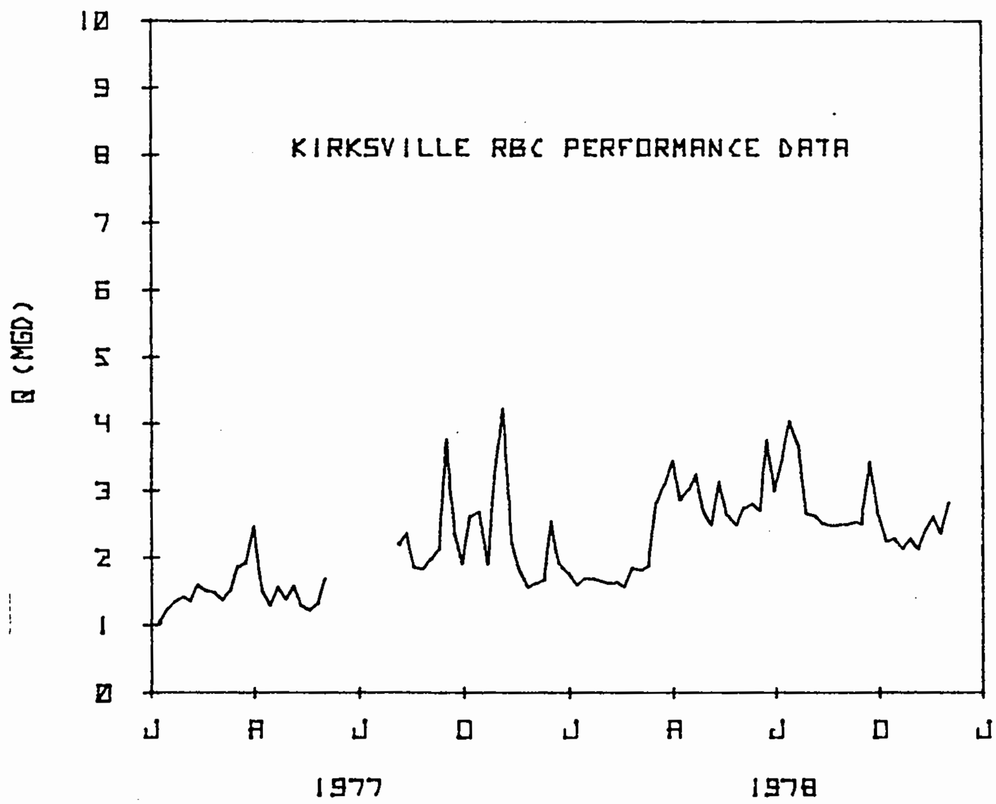
No soluble BOD data was available so the ratio of BOD/SS was used to give a relative value for the soluble nature of the waste. It was felt that with this ratio being maintained throughout the study period at a relatively constant value a relatively constant waste type would be entering the plant. Municiple waste characteristics would not normally be expected to change drastically and BOD/SS was used to indicate possible discharges of other types of wastes into the system. The highly soluble waste received at the plant in February 1978 was thought to come from a creamery in the city. Several other isolated incidents of high BOD/SS ratios occurred throughout the study period, but February 1978 was isolated because the BOD/SS ratio remained high for a full four week period, providing a sufficient period for results to show any effect on removal efficiency that might occur from the highly soluble organic loading.

Figure 7 shows the variation of weekly flow and HLA values. Flow increased gradually from 1 MGD in January 1977 to over 2.5 MGD at the end of the study period, still at half design flow. This gradual increase was related to growth of the area and to more of the collection system coming on line to the new plant. Peak flows that occurred during the study period were generally related to seasonal variations in rainfall and subsequent infiltration. Average flow for September to December 1977 was 2.5 MGD and varied from 1.6 to 4.2 MGD during that period. Average flow for August to December 1978 was 2.5 MGD and varied only between 2.1 and 3.4 MGD. Flow during February 1978 was relatively constant at 1.7 MGD ranging from 1.6 to 1.8 MGD. HLA exhibited similar trends in the data. Values of HLA ranged from 0.75 to 3.5 gal/d/sq ft and were quite variable at times as was flow. An average value of HLA for September to December 1977 was 1.3 gal/d/sq ft, for August to December 1978 was 1.3 gal/d/sq ft, and for February 1978 was 1.2 gal/d/sq ft. Variations of HLA during these same periods were 0.8 to 2.2 gal/d/sq ft, 1.1 to 1.8 gal/d/sq ft, and 1.1 to 1.3 gal/d/sq ft respectively.

Figure 8 shows influent BOD and SS weekly average variations with time. Again no soluble BOD data was available from Kirksville but the ratio of influent BOD/SS was used as an indicator of the soluble nature of the waste. BOD/SS averaged 1.24 for the study period less February 1978. During February 1978 the ratio of BOD/SS averaged 2.03. This four week period of high BOD and low SS was seen to represent the lowest BOD removal efficiency of the whole study period. Influent BOD varied from 113 to 441 mg/l during the study period and averaged 237 mg/l. Influent BOD averaged 192 mg/l for the period September to December 1977, 262 mg/l for August to December 1978, and 383 mg/l for February 1978. Influent BOD ranged between 127 and 257 mg/l, 165 and 343 mg/l, and 299 and 441 mg/l during the three periods respectively.

Suspended solids did not show the variation that BOD did and ranged from 103 to 312 mg/l, averaging 191 mg/l. Influent SS generally ranged between 100 and 200 mg/l except for the period of May 1977 where SS steadily rose from 242 to 312 mg/l. This was the period when digester problems occurred and the solids build up in the system was thought to be the reason for the rise in influent SS.

Figure 9 shows the variation of weekly OLA and BOD removal for the study period. OLA reflects both hydraulic and influent BOD variations and was seen to vary widely throughout the study period. OLA varied from 1.2 to 4.9 lb/d/1000 sq ft and averaged 2.5 lb/d/1000 sq ft for the whole study period. OLA averaged 2.0 lb/d/1000 sq ft and varied from 1.2 to 3.6 lb/d/1000 sq ft during the period September to December 1977. OLA for August to December 1978 varied between 1.9 and 3.9 lb/d/1000 sq ft and averaged 2.8 lb/d/1000 sq ft. February 1978 OLA values averaged 3.7 lb/d/1000 sq ft, and ranged from 3.2 to 4.2 lb/d/1000 sq ft. BOD removal followed OLA closely for nearly all of the study period except during February 1978 when the highly soluble organic loading was applied to the plant.



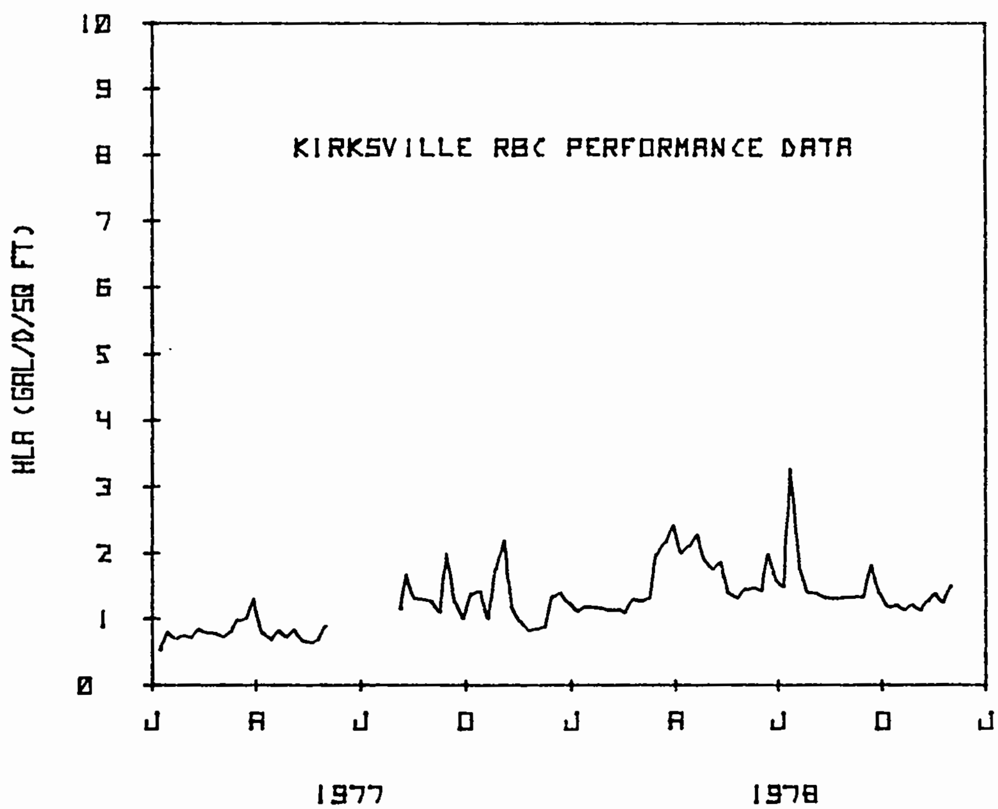
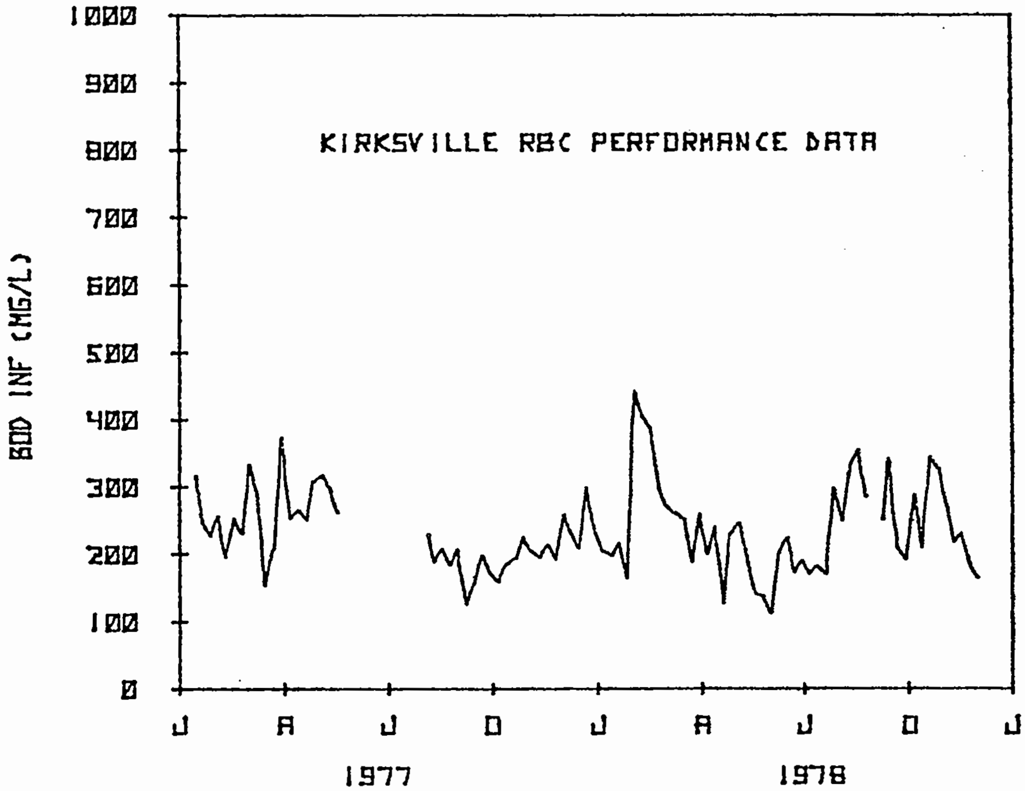


FIGURE 7 WEEKLY VARIATIONS OF HYDRAULIC PARAMETERS



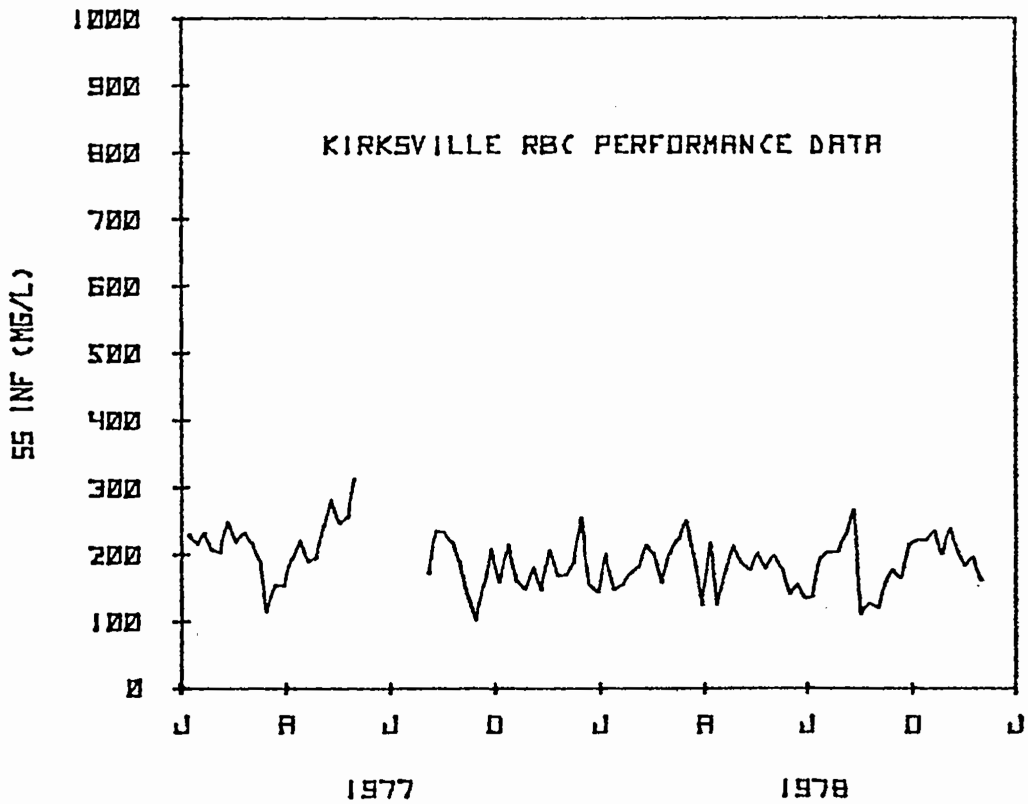
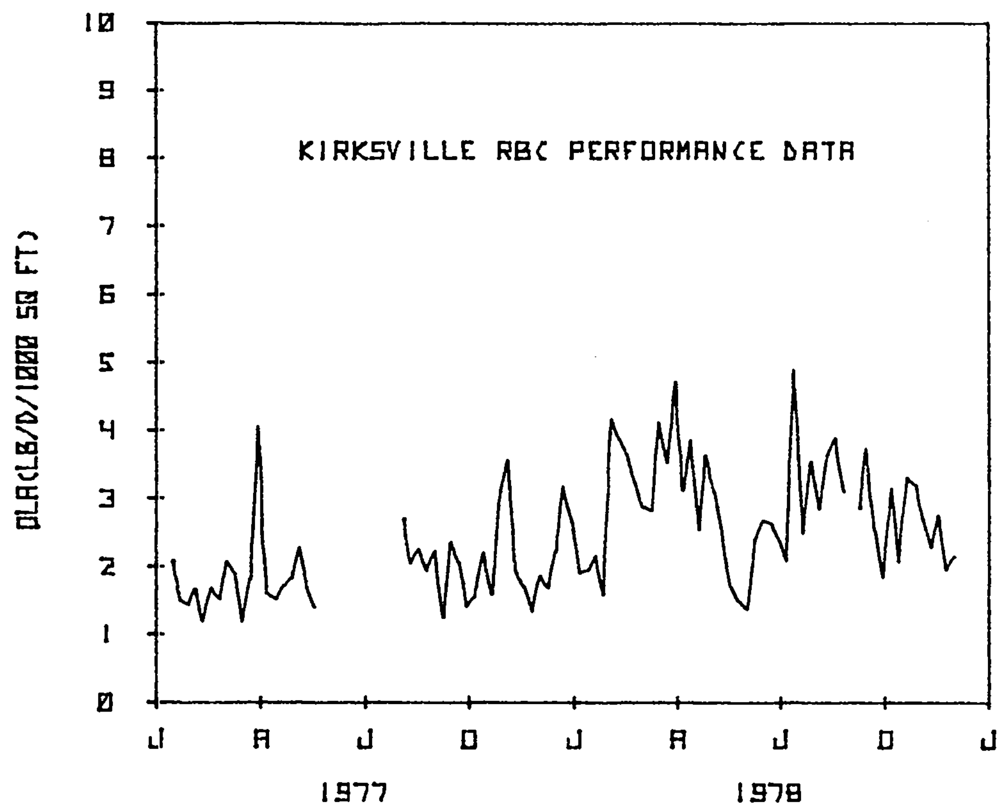


FIGURE 8 WEEKLY VARIATIONS OF WASTEWATER INFLUENT PARAMETERS



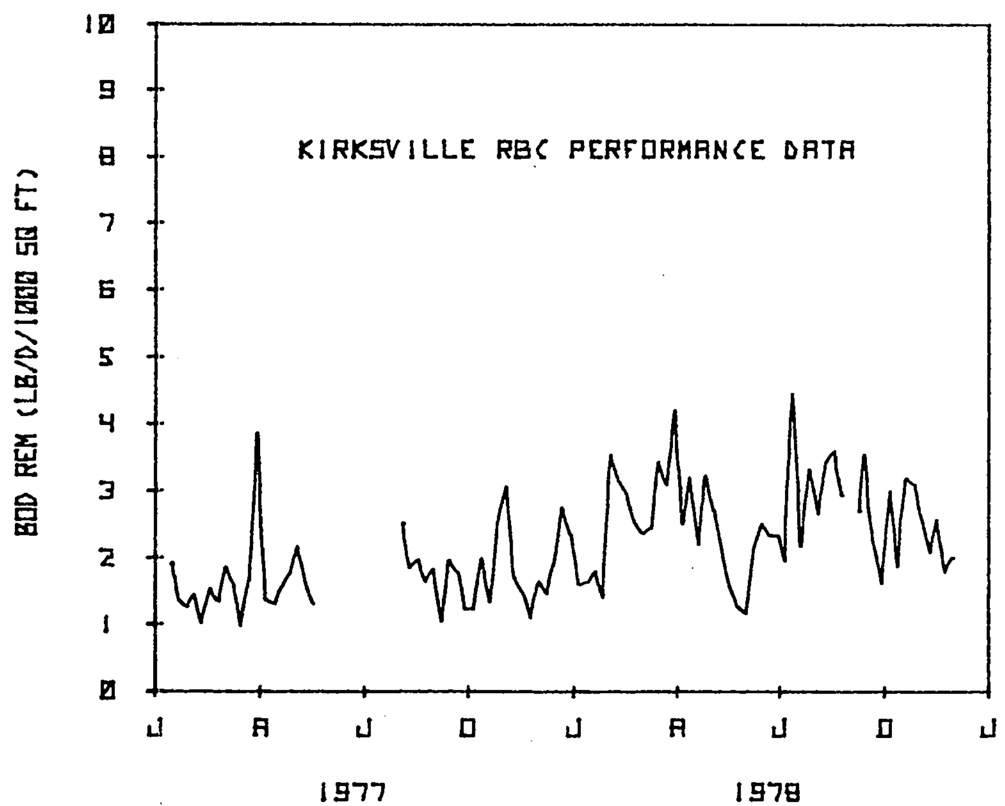
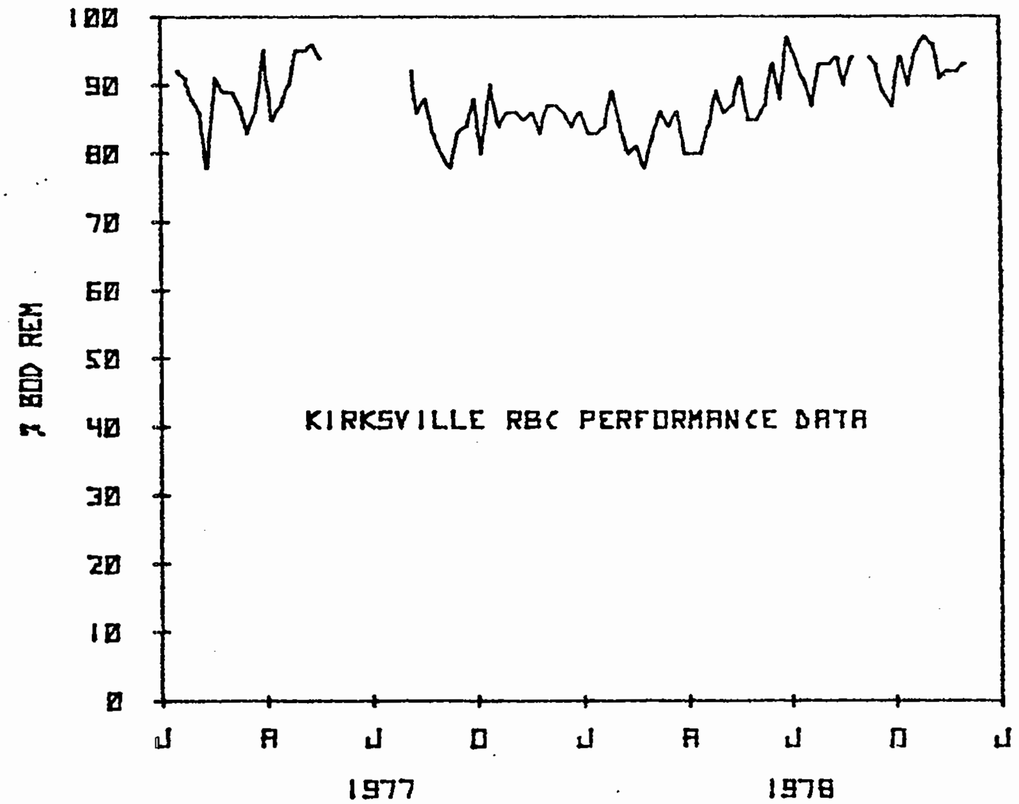


FIGURE 9 WEEKLY VARIATIONS OF SURFACE LOADING AND REMOVAL PARAMETERS

Figures 10 and 11 represent removal efficiency as percent BOD and SS removal and resulting effluent quality in mg/l of BOD and SS. Percent BOD removal generally ranged between 80 and 95 percent during the study and averaged 88 percent for the whole period less February 1978. February 1978 weekly average BOD removal values ranged between 78 and 84 percent and averaged only 81 percent, the lowest of any four week period during the study. Low BOD removal during February 5-11, 1977 was due to a reported influent BOD concentration of only 23 mg/l which yielded 22 percent BOD removal. Average removal without this value would be 90 percent for the period. February 25 to March 2, 1978 was also a low BOD removal efficiency period averaging little better than 81 percent. Again low removal efficiency was due not to high loading on the system, but to low influent BOD concentrations reported. Influent BOD's as low as 83 mg/l resulted in very low removal efficiencies and thus lowered the average removal value for that period. BOD removal for Septmeber to December 1977 was fairly constant ranging from 80 to 95 percent and averaged 90 percent. August to December 1978 yielded BOD removals ranging from 87 to 98 percent, averaging 94 percent for the period.

SS removal remained over 90 percent for most of the study period but was seen to deteriorate after solids built up in the system as a result of digester problems in the spring of 1977. Despite the digester problems however, removal remained at an average of 91 percent during the latter part of 1977. Surge hydraulic loadings were also shown to affect SS removal but percent SS removal during the variable flow period of September to December 1977 still averaged 90 percent and ranged from 80 to 95 percent. August to December 1978, the period of constant flow and varying organic loading, produced a range of SS removal to 87 to 98 percent with an average removal of 94 percent. SS removal ranged from 86 to 90 percent and averaged 88 percent for February 1978.

Effluent quality as shown in Figure 11 provided valuable information as to the trouble areas of the system. Effluent BOD averaged 24 mg/l for the study period less February 1978. Effluent quality in February 1978 deteriorated to an average 70 mg/l with peaks reaching 75 mg/l during the middle two weeks of the month. Effluent quality did not completely recover



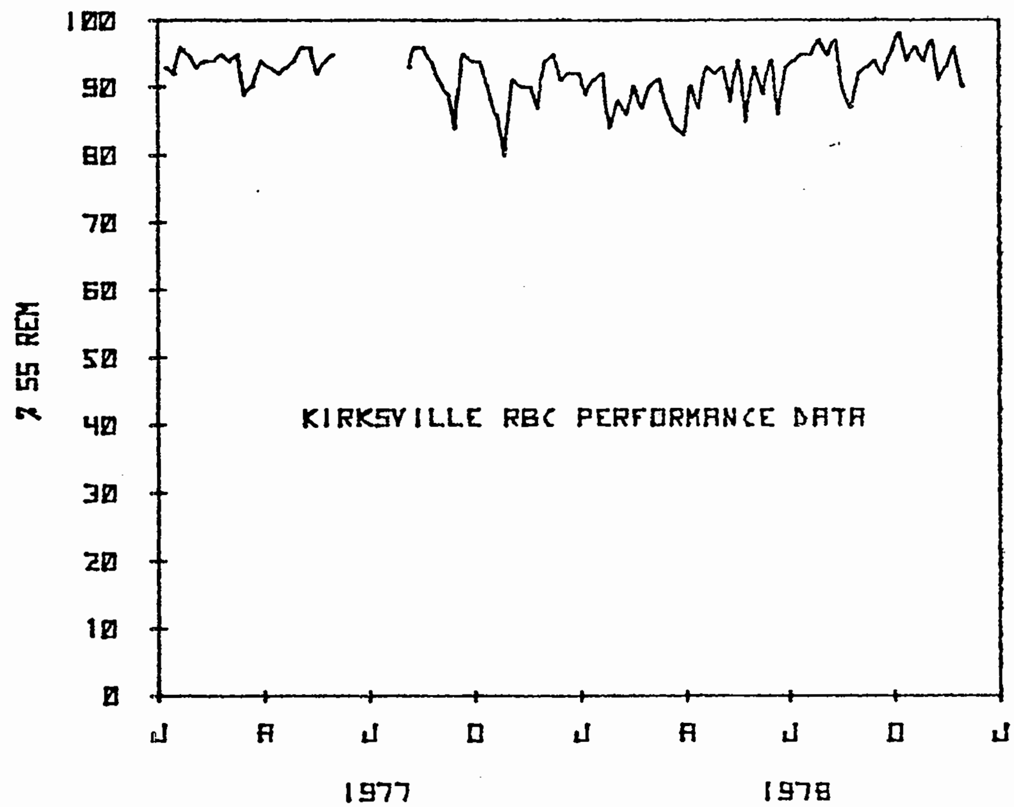
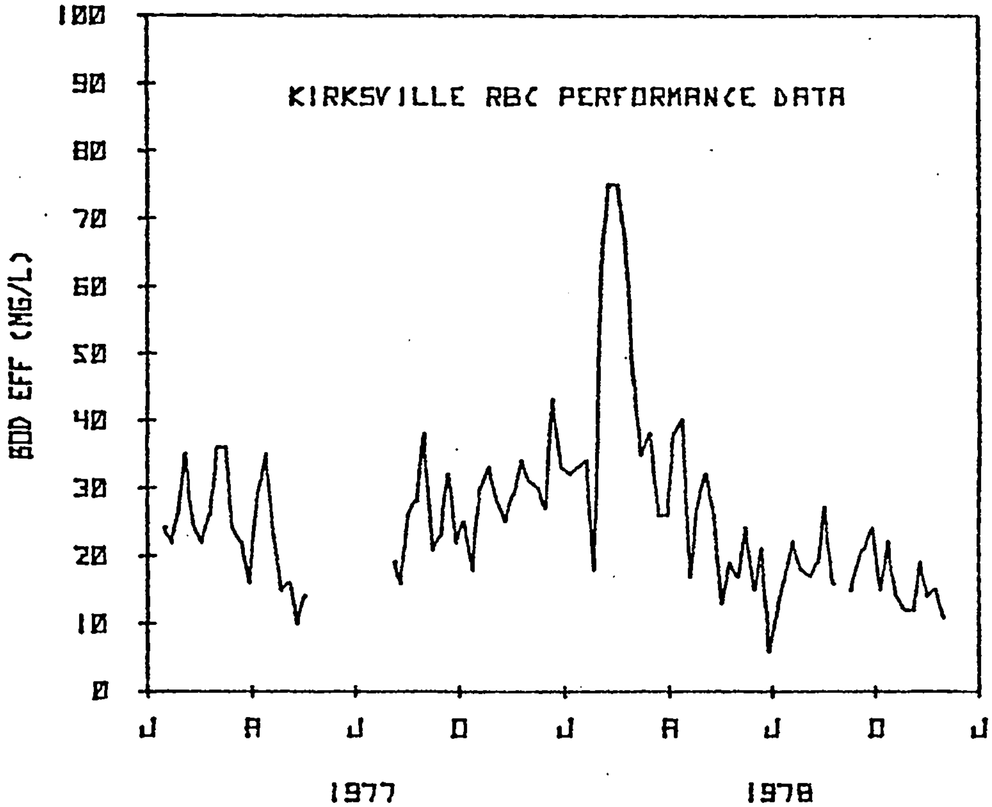


FIGURE 10 WEEKLY VARIATIONS OF REMOVAL EFFICIENCY PARAMETERS



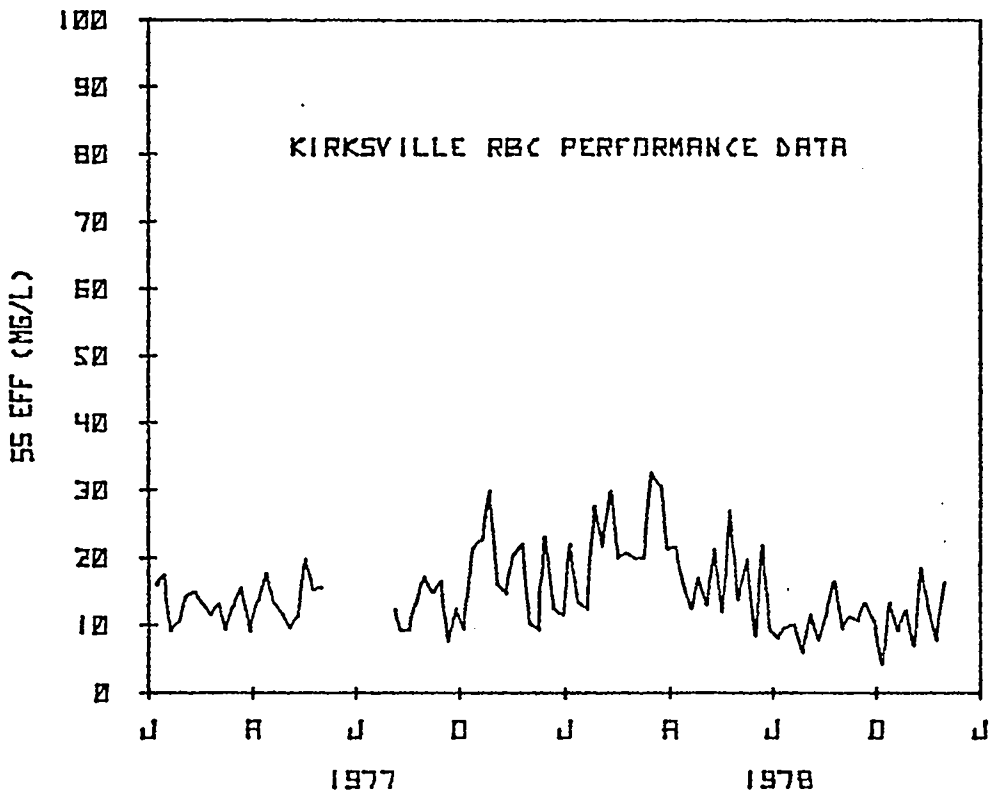


FIGURE 11 WEEKLY VARIATIONS OF EFFLUENT QUALITY PARAMETERS

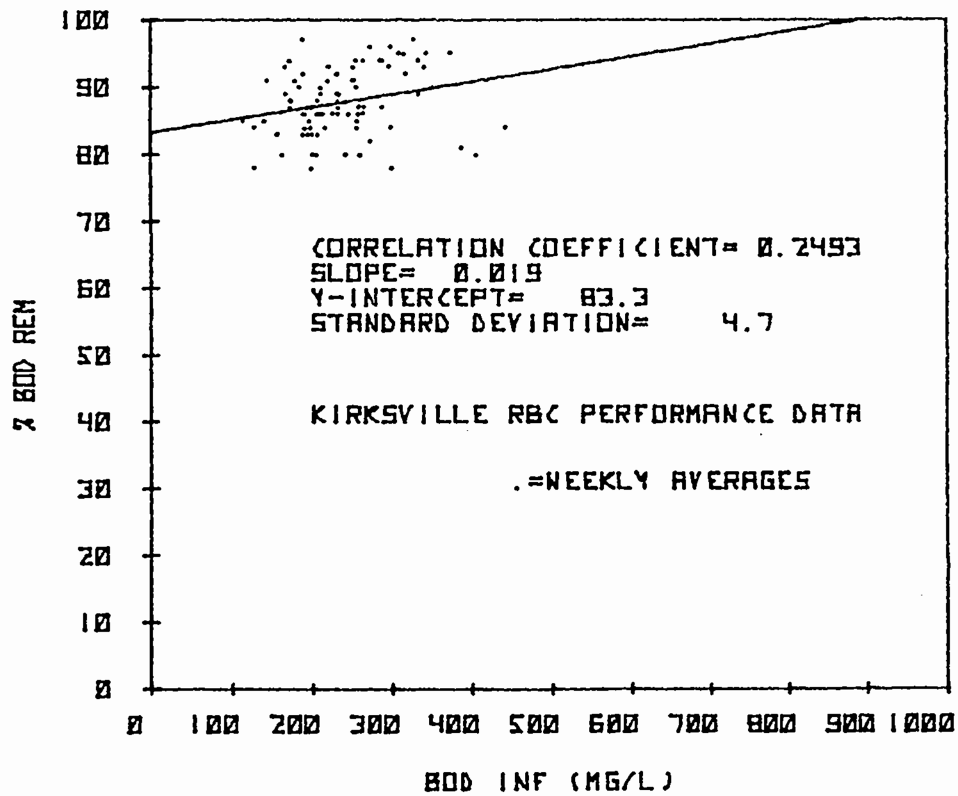
identified in the discussions above. All values are average weekly values from data found in the Kirksville treatment plant monthly reports.

TABLE 3 RBC LOADING CHARACTERISTICS

PARAMETER	S-D 1977	February 1978	A-D 1978
		PERIOD	
Q (MGD)			
Average	2.5	1.7	2.5
Range	1.6-4.2	1.6-1.8	2.1-3.4
HLA (gal/d/sq ft)			
Average	1.3	1.2	1.3
Range	0.83-2.2	1.1-1.3	1.1-1.8
BOD/SS	1.24*	2.03	1.24*
BOD Influent (mg/l)			
Average	192	383	262
Range	127-257	229-441	165-343
SS Influent (mg/l)			
Average	191*	191*	191*
OLA (lbBOD/d/1000 sq ft)			
Average	2.0	3.7	2.8
Range	1.2-3.6	3.2-4.2	1.9-3.9
%BOD Removal			
Average	84	81	93
Range	78-90	78-84	87-97
%SS Removal			
Average	90	88	94
Range	80-95	86-90	87-98
Effluent BOD (mg/l)			
Average	29	70	17
Range	18-43	63-75	11-25
Effluent SS (mg/l)			
Average	16	23	11
Range	8-30	20-30	5-18

*Average for the entire study period

The relationships between BOD removal and loading parameters were then investigated. Figure 12 shows the relationship between percent BOD removal and influent BOD. Without daily extremes a general increase in percent removal resulted from an increase in influent BOD. The correlation coefficient was extremely low however, and the relation did not prove to be an adequate estimator of BOD removal efficiency. With daily extreme values added the curve approached the theoretical curve shown earlier in Figure 5, with an initial increase in removal to a limiting value of influent BOD of approximately 100 mg/l, beyond which an approximately constant removal of 90 percent resulted, independent of influent BOD concentration to a value of 700 mg/l.



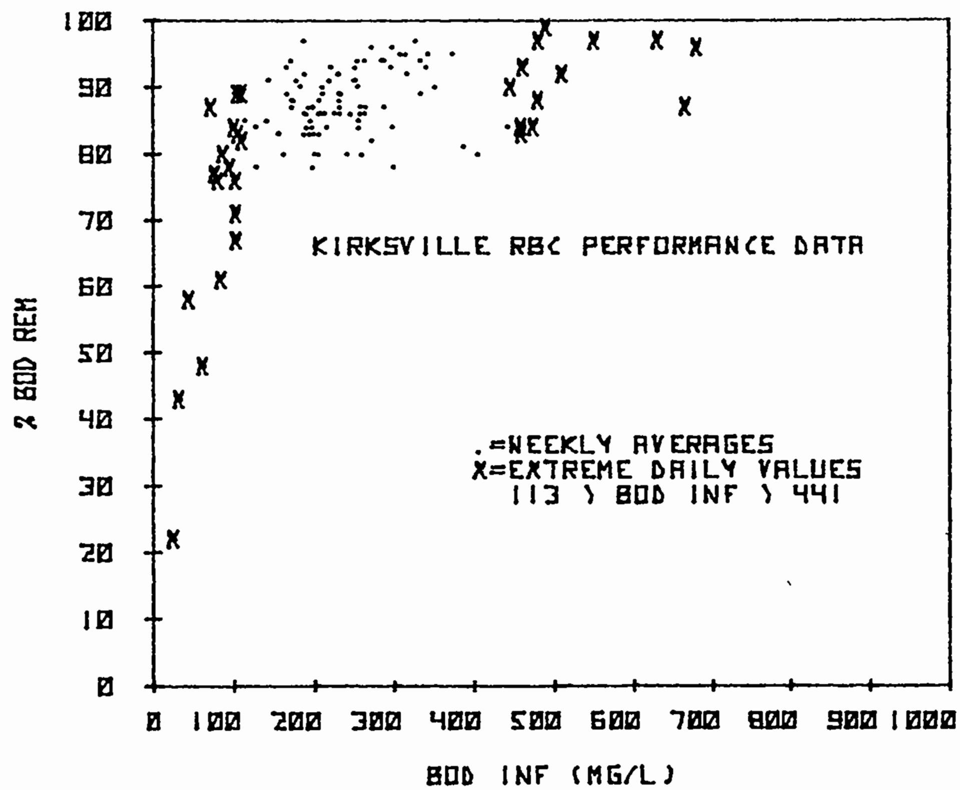


FIGURE 12 PERCENT BOD REMOVAL VERSUS INFLUENT BOD CONCENTRATION, WITH AND WITHOUT DAILY EXTREME VALUES

Similar results were observed for the plot of percent BOD removal versus OLA shown in Figure 13. A slight increase in BOD removal with increased OLA was seen in the plot of weekly averages; but again correlation was very poor between variables, making the relation impractical for predictive purposes. With the addition of extreme daily values a rapid increase in percent removal occurred to approximately 90 percent at 1 lb BOD/d/1000 sq ft, beyond which percent removal remained constant for values of OLA to 9.6 lb BOD/d/1000 sq ft.

Figure 14 shows percent BOD removal as a function of HLA. Generally, increased HLA produced decreasing percent BOD removal, but, again, correlation between variables was extremely low. When daily extreme values were added, correlation of the variables increased; but, correlation was still very low and no definitive relation was evident from the plot. HLA was found to be an imprecise indicator of RBC performance on a percent removal basis and other relationships were investigated to see if they would be more descriptive of RBC removal efficiency.

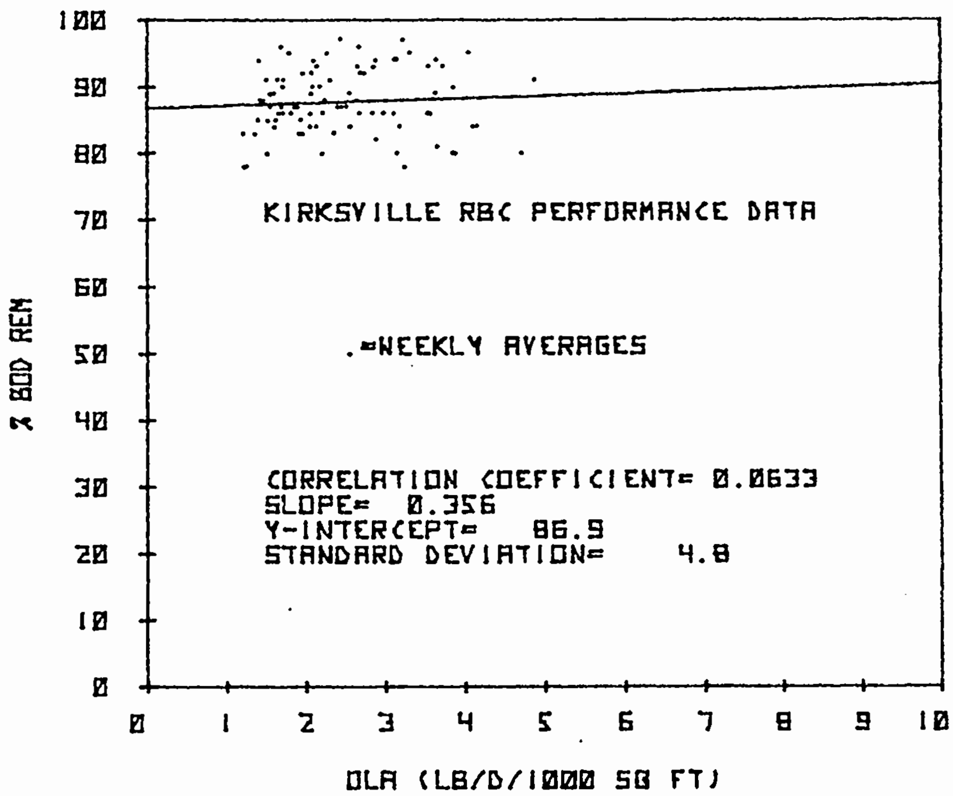
BOD removal on a pound/d/1000 sq ft basis was plotted against HLA for specific influent BOD concentration ranges as shown in Figure 15. These ranges were chosen to produce the lines of best fit and were shown to describe the data extremely well. Daily extreme values were added to yield information on whether peak loadings would cause deviations from weekly average plots. Table 4 gives the least squares statistical results for the data plotted for Figure 15. It can be seen that the data correlated well for all ranges of influent BOD with and without daily extreme values.

Figure 16 is another form of Figure 15, being BOD removal as lb/d/1000 sq ft versus influent BOD for a given HLA range. Ranges of HLA were again chosen to yield maximum correlation coefficients and daily extreme values showed no significant deviation from monthly average trends. Table 5 shows the statistical data for the set of curves in Figure 16. Again the data correlated extremely well and could be used as an indicator of RBC performance for RBC analysis and design.

Figure 17 shows BOD removal versus OLA as lb/d/1000 sq ft and exhibits the same relationship reported in the literature. A constant rate of removal resulted for general weekly average values from 1.2 to 5.0 lb BOD/d/1000 sq ft and was maintained with the addition of daily extreme values to a loading of 9.6 lb BOD/d/1000 sq ft. No leveling off of the relationship was observed at the high loading conditions, suggesting that the Kirksville plant was mass limiting even at these extreme loadings. With the high correlation coefficient of 0.9869 for weekly average values, this relationship can be used as an estimator of RBC BOD removal on a lb/d/1000 sq ft basis for a given organic loading. The most important variable in analysis of RBC performance, the one most indicitive of BOD removal efficiency, was OLA. For this reason OLA should be used for design and analysis of RBC's rather than HLA or detention time in the system.

CONCLUSIONS

The results of the described research and review of the literature enabled the following conclusions to be made:



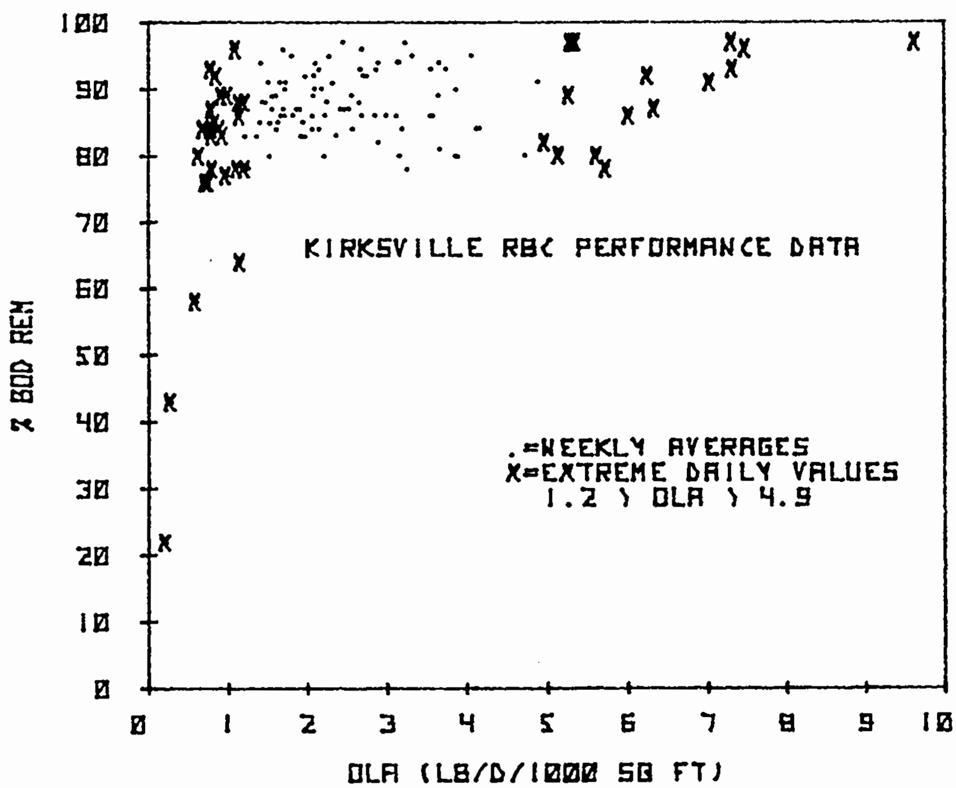
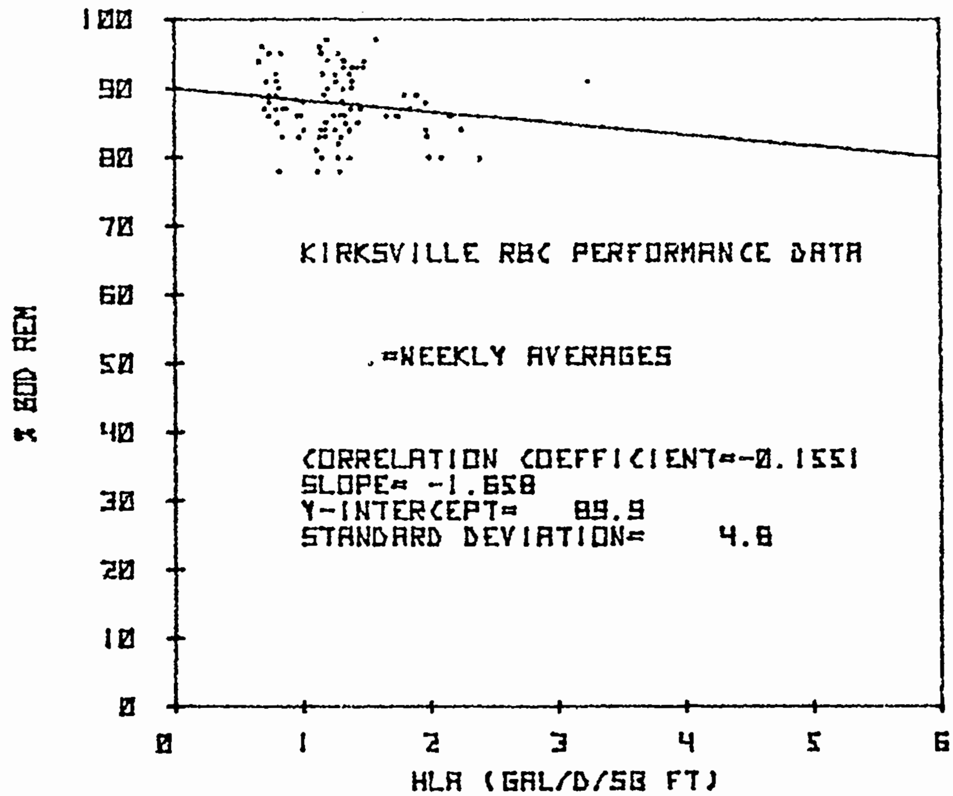


FIGURE 13 PERCENT BOD REMOVAL VERSUS DLA, WITH AND WITHOUT DAILY EXTREME VALUES



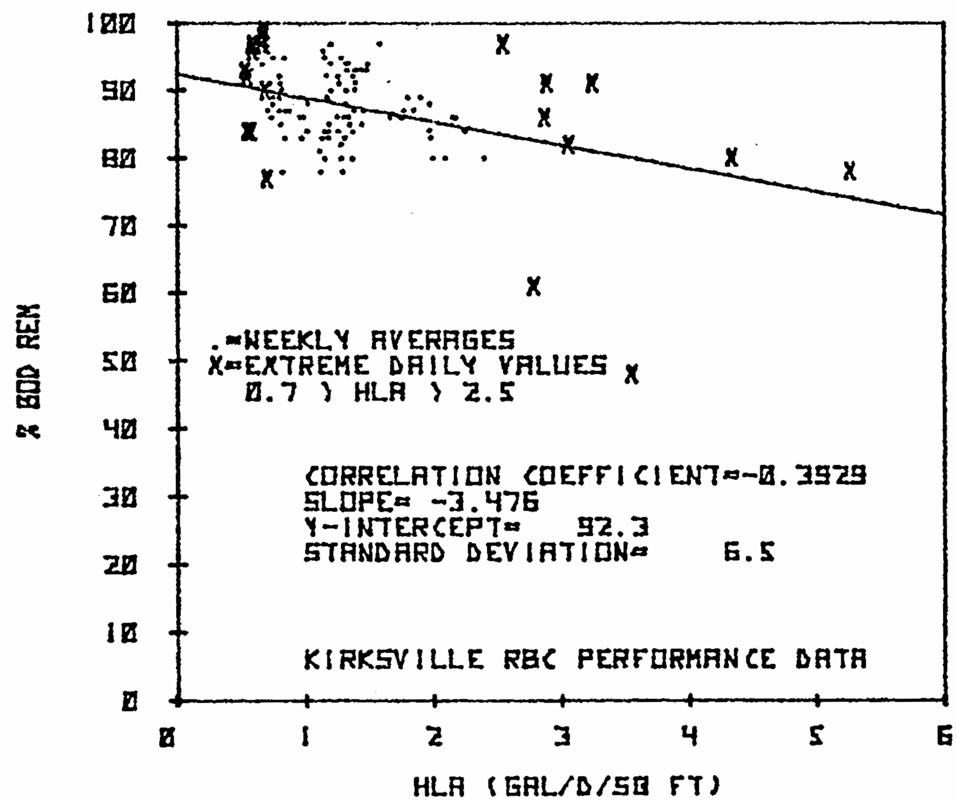


FIGURE 14 PERCENT BOD REMOVAL VERSUS HLA, WITH AND WITHOUT DAILY EXTREME VALUES

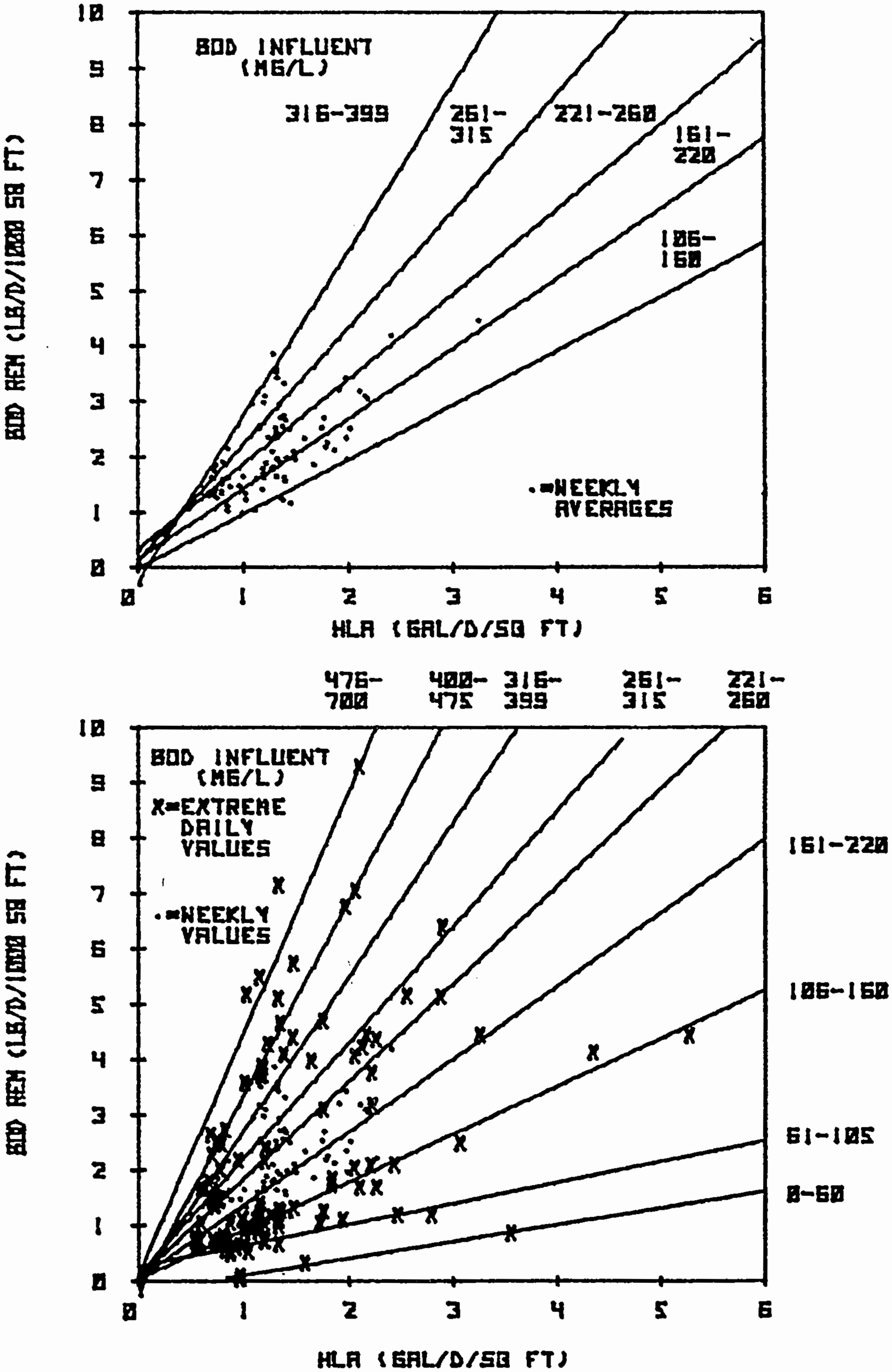


FIGURE 15 BOD REMOVAL (LB/D/1000 SQ FT) VERSUS HLA FOR A GIVEN INFLUENT BOD CONCENTRATION RANGE, WITH AND WITHOUT DAILY EXTREME VALUES

TABLE 4 BOD REMOVAL VS HLA CURVE STATISTICAL DATA

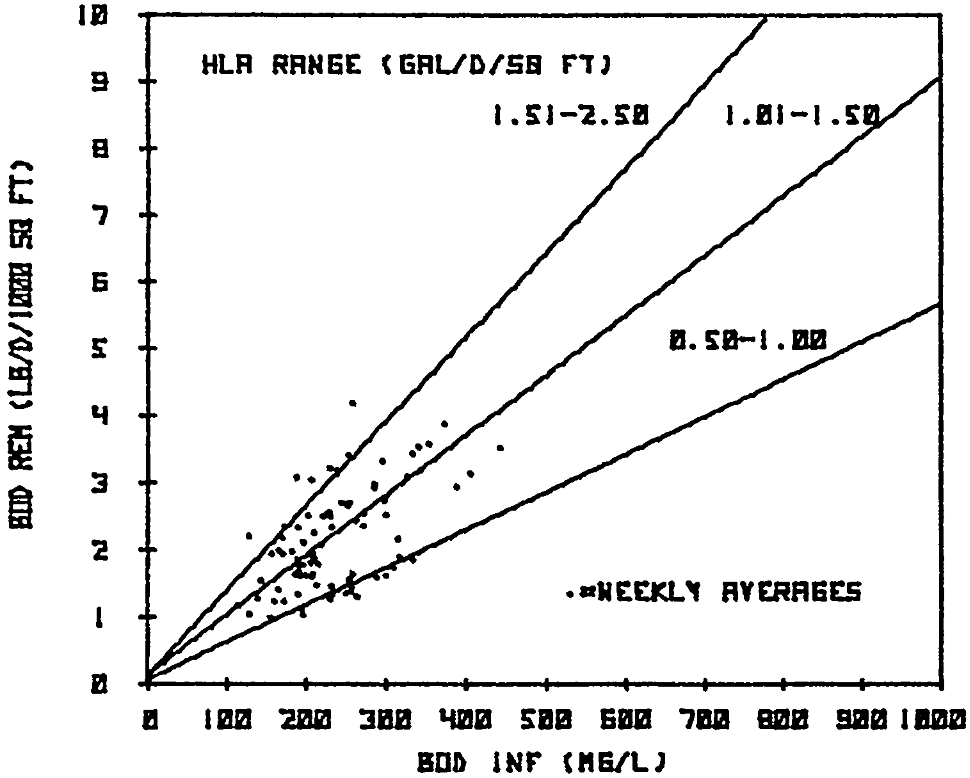
BOD INFLUENT RANGE	WITHOUT DAILY EXTREMES			WITH DAILY EXTREMES		
(mg/l)	R	M	Y	R	M	Y
0-60				0.9937	0.30	-0.20
61-105				0.9927	0.38	0.26
106-160	0.9594	0.98	-0.02	0.9786	0.87	0.04
161-220	0.9572	1.26	0.14	0.9690	1.32	0.03
221-260	0.9637	1.53	0.32	0.9707	1.76	0.05
261-315	0.9341	2.10	0.00	0.9842	2.09	0.08
316-399	0.9717	2.97	-0.03	0.9533	2.78	-0.13
400-475				0.9875	3.52	-0.25
476-700				0.9072	4.34	0.04

WHERE: R=CORRELATION COEFFICIENT
M=SLOPE OF LINE OF BEST FIT
Y=Y-INTERCEPT OF LINE OF BEST FIT

TABLE 5 BOD REMOVAL VS BOD INFLUENT CURVE STATISTICAL DATA

HLA RANGE	WITHOUT DAILY EXTREMES			WITH DAILY EXTREMES		
(GAL/DAY/SQ FT)	R	M	Y	R	M	Y
0.50-1.00	0.8778	0.006	0.06	0.9489	0.006	-0.04
1.01-1.50	0.9162	0.009	0.12	0.9591	0.009	-0.06
1.51-2.50	0.7311	0.013	0.10	0.9649	0.016	-0.51
2.51-4.50				0.9582	0.023	-0.07

WHERE: R=CORRELATION COEFFICIENT
M=SLOPE OF LINE OF BEST FIT
Y=Y-INTERCEPT OF LINE OF BEST FIT



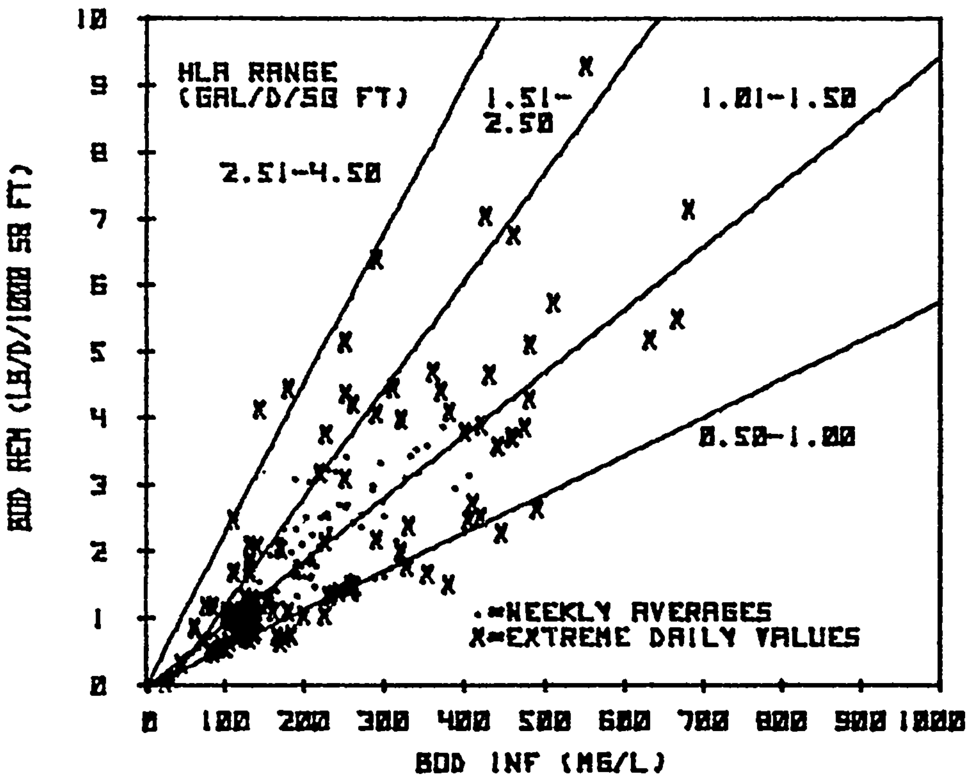
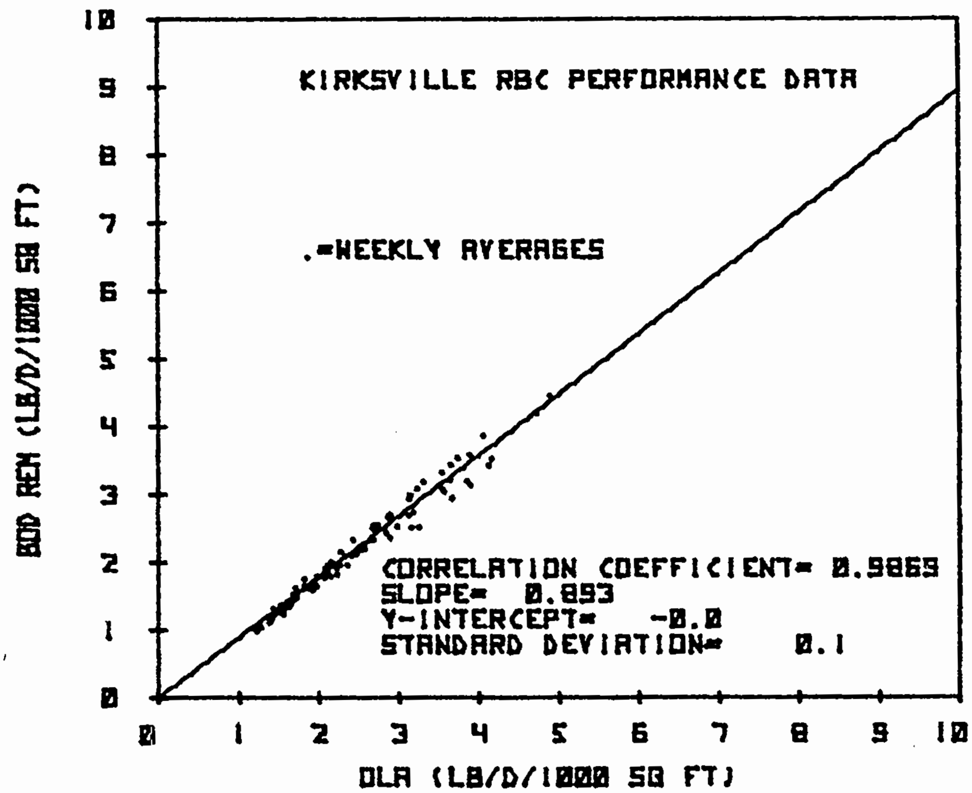


FIGURE 16 BOD REMOVAL (LB/D/1000 SQ FT) VERSUS BOD INFLUENT FOR A GIVEN HLA RANGE, WITH AND WITHOUT DAILY EXTREME VALUES



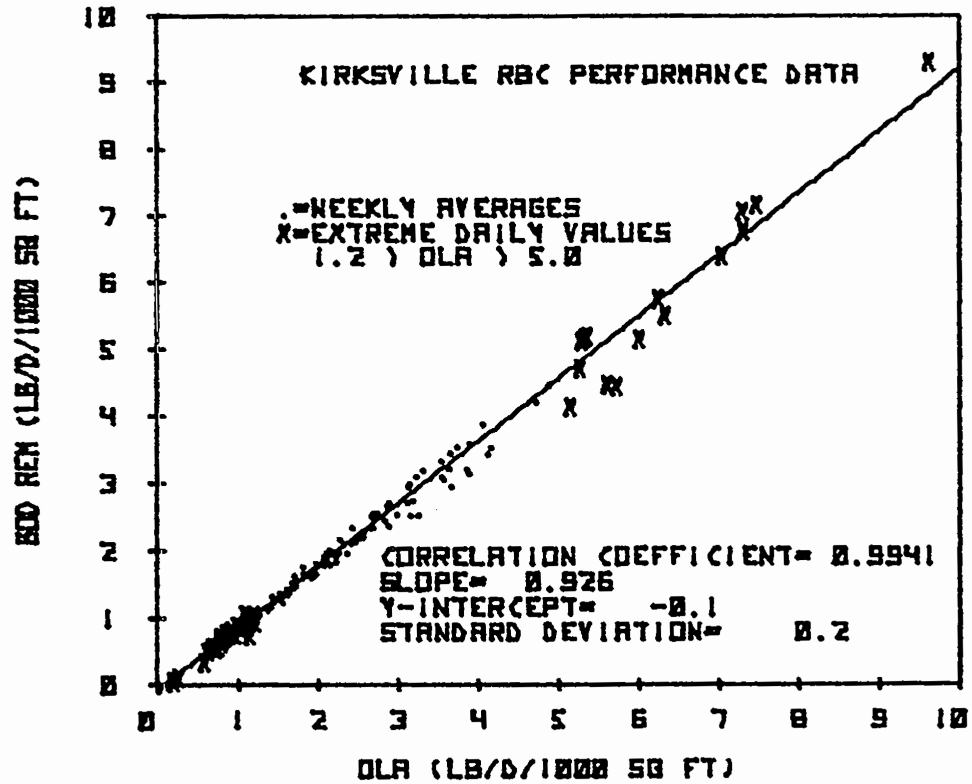


FIGURE 17 BOD REMOVAL (LB/D/1000 SQ FT) VERSUS OLA, WITH AND WITHOUT DAILY EXTREME VALUES

1. The Kirksville RBC units performed efficiently under normal organic and hydraulic loading conditions, giving an average BOD removal efficiency of 88 percent with an average BOD of 24 mg/l over OLA values from 1.2 to 4.9 lbs BOD/day/1000 sq ft and HLA values from 0.5 to 3.2 gals/day/sq ft.

2. High soluble organic loadings adversely affected RBC performance, giving an average BOD removal of 81 percent with an average BOD of 70 mg/l at an OLA of 3.7 lbs BOD/day/1000 sq ft and an HLA of 1.2 gals/day/ sq ft.

3. Highly variable hydraulic loadings reduced the treatment efficiency of the RBC units as a result of reduced contact time and surges on the final clarifiers.

4. By itself, HLA was not a satisfactory indicator of RBC performance; but could be combined with organic concentration to yield the OLA which correlated readily with RBC operations.

5. The best indicator for RBC operations appeared to be the soluble BOD/day/1000 sq ft loading rate.

6. The lack of adequate data prevented the development of more precise design criteria for RBC units but the results of this study helped to indicate the data required for proper evaluation.

ACKNOWLEDGEMENTS

We would like to thank the City of Kirksville for their help in providing the data necessary for the study. We would also like to thank Bob Lamberton of Larkin and Associates for his help in providing background information on the design and operation of the Kirksville plant. Also Jane Boatwright, Secretary of the Environmental Health Engineering Department, deserves a great deal of thanks for all her help and patience in typing this paper.

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ROTATING BIOLOGICAL CONTACTOR FOR THE TREATMENT OF WASTEWATER IN INDIA

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INTRODUCTION

There is a long standing need in India of providing satisfactory, low cost and simple sewage treatment facilities for isolated houses, institutions, hotels, small and medium sized communities. Such facilities should require occasional inspection and maintenance, flexibility in construction and operation, relatively unskilled supervision, maintenance and operation, and should occupy limited area. They should be capable of treating wide variation of flow and organic load producing effluent of desirable standards of purity with respect to BOD and suspended solids. The Rotating Biological Contactor (RBC) or Bio-disc system, a secondary biological treatment process, is claimed to have these advantages which appear to be suitable to Indian conditions.

The RBC or Bio-disc unit consists of a series of closely spaced circular discs mounted on a horizontal rotating shaft. The shaft along with the discs are fixed in a semi-circular cylindrical tank through which the wastewater flows, with the water level just below the shaft. While rotating at a low speed, the disc surface is alternately exposed to the atmosphere and wastewater. The disc serves as media for growth and adhesion of biological slime, device for bringing the slime and film of water in contact with air, and creating mildly turbulent mixing conditions within the tank contents. As the disc rotates, the biological slime on any sector of the disc is alternately dipped into the wastewater, where the slime metabolises non-settleable and dissolved organic matter and aerated with each revolution of the disc. The settleable organic matter and the sloughed film due to excess growth from the disc, passes along with the effluent as suspensions which are removed in the subsequent stage in the secondary settling tank.

REVIEW OF LITERATURE

The idea of RBC originated in USA in 1928, and it was referred to as 'Biological Wheel' and in Germany as 'Immersion Drip Filter'[1]. Further developments in the process took place during the last two decades mainly in Germany and USA. Large number of plants are now in vogue in Europe and USA as reported by Anthonie[2]. Laboratory and pilot plant studies in understanding the kinetics and evolving design criteria are reported by Antonie[3,4], Torpey et al[5], Khan and Siddiqi[6], Raman and Khan[7,8], Pescod and Nair[9], Pretorius[10], Steels[11], Bruce and Merkens[12], Clark et al[13] and Kluge and Kipp[14].

The first municipal wastewater treatment plant in Pewaukee, USA using RBC process constructed by Autotrol Corporation on full scale was in operation since 1971 to treat a flow of 1179 m^3/day (0.47 mgd); and the first large scale RBC treatment unit with a capacity of .20 mgd for upgrading the existing treatment unit is in operation in Philadelphia.

Based on the studies reported in USA, Europe and Thailand, the general requirements for design and operation are as follows: The rotational speed of the disc is limited to 1 to 10 rpm and the clear spacing between the discs is kept at 2 to 5 cms with a gap of about 5 cms at the bottom. A Detention time of 10 to 90 minutes, organic disc loading rate of 3 to 16 gms/m^2/day, hydrauloc loading rate of 0.04 to 0.11 m^3/day/m^2 of disc area, power consumption (depending on capacity) of 0.3 to 2.6 kwh/kg BOD removed and power requirements of 0.4 to 1 watt/m2 of disc area are the other relevant parameters, while the BOD removal efficiency ranges from 80 to 95% at wastewater temperatures of 10°C to 20°C.

OBJECTIVE OF STUDY

The RBC as a treatment device is not yet in use in India, eventhough it has great potential due to its compactness and simplicity of operation and favourable temperature conditions. As such, it is felt that the feasibility of its use in India should be investigated under laboratory and field conditions. The studies aim to achieve the objectives of formulating relationship amongst the treatment efficiency, organic loading, hydraulic loading, rotational speed, power consumption, and evolve compact units for various treatment capacities.

MATERIALS AND METHODS

Initially, laboratory model studies were carried out with synthetic sewage and settled sewage, followed by studies on pilot plant treating raw municipal sewage.

A. Laboratory Model Using Synthetic Sewage and Domestic Sewage

The laboratory model set up of RBC as shown in Fig. 1 consisted of 20 cm diameter, 4 mm thick, as-bestos cement discs (with perforations) with clear spacing of 2 cms. They were centrally mounted to a horizontal shaft rotated by a fractional horse power motor (¼ HP) fitted with reduction gear and adjustable belt drive for varying the rotational speed (3,5 and 8 rpm). The discs along with shaft assembly were mounted to a perspex tank (of cross section similar to Imhoff tank). About 40 per cent of the disc area was submerged in wastewater. The bottom portion of the tank acted as a settling tank cum sludge storage tank. The reaction-cum-settling chamber was partitioned into small compartments to contain varying number of discs.

It was possible to feed separately each of the compartment and work independently of the others with separate inlet and outlet arrangements. Continuous feeding was accomplished by electrolytic pump connected to feed bottle containing synthetic or raw settled sewage. Fig. 2 shows the set up of the laboratory model experiment. The sloughed material and the settled organic material were periodically removed from bottom storage chamber. The feed was varied for various hydraulic and organic loading rates. The influent and effluent samples were collected at particular time intervals regularly and analysed for the usual parameters like temperature, pH, BOD, COD, S.S, TDS, NH_3-N. The units were continuously dosed at loading rates varying from 6.2 to 42 gms of BOD per square meter of disc area per day. The synthetic sewage was prepared by dissolving a high protein cereal and milk powder in water to get a BOD_5 of about 250 mg/l.

Later, the experiments were repeated with settled domestic municipal sewage, and the unit was operated at organic load rates ranging from 6.2 g/day/m^2 to 31.0 g/day/m^2 of disc area, at rotational speeds of 3,5 and 8 rpm.

Further studies were carried out using cylindrical semicircular chamber fitted with shaft and some discs with separate settling compartment, the total capacity remaining the same as that of Imhoff type tank.

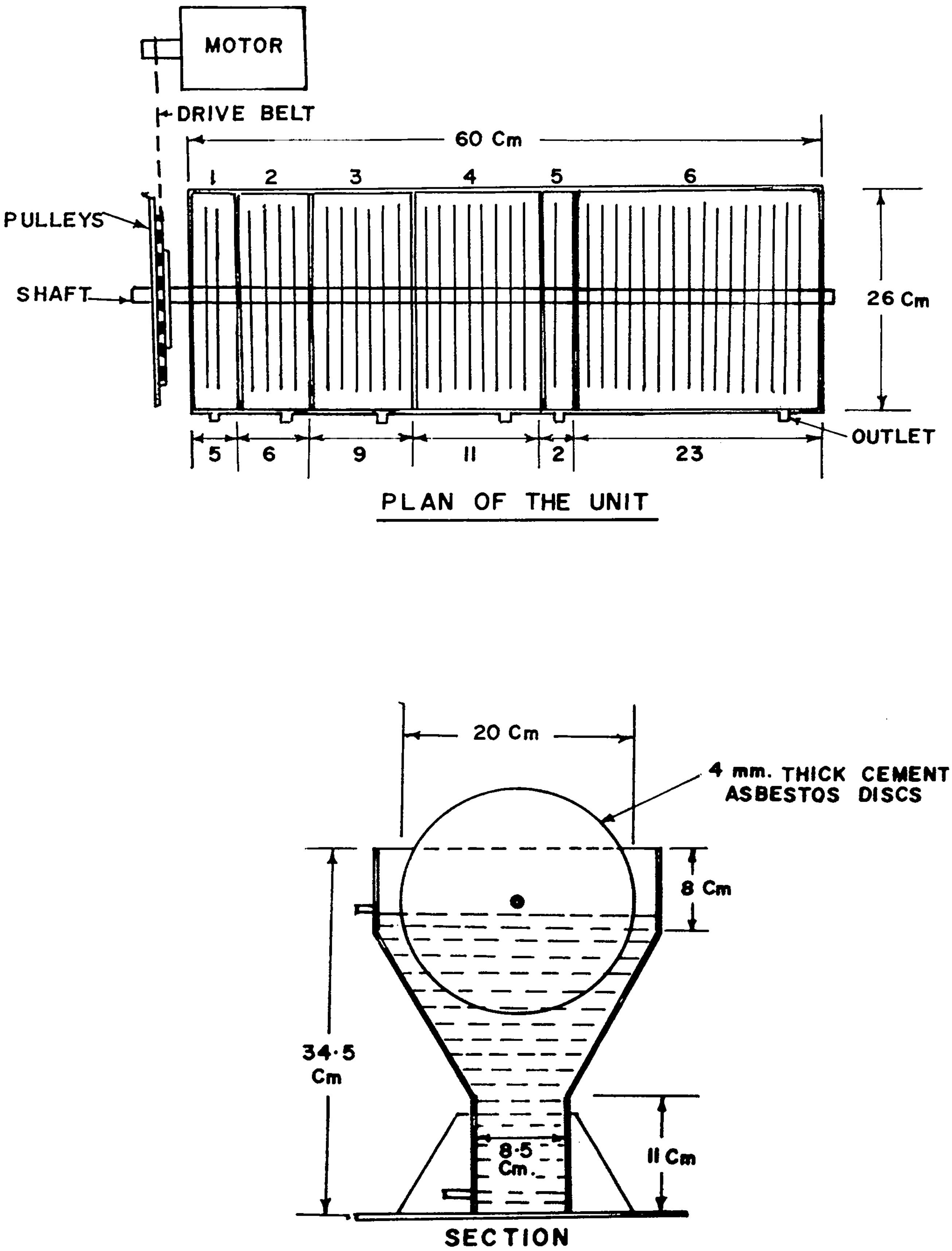


FIG. 1 : LABORATORY UNIT OF ROTATING BIOLOGICAL CONTACTOR.

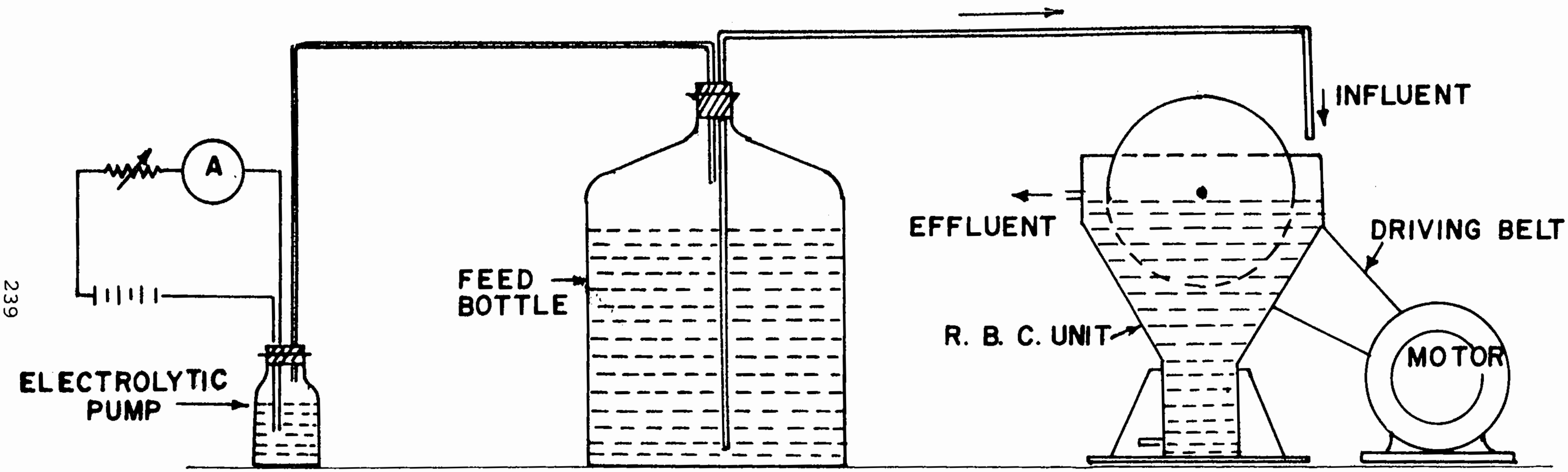


FIG. 2 : ROTATING BIOLOGICAL CONTACTOR SHOWING FEEDING ARRANGEMENT

B. PILOT PLANT.

(i) Large size composite RBC cum Settling Tank with A.C. Sheets.

The studies were carried out initially on a large size pilot plant of Imhoff tank type with the settling chamber below the reactor. The discs were made from 1.22m x 1.22m square pieces of asbestos cement sheets (136 numbers) by chopping off the corners to give octagonal shape with surface area of 1.226 square metres. They were mounted on a mild steel shaft rotated at 5 rpm by 5 H.P. motor fitted with reduction gear and chain drive. The settled domestic sewage was fed at the rate of 47.67 m^3/day for a continuous period of 10 to 12 hours in a day. The studies had to be discontinued after 4 months operation due to some structural failure of the discs.

(ii) Small size Plant with PVC circular sheets.

Fig. 3* shows the details of the set up of the RBC where the secondary settling chamber is a separate unit. The tank located in NEERI campus, Nagpur, consisted of semicircular mild steel tank of 1.22 metre diameter and 1.52 m long. 40 plane discs made of PVC of 1 meter diameter were fixed centrally to a mild steel shaft of 3.7 cm diameter and they were spaced longtudinally at clear interval of 2.5 cms. The clearance from the bottom of the tank to the bottom edge of disc was kept at 7.5 cms. The shaft was rotated at 5 rpm by a 1.5 kw (2 H.P.) electric motor fitted with reduction gear and belt drive. The unit was fed by raw sewage tapped from a distribution chamber where raw municipal screened sewage was pumped. The effluent from the disc unit passed on to the rectangular settling tank (of mild steel) of size 1.22m x 0.91m x 0.508 m depth with a liquid capacity of 0.5 m^3. The settleable organic material and the 'humus' or sloughings from the film attached to the discs were removed in the settling tank. The feed to the system was for 8 to 12 hours continuously in a day, and the remaining hours of the day, the RBC was working without any feed. The flow measurements were carried out by V-notch attached to a separate chamber after the settling tank. Occasionally, volumetrically also the flow was computed. The performance of RBC was observed under two conditions viz. open type and enclosed type covered by a mild steel perforated cover (fig. 4). Later, some modifications were made in the outlet for the settling tank (vide Fig. 4) by providing serrated weirs instead of 5 cms. diameter circular opening initially.

The studies were carried out first with open reactor and secondary settling tank with a single pipe outlet and later the same units were modified with the reactor closed by a ventilated semi-circular lid and the settling tank provided with serrated weir outlets.

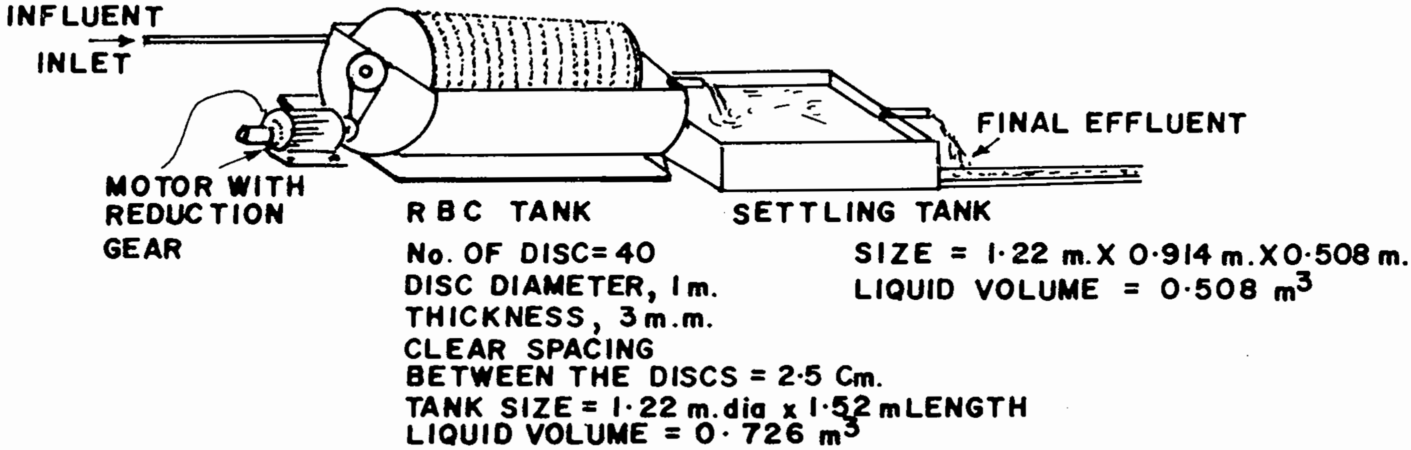


FIG. 3 : ROTATING BIOLOGICAL CONTACTOR PILOT PLANT WITH P V C DISCS

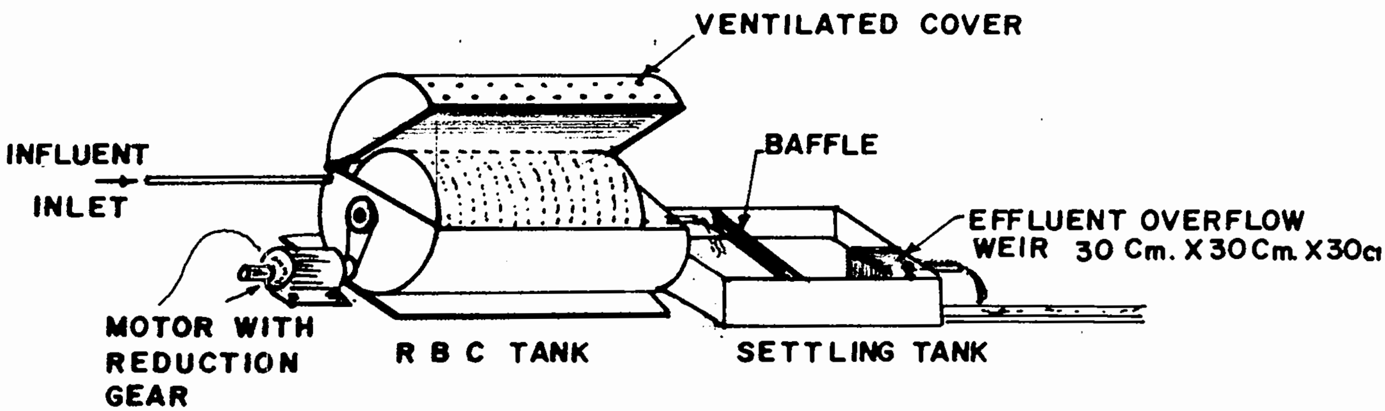
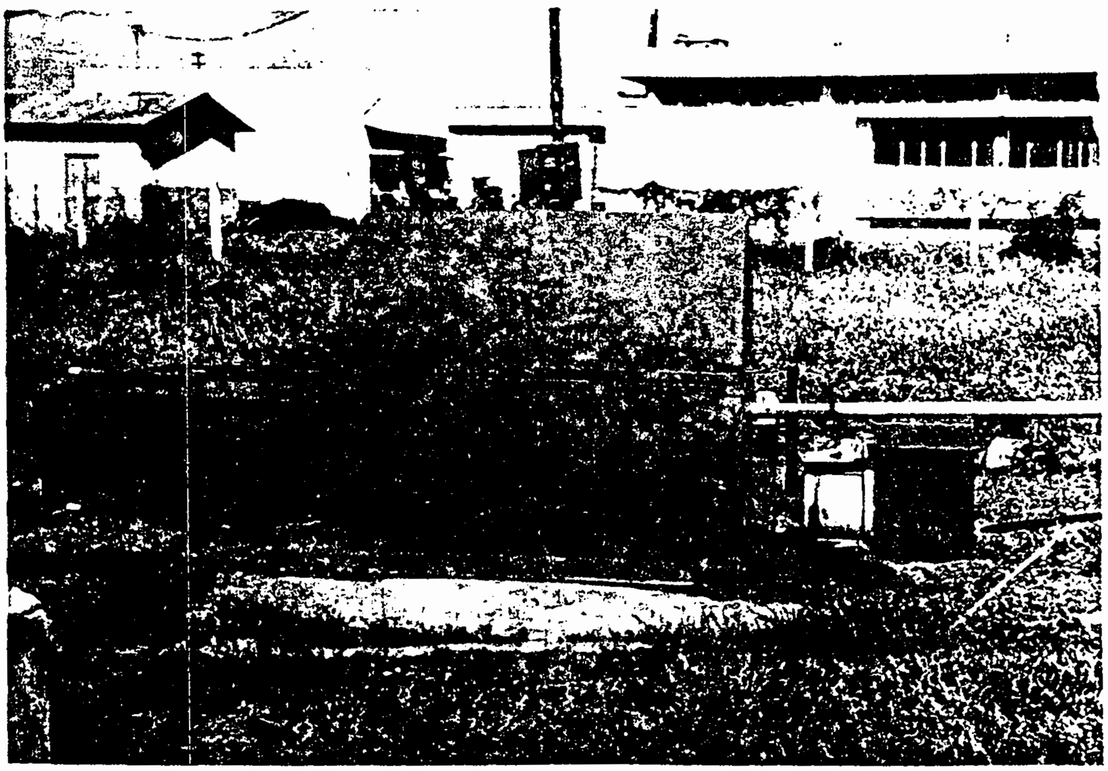
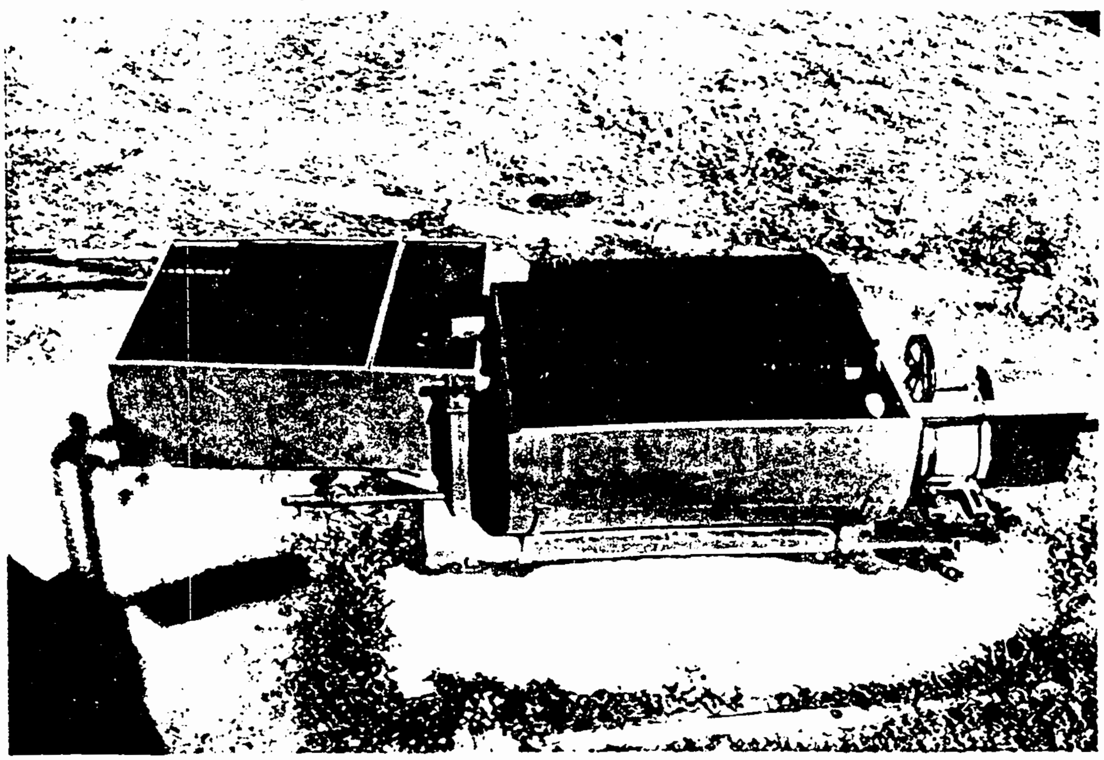


FIG. 4 : R B C PILOT PLANT WITH COVER AND WITH MODIFIED SETTLING TANK

Photograph : 1 Rotating Biological Contactor showing Biological Slime on Discs.



Photograph:2 RBC Pilot Plant with modified settling tank and ventilated cover.



Photograph : 3 RBC Pilot Plant with cover removed.

OBSERVATIONS AND DISCUSSIONS.

A. Laboratory RBC Model Composite (Imhoff tank type) Reactor.

i) Using Synthetic Sewage

In the Imhoff tank type reactor with settling chamber below the disc chamber, it took about 7 to 10 days to reach for the optimum growth of biological film on the disc surfaces and steady state conditions. Table 1 shows the performance data of the reactor with reference to hydraulic load, temperature, COD, BOD and S.S. of the influent and effluent, and the biomass accumulation on the discs. The organic loading rates (in terms of BOD_5) ranged from 6.2 to 42 gms/day/m^2 of disc area (1.2 to 8.7 lbs/day/1000 square feet of disc area). The efficiencies varied from 55 to 88 percent removal with reference to BOD_5 in the decreasing order of organic loadings. The BOD of the influent varied from 233 to 276 mg/l while the S.S. of the influent was 196 mg/l. The S.S. in the effluent ranged between 18 to 25 mg/l.

ii) Using Settled Domestic Sewage.

The same composite unit was later charged with settled domestic sewage for different loading rates working at three different rotating speeds namely 3 rpm, 5 rpm and 8 rpm. Different organic loadings were achieved by utilising different chambers having different numbers of discs and adjusting if necessary the hydraulic flow rate. It took about 5 to 7 days for the biological slime to grow and attain a steady state. Table 2 shows the average performance characteristics of the unit. The average wastewater temperature was 28°C. It is seen from Table 2 that 90 per cent overall BOD_5 removal could be obtained with organic loading rate of 10 g/m^2/day, when the unit was operated at 5 rpm. The BOD_5 of the final effluent after settling was always less than 23 mg/l (8 to 23 mg/l) during the period of study, while the BOD_5 of the influent to the reactor varied from 102 to 130 mg/l, for the three different speeds considered. The suspended solids present in the settled effluent varied from 8 to 15 mg/l.

Fig. 5 & Table 2 & 3 show the average overall performance of the disc system at the three speeds studied viz. 3,5 and 8 rpm. There was marginal increase in the efficiency of removal of BOD with increase in speed for the same organic loading rates.

iii) Laboratory RBC Unit with Semi-Circular reactor and Separate Settling Tank.

A comparative study for a brief period was made regarding the performance of semicircular tank reactor with separate settling tank and composite Imhoff tank reactor using settled domestic sewage. It is seen from Table 4 that the efficiency of BOD_5 removal of 83% could be obtained at an organic loading rate of 14.9 g/m^2/day for the semi-circular unit,

TABLE- 1 : LABORATORY STUDY DATA ON THE PERFORMANCE OF RBC UNIT (IMHOFF TANK TYPE, COMPOSITE) FOR THE TREATMENT OF SYNTHETIC SEWAGE (1970-1972)

PARAMETERS		EXPERIMENT NO.[@]		
		1	2	3
No. of discs		6	4	1
Surface area (m^2)		0.39	0.26	0.06
Hydraulic loading, m^3/m^2/day		0.0256	0.0385	0.166
Detention time (hrs.)		8.4	6.0	1.9
Organic loading, g/m^2/day		6.0	10.6	42.4
BOD (mg/l)	Influent	243	276	276
	Effluent	28	33	124
	% reduction	88.4	87.8	54.6
COD (mg/l)	Influent	438	466	466
	Effluent	72	60	195
	% reduction	84.0	86.3	58.0
Effluent suspended solids, mg/l.		12.0	12.0	23.0
Biomass, VSS g/m^2 of disc surface		18	25	25

@ For each experiment No. at least 10 observations were recorded and the average given.

* Disc speed - 3 RPM

* Hydraulic flow for 24 hours in all the experiments was 10 litres.

* Liquid temperature varied between 21-22°C.

* Influent Suspended solids 196 mg/l.

TABLE -2 : PERFORMANCE OF THE LABORATORY RBC UNIT* (IMHOFF TANK TYPE) TREATING SETTLED SEWAGE+ (1975)

DISC RPM	3			5			8		
** EXPT. No.	1	2	3	1	2	3	1	2	3
Hydraulic load, $m^3/m^2/day$	0.051	0.099	0.287	0.047	0.066	0.115	0.051	0.099	0.288
Organic load $g/m^2/day$	6.64	11.76	29.29	6.20	10.00	14.88	6.64	12.20	31.00
BOD_5 mg/l									
Influent	130	118	102	130	150	129	131	122	108
Effluent	10	15	29	9	14	23	8	8	22
% reduction	92	87	71	93	90	83	94	93	79
COD mg/l									
Influent	290	261	218	278	309	265	304	269	234
Effluent	48	72	96	47	66	80	59	64	90
% reduction	83	72	55	83	78	70	80	76	61
pH									
Influent	7.3	7.3	7.2	7.3	7.3	7.3	7.3	7.1	7.2
Effluent	7.7	7.6	7.4	7.6	7.6	7.5	7.6	7.5	7.5
Suspended Solids, mg/l									
Influent	91	80	45	74	63	80	86	76	51
Effluent	11	9	18	8	5	12	10	8	15
% reduction	88	89	60	89	92	85	88	89	70
Biomass VSS, g/m^2	19.02	28.16	26.68	20.80	23.36	17.92	16.00	23.00	18.00

* Average Liquid temperature was 28°C.

\+ Municipal Sewage

** Number of Discs 6,4,1, for items 1, 2 and 3 in each column.

TABLE - 3 : WEIGHT OF BIOMASS DEVELOPED ON THE DISCS AT DIFFERENT DISC SPEEDS AND LOADS (USING SETTLED DOMESTIC SEWAGE) - LAB MODEL.

Disc Speed RPM	Organic Loading Rate, $g/m^2/day$	Biomass* VSS, g/m^2 of disc area
3	6.7	19.20
	11.7	28.16
5	6.3	20.80
	10.0	23.36
8	6.7	16.00
	12.2	23.00

* A month after start of operation.

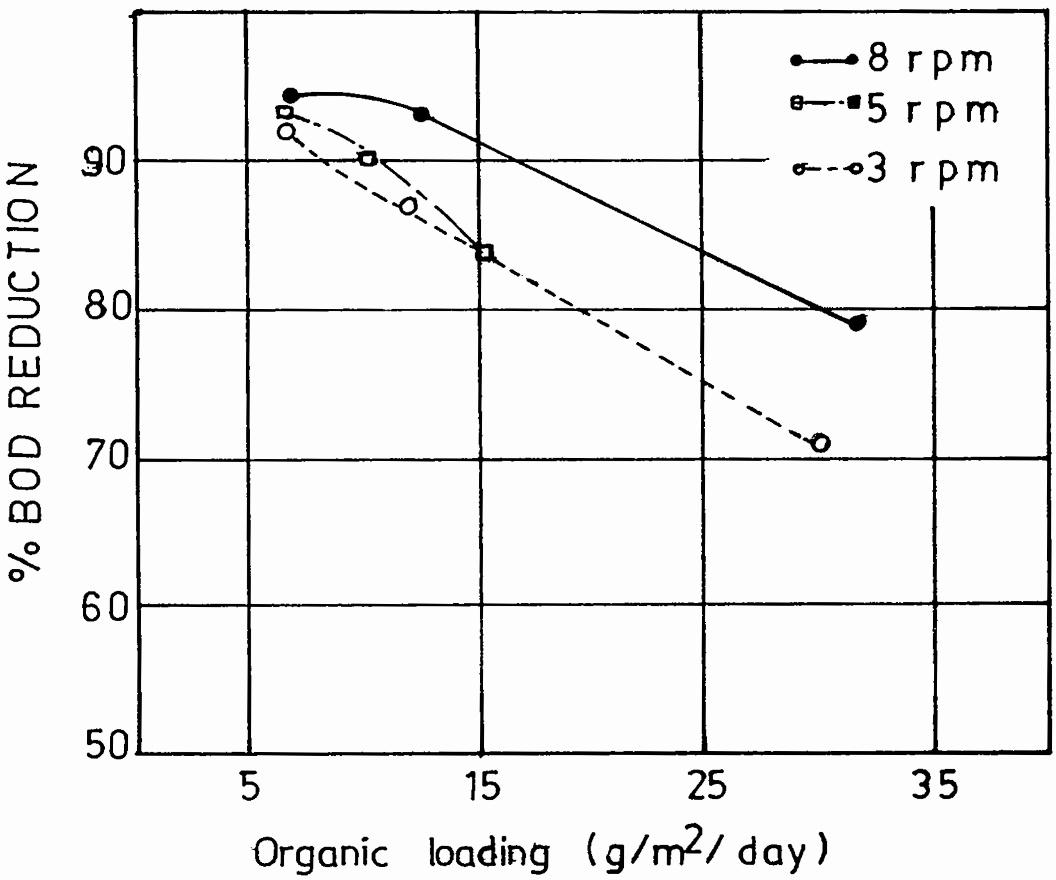


FIG. 5 ORGANIC LOADING VS. PERCENTAGE BOD REDUCTION AT 3,5 AND 8 RPM

TABLE - 4: COMPARATIVE STUDIES USING SETTLED SEWAGE WITH IMHOFF TYPE AND SEMI CIRCULAR CHAMBERS WITH SEPARATE SETTLING TANK (OCT. 1975 TO MARCH 1976) AVERAGE VALUES+.

PARAMETERS	EXPERIMENT No.	
	1*	2**
No. of Discs	2	2
Surface area, m^2	0.13	0.13
Chamber(s) capacity(litres)	2.05	2.18
Hydraulic loading rate, m^3/m^2/day	0.115	0.115
Detentime time, (hrs.)	3.3	3.4
Organic Loading rate, g/m^2	14.9	14.9
B.O.D. mg/l		
Influent	130	130
Effluent	31	22
% reduction	76	83
C.O.D. mg/l		
Influent	307	307
Effluent	87	70
% reduction	71	76
Effluent Suspended Solids mg/l	15	10
Biomass, g/m^2	62.8	76.8

* Imhoff type of Tank with Discs.

** Semicircular tank with Discs followed by a semicircular settling tank.(Vol. 0.950 + 1.230 = 2.180 litres)

\+ Average value of 21 observations.

\- Disc speed, 5 RPM

\- Average liquid temperature, 24°C.

\- Total flow to the unit for 24 hours, 15 litres.

while for the composite Imhoff tank type, the efficiency of BOD_5 removal was 76%, under identical conditions. The removal of the S.S. in the former type was 83% while for the composite type, was 76%.

B. Pilot Plant Studies

i) Pilot Plant with Asbestos Cement Sheets Using Settled Domestic Sewage.

The pilot plant composite RBC reactor (with settling tank below) was operated at a rotational speed of 5 rpm for feed rate of 47.67 m^3/day of settled domestic sewage. While the RBC reactor was operating thoughtout the day, the sewage was fed only for 10 to 12 hours continuously during the day. An overall BOD_5 reduction of 77 percent was observed, while there was 50 percent removal of ammonia nitrogen and 30% removal of Phosphate (vide Table 5). The studies had to be discontinued after three months of continuous operation due to structural defects noticed in the disc assembly.

ii) Pilot Plant with PVC circular Disc Using Raw Domestic Sewage.

The raw domestic municipal sewage pumped from a sump receiving sewage from a municipal manhole was fed by gravity to pilot RBC plant with PVC discs (vide Fig. 3 and 4) through a distribution chamber. The RBC reactor disc were operated continuously throughout the day, while the flow of sewage was restricted to 8 to 10 hours during the day time, due to operational difficulties. For a period of 2 years, the RBC reactor was operated by keeping it open to the atmosphereand at a constant speed of 5 rpm. The speed of 5 rpm was selected based on laboratory studies and with reference to reduced power consumption.

The PVC discs were initially found to be smooth and the biological film adhering to the disc surface was not apparently 'thick'. After 3 months of operation with smooth discs, the disc surface was showered and coated with fine sand which was fixed to the discs by 'Wavin PVC Cement'. Table 6 give the performance data of the pilot plant studies for varying conditions viz. open reactor tank with smooth PVC disc and roughened with coated sand, closed reactor tank with modifications to the outlet of settling tank. The hydraulic flow rate ranged from 4.54 to 5.77 m^3/day corresponding to an organic loading rate of 15.2 to 27.7 g/m^2 of disc area per day. The BOD_5 of influent was almost consistent during the period of the day and as such samples were collected at a time 2 to 3 hours after the pumping of sewage started. The effective detention time for the stated ranges of flow in the RBC reactor tank was 1.6 to 1.26 hours, while for the secondary settling tank it was 1.1 to 0.9 hours. The influent BOD during the period of the studies varied from 218 mg/l to 308 mg/l.

TABLE -5: SHOWING THE DATA OBTAINED FROM RBC PILOT PLANT STUDIES FOR THE TREATMENT OF SETTLED DOMESTIC SEWAGE USING DISCS MADE UP OF ASBESTOS CEMENT SHEETS.

PARAMETERS	AVERAGE VALUES AND RANGE
Temperature ^0C.	(20-30)
Total Flow (10-12 hrs/day), m^3	47.67
Hydraulic Load, $m^3/m^2/day$	0.143
Organic Load, $g/m^2/day$	16.1
<u>B.O.D. mg/l</u> :	
Influent	114 (86-135)
Effluent	26 (15-45)
% reduction	77
<u>C.O.D. mg/l</u> :	
Influent	280 (210-360)
Effluent	54 (30-90)
% reduction	80
<u>NH_3-N, mg/l</u> :	
Influent	17.0 (13.6-19.0)
Effluent	8.5 (6.5-10.4)
% reduction	50
<u>PO_4, mg/l</u> :	
Influent	11.5 (8.5-14.0)
Effluent	8.0 (6.2-11.0)
% reduction	30

* The table gives the average values of 25 observations during the operation of RBC from January, 1974 to April, 1974.

* No. of Discs mounted on two shafts : 136.

*Surface Area of each disc (on both sides) : 2.45 m^2

* Total surface area of 136 discs : 333.2 m^2.

* Disc speed, 5 RPM.

TABLE-6: DATA ON THE PERFORMANCE OF RBC PILOT PLANT WITH PVC DISCS FOR THE TREATMENT OF RAW DOMESTIC SEWAGE.

PARAMETERS		PERIOD OF OPERATION				
		Jan. '76 to Apr. '76	Oct. '76 to Feb. '77	Feb. '77 to June '77	Aug. '78 to May '79	June '79 to Aug. '79
		1*	2**	3**	4**	5***
RBC tank Temp. °C.		24-27	20-28	26-31	29-35	29-34
pH of liquid in RBC tank		7.4-7.5	7.5-7.8	7.6-7.7	7.5-7.9	7.6-7.9
Total flow/day (8-12 hrs/day) (m^3)		5.77	5.58	4.49	4.54	4.99
Hydraulic loading rate, $m^3/m^2/day$		0.090	0.087	0.070	0.071	0.078
Organic loading rate, $g/m^2/day$		27.7	25.6	22.3	15.2	20.7
BOD, mg/l	Inf.	308	295	319	216	255
	E_1	98(68%)	32(89%)	75(77%)	48(78%)	86(66%)
	E_2	70(77%)	24(91%)	55(83%)	40(82%)	45(82%)
COD, mg/l	Inf.	573	588	551	600	662
	E_1	217(58%)	105(82%)	127(77%)	134(78%)	199(70%)
	E_2	132(73%)	87(85%)	90(84%)	98(84%)	75(88%)
Suspended solids mg/l	Inf.	109	405	355	314	466
	E_1	54(50%)	44(89%)	38(89%)	68(78%)	152(59%)
	E_2	36(67%)	24(94%)	26(93%)	22(90%)	44(90%)
NH_3-N, mg/l	Inf.	-	37	38	40	32
	E_1	-	22(41%)	30	23(43%)	16(50%)
	E_2	-	15(60%)	26	18(55%)	16
NO_3-N, mg/l	Inf.	-	Nil	Nil	Nil	Nil
	E_1	-	1.0	2.0	1.2	2.1
	E_2	-	1.3	2.3	1.3	4.2
Bio-mass on Disc, g/m^2		27.7	34.7	50.0	78.3	63.0

No. of Discs of 1m. dia : 40; Total surface Area : 64.138 m^2
Size of semi-circular contact tank: 1.22m dia, 1.52m length, Depth : 0.72 m^3
Size of Settling tank: 1.22m x 0.91m x 0.508m; volume: 0.50m^3
Inf: Raw Domestic Sewage Influent
E_1 : Effluent from RBC Contact tank
E_2 : Final Effluent after settling tank : (%) Percentage Efficiency
*: The RBC plant was operated with smooth PVC discs, without cover.
***: The RBC plant was operated with ventilated cover and modified settling tank, using sand coated discs(column 5).
(The table gives the average values of 10-12 observations).

**: The RBC plant was operated with sand coated PVC discs without cover.

There was gradual reduction of efficiencies of removal of BOD_5 and S.S as the organic feed loading rate was increased from 15g/m^2 of disc area/day to 28g/m^2 of disc area/day. The overall BOD_5 removal varied from 77 to 91 per cent and S.S. removal varied from 67 to 94 percent in the decreasing order of organic loading rate. The removal of BOD_5 in the reactor alone ranged from 68 to 89 per cent. Ammonia nitrogen to an extent of 50 to 60 percent was removed in the system.

During the period of studies at various seasons of the year the temperature of sewage water varied from 20°C to 35°C. The whole plant was worked under natural field conditions without any control over temperature, organic strength etc. Only the hydraulic flow rate was regulated.

As can be seen from Table 5, the efficiency of removal of BOD_5 in the reactor with smooth PVC discs was lower by about 15 to 18 percent compared to the reactor with sand coated PVC discs. Under similar conditions with sand coated PVC discs, the removal of BOD in RBC reactor which was open was higher by about 10 to 15 percent to that of reactor which was closed by ventilated lid. It was to some extent improved with respect to BOD and SS removal after modification of the outlet arrangements using serrated weirs. Settling tank with modified outlet arrangements was being used for the closed reactor, while earlier the same reactor without closed lid was operated along with settling tank with single pipe outlet.

It may be noted that the sewage that was fed to the system, had only preliminary treatment of screening and grit removal and no primary settling. Perhaps, the performance efficiency of the system could be bettered by charging with primary settling sewage. Otherwise, by providing a second stage smaller reactor unit in series with the existing one, the efficiency could possibly be improved.

ORGANISMS IN THE SLIME

After the start up of the plant, about 7 to 10 days of continuous operation was required to build up an optimum growth of biological slime adhering to the disc and to reach a steady state condition producing an effluent of desired quality.

The microscopic examination of slime revealed mixed culture of Protozoans, Rotifers, Nematodes, Filamentous bacteria, Fungi and Algae, while particular species dominated under different conditions of working of the reactor. Identified organisms are shown below

PROTOZOA

Paramaecium cdudatum
Vorticella sp.
Epistylis sp.
Ameoba proteus
Glaucoma sp.
Aspidisca costata
Carchesium sp.
Opercularia sp.
Lionotus sp.
Podophyra sp.
Chilodonella uncinata
Colpoda sp.
Euplotus sp.

ROTIFERS

Rotaria rotatoria
Lecane sp.
Philodina sp.

NEMATODES

Rhabditis Larvae
Doriolamus sp.

ALGAE

Oscillatoria sp.
Spirulina sp.
Phormidrum sp.
Chlamydomonas sp.
Selenastrum sp.
Chlorella sp.
Actinosphaerium sp.
Anacystis sp.
Synedra sp.
Denticula sp.
Diatoma sp.
Tabellaria sp.
Nitzschia sp.
Navicula sp.

FUNGI

Fusarium sp.

FILAMENTOUS BACTERIA

Sphaerotilus natans

Laboratory studies with synthetic sewage showed abundance of filamentous bacteria, Sphaerotilus natans. When the pilot plant reactor was operated with raw sewage under ventilated cover, the biological organisms present in the slime adhering to the disc showed abundance of filamentous fungus, Fusariuem sp. and a small proportion of algae. Under conditions of pilot plant reactor completely exposed to the atmosphere (without the ventilated cover) and operating with raw sewage, algae was also found in abundance in the slime.

The average concentration of bio mass present on the discs of the pilot plant varied from 35 to 78 g/m^2 of disc area. On an average, the total weight of the biomass adhering to the surfaces of 40 discs in the pilot plant, was estimated to be about 4 kgs; while the total suspended solids present in the mixed liquor in the reactor was found to be about 0.25 kg, which works out to about 6 percent of the total biomass present in the discs. The total quantity of sludge accumulated in the settling tank for a fixed period was measured and worked out to about 0.4 gm per gm of BOD applied.

LOADING AND POWER REQUIREMENTS

From the results of the performance of the pilot plant, it is possible to obtain efficiencies of purification ranging from 82 to 90 percent for loading rates of 16 to 20 g/m^2 of disc area per day when screened and degritted raw domestic sewage (pumped from municipal manhole) is applied to the system.

Power consumption as measured by Wattmeter connected to the drive motor of 1.5 kw capacity, worked out to 1 to 1.25 kw/hr for one kg. of BOD removed. It is to be noted that the power rating of the motor was much higher than that required, and as such, the power consumption seemed to be relatively higher than that reported elsewhere, even-though the consumption will only be slightly less than that for conventional activated sludge process. The plant could very well operate with a 0.5 kw motor instead of 1.5 kw motor.

COST ASPECTS

The break up of the cost of different components of the pilot RBC fabricated at NEERI are as follows:

1. Geared Motor	...	Rs. 5,890.00	
2. 40 Nos. circular PVC disc of 1m dia	...	Rs. 4,680.00	
3. Settling tank piping & fabrication	...	Rs. 4,500.00	
		Rs. 15,070.00	
Contingencies		Rs. 930.00	
	Total	Rs. 16,000.00	(as in the year 1976).

The cost of the system is equivalent to about 2000 US dollars, for treating a flow of 0.5 m^3/hour (110 gallons per hour) of raw sewage with BOD of 250 to 300 mg/l and working at an efficiency range of 82 to 90%.

SUMMARY AND CONCLUSION

1. The Rotating Biological Contactor (RBC) or Bio-disc due to its compact construction, simplicity of operation and favourable climatic conditions has great potential for use in the treatment of wastewaters in India. Accordingly, studies were conducted under laboratory and field conditions with the objective of its feasibility for use in India, and formulating relationship amongst the treatment efficiency, organic and hydraulic loadings, rotational speed and power consumption.

2. Performance studies were carried out in a laboratory Rotating Biological Contactor using synthetic and domestic sewage for nearly two years period, for two types namely, composite Imhoff tank type reactor with settling tank below the reactor, and RBC reactor with separate settling tank.

3. Later, studies were carried out using raw degritted and screened municipal sewage, on the performance of RBC pilot plant with 40 PVC circular discs of one meter diameter followed by a rectangular settling tank. The performance characteristics were studied under two conditions viz. when the reactor was open, and when the reactor was enclosed by ventilated cover.

4. The operation of the pilot plant was carried out under field conditions, and there was no control as regards the strength of wastewater and temperature. The RBC was operating continuously, while the sewage flow was limited to 8 to 10 hours during the day time, feeding the reactor at a constant rate of 0.5 m^3/hour (variation ± 10 percent) under ambient conditions and temperature of wastewaters varying from 20°C to 35°C.

5. Under the tropical climatic conditions prevailing at Nagpur, India, the RBC pilot plant achieves overall efficiencies of removal of 82 to 90 percent, for BOD_5 at 20°C, and 90 to 93 percent for Suspended solids at an organic loading rates of 16 to 20 g/m^2 of disc area/day and hydraulic loading rate of 0.07 to 0.08 m^3/m^2/day when the applied feed of raw screened and degritted domestic sewage has BOD concentration of 250 to 300 mg/l, suspended solids concentration of about 400 mg/l and the temperature of sewage in the reactor varied from 20° to 33°C for different seasons of the year. Other design parameters include the submergence of 45 percent for the discs, rotational speed of 3 to 5 rpm, effective detention time in the reactor of 1.5 hours.

6. Such a compact unit occupying an overall area of 1½ Meter by 4½ Meter can as well serve a population of 50 to 100 persons depending on the water consumption. A 0.5 K.W. motor with reduction gear and belt drive would suffice to rotate the discs.

7. The desludging from the settling tank need to be carried out only once in a week or two weeks. The quantity of sludge produced is estimated at 0.4 kg per kg of BOD removed in the system.

8. There were no major mechanical troubles in the RBC Pilot Plant during its continuous operation for two years. The bearings have to be occasionally lubricated or replaced.

9. The cost of the PVC discs (in India) form nearly 35 to 40 percent of the total cost of the RBC system. The cost can be reduced by using alternative cheap materials like split bamboo, aluminium sheet etc. for discs. By using PVC discs, the per capita capital cost works out to Rs. 160 for the unit serving 100 persons.

10. The power consumption varied from 1 to 1.2 KWH/kg BOD removed, which could be brought down by using a lower horse power rating motor drive.

11. Further field studies on pilot plant scale should be continued with alternative materials for disc, feasibility of use of wind mills as drive mechanism for rotation of discs, increasing loading rates, additional surfaces, methods to decrease power consumption, and using the RBC as based on extended aeration principle for reduction of sludge volume and its easy disposal.

12. Open type of RBC gives higher efficiency for BOD removal than that for the enclosed type.

13. The RBC can be successfully operated with screened, degritted sewage, avoiding primary settling tank and the problem of disposal of primary sludge. The efficiency can be improved by working the RBC in stages.

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HIGH SALINITY WASTEWATER TREATMENT
USING
ROTATING BIOLOGICAL CONTACTORS

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Introduction

Design of wastewater treatment facilities for small, offshore communities presents unique engineering problems associated with conventional biological treatment. Islands are often inhabited by seasonal populations isolated from the basic services available on the mainland. Limited freshwater supplies are frequently conserved by using seawater as a carriage medium for sanitary wastes. The salinity of the domestic wastewater can vary from 0% to 3.5% depending on the source of the waste at a given time. To further complicate design problems, systems must be able to adapt to varying flows, little maintenance and intermittent nature of island generated electrical energy.

The inherent flexibility in the operation of the rotating biological contactor (RBC) process makes it an ideal candidate for this set of conditions. As part of an evaluation of the ability of an RBC unit to treat saline domestic wastewater, microbial communities attached to disk surfaces were studied. Disk populations growing in freshwater and seawater based sewage were compared as a function of their distance from the influent end of the unit and the hydraulic loading rate.

Most previous research on saline domestic wastewater treatment has been conducted on the activated sludge process. Using a bench scale unit

Ludzack and Noran (1) found that free swimming ciliates survived in a continuous high chloride artificial waste made from NaCl (3.5% salinity). When the chloride content alternated between high and low concentrations, hypotrichs and stalked ciliates occurred, but the latter were small and inactive. The activated sludge took from one to five weeks to adapt to steady state conditions during saline treatment. Kincannon and Gaudy (2) fed artificial waste of 3.0% salinity to a batch activated sludge system. They found a 30% decrease in substrate removal efficiency. When the salinity increased to 4.5%, operation was impaired. They hypothesized that a change in species composition might occur in saline treatment. In subsequent research, Kincannon and Gaudy (3) used a continuous flow activated sludge unit to assess the treatment capabilities of organisms exposed to an artificial waste containing NaCl (salinity 3.0% and 4.0%). After a period of acclimation of up to two days, the system achieved excellent removal capacity. In another bench scale study of an activated sludge treatment plant, Burnett (4) conducted research using domestic waste mixed with seawater. Acclimation occurred within thirteen days after changing from freshwater to seawater based sewage (salinity 3.2% to 3.8%). The system operated well under these conditions. He noted a rapid die-off of rotifers and stalked ciliates and motile ciliates with increasing salinity. After the first few days of acclimation, motile ciliates reappeared, but the stalked ciliates and rotifers remained absent. Kessick and Manchen (5) experimented with another bench scale activated sludge system and found that freshwater seed bacteria could adequately treat the soluble fraction of the domestic sewage containing artificial sea salts (salinity 3.6%). Recently Tokuz and Eckenfelder (6) observed a similar treatment capacity between a bench scale activated sludge unit fed artificial waste at 0% salinity and one at 5.0% salinity.

Research on the effect of salt concentration on the treatment capacity of trickling filters has been more limited. Stowell (7) reported that an intermediate type high rate filter with a 1:1 recirculation ratio achieved BOD removals up to 90% when treating waste from San Quentin Prison (salinity 1.1% to 1.5%). Lawton and Eggert (8) found that a bench scale trickling filter treating an artificial waste was able to recover in one day after the salinity was increased to 2.0% with NaCl. Mills and Wheatland (9) found no change in the removal efficiency of a bench scale trickling filter after adding NaCl (final salinity 1.2%) to an artificial waste. In this study an intermittent application of waste of salinity 1.2% to 3.6%, did cause a decreased efficiency.

Rotating biological contactors are a relatively new waste treatment system compared with the activated sludge and trickling filter processes. Therefore research into the various applications of the unit has been rapidly expanding. The major work done on RBC saline waste treatment is reported by Mikucki and Poon (10), Poon and Mikucki (11) and Poon, Chao and Mikucki (12). A pilot plant study was conducted using a mixture of domestic waste and an artificial supplement and seawater with a final salinity of 2.1%. This system worked well in removing BOD after a few days of acclimation. Poon, Chao and Mikucki (12) reported that the growth on the disks contained estuarine forms of filamentous fungi and algae.

A review of the literature to date reveals that little information is available on the microbial populations inhabiting saline waste treatment systems. Populations for our evaluations were obtained from an RBC pilot plant fed settled domestic sewage mixed with artificial sea salts to a salinity of 1.0% to 2.0%. A qualitative analysis was conducted using photomicroscopy to assess the difference in community structure between freshwater and salt water based samples. To avoid lowering the BOD and to closely simulate seawater based sewage, a small volume of highly concentrated artificial seawater was added directly to the waste.

The qualitative differences between the populations occurring on the disk surfaces in the freshwater and salt water based sewage appeared minimal. Under both conditions the community structure consisted of a primary substrate composed of filamentous organisms and a zoogleal mass, which supported active populations of rhizopods and stalked ciliates. At an hydraulic loading rate of one gallon per day per square foot (gpd/ft^2) stalked ciliates occurred in all four compartments of the units. At a rate of 2 gpd/ft^2 these organisms did not appear until the second compartment.

Materials and Methods

A pilot plant consisting of four separate rotating biological contactors was assembled in the Durham, New Hampshire sewage pumping station. Each RBC consisted of a tank made from a plexiglass half cylinder four feet long and eight inches in diameter. Each was divided by plexiglass plates into four separate compartments. Wastewater flowed from one compartment to the next over notched weirs in the dividing plates. A horizontal stainless steel shaft supported 64 disks, each with a seven inch diameter, for a total surface area of 34 square feet. Two of the units had disks made of polyethylene, while the other two units had polyurethane sealed masonite disks. The disks in all units were equally spaced, with 16 disks per compartment. All units were rototated at 12 revolutions per minute yielding a peripheral speed of 0.37 feet per second. The temperature in the pump station decreased during the course of the experiments and the wastewater temperature ranged from 21^{o} C to 13.5^{o}C. The units were exposed to a constant low level incandescent light of less than 10 footcandles. The hydraulic loading rate for one unit with each type of disk was 1 gpd/ft^2; total retention time of wastewater in these RBC's was 3.6 hours. The other two units had an hydraulic loading rate of 2 gpd/ft^2 and a retention time of 1.8 hours. All disks had approximately 40% of their surface area submerged in the wastewater at a given time.

Raw sewage was taken from the pump station channel just prior to the bar rack and comminutor. The liquid was filtered through a ½ inch wire mesh and a 1/8 inch wire mesh to remove large particulates. It was then pumped in ½ inch plastic hose to a 30 gallon plastic primary clarifier (retention time 3.5 hours). Sludge from the clarifier was removed every two to three days. The flow of the settled sewage was pumped into four separate lines of ¼ inch polyethylene tubing, each of which supplied an individual constant head feed tank. Influent was delivered directly to the first compartment of each RBC unit by gravity flow through 1/8 inch polyethylene tubing from the overhead feed tank. Flow rate was controlled by stopcocks inserted in the influent lines.

The system described above was modified slightly during salt water operation. Concentrated artificial seawater (salinity 13.3%) was made daily by preparing a saturated solution of Utility Marine Mix[1] and tap water. The salt water flowed by gravity from two 30 gallon reservoirs (plastic) into the lines carrying settled sewage to the constant head feed tanks. Flow rate was regulated by an in-line valve. The salinity of the salt water based sewage ranged from 1.0% to 2.0%.

During the startup period the units received freshwater settled sewage. After approximately two weeks the units reached steady state operation as assessed by BOD sampling. The disk populations were removed and observed as outlined below. The system was then switched to the salt water based waste. After one week steady state was achieved and samples were removed and observed. The system was returned to the freshwater based sewage and the entire procedure repeated.

BOD tests were done on the soluble fraction of the waste to determine influent and effluent quality. Samples were filtered through Whatman 40 filter paper or the equivalent. For freshwater based sewage the procedure outlined in Standard Methods (13) was followed. For salt water based sewage Martin's (14) procedure was used with a dilution water of 0% salinity. Dissolved oxygen was measured with a YSI 51A Oxygen Probe and Meter[2].

Samples for microscopic examination were scraped from the surface of one disk in each compartment of the RBC's. The scrapings weighed 2 to 3 grams. We assumed that each compartment was completely mixed and therefore the microbial population would be fairly uniform regardless of position within the compartment. Samples were immediately transferred to sterilized plastic bags. A few milliliters of wastewater from the sampled compartment was added to the bag. The samples were sealed and stored in a refrigerator in the laboratory at 4.4°C until examined. Observations usually occurred within 24 hours, however some samples were held up to one week with no visible change in composition upon microscopic examination. Random samples taken from each bag were placed on a glass slide and covered with a glass cover slip. In some cases one or two drops of 10% $MgCl_2$ were added to slow the protozoans. All fields on a slide were examined with an Olympus BHA microscope. Observations were recorded in the form of written notes and photomicrographs. Kodak Ektachrome ASA 64 daylight film was used with an LBD and two LD 45 filters for all photomicroscopy. Resources used in identifying organisms included Kudo (15), Bergey's Manual (16) and Barnes (17). Identifications were subsequently confirmed by specialists in protozoology and microbiology.

Results

The soluble influent BOD averaged 150 mg/l. It showed a typical diurnal pattern for a small town; lower concentration in the daytime and higher at night . During steady state operation on freshwater and salt water based sewage the soluble effluent BOD from the RBC's was consistently below 30

[1] Utility Chemical Company, Paterson, New Jersey
[2] YSI Company, Yellow Springs, Colorado

mg/l.

Tables 1 and 2 list the general types of organisms found on the disks during steady state conditions when the feed was freshwater and salt water based sewage, respectively.

The populations present during freshwater and salt water operations were very similar. A filamentous organism grew on all the disks, forming a dense mat varying in color from white to dark brown. Growth was patchy in the last two compartments of the RBC's. The other predominate population was a zoogleal mass of bacteria. Both the filaments and the zoogleal mass were embedded in a thick mucilage. Zooflagelates and rhizopods (amoebae) were also observed in all compartments, under both conditions, as were fungal fruiting bodies and unidentified cysts. The fruiting bodies and cysts were found sporadically. Fewer rotifers and nematodes were observed during the saline conditions.

At an hydraulic loading rate of 1 gpd/ft^2 motile and nonmotile peritrichs were present throughout the entire unit. The nonmotile peritrichs included solitary and colonial stalked forms. At the higher loading rate (2 gpd/ft^2) peritrichs did not appear in the first compartment, but were present in all the remaining compartments.

There was no difference observed between the populations grown on the polyethylene and masonite disks.

Discussion

"There has been, and still is, too much engineering and too little microbiology in this field of environmental sanitation" (18). Fifteen years later this generalization is still applicable. One reason for the lack of microbiological research is the difficulty encountered in isolating and identifying specific organisms occurring in wastewater treatment systems. Though *Sphaerotilus* (19, 20), *Beggiatoa* (20), and fungi (12) have all been reported as major components of RBC microbial communities, there have been no reported instances of successful culturing of these organisms from the disks. Identification of genera by microscopic examination is not a foolproof microbiological method. Therefore we have not attempted to specify the type of filament growing on the disks in our experiments. Research is currently underway in our laboratory to isolate and culture the filaments and make a positive identification of the species.

The identification of zoogleal bacteria and specifically *Zooglea ramigera* is difficult, though it has been reported growing in RBC's (19). Crabtree and McCoy (21) conclude that *Z. ramigera* may consist of a heterogenous population of bacteria. This view is supported by Unz and Dondero (22), who report that two forms of zoogleal mass exist. The most predominant and "biochemically active" form is composed of nonzoogleal bacteria (22). In our research we have not attempted to identify the species of bacteria present in the zoogleal mass. We can only confirm that a zoogleal type of growth occurs on the disks.

We know relatively little about the types of organisms growing on the

Table 1. Organisms Present in RBC's Treating Freshwater Based Sewage

Hydraulic Loading Rate gpd/ft^2	Compartment 1	Compartment 2	Compartment 3	Compartment 4
1.0	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers Plants Oomycetes	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers Plants Phycomycetes
2.0	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Metazoans Nematodes	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers Unidentified cysts	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Unidentified cysts

Table 2. Organisms Present in RBC's Treating Salt Water Based Sewage

Hydraulic Loading Rate gpd/ft^2	Compartment 1	Compartment 2	Compartment 3	Compartment 4
1.0	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers Plants Phycomycetes Unidentified cysts	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers Unidentified cysts
2.0	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers Unidentified cysts	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Metazoans Nematodes Rotifers Unidentified cysts	Monerans Filaments Zoogleal mass Animals Protozoans Zooflagellates Rhizopods Holotrichs Peritrichs Unidentified cysts

RBC disks, and their community interactions. The disks provide a surface to which filamentous organisms adhere. It appears from our observations that the filaments provide a primary substrate for all the other members of the community including nonfilamentous bacteria, protozoans, rotifers and nematodes. The primary substrate serves two functions. 1. As a habitat, the filaments provide a refuge for bacteria from large celled predators (24) as well as a substrate to protect the other microorganisms from the shearing force of the fluid as the disks are rotated. 2. The filaments also are a food source for some of the protozoans (25). The relationships between filamentous forms and their associated inhabitants are noted by several authors (25, 26, 27).

The protozoans may play several roles in the community structure. We have adapted the theories developed by Reid (28) to explain protozoan activity on the RBC disks. Protozoans are mainly carnivorous. As primary carnivores they prey on free bacteria, helping to maintain bacterial activity by controlling the size of the bacterial population. As secondary carnivores, some protozoans eat the primary carnivores preventing excess predation. Protozoans also assimilate bacterial metabolic by-products, contributing to removal efficiency. This simple model of the community structure of the disk growth provides a basis for future research on RBC microbial ecology.

In our observations we found little difference in the organisms present during freshwater and salt water based sewage treatment. Only the rotifers and nematodes appeared to be present in reduced numbers under saline conditions. In two previous bench scale activated sludge experiments on saline sewage, rotifers were absent (4), and stalked ciliates were absent (4) or reduced in size and inactive (1). We are unable to explain the reasons for the inhibitions of nematodes and rotifers and the ability of stalked ciliates to survive under saline conditions in our RBC experiments. It is possible that salinity tolerance is species specific (25) and/or that some inherent characteristic of the RBC process may mitigate the inhibition of the stalked ciliates. Further research should be conducted in these areas.

The bactericidal action of the seawater on non-marine species is well known (29, 30, 31). In previous experiments on saline biological waste treatment it has been assumed that marine and estuarine organisms would predominate due to the die-off of freshwater forms (4,12, 32). Many factors contribute to the die-off of non-marine microorganisms in seawater. The two major factors, Jones (33) concludes, are the low nutrient concentrations in seawater, and heavy metal toxicity. In continuous culture in seawater, E. coli can outcompete marine bacteria if substrate concentrations are similar to those in sewage (34). In saline sewage the negative effects of seawater are mitigated, as the nutrient concentration is high and the heavy metals are complexed by organics (33). We observed no change in species composition under saline conditions. We think that the freshwater sewage bacteria are able to survive because the environment they "perceive" is not inhibitory. We are presently conducting further research on heavy metal complexation in saline sewage.

The existence of active peritrich populations on our disks indicates that good treatment was occurring (28, 35). This is supported by the low soluble BOD's we measured (less than 30 mg/l). The organisms we observed were similar to those found by other researchers in activated sludge (36, 37), trickling filters (38, 39) and RBC's (19, 20) which also exhibited good BOD removals. Our microbial populations did not show a distinct succession along the length of the unit, which had been observed by others (19, 20). There may be a change in the bacterial species which we could not determine in these experiments. Identification and quantification of organisms and their position along the unit should be examined as it affected by RBC operating parameters.

Our findings confirm those of Poon and Mikucki (11) that RBC's can achieve BOD removal of greater than 30 mg/l when treating saline domestic wastewater. RBC's do offer a possible treatment alternative for small offshore communities which use seawater as a carriage medium for sanitary wastes. We found that microbial populations on the disks during the treatment of salt water based sewage were similar to their freshwater counterparts. Under both conditions peritrichs were present, indicating a healthy and active microbial growth. We developed a simple explanation of the interrelationships among disk microorganisms. Many questions remain unanswered and we hope that continued research on the microbial populations and their ecology inRBC's will help provide the solutions.

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FULL-SCALE ROTATING BIOLOGICAL CONTACTOR FOR SECONDARY TREATMENT AND NITRIFICATION

BY

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INTRODUCTION

A full-scale RBC system, designed to achieve effluent limitations of 10 mg/l BOD_5 and 2 mg/l NH_3-N for domestic wastewater, performed at less than design expectations under both summer and winter conditions. This RBC system constitutes the biological treatment portion of a 6 MGD wastewater treatment plant serving a major US Army installation (effective population of 40,000) and consists of 36 stages arranged with 6 treatment trains of 6 stages each. The design hydraulic loading at 6 MGD is 1.33 gpd/sq ft of media surface. During the summer (wastewater temperature of 26^{o}C and average plant flow of 3.7 MGD), the treatment plant effluent BOD_5 and NH_3-N levels were higher than NPDES permit limitations and design expectations. High levels of effluent BOD_5 resulted primarily from oxygen demand of suspended solids and nitrification in the BOD bottle. In fact, effluent soluble-BOD_5 was consistently measured at less than 5 mg/l. High levels of effluent NH_3-N resulted from DO limiting conditions (less than 1 mg/l) in several RBC stages and from relatively low pH (less than pH 7.0) in latter RBC stages. During the winter (wastewater temperature of 13^{o}C and average plant flow of 4.6 MGD), RBC performance actually improved; DO limiting conditions did not exist. The RBC system effectively removed soluble organic material during the winter and was more effective in removing NH_3-N than during the summer. NH_3-N levels of design expectation were, however, still not met. Approximately 370 lbs/day of

NH_3-N were removed during the summer evaluation period while approximately 500 lbs/day of NH_3-N were removed during the winter -- a 36 percent improvement. This improved performance occurred during the winter even though hydraulic and organic loads were higher and biological activity was considered relatively low due to lower wastewater temperature.

Conclusions from these studies were that low DO levels in initial RBC stages and low pH levels in latter RBC stages adversely affected biological activity. The importance of evaluating performance in each stage of an RBC system was shown necessary to effectively judge design criteria and pinpoint operational problems. In addition, both soluble (filtered) and carbonaceous (nitrification suppressed) BOD_5 should be used during evaluations, particularly where systems are designed for nitrification. Finally, the dependence of nitrification on prior soluble-BOD_5 removal was highly evident. Recommendations for future RBC system designs include use of supplemental aeration to overcome limiting DO levels and chemical feed to maintain optimal pH levels.

BACKGROUND

The 6 MGD, domestic wastewater treatment plant had been upgraded from a trickling filter system for secondary treatment to a rotating biological contactor (RBC) system for secondary treatment and nitrification. Effluent permit parameters and limitations for the plant discharge are listed in Table 1. The upgraded treatment plant flow diagram and RBC system are shown in Figures 1 and 2, respectively. As indicated, the upgraded plant consists of a bar screen, Parshall flume, comminutor, aerated grit chamber, primary clarifiers, RBC system w/pump station, secondary clarifiers, chlorine contact chambers and a step aerator. Two anaerobic digesters (high-rate and secondary), together with a vacuum filter and sludge drying beds, are used for sludge handling and disposal. Secondary clarifier sludge (recirculated flow), digester supernatant and vacuum filter filtrate and washings are all returned to the head of the plant. The plant has a maximum hydraulic capacity of 18 MGD. The RBC system consists of 36 stages arranged in a configuration of 6 treatment trains with 6 stages each; 3 of regular density and 3 of high density media (see Figure 2). Although each unit operation of the treatment plant was evaluated during both summer and winter conditions, this paper addresses only the RBC system.

LITERATURE REVIEW

RBC Treatment Process

The RBC process consists of a series of plastic disks of which 40 percent of the surface area is immersed in wastewater (see Figure 3). As the disks rotate, the entire media surface develops a culture of microbiological organisms. The organisms adhere and multiply to form a uniform growth referred to as a fixed-film. The biomass supported by the plastic media picks up a thin layer of nutrient laden water as it rotates through the wastewater. The film of water trickles over the microorganisms which remove dissolved organics and oxygen. The rotation of the media through the wastewater not only allows for aeration and mixed liquor, but also provides shear forces which cause sloughing of excess growth.

RBC units commonly operate in series with the number of units depending on the organic and/or hydraulic load to be treated. The function of the first stages is to remove organic material, with subsequent stages removing ammonia in cases where nitrification is used to meet effluent NH_3-N standards. Nitrification usually does not begin until the soluble-BOD_5 level and corresponding large populations of heterotrophic organisms have been adequately lowered. The actual reason that heterotrophs and autotrophic nitrifiers do not co-exist in equal quantities throughout successive RBC stages is not clearly understood, but it is reported by some that the activities of the two populations do not occur simultaneously.[1,2,3] However, others report that optimal pH levels for nitrification (e.g., pH 7.0 -8.5) appear to favor initiation of nitrification simultaneously with low-level soluble-BOD_5 removal; whereas, nitrification at suboptimal pH levels (e.g., pH 6.6-7.0) is not initiated until soluble-BOD_5 removal is complete.[4,5] The amount of nitrification achieved has been correlated to the hydraulic loading of the system, usually expressed as the volume of wastewater applied to a square measure of surface area per day. One to 4 gpd/sq ft have often been used as standard loading rates for pilot plants[5] and full scale wastewater treatment facilities.[2,6,7]

The change in hydraulic load also changes the organic load as more food is introduced to the active component of the waste treatment system. It has recently been suggested that shortcomings observed in the quality of treatment by RBC units was due to excessive organic loading, while operating at less than hydraulic design capacity.[2] The question of which parameter, hydraulic loading or organic loading, to use for proper design and operation of an RBC process has not been resolved.

As with other biological processes, sufficient dissolved oxygen (DO) must be available in the wastewater within the RBC system to insure adequate treatment for BOD_5 removal and nitrification. Wastewater DO levels of 1 to 2 mg/l are generally considered to be the minimum requirement to avoid DO limiting conditions. Frequently, RBC systems have been designed to provide oxygen mass transfer via disk rotation through the wastewater and air. However, in some cases, this has been considered a shortcoming of the process since supplemental oxygen must sometimes be provided to prevent DO limiting conditions.[2]

Oxygen Demand of Wastewater

A major criterion used to determine the extent of pollution of receiving waters is the measurement of oxygen required for the stabilization of organic matter present in the system. The total amount of oxygen necessary to stabilize a waste is referred to as the oxygen demand. The ultimate oxygen demand includes not only the amount of oxygen required to stabilize oxidizable carbonaceous materials, but also that which is required to microbially transform ammonia-nitrogen to nitrate-nitrogen. For untreated domestic sewage there is little oxygen demand by nitrifier populations for the first 8 days of stabilization.[8] Therefore, the BOD_5 test is normally considered as representing the oxygen damand of carbonaceous material. However, total BOD_5 is a poor indication of treatment where a significant population of nitrifying bacteria are present. For sewage that has received secondary treatment and nitrification, conversion of ammonia to nitrate in the BOD bottle may significantly increase the BOD_5 measurement and erroneously indicate a lesser degree of treatment than that actually received.

In the RBC system, as other biological treatment processes, nitrifying organisms may be interspersed among the heterotropic population which utilize carbonaceous materials. The relative concentrations of both populations, at any specific point in the treatment train, are a consequence of the nutrient supply, environmental conditions (pH and temperature), and the degree of treatment received. Therefore, to adequately assess treatment performance, it is necessary to know how much of the observed oxygen demand in the BOD_5 test was required to stabilize the carbonaceous materials. The presence of autotrophic bacteria complicates the BOD_5 measurement at the end of the treatment train and gives "false positives" when testing for regulatory compliance.

The purpose of biological treatment relative to carbonaceous material is the conversion of soluble organics to particulate bacteria.[9] However, the unfiltered BOD_5 test represents a measure of soluble as well as insoluble organic matter and NH_3-N oxidation. Biological treatment need not be applied to removal of colloidal and suspended organics. Suspended solids that contribute to oxygen demand can be removed by physical-chemical processes such as gravity settling and filtration. The practical consequence is that optimal treatment for removal of oxygen demand may be removal of suspended solids and not biological treatment. The use of filtered and unfiltered BOD_5 tests should indicate relative fractions of oxygen demand as originating from soluble or particulate material. In addition, the filtered BOD_5 test should not undergo nitrification, because initial nitrifying populations in the BOD bottle would be reduced to insignificant levels and the BOD bottle is subsequently seeded with raw sewage (i.e., heterotrophic bacteria). Thus, the unfiltered BOD_5 test is an unreliable parameter from which to judge biological treatment performance.

Nitrogen Control

The principle of biologically induced nitrogen removal in wastewater treatment facilities is wholly based on the activity of populations of autotrophic nitrifying and denitrifying bacteria and their capability to sequentially oxidize and reduce nitrogen from ammonia to nitrate to nitrogen gas. Nitrification is the oxidation of NH_3-N to nitrate, and denitrification is the reduction of nitrate to nitrogen gas. Different types of microorganisms are required for each action. The extent of their use in wastewater treatment depends upon the end objective. Nitrification is used to control wastewater effluent levels of ammonia, but both nitrification and denitrification must be used to control total nitrogen levels in wastewater effluents. Although process technology for ammonia-nitrogen removal includes breakpoint chlorination, ammonia stripping, ion exchange, and nitrification/denitrification, this paper deals only with nitrification.

In addition to nitrification/denitrification, microorganisms other than the nitrifiers and denitrifiers require nitrogen for growth. The amount of nitrogen assimilated during oxidation of carbonaceous material has been generally placed at 5 percent of the oxygen demand (i.e., BOD to N = 20 to 1).[10] The consequence is two fold: (1) nitrogen must be present for biological oxidation of carbonaceous material, and (2) removal of ammonia-nitrogen during biological treatment of wastewaters may be due to assimilation, not necessarily due to nitrification.

The importance of nitrogen control in wastewater effluents is its impact on receiving waters. As ammonia becomes oxidized to nitrate, the dissolved oxygen level of water is decreased. Ammonia-nitrogen at concentrations of 0.25 to 0.30 mg/l are lethal to fish within 14 to 21 days.[11] Nitrate is readily available for assimilation by plant life, causing algal blooms when present in too large a quantity.[8] Also, nitrate can cause methemoglobinemia in infants when contaminated water is used as a drinking water supply.[12]

Nitrification

The two microbial genera usually associated with nitrification are Nitrosomonas and Nitrobacter. Both genera of organisms are autotrophic nitrifying bacteria indicating that energy for growth is derived from the oxidation of inorganic nitrogen. The oxidation of ammonia to nitrate is a two step process requiring both organisms for the conversion. Nitrosomonas transforms ammonia to nitrite while Nitrobacter further oxidizes nitrite to nitrate. The overall oxidation of ammonia by these organisms is given by the following equation:

$$NH^{+}_{4} + 2O_2 + 2\ HCO^{-}_{3} \xrightarrow[\text{Nitrobacter}]{\text{Nitrosomonas}} NO^{-}_{3} + 2\ H_2CO_3 + H_2O \qquad (1)$$

As ammonia is oxidized, carbonate is utilized, As nitrate formation occurs, carbonic acid is produced. This microbiologically induced change in the carbonate buffering system results in the destruction of alkalinity at a rate of 7.1 mg (as $CaCO_3$) per mg of ammonia oxidized.[4] As the nitrification process reduces the alkalinity and increases the carbonic acid concentration, the pH of the wastewater may drop as low as pH 6.0, and adversely impact the rate of nitrification. This decrease in pH can be minimized by aeration to strip CO_2 from the wastewater, or by insuring the presence of excess alkalinity.[4]

Primary environmental conditions for optimal rates of nitrification are pH and temperature. The reported pH optima cover a wide range, but the consensus is that as the pH decreases, the rate of nitrification declines. Sawyer, et al,[13] and Engel and Alexander[14] have reported pH optima for nitrification between 8.0 and 9.0, and 7.0 and 9.0, respectively. Painter[15] has stated that nitrification processes cease at or below pH 6.3 to 6.7. Poduska and Andrews[16] have shown that abrupt changes in pH from 7.2 to 5.8 markedly reduced the ammonia oxidation by nitrifiers while the reversal in pH restored the original nitrification rate.

Temperature optima for nitrification are generally reported by various authors at about 30°C with a range of 28-35°C.[17,18,19,20,21,22] Temperature influences heterotrophic and autotrophic microorganisms, thereby affecting secondary treatment and nitrification efficiencies. The nitrification rate is more temperature sensitive than the rates for organic removal.[9] Nitrification rates decrease about 50 percent for each 10°C drop in wastewater temperature below about 30°C.[4] For example, the nitrification rate at 10°C would be about half that of 20°C. Secondary treatment efficiency is less likely to be affected by temperature changes, probably due to microbial population diversity and other system constraints. Organic removal rates for fixed-film processes should decrease about 25 percent for each 10°C drop in wastewater temperature below about 30°C. For example, the rate of biological activity in a trickling filter process, at 10°C would be about 75 percent of that of 20°C. However, the actual temperature effect on a biological process is probably characteristic only of that system.

MATERIALS AND METHODS

RBC Process

The RBC system evaluated was designed to remove BOD_5 (secondary treatment) and NH_3-N (nitrification) to 10 mg/l and 2 mg/l, respectively, at a 6 MGD design flow during both summer and winter conditions. These design parameters were chosen on the basis of NPDES permit limitations in effect during the design phase of the wastewater treatment plant upgrade program. (The seasonal variance of the NPDES permit, as shown in Table 1, became effective during construction of the upgraded facility and, essentially, relaxed the requirements for winter operation.) The system, shown in Figure 2, consists of six trains of RBC's with six stages per train (i.e., a total of 36 RBC units). The first three stages were designed primarily for BOD_5 removal and consist of 18 standard shafts of 100,000 sq ft of surface area each. The last three stages were primarily designed for NH_3-N removal and consist of 18 high-density media shafts of 150,000 sq ft of surface area each. At a 6 MGD design flow, the overall hydraulic loading of the system is 1.33 gpd/sq ft. The RBC units were manufactured by Autotrol Corporation (Bio-Surf process). The BOD_5 removal portion of the RBC system was designed based on hydraulic loading (gpd/sq ft) versus BOD_5 removal (percent) curves. A BOD_5 influent concentration (based on total or unfiltered BOD_5) to the RBC system of 140 mg/l was used for the design. The NH_3-N removal portion of the system was designed using specific removal rates of NH_3-N concentration and was based on an influent NH_3-N concentration to the nitrification phase of the RBC system of 15.8 mg/l NH_3-N. Based on a dye study, the hydraulic detention time of the RBC system was 2 hours and 30 minutes at a flow rate of 5.5 MGD (influent plus recirculated and sidestream flows).

Sampling and Analyses

Treatment train No. 4 (see Figure 2) was used as the primary train to evaluate performance of the overall RBC process. During both summer and winter studies, grab samples of the RBC influent and wastewater in each of the 6 stages of train No. 4 were collected at various times during the studies to determine changes in wastewater characteristics through the system. Temperature and DO data were taken at each sample period using a YSI, Model 57 DO meter. Twenty-four hour, flow proportioned composite samples were also collected at the RBC system influent, effluent and the treatment plant effluent. Sample point locations are shown in Figures 1 and 2.

Twenty-day BOD versus time curves for the STP effluent were developed from 24-hour flow-proportioned composite samples. BOD values were measured every day for the first 10 days and every other day, thereafter. Nitrification was suppressed by the addition of ammonium chloride in order to determine the BOD exerted by carbonaceous and nitrogenous substances.[23]

All analytical chemistry procedures were conducted by the Environmental Chemistry Division of the US Army Environmental Hygiene Agency. A mobile laboratory was set up at the installation for performance of the requisite laboratory work. Sampling and analyses were conducted in accordance with Standard Methods for the Examination of Water and Wastewater, 14th Edition[24] or Methods for Chemical Analysis of Water and Wastes.[25] Tests for soluble-BOD_5 and TOC were conducted on the filtrate of samples filtered through a 0.45 μm filter.

RESULTS AND DISCUSSION

The RBC system influent and effluent characteristics and critical wastewater treatment plant effluent values are summarized in Tables 2 and 3 for both summer and winter studies.

During the summer study, the mean treatment plant effluent BOD_5 level was 11 mg/l, one mg/l higher than the monthly permit limitation. To reduce the BOD_5 level below 10 mg/l did not necessarily mean that additional RBC surface area was required. In fact, there was more than sufficient carbonaceous surface area available for BOD_5 removal. Detailed analysis of the RBC system wastewater characteristics and an evaluation of the BOD_5 test procedure confirmed this. As shown in Figure 4, soluble BOD_5 and TOC values were reduced to essentially constant values within the initial three RBC stages. Hence, the other three stages were available as extra capacity for soluble-BOD_5 removal and nitrification. The RBC system effluent soluble-BOD_5 averaged less than 5 mg/l. Detailed evaluation of soluble-BOD_5 removal through the RBC system can be obtained from Table 5. For example, 9 mg/l (24-15 mg/l) of soluble BOD_5 were removed from stage 1 at an average flow rate of 4.5 MGD.

As shown in Figure 4, wastewater DO concentrations were about 1 mg/l or less in stages 1 through 4. Hence, the oxidation rate of organics was limited by DO level. (The DO levels observed were that of the bulk liquid. Since most oxidation of organics takes place at the media interface, the DO level at the interface is believed to be much less than observed values; therefore, the media interface DO level would limit oxidation rates.) Another factor indicating that DO limiting conditions existed was the nature of the biomass growing on the first two stages of the RBC trains. A white biomass, indicative of the autotrophic, sulfur bacteria, _Beggiatoa_, was predominant. _Beggiatoa_ utilize hydrogen sulfide and sulfur as energy sources in the presence of oxygen, according to the following equations:

$$2H_2S + O_2 = S_2 + 2H_2O \tag{2}$$

$$S_2 + 3O_2 + 2H_2O = 2H_2SO_4 \tag{3}$$

Although _Beggiatoa_ exist under aerobic conditions, anaerobic conditions must be present for the formation of hydrogen sulfide and sulfur. Because of the low DO conditions of the bulk liquid, oxygen transfer to the fixed film was severely limited. Therefore, the bacteria within the film interface were most likely anaerobic leaving only the bacteria on the outside surface of the media interface aerobic. As such, the anaerobic bacteria produced the hydrogen sulfide and provided an energy source for the _Beggiatoa_ to thrive at the media interface. When plenty of hydrogen sulfide is present, _Beggiatoa_ store the sulfur in the cells, giving the organism the distinctive milky appearance as was seen on the first two RBC stages (see Plate 1).*[26] In later stages, where the oxygen level of the bulk liquid increased, the _Beggiatoa_ became visually non-existent (see Plate 2).* The predominance of _Beggiatoa_ noted, not only aggravates soluble-BOD_5 removal; but, because of sulfuric acid formation, the pH dependent nitrifying bacteria (optimum pH 8.0 - 8.5) are aggravated as well.

The RBC system analysis showed that very little soluble BOD_5 remained after treatment. This obviously indicates that other factors contributed to the treatment plant effluent BOD_5. In fact, both suspended solids and ammonia were found to exert a five-day oxygen demand in this case (see Table 4). More biological treatment may not be needed to overcome this, because the purpose of biological treatment is conversion of soluble organics to CO_2, H_2O, and particulate matter (i.e., suspended solids).[9] Removal of suspended solids may be called for to reduce the effluent BOD_5. Carbonaceous BOD_5 (nitrification

*Not reproducible in these proceedings.

suppressed) versus total BOD_5 indicated that ammonia nitrogen was also a factor in exerting an oxygen demand in the treatment plant effluent. Since part of the RBC process involves nitrification, relatively large numbers of nitrifying bacteria were present in the plant effluent and in the BOD bottles. Hence, significant nitrification occurred within the first five days. Data in Table 4 show that nitrification accounted for about 7 mg/l (11-4 mg/l) of the BOD_5. Figure 5, an example of the BOD versus time data collected, also shows significant nitrification. Consequently, suspended solids (2 mg/l BOD_5) and nitrification (7 mg/l BOD_5) accounted for most of the BOD_5 (11 mg/l) in the wastewater treatment plant effluent. While the effluent did not strictly meet the BOD_5 limitations, the reasons for this were oxygen demand of suspended solids and NH_3-N oxidation. The RBC system was, therefore, performing adequately for soluble-BOD_5 removal at the existing conditions (26°C and 3.7 MGD). It is important to note, however, that at the 6 MGD design flow rate, adequate BOD_5 removal could not be assumed. In fact, although adequate soluble-BOD_5 removal occurred, it must be understood that the system can be considered to have been operating at suboptimal levels for soluble-BOD_5 removal because of DO limiting conditions. If supplemental oxygen were provided to the system at the same existing conditions, effluent soluble-BOD_5 levels would still be about the same except that less surface area would be required to remove the same amount of soluble-BOD_5. Then, more surface area would be available for nitrification.

During the summer study, the treatment plant effluent NH_3-N level was significantly higher than the monthly effluent limitation (Tables 1 and 3). The observed level of ammonia-nitrogen in the effluent was 6.2 mg/l while the limitation (and design expectation) is 2.0 mg/l. As with BOD_5, the shortcomings in meeting effluent limitations for ammonia-nitrogen are not obvious. Nitrification can be affected primarily by pH and temperature, but also by the sequence of organic removal (i.e., nitrification does not begin to any significant degree at neutral pH levels until soluble-BOD_5 has been oxidized).[5] Hence, to evaluate RBC performance for NH_3-N removal, progression of treatment within the RBC stages as well as environmental conditions must be assessed and discussed. NH_3-N data are presented in Table 5 and shown in Figure 6. Nitrification began in RBC stage 4 and continued through stage 6. Nitrification data is also supported by TKN and NO_2/NO_3 (Figure 6) and alkalinity changes (Figure 7). NH_3-N oxidation rates were also believed to be limited by low DO levels in some of the RBC stages, as discussed earlier for BOD_5 removal. In addition, NH_3-N oxidation rates were limited by low pH levels shown in Figure 6. Optimum pH for nitrification is about pH 8.0-8.5.[8,9] whereas observed values were pH 6.5-6.7 in latter RBC stages. (The summer wastewater temperature of 26°C was believed to be near optimum for nitrification.)

During the winter study, the RBC system was operating under markedly different conditions than during the summer. Wastewater temperature was, of course, much lower, averaging 13°C, and the hydraulic and organic loadings were significantly higher (see Tables 2 and 3). These factors would expectedly result in poorer RBC performance than noted during the summer. To the contrary, as shown in Table 3, the average wastewater effluent BOD_5 level during the winter was about the same as the summer BOD_5 level (10 mg/l during the winter and 11 mg/l during the summer). Also, average NH_3-N levels were actually lower (5.1 mg/l during the winter and 6.2 mg/l during the summer), albeit NH_3-N design expectations were still not met. The only positive effect during the winter study that could be attributed to this noticeable improvement in RBC system performance was the wastewater DO levels in the RBC system. Because of the low wastewater temperatures, adequate oxygen mass transfer occurred and, as shown in Figure 8, DO limiting conditions were non-existent. Also, the <u>Beggiatoa</u> bacteria had disappeared.

As shown in Figures 8 and 9, concentration profiles representing soluble organic removal and ammonia oxidation through the RBC system are not significantly different than those from the summer study. Although soluble organic removal may have been slightly more sluggish, adequate removal still occurred such that nitrification began in the fourth stage. Wastewater pH was, again, too low to insure optimum nitrification (see Figure 10).

The effect of adequate wastewater DO levels in the RBC system cannot, in this case, be fully appreciated by merely comparing summer vs winter treatment plant effluent and RBC system characteristics. These comparisons only show

7. Low pH levels can be easily corrected by chemical feed to maintain a pH level of 7.0 or higher.

8. BOD_5 analyses should include soluble-BOD_5 from each RBC stage when evaluating RBC system performance; likewise, BOD_5 analyses should include carbonaceous and nitrogeneous oxygen demands.

ACKNOWLEDGMENTS

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ABBREVIATIONS

BOD	Biochemical oxygen demand is the amount of oxygen required by microorganisms to oxidize dissolved organics in a wastewater.
BOD_5	BOD measured after 5 days
BOD_5-S	BOD of a filtered sample (BOD_5 of soluble organic material)
BOD_{20}	BOD measured after 20 days
$CaCO_3$	Calcium carbonate
$CBOD_5$	Carbonaceous BOD_5. Oxidation of any NH_3-N present in the sample is chemically inhibited.
CO_2	Carbon dioxide
DO	Dissolved oxygen
Eff	Effluent
FC	Fecal coliform
gpd/sq ft	Gallons per day per square foot
HCO_3^-	Bicarbonate ion
H_2CO_3	Carbonic acid
H_2S	Hydrogen sulfide
H_2SO_4	Sulfuric acid
Inf	Influent
l	liter
lbs	pounds
mg	milligram
MGD	Million gallons per day
ml	milliliter
μm	Micrometer
μmho/cm	Micromhos per centimeter
N	Nitrogen

NH^+_4	Ammonium ion
NH_3-N	Ammonia expressed as nitrogen
NO_2/NO_3	Nitrite plus nitrate expressed as nitrogen
NO^-_3	Nitrate ion
O_2	Oxygen
pH	Negative logarithm of hydrogen ion concentration
RBC	Rotating Biological Contactor
S_2	Elemental sulfur
SS	Suspended solids
T Alk	Total alkalinity
Temp ^{o}C	Temperature in degrees Celsius
TKN	Total Kjeldahl nitrogen
TOC-S	Total organic carbon of a filtered sample (soluble TOC)
YSI	Yellow Springs Instruments

TABLE 1. NPDES PERMIT PARAMETERS AND LIMITATIONS*

	Monthly Average Summer	Monthly Average Winter
pH	6.0 - 9.0	6.0 - 9.0
Chlorine Residual	Min conc to comply w/FC limit	Min conc to comply w/FC limit
Fecal Coliform (FC)	200/100 ml	200/100 ml
Suspended Solids (SS)	30 mg/1	30 mg/1
Five-day Biochemical Oxygen Demand (BOD_5)	10 mg/1	20 mg/1*
Ammonia Nitrogen (NH_3-N)	2.0 mg/1	5.0 mg/1*
Dissolved Oxygen (DO)	Greater than 6.0 mg/1	Greater than 8.5 mg/1

* Limitations at the time of design did not allow for winter variance.

TABLE 2. RBC INFLUENT AND EFFLUENT CHARACTERISTICS

	15 - 21 AUGUST 1978 Avg. Wastewater Flow = 4.5 MGD*		23 - 29 JANUARY 1979 Avg. Wastewater Flow = 5.2 MGD*	
	Influent	Effluent	Influent	Effluent
Conductivity (μmho/cm)	960	930	922	873
T Alk	158	90	174	97
SS	69	63	110	89
BOD_5	72	61	126	96
$CBOD_5$	48	28	---	---
BOD_5-S	21	4	33	5
TOC - S	23	11	24	9
TKN	21	8.9	25.1	11.1
NH_3-N	16.0	6.2	16.3	4.8
NO_2/NO_3-N	0.05	8.9	0.89	9.29

* STP Influent flow + recirculated flow.
All units are mg/l unless otherwise noted.

TABLE 3. SEWAGE TREATMENT PLANT EFFLUENT VALUES

	15 - 21 AUGUST 1978	23 - 29 JANUARY 1979
pH	6.7 (median)	7.0 (median)
BOD_5	11 mg/l	10 mg/l
NH_3-N	6.2 mg/l	5.1 mg/l
SS	9 mg/l	13 mg/l
Flow	3.7 MGD	4.6 MGD

TABLE 4. TREATMENT PLANT EFFLUENT BOD LEVELS

	BOD_5	BOD_{20}
Filtered (soluble) BOD	2	No data
Unfiltered (total) BOD	11	45
Suppressed Nitrification (carbonaceous only) BOD	4	8

Analysis:

BOD_5: 2 mg/l were due to soluble organics; 4 mg/l were due to carbonaceous material (2 mg/l soluble + 2 mg/l suspended); 7 mg/l (11-4 mg/l) were due to nitrification.

BOD_{20}: At 20 days, 37 mg/l (45-8 mg/l) were due to nitrification.

TABLE 5. RBC SYSTEM DATA SUMMARY*

AUGUST (WASTEWATER TEMP $\cong$ 26°C)

Avg Flow: 3.7 MGD (plant flow) + 0.8 MGD (recirculated flow)

	pH	T Alk	DO	BOD_5-S	TOC-S	TKN	NH_3-N	NO_2/NO_3-N
RBC Influent	6.8	159	3.4	24	24	23	16	<0.04
Stage 1	6.8	159	1.4	15	18	22	16	<0.04
2	6.8	158	0.7	9	14	20	16	0.2
3	6.75	149	0.8	6	11	19	15	1.4
4	6.6	126	1.3	4	10	14	12	4.4
5	6.55	104	1.9	3	9	11	8.3	7.7
6	6.5	93	2.2	3	9	9	6.6	9.7

JANUARY (WASTEWATER TEMP $\cong$ 13°C)

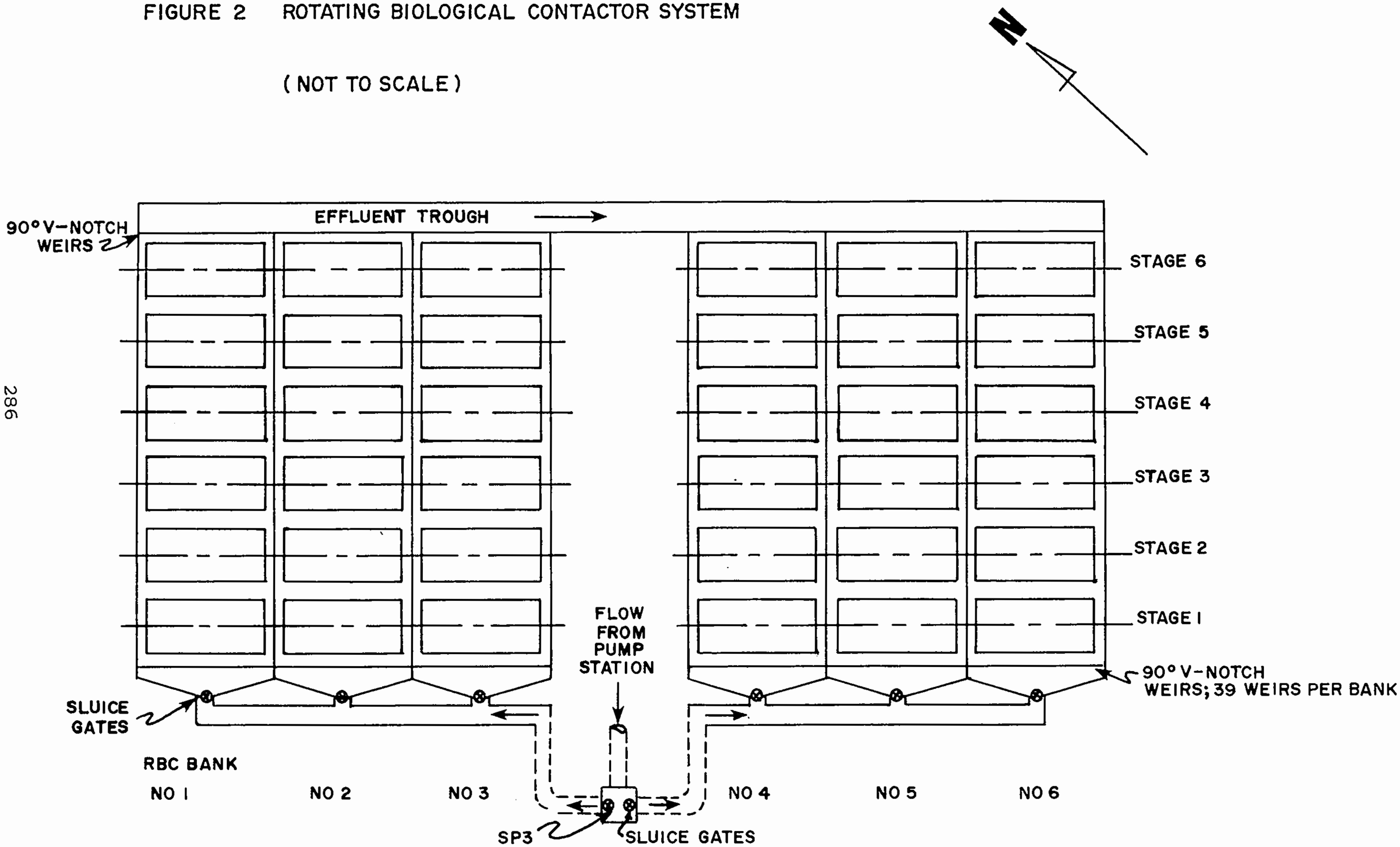
Avg Flow: 4.6 MGD (plant flow) + 0.6 MGD (recirculated flow)

	pH	T Alk	DO	BOD_5-S	TOC-S	TKN	NH_3-N	NO_2/NO_3-N
RBC Influent	7.1	163	5.7	32	23	24	16	0.9
Stage 1	7.1	163	3.3	24	17	23	15	0.6
2	7.1	161	2.7	12	13	21	14	0.8
3	7.1	152	2.9	10	11	21	14	1.8
4	6.95	124	3.5	7	10	16	9.1	5.3
5	6.8	106	4.2	6	9	14	6.7	7.7
6	6.75	87	4.8	5	8	12	3.9	10.2

* Mean of 5 sets of grab samples collected at various times for each study period. pH values represent median values. Units expressed as mg/l unless otherwise noted.

FIGURE 2 ROTATING BIOLOGICAL CONTACTOR SYSTEM

(NOT TO SCALE)



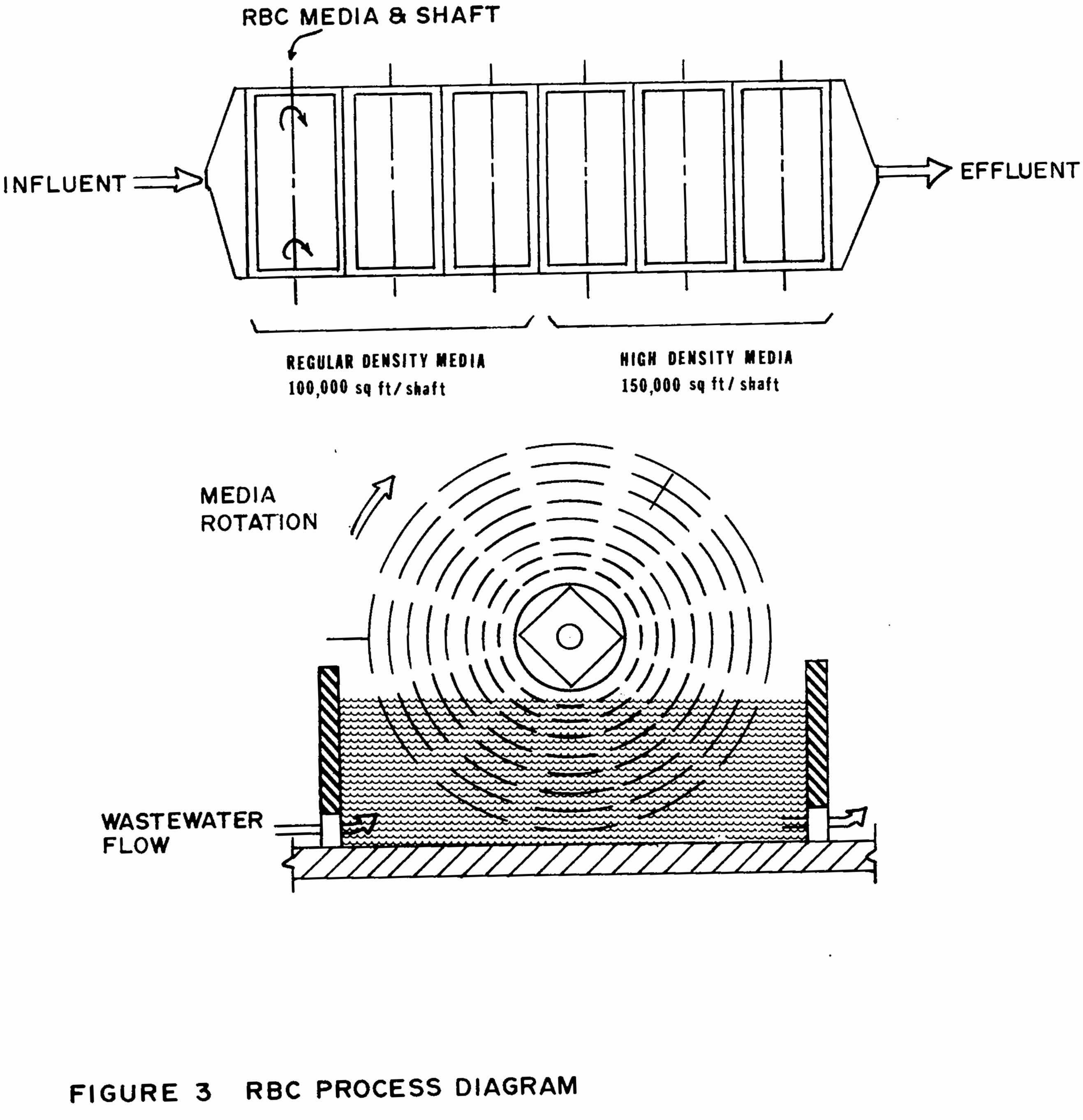


FIGURE 3 RBC PROCESS DIAGRAM

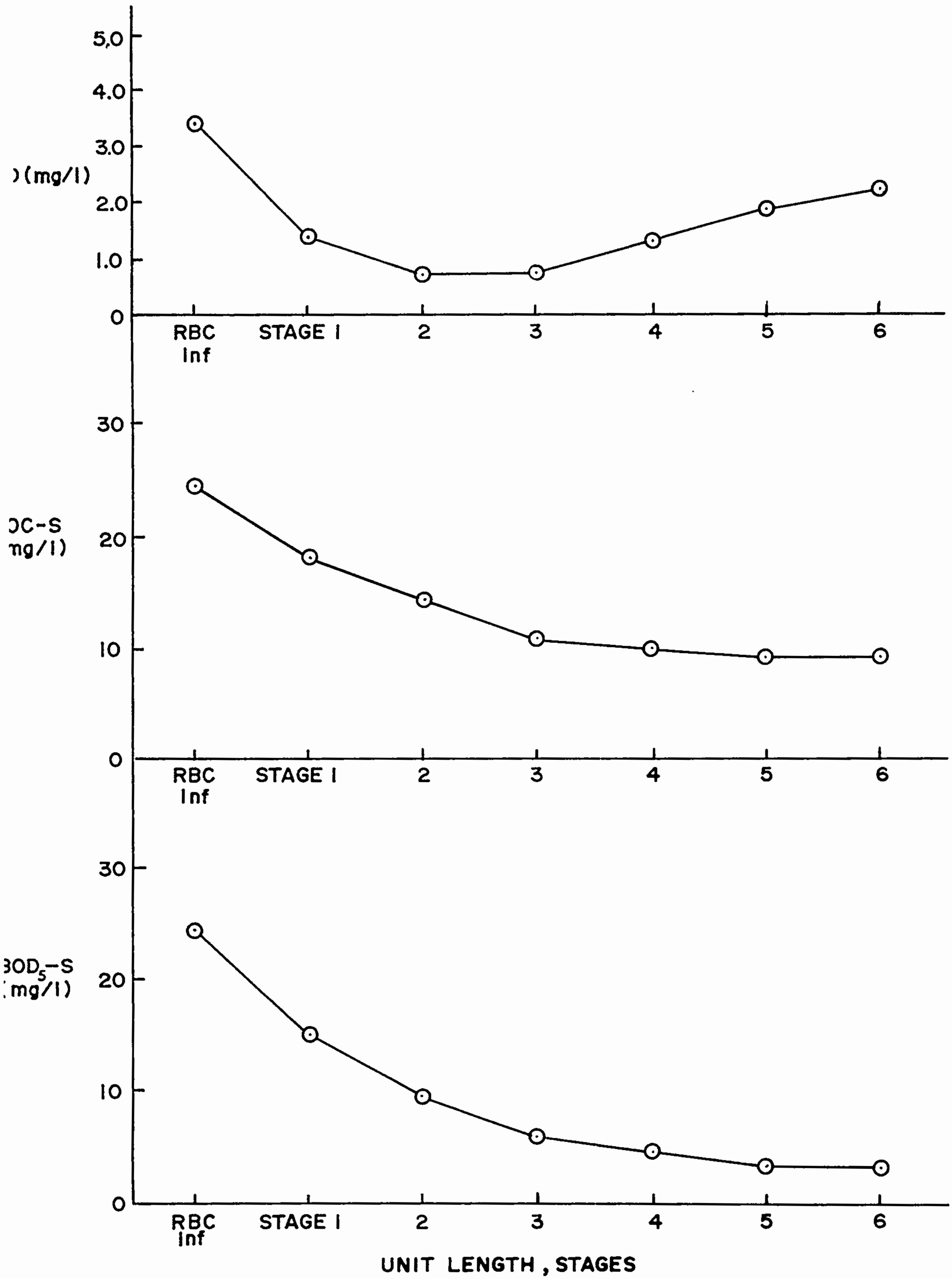
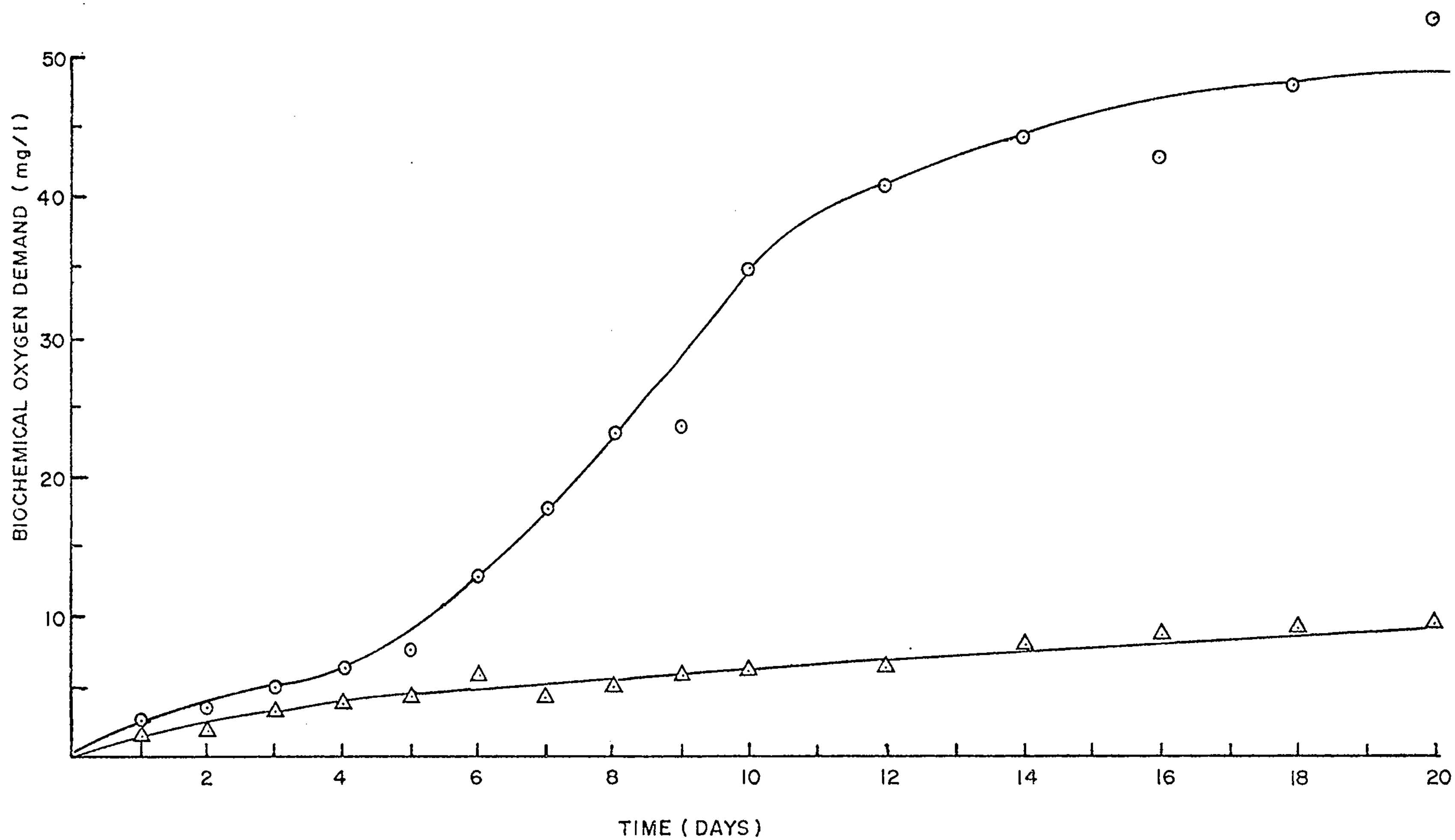


FIGURE 4. SOLUBLE BOD_5, SOLUBLE TOC AND DO LEVELS THROUGH THE RBC SYSTEM, SUMMER STUDY

FIGURE 5 BIOCHEMICAL OXYGEN DEMAND VERSUS TIME, SAMPLE POINT 6 - STP EFFLUENT

⊙ — STANDARD BOD PROCEDURE - SEEDED W/RAW INFLUENT

△ — NH_3-N SUPPRESSION (C BOD) USING 0.1M NH_4CL - SEEDED W/RAW INFLUENT



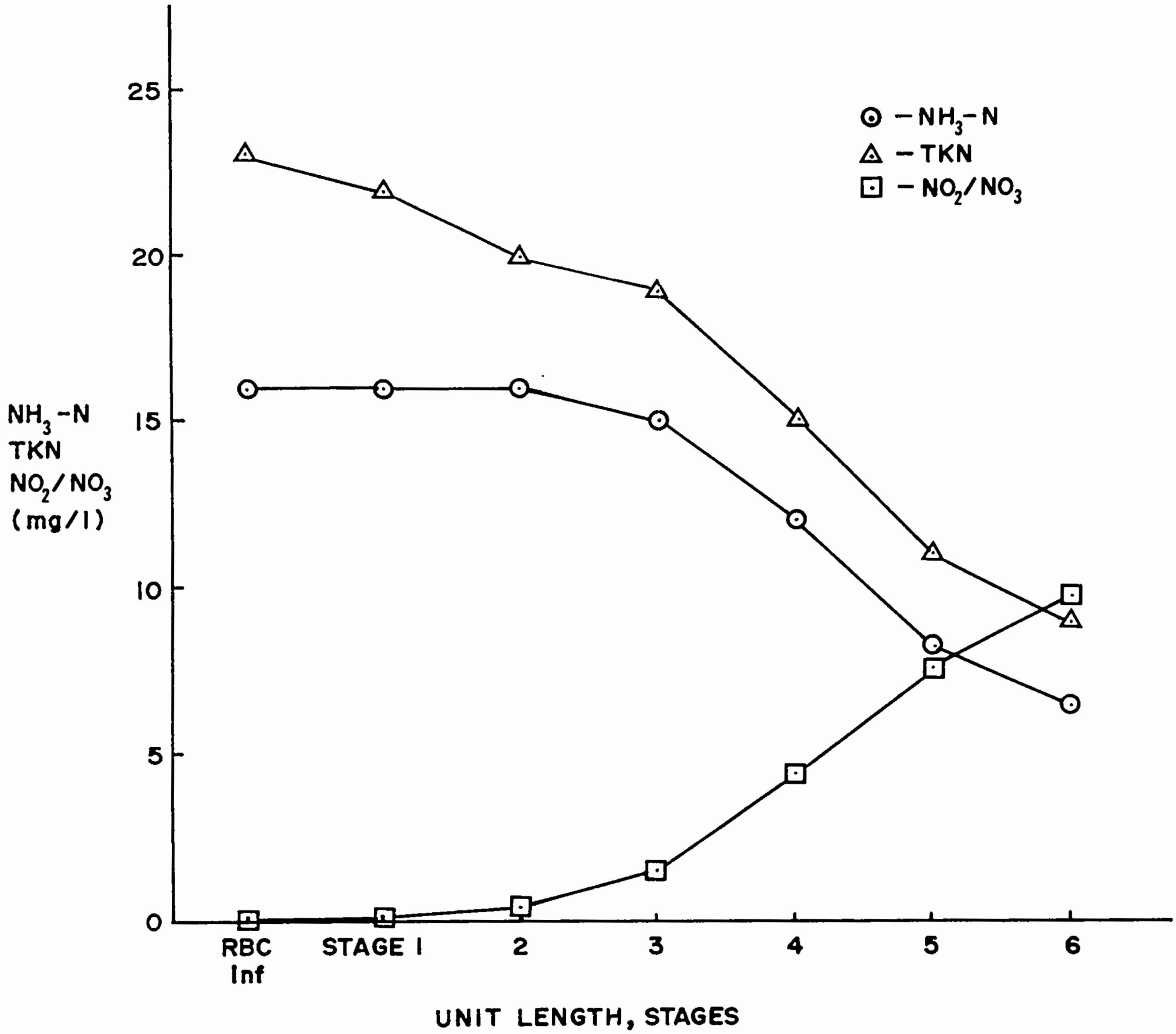


FIGURE 6 NH_3-N, TKN AND NO_2/NO_3 LEVELS THROUGH THE RBC SYSTEM, SUMMER STUDY

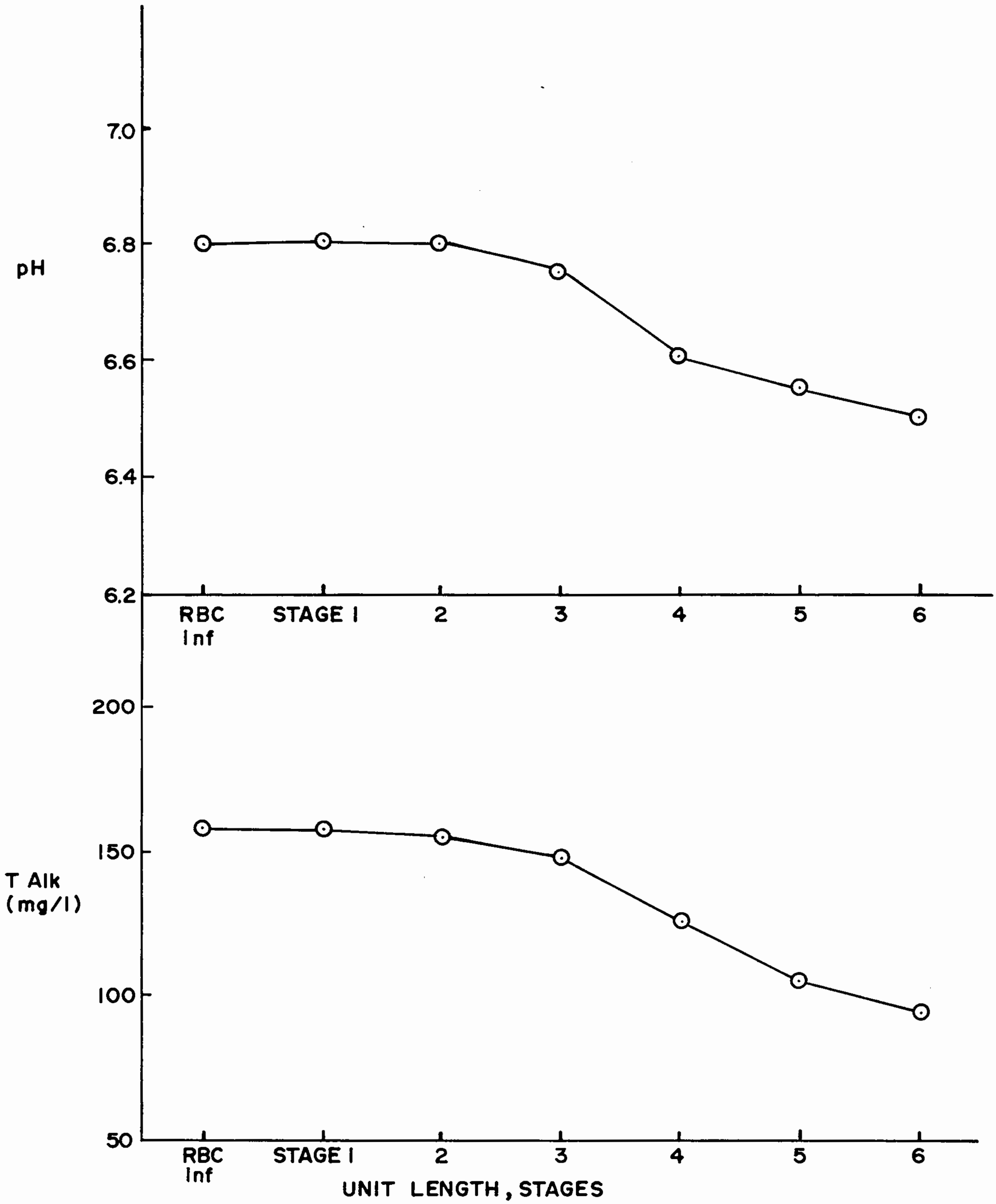


FIGURE 7 CHANGES IN ALKALINITY AND pH THROUGH RBC SYSTEM, SUMMER STUDY

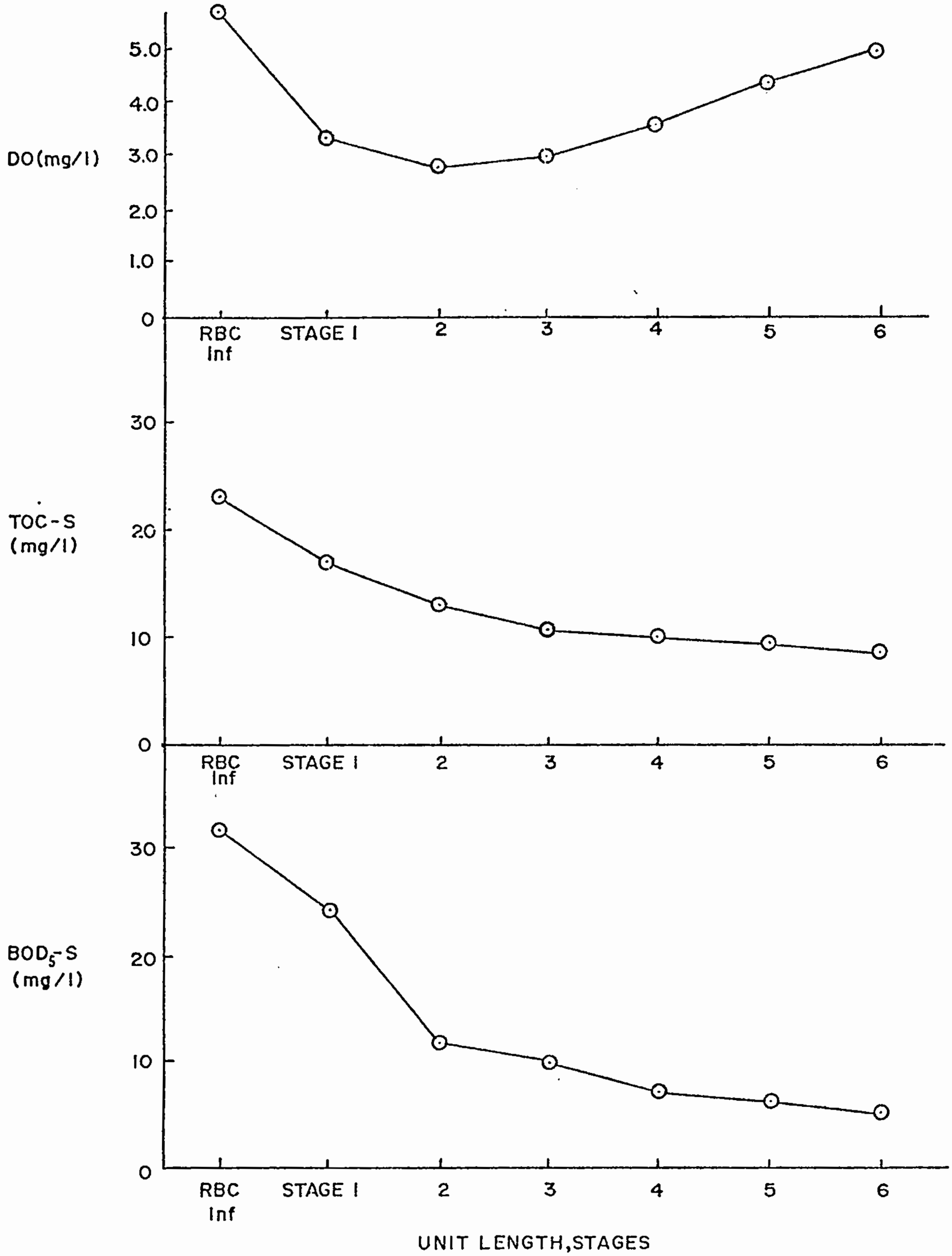


FIGURE 8 SOLUBLE BOD_5, SOLUBLE TOC AND DO LEVELS THROUGH THE RBC SYSTEM, WINTER STUDY

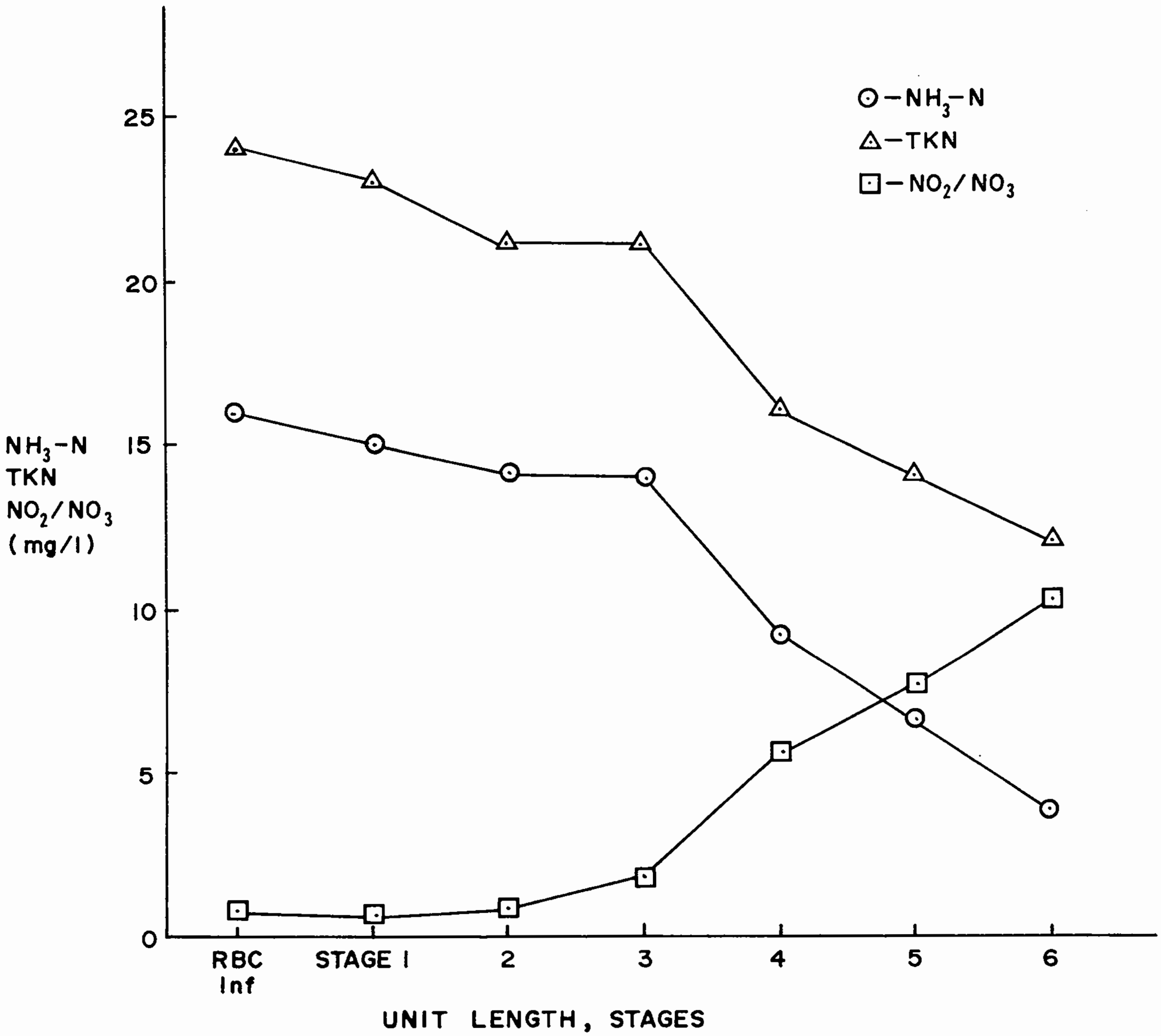


FIGURE 9 NH_3–N, TKN AND NO_2/NO_3 LEVELS THROUGH THE RBC SYSTEM, WINTER STUDY

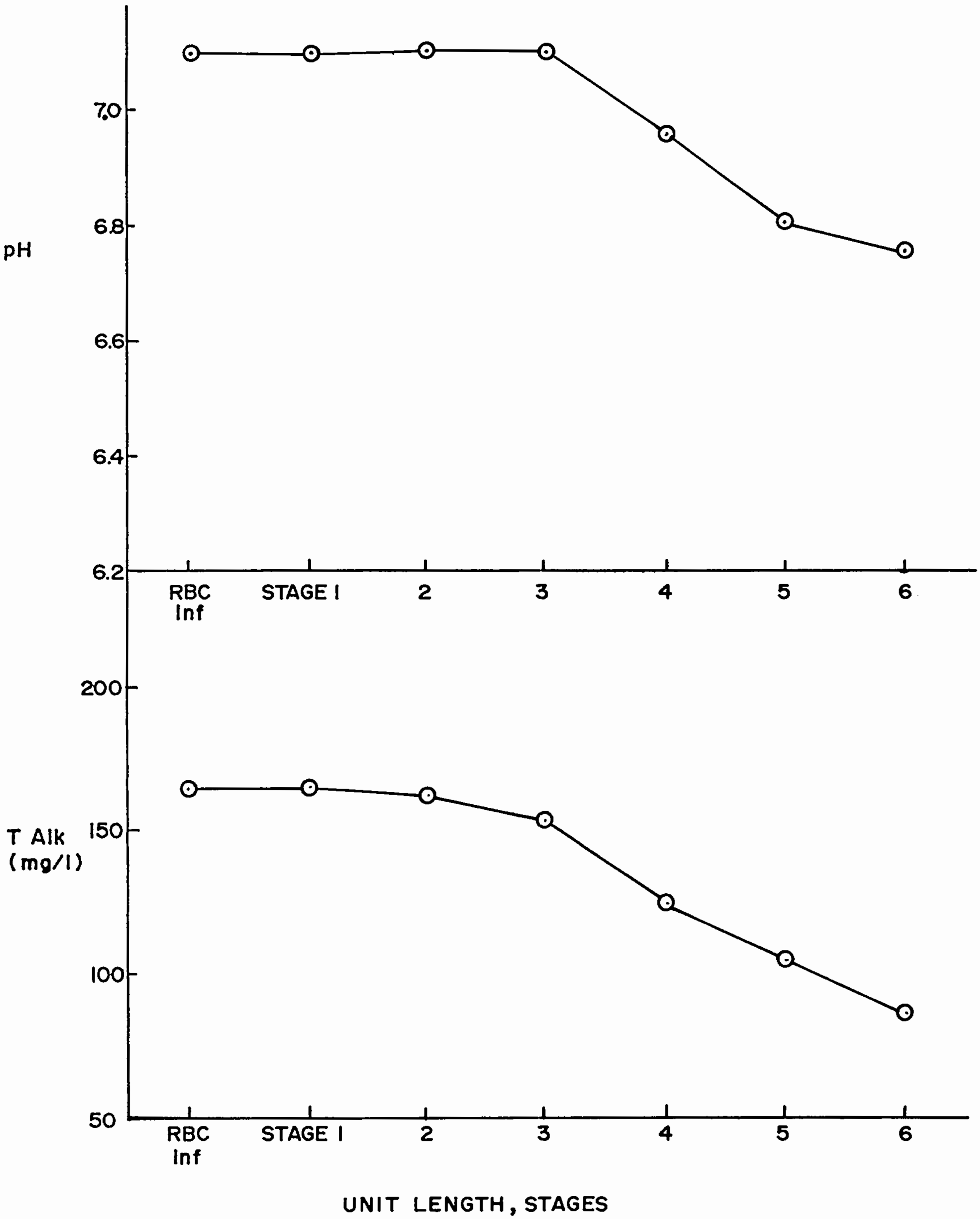


FIGURE 10 CHANGES IN ALKALINITY AND pH THROUGH THE RBC SYSTEM– WINTER STUDY

NITROGEN AND PHOSPHORUS REMOVAL WITH ROTATING BIOLOGICAL CONTACTORS

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Introduction

As of September, 1979, over 200 municipalities in the United States and Canada had chosen Autotrol's Bio-Surf Process for wastewater treatment. Of these, 25 percent were installed to remove BOD and to nitrify the sewage. Another 5 percent were installed to nitrify effluent from existing secondary treatment plants to meet new discharge limitations. About 10 percent were installed to include phosphorus removal and one of the largest is designed for denitrification.

Those installations concerned with nutrient removal range in size from 0.1 to 37 MGD. Average flow for the 70 plants was 3 MGD, with 25 plants within 2 MGD of this agerage. This wide and relatively uniform distribution of plant sizes attests well to the application of the Bio-Surf Process to communities of all sizes, while at the same time offering a non-complex system using a minimum amount of energy.

The first portion of this paper will discuss two case histories of nitrification and phosphorus removal with the Bio-Surf Process in the United States. The latter portion of this paper will be presented by my colleague from Nippon Autotrol, Mr. Hiroshi Iemura. His discussion will include information on several installations designed for nitrification and denitrification. Full-scale performance since 1976 and a new design rationale for methanol addition resulting from this experience should be of keen interest to this audience.

Case No. 1 - Gladstone, Michigan

The first case history illustrates application of the Bio-Surf Process designed for BOD removal from primary clarifier effluent. Nitrification was not required, but is achieved at no extra cost because of the temperature corrections made for BOD removal at low winter temperatures. Phosphorus removal was required and is achieved by addition of alum and anionic polymer to the Bio-Surf effluent prior to secondary clarification.

This plant is located in Gladstone, Michigan on the southern exposure of the Upper Pennisula to Lake Michigan. The existing real estate was narrow and little space was available for treatment expansion between the existing primaries and the lake. This situation and the desire to maintain process simplicity led to the installation of six rotating biological contactors in two parallel trains. This arrangement allowed placement immediately adjacent to the rectangular primaries under an expanded common building, while at the same time allowing for construction of two new secondaries next to the lake.

Startup of this facility designed for treatment of 1 MGD began in 1974 and has provided a most significant contribution to Autotrol's knowledge of the Bio-Surf Process and to current design rationales based on full-scale, rather than pilot-scale, performance.

Table A summarizes monthly average Process performance for the year of 1977. This performance is very typical of performance since startup to the present day, showing effluent BOD and suspended solids well within discharge requirements of 35 mg/l and a minimum of 80 percent phosphorus removal.

The data in Table I were divided into the summer and winter periods (June through October and November through May) to illustrate performance at the different wastewater temperatures of 63°F (17.2°C) and 49°F (9.4°C). It is quite obvious that only nitrification is affected, as would be expected by the transition from warm to cold conditions, whereas BOD, phosphorus and suspended solids removals are very consistent at 94, 87 and 81 percent, respectively.

Table B summarizes monthly operational costs with respect to chemicals, electrical power and miscellaneous utilities for the same seasonal and annual periods. Little seasonal differences are seen for alum and polymer costs per million gallons treated per day, whereas relatively large differences are seen for chlorine and utilities. Higher chlorine consumption in summer is understandable from decreased solubility and greater reactivity considerations.

The increase in utilities costs in winter reflect the purchase of larger amounts of natural gas to maintain anaerobic digester temperatures at proper levels.

Seasonal treatment costs per million gallons are quite uniform in either case, however, varying only about 10 percent from the annual average.

Cost analysis for nutrient removals can be looked at in several ways. However, since this plant was designed and constructed for secondary treatment and phosphorus removal, all costs should be so analyzed. In that event, the alum and polymer costs are those principally related to phosphorus removal and, on an annual basis, the City of Gladstone is expending seventy-six cents per pound of phosphorus removed per million gallons treated. The balance of the costs for chlorine, power and utilities are largely associated with BOD and suspended solids removal. For 1977, these costs were 4.1 cents per pound removed, for a total of 80.1 cents for removal of the three pollutants.

Nitrification could be regarded as being obtained at no cost. However, the records are complete enough to allow differentiation between the removal of each "nutrient" category. Field evaluations by Autotrol reveal that soluble BOD oxidation is essentially complete with 50 percent of the equipment in summer and 67 percent in winter. Nitrification occurs on the remainder in each case. In addition, electrical energy consumption has been measured at Gladstone with polyphase wattmeters in each season. With the above information one can compute daily energy costs for nitrification with rotating biological contactors for installations similar to Gladstone. Calculations summarized in Table C show that less than five cents of electrical energy are required for nitrification of one pound of ammonia nitrogen per million gallons treated per day with the Bio-Surf Process at Gladstone.

Table A

Monthly Bio-Surf Process Performance - Gladstone, Michigan

- 1977 -

	Summer	Winter	Annual
Q, MGD	0.748	0.707	0.724
^{o}F	63	49	55

	Summer mg/l			Winter mg/l			Annual mg/l		
	Raw	FE	%R	Raw	FE	%R	Raw	FE	%R
BOD_5	127	7	94.5	155	8	94.8	143	8	94.4
NH_3-N	15.8	1.1	93.0	15.0	4.9	67.3	15.4	3.5	77.3
P	6.5	1.2	81.5	6.1	1.2	80.3	6.3	1.2	81.0
TSS	131	15	88.5	155	17	85.2	122	16	86.9

Table B

Monthly Operational Costs - Gladstone, Michigan

- 1977 -

Dollars Per Million Gallons Treated*

	Summer	Winter	Annual
Alum	28.60	28.15	28.33
Chlorine	6.00	3.45	4.51
Polymer	4.31	4.14	4.21
Power	50.76	59.25	55.71
Utilities	3.79	16.92	9.98
Total	$ 93.46	$ 111.91	$ 102.74

*Alum, chlorine and polymer costs were $0.041, $0.170 and $1.90 per pound, respectively.

Table C

Estimated Nitrification Energy Costs - Gladstone, Michigan

- Primary Clarifier Effluent -

Per Million Gallons Treated

	Summer	Winter	Annual
LBS/Day	117	81.0	95.5
KWHR/LB	1.13	1.08	1.11
KWHR $/LB*	0.045	0.045	0.044

*Calculated using $0.040/KWHR

Case No. 2 - Cadillac, Michigan

The second most popular application of the Bio-Surf Process for nutrient removal is nitrification of secondary effluent following pretreatment with alternative processes. At Cadillac, Michigan, an existing activated sludge plant required upgrading for nitrification and phosphorus removal. The final design incorporated ferric chloride addition for phosphorus removal ahead of the nitrification section.

Eight rotating biological contactors began operation in 1976 to meet discharge requirements of 1.5 mg/l NH_3-N during the summer months of June through October. Performance did not meet requirements and thorough investigation culminating in tracer studies revealed hydraulic shortcircuiting because of excessive underflow openings in the baffles separating the reactors. Performance improved measureably after corrections were made to improve staging and effluent requirements have since been met consistently.

Current annual operation consists of shutting down the RBC shafts on November 1st and re-starting on April 15th. This was done to take advantage of the annual opportunity to reduce laboratory and maintenance schedules and also to reduce overall energy consumption. Performance in this mode of operation for 1979 is summarized in Table D. As seen in Table D, both the Bio-Surf and final effluent averages are below the 1.5 mg/l requirement for the 5-month period. It is also evident that the secondary clarifier effluent NH_3-N declined considerably toward mid-year, then abruptly increased in October by approximately 90 percent, from 5.9 to 11.0 mg/l. This pattern reflects eventual nitrification in the activated sludge basin. This results from recirculation of Bio-Surf nitrifiers removed by the final sand filters back to the head end of the plant for separation by the primary clarifier prior to eventual digestion. The abrupt decrease in this pre-nitrification cannot be explained from plant records, but both the Bio-Surf and the final effluent quality remained within the 1.5 mg/l specification.

An interesting sidelight to these 1979 data was made possible this past year by monitoring nitrifier growth on the biological contactors from startup on April 15th through shutdown on November 1st. This data was developed from hydraulic load cells under the idle end bearing of the first and last shaft in one of the tanks. Hydraulic pressure readings were taken during rotation on a frequent basis in conjunction with normal influent and effluent sampling routines. These pressure readings were converted to dynamic biofilm thickness equivalents with Autotrol formulae and Table E summarizes biofilm response to influent NH_3-N following flow startup. Growth was evident within two weeks and was maximized to biological equilibrium by mid-June. A decline began shortly afterward in response to mid-month minimums in daily influent concentrations and a relatively flat, but gradually declining profile was maintained until mid-September. At this time, the partial nitrification occurring in the activated sludge unit abruptly failed, and by the latter part of October a slight uptrend in biofilm weight was evident.

Several factors no doubt contributed to this pattern of nitrifier growth. Although influent NH_3-N must be by far the major factor, it is normal for predator growth to respond to bacterial growth and to reduce the population until equilibrium is established between influent strength, bacterial kinetics and predation efficiency. This phenomenon has been observed many other times in our BOD removal studies, but these data represent the first documentation in a low BOD, secondary nitrification situation. It will be most interesting to see if a similar pattern develops in 1980.

Table D

Nitrification of Secondary Effluent - Cadillac, Michigan

- 1979 -

Month	Q, MGD	T^oF	NH_3-N, mg/1 S.C.E.	NH_3-N, mg/1 B.S.E.	NH_3-N, mg/1 F.E.
June	1.62	63	13.1	1.40	1.00
July	1.61	65	8.9	0.67	0.43
August	1.61	67	8.6	0.76	0.57
September	1.50	66	5.9	0.60	0.42
October	1.51	63	11.0	1.52	1.22
	1.57	65	9.5	0.99	0.73

Table E

Bio-Surf Biofilm Response to NH_3-N

Cadillac, Michigan - 1979

Date		30-day NH_3-N, mg/l Influent	Effluent	Biofilm Thickness Inches S-1	S-4
April	15*	-	-	0.000	0.000
	30	16.5	16.3	0.008	0.006
May	15	-	-	0.014	0.010
	30	13.1	1.5	0.022	0.016
June	15	-	-	0.026	0.016
	30	13.1	1.4	0.014	0.010
July	15	-	-	0.014	0.013
	30	8.9	0.7	0.014	0.013
Aug.	15	-	-	0.016	0.015
	30	8.6	0.8	0.014	0.013
Sept.	15	-	-	0.013	0.013
	30	5.9	0.6	0.007	0.005
Oct.	15	-	-	0.004	0.003
	30	11.0	1.5	0.007	0.007

*Flow began.

Table 1

SUMMARY OF RBC INSTALLATIONS IN JAPAN

WASTEWATER	SITES	%	SHAFTS	%	FLOW 1000 m^3/d	FLOW MGD
DOMESTIC	309	40	495	44	98.7	26.1
FOOD	188	24	377	11	25.5	6.7
INDUSTRIAL	196	25	384	25	82.3	21.7
GARBAGE	66	8	110	6	12.8	3.4
ANIMAL	23	3	48	1	3.2	0.8
	782	100	1,414	100	222	58.7

Table 2

BIO-SURF NITROGEN REMOVAL INTALLATIONS IN JAPAN

NAME	WASTE SOURCE	FLOW m^3/d	FLOW MGD	NH_3-N, MG/L INFL.	NH_3-N, MG/L EFFL.	START
MIYASAKI CITY	Garbage Dump	350	0.092	200 (TN)	50 (75%)	1976
NIPPON CHEMICAL	Brine	2,000	0.528	200		1977
IWAKI CITY	Garbage Dump	200	0.053	200	20 (90%)	1978
CHIBA PREF.	Domestic	65	0.017	20	3 (85%)	1979
YOKKAICHI CITY	Garbage Dump	500	0.132	250	6	1979
HAMMATSU CITY	Garbage Dump	400	0.106	55 (TN)	10	1980

Table 3

NITRIFICATION AND DENITRIFICATION DESIGN CRITERIA

CASE NO. 1 - MIYAZAKI CITY

LOADING RATES

A. NITRIFICATION SECTION

BOD_5:	0.9 g/m^2-D (0.18 #/1,000 ft^2-D)
Hydraulics:	18.7 l/m^2-D (0.46 gpd/ft^2-D)
NH_3-N:	3.7 g/m^2-D (0.76 #/1,000 ft^2-D)
Time:	6.9 Hours

B. DENITRIFICATION SECTION

Hydraulics:	42.2 l/m^2-D (1.04 gpd/ft^2-D)
Methanol:	1.9 times NO_3-N Concentration
NO_2-N + NO_3-N:	7.4 g/m^2-D (1.52 #/1,000 ft^2-D)
Time:	9.9 Hours

C. RE-AREATION SECTION

BOD_5:	20.0 g/m^2-D (4.09 #/1,000 ft^2-D)
Hydraulics:	269 l/m^2-D (6.62 gpd/ft^2-D)
Time:	0.57 Hours

Table 4

BIO-SURF PERFORMANCE - MIYAZAKI CITY

- 1977 -

MONTH	DESIGN Q %	°C/F	NH_3-N, MG/L IN	NITR.	DENITR.	NO_3-N, MG/L IN	NITR.	DENITR.
Jan.	21	9/48	101	Tr	Tr	1.9	102	15
Feb.	19	8/46	104	Tr	Tr	1.6	108	Tr
Apr.	34	22/72	128	28	21	22	26	11
May	47	24/75	132	4	4	1	70	13
Jun.	53	30/86	119	4	3	1	68	5
Jul.	43	30/86	138	3	3	1	67	14
Nov.	71	23/73	128	4	3	1	77	2
May/Nov.	54	27/81	129	4	3	1	70	8
PEAK FLOW OCCURRENCE								
Nov.	153	20/68	97	0	0	1	59	0

Table 5

BIO-SURF PERFORMANCE - MIYAZAKI CITY

- 1977 -

MONTH	BOD, MG/L IN	BOD, MG/L NITR.	BOD, MG/L DENITR.	TSS, MG/L IN	TSS, MG/L NITR.	TSS, MG/L DENITR.	$CaCO_3$, MG/L IN	$CaCO_3$, MG/L NITR.	$CaCO_3$, MG/L DENITR.
Jan.	8	5	3	7	4	1	1,020	324	503
Apr.	18	5	8	11	27	2	1,068	169	192
May	23	6	14	48	27	4	1,082	345	384
Jun.	30	3	10	72	240	4	1,104	354	475
Jul.	19	6	4	37	198	9	-	-	-
Aug.	-	9	4	38	-	3	-	-	-
Nov.	24	2	-	101	92	13	-	-	-
Avg.	20	5	7	45	98	5	1,067	298	389
PEAK FLOW OCCURRENCE									
Nov. 16	8	5	5	264	53	10	783	337	484

Bio-Surf Process in Japan

Historical

The RBC process was introduced to Japan approximately 10 years ago in 1970. Initial emphasis was directed to secondary treatment of food processing wastewater, as there were many small companies supplying Japan and foreign countries with canned fruit, mandarin oranges being only one example. Tertiary treatment started six years later in 1976.

The growth of the RBC process in Japan has been very rapid. Table 1 summarizes the number of locations, reactors and the quantity of flow for the five major wastewater categories identified as domestic, food processing, industrial (such as pulp and paper), garbage dumps and animal breeding. The number of firms actively pursuing this market is approximately 20, with only two or three of significant corporate size. Nippon-Autotrol is the major active company with a manufacturing, R & D and sales organization, accounting for 20 percent of the installations, 26 percent of the rotating contactors and 25 percent of the quantity treated.

There are three basic reasons for the rapid growth; (1) low energy and small space, very appealing to industrial firms; (2) maintenance and process simplicity; and (3) impetus provided by the government in the form of the Japan Sewage Works Agency recognizing applicability to municipal wastewater.

At the present time there are no regulations restricting discharge of nitrogen or phosphorus in wastewaters. However, problems associated with nutrient discharge intensify with each summer season and many Japanese are becoming very concerned with the implications for the immediate and long range future.

One of these problems is called the 'Red Tide', which occurs annually in many of the inland seas of Japan. The microorganism responsible for this phenomenon, resulting in significant fish and shellfish kills, has stimulated much research. The identity and life mechanisms have not yet been fully defined, but nitrogen and phosphorus removal will no doubt play a major role in control measures.

Another problem is referred to as "Flower of Water", or water bloom, in Lake Sagami and Lake Biwa. These lakes provide drinking water to the metropolitan areas of Tokyo and Osaka and water flavor is noticeably affected.

A third problem has been associated with the rice growing industry. Poor rice yields have been traced to high nitrogen and phosphorus levels, with the result being excessive stalk growth versus the desired kernel growth.

The fourth problem is linked to the newly-defined limits of available drinking water supplies. Water re-use is now being promoted and is gaining wider acceptance each year. Wastewater from hand washing, kitchens and cooling towers are being collected separately, treated biologically with RBC units and sand filters, and then chlorinated in order to reuse for sanitary flushing, car washing and lawn irrigation, etc.

Table 2 summarizes information relative to six Bio-Surf nitrification and/or removal installations, five of which are in operation at the present time. Flow ranges from 65 to 1000 m^3/D (17,000 to 528,000 gpd), with NH_3-N concentrations in the range of 20 to 200 mg/1. Runoff from garbage dumps obviously is the major problem area for the present, as 4 of the 6 plants shown are directed to this problem.

Of the six plants, Miyazaki City, Iwaki City and Chiba Prefecture use methanol as the source of carbon for denitrification. Nippon Chemical is only a nitrification application for the present. Yokkaichi City utilizes BOD from the incoming wastewater as a source of carbon, and this coupled with 3-1 effluent recycle rates effectively eliminates methanol costs for denitrification.

Hammatsu City is now under construction and will incorporate the new Aero-Surf Process for both nitrification and denitrification.

The following discussion will concentrate on the Miyazaki City application, since it is the first installation and has been in operation since October of 1976.

Case No. 1 - Miyazaki City

This Bio-Surf plant is installed adjacent to a land area created by garbage dumping. Leachate from this landfill flows to a river and at a point downstream is withdrawn as river water for rice field irrigation. The nitrogen content of this river water was found to be the principal cause of the previously discussed poor rice yield and several studies were made to select a process to solve the problem. Unit processes evaluated were activated sludge, trickling filters, lagoons, stripping and the RBC process. Simplicity of construction, operation, maintenance and low energy proved to be the decisive factors and the Bio-Surf Process was selected.

The leachate is characterized as a low BOD (10 - 20 mg/1), low suspended solids (2 - 30 mg/1), normal pH (7.6 - 7.9), low phosphorus (non-detected) and high ammonia (100 - 120 mg/1). Alkalinity was sufficient for nitrification (800 - 1,100 mg/1), and temperature was expected to be at least 15°C (59°F) year round.

Design loading criteria for BOD_5, hydraulic flow, NH_3-N, equivalent detention time, methanol dosage, NO_2-N and NO_3-N are summarized for the total installation in Table 3. Influent BOD_5, total nitrogen, and suspended solids values were defined as 50, 200 and 100 mg/1. Respective final effluent values were defined as 20, 50 and 25 mg/1.

Figure 1 illustrates the process flow and key sampling points for the Miyazaki installation. Two 2-stage Bio-Surf shafts in series were provided for BOD removal and nitrification. Each shaft was 7.5-m (25.6 ft.) in length and media diameter was 3.6-m (11.8 ft.), for a total surface area of 18,680 m^2 (200,810 ft^2).

A single 4-stage Bio-Surf shaft of special design for completely submerged operation with 8,490 m^2 (91,265 ft^2) was provided for denitrification. For reaeration, and residual methanol oxidation, a small 4-stage Bio-Surf shaft was provided. Media diameter is 2.0 m (6.6 ft) on a 4.5 m (14.8 ft) shaft.

The single and final clarifier provided an overflow rate of 15.8 m^3/m^2-D (388 gal/ft^2-D) and a detention time of 4.7 hours. Underflow solids are transported by truck for disposal elsewhere.

Plant startup occurred on October 15, 1976 at a flow of 157 m^3/D (0.041 MGD), or 45 percent of design. Temperature was 20°C (68°F), BOD was about 10 mg/1, NH_3-N was 118 mg/1, and pH was 7.9 units. After two weeks small amounts of NO_2-N were detected and startup seemed to be progressing very well. Temperature dropped within another two weeks to 12°C (54°F) and, disappointedly, no NO_2-N was detected.

A review of the design data pointed out that lack of phosphorus may be a factor. On November 29th, phosphoric acid was added to a 3.2 mg/1 as P concentration. Three days later NO_2-N rapidly increased to a maximum value of 39 mg/1. Six days later the NO_2-N began to decrease and by December 20th, nitrification was essentially complete.

Denitrification followed and was complete in 30 days. Figure 2 illustrates the startup NH_3-N, NO_2-N and NO_3-N concentration profiles during the 4-month October 1976 through January 1977 period. It is clearly evident from this graph that satisfaction of the phosphorus deficiency was the key factor to proper nitrification, as all other factors were in the usual order (temperature, dissolved oxygen, etc.).

Table 4 summarizes nitrification performance for the January-November, 1977 period. Plant flow gradually increased to 71 percent of design but the NH_3-N concentration was at or slightly in excess of the 100-120 mg/1 design for most of the year. Nitrification was greater than 96 percent, particularly in the May-November period, with final effluent NH_3-N concentrations averaging 3 mg/1.

Denitrification was also very successful and the total nitrogen effluent requirement of 50 mg/1 was easily met. Although data is sparse for nitrogen forms other than ammonia or the oxides, evidence as early as January, February and May revealed less than 20 mg/1 was being discharged routinely.

Figure 3 illustrates NH_3-N removal as a function of NH_3-N loading. Dots represent full-scale performance at Miyazaki. Squares denote pilot plant data developed elsewhere on municipal wastewater in Japan. The correlation is extremely good and signifies no difference despite the great difference in wastewater source.

Figure 4 illustrates ammonia nitrification as a function of influent BOD_5 concentration. It is quite evident that removal decreases rapidly when BOD_5 concentration exceeds 35 mg/1. This represents a confirmation of information held during the design phase in 1975 that the converse was true (that high degrees of <u>nitrification</u> would occur in the region of 30 mg/1 BOD_5).

Figure 5 illustrates NO_2-N and NO_3-N removal in the denitrification unit process as a function of loading to the submerged Bio-Surf shaft. The data is again contrasted with pilot data on municipal waste (Dots are full-scale, squares and triangles are pilot scale.) From Figure 5 it is clear that concentration is not a factor, as 90 percent removal is attained at comparable loading rates despite concentration differences of two or four-fold.

Figures 6 and 7 provide an illustration of supplementary data with respect to nitrification and denitrification. Samples were taken from the individual four stages of treatment in each case and various analytical parameters are plotted to show the fate of each during progressive degrees of treatment. These data are somewhat limited in that samples were taken on only one day in each case. However, the data is believed quite representative.

It is evident from Figure 6 that nitrification is nearly complete after only two of the four stages of treatment. That some denitrification was achieved is evident from the total nitrogen line which shows a significant decrease occurred during nitrification. This is not surprising, as other Bio-Surf studies have shown that simultaneous nitrification-denitrification reactions occur in the early stages of treatment when soluble BOD is available as a carbon source.

Similarly, Figure 7 shows denitrification is essentially complete by the second stage of treatment. Dissolved oxygen concentrations historically are zero in Stage One and thus support the small amount of denitrification that is shown to occur there. These and other stage data confirm that residual dissolved oxygen entering the first stage with the NO_3-N is preferentially consumed by the heterotrophic bacteria and that not until Stage Two is NO_3-N the sole source of oxygen.

The Miyazaki plant demonstrated its ability to perform under high hydraulic overload on November 16, 1977. Heavy rains resulted in a flow of 534 m^3/D (0.141 MGD), or 152 percent of the design maximum flow of 350 m^3/D. Despite this hydraulic overload, effluent BOD_5, NH_3-N, NO_3- N and TSS values were 5, 0, 0 and 10 mg/l, respectively.

Case No. 2 - Yokkaichi City

This is a similar application for nitrification and denitrification at a garbage dump, although both hydraulic and nitrogen loads are significantly higher (See Table 2).

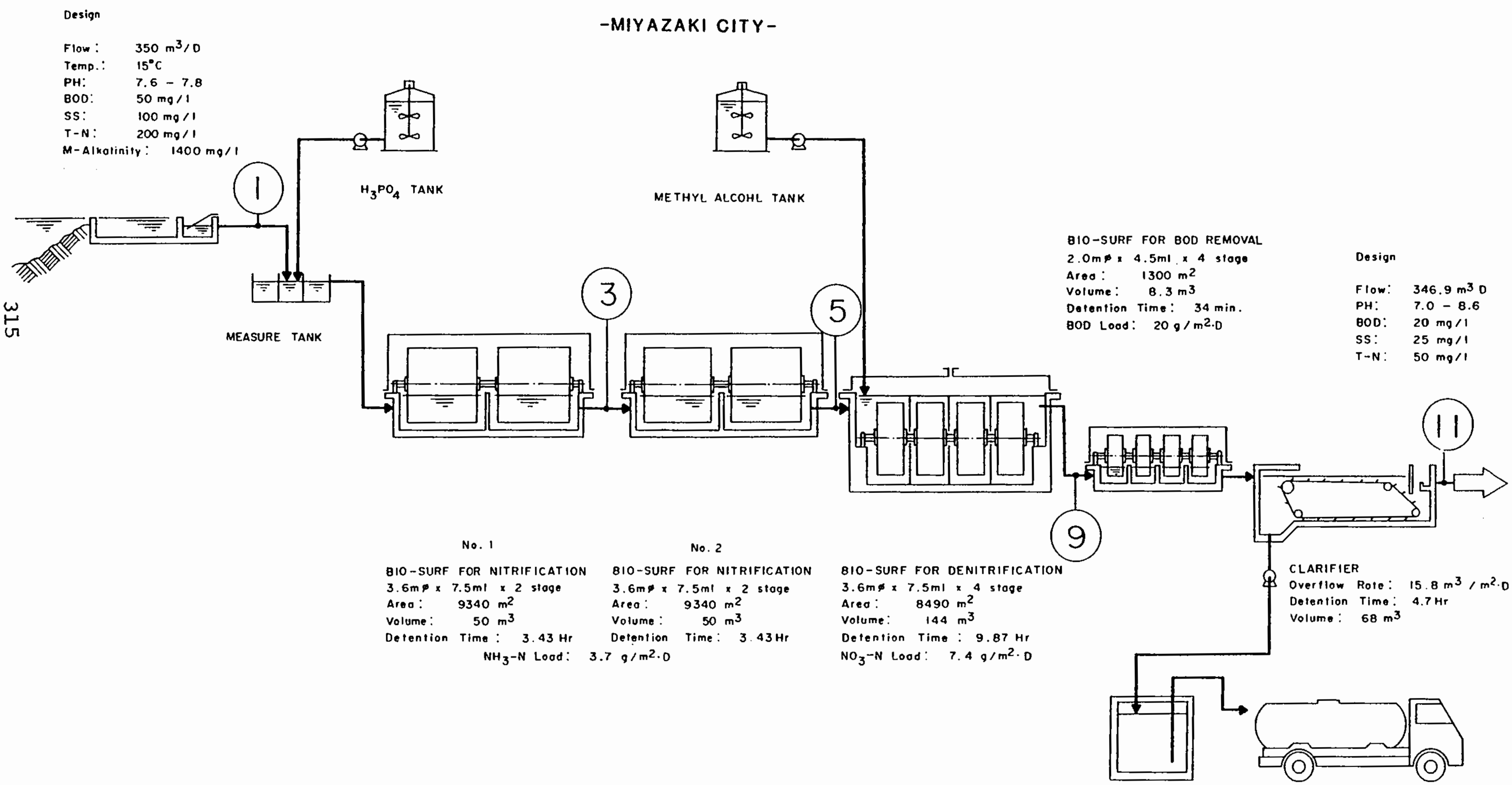
Experience gained at Miyazaki City and elsewhere resulted in a new design rationale to minimize or eliminate entirely the methanol requirement as a carbon source for denitrification. Bio-Surf nitrified effluent at this installation will be recirculated and combined with raw sewage in a submerged Bio-Surf reactor. It is predicted that a recycle ratio of 3 parts of nitrified effluent to one part of raw sewage will essentially eliminate and significantly reduce the need for methanol.

The installation consists of two systems in series, with the first half employing two fully-submerged, anoxic reactors followed by five standard submergence aerobic reactors for the nitrified effluent recirculation scheme. The latter half of the installation has a fully-submerged anoxic shaft followed by a standard submergence aerobic reactor for reaeration and BOD polishing. Table 6 summarizes design loadings and Figure 8 illustrates the process schematic for this plant designed to produce a final effluent with zero concentrations of BOD, NH_3-N and NO_3-N.

FIGURE 1

PROCESS SCHEMATIC AND SAMPLING POINTS

-MIYAZAKI CITY-



START-UP PROFILES-MIYAZAKI CITY

FIGURE 2

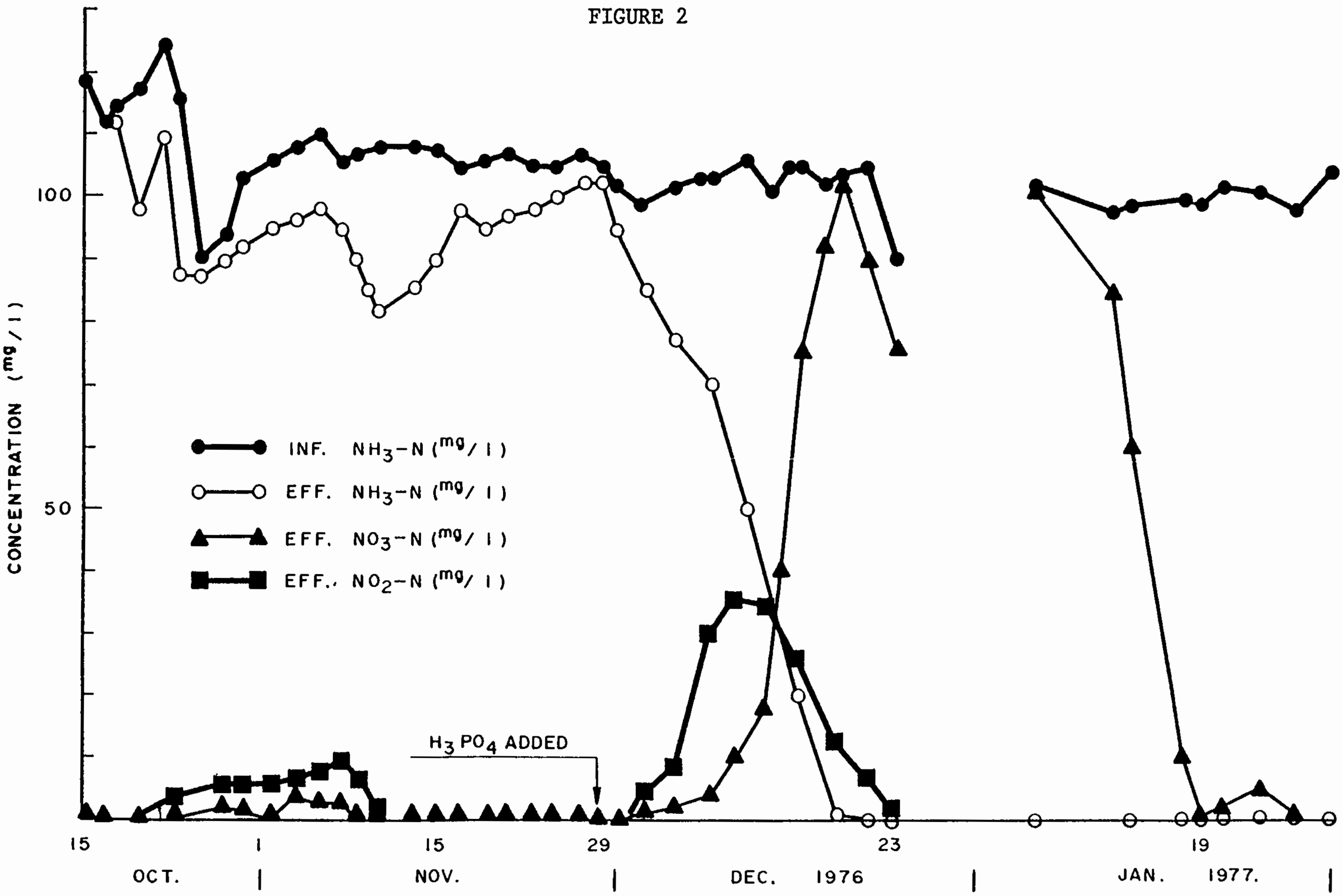


FIGURE 3

BIO-SURF NITRIFICATION

MIYAZAKI CITY VS. MUNICIPAL PILOT DATA

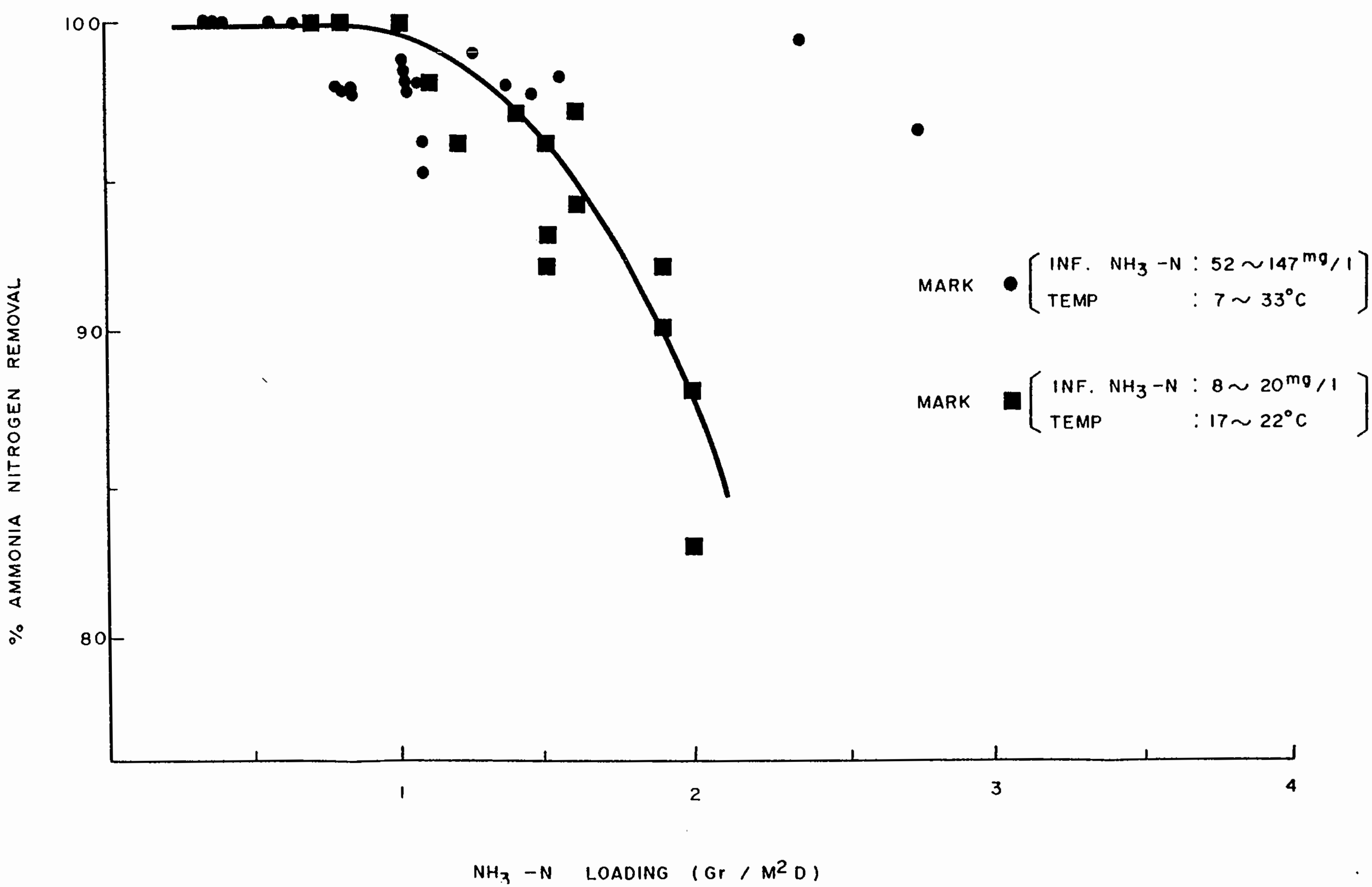


FIGURE 4

AMMONIA REMOVAL VS. TOTAL BOD

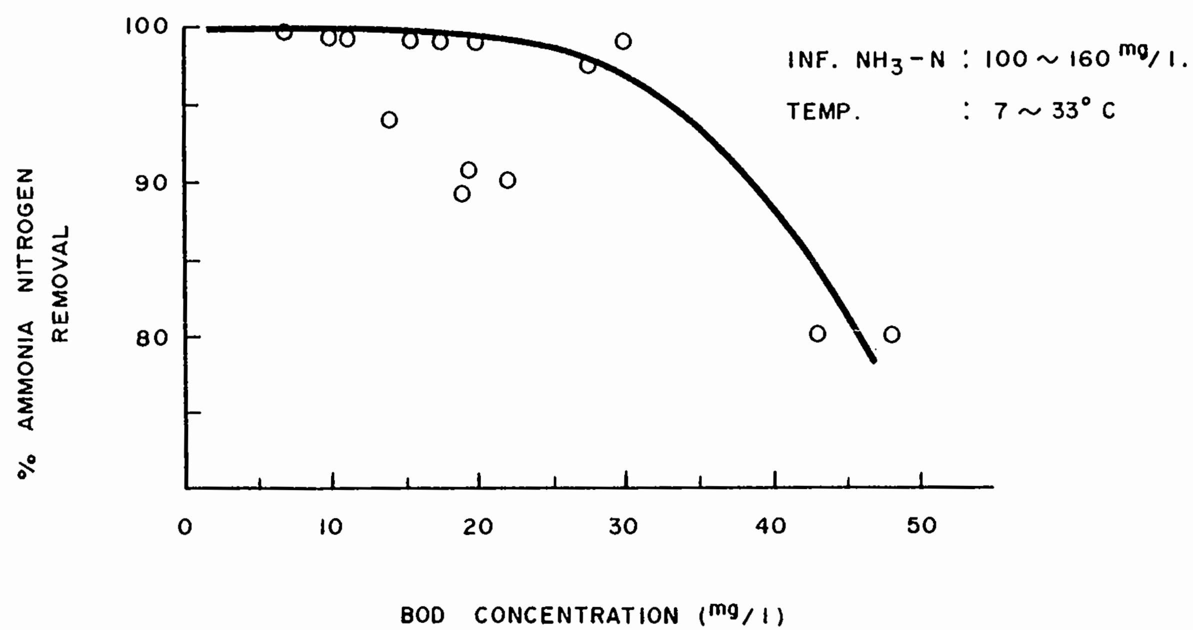
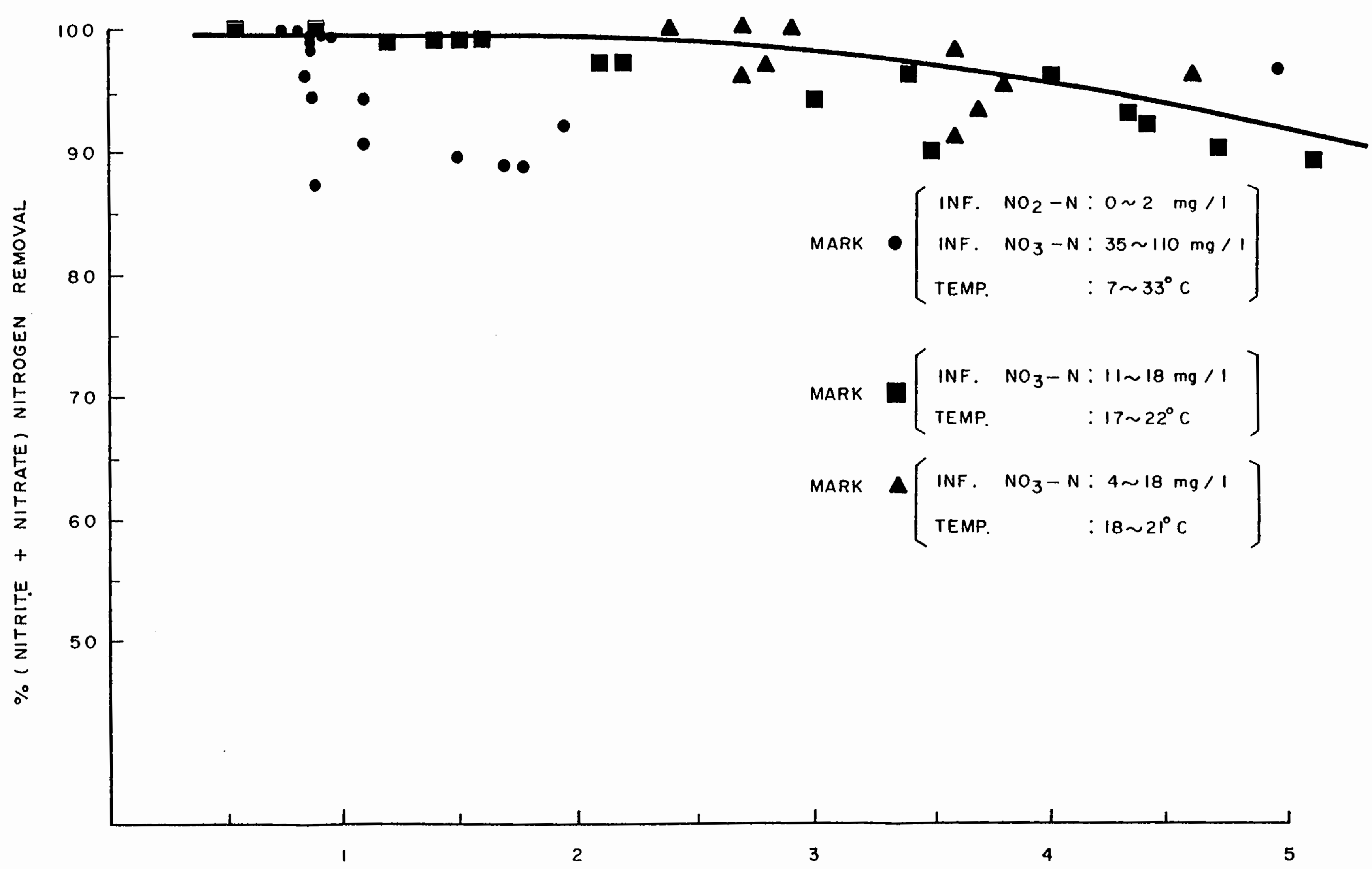


FIGURE 5

BIO-SURF DENITRIFICATION

MIYAZAKI CITY VS. MUNICIPAL PILOT DATA



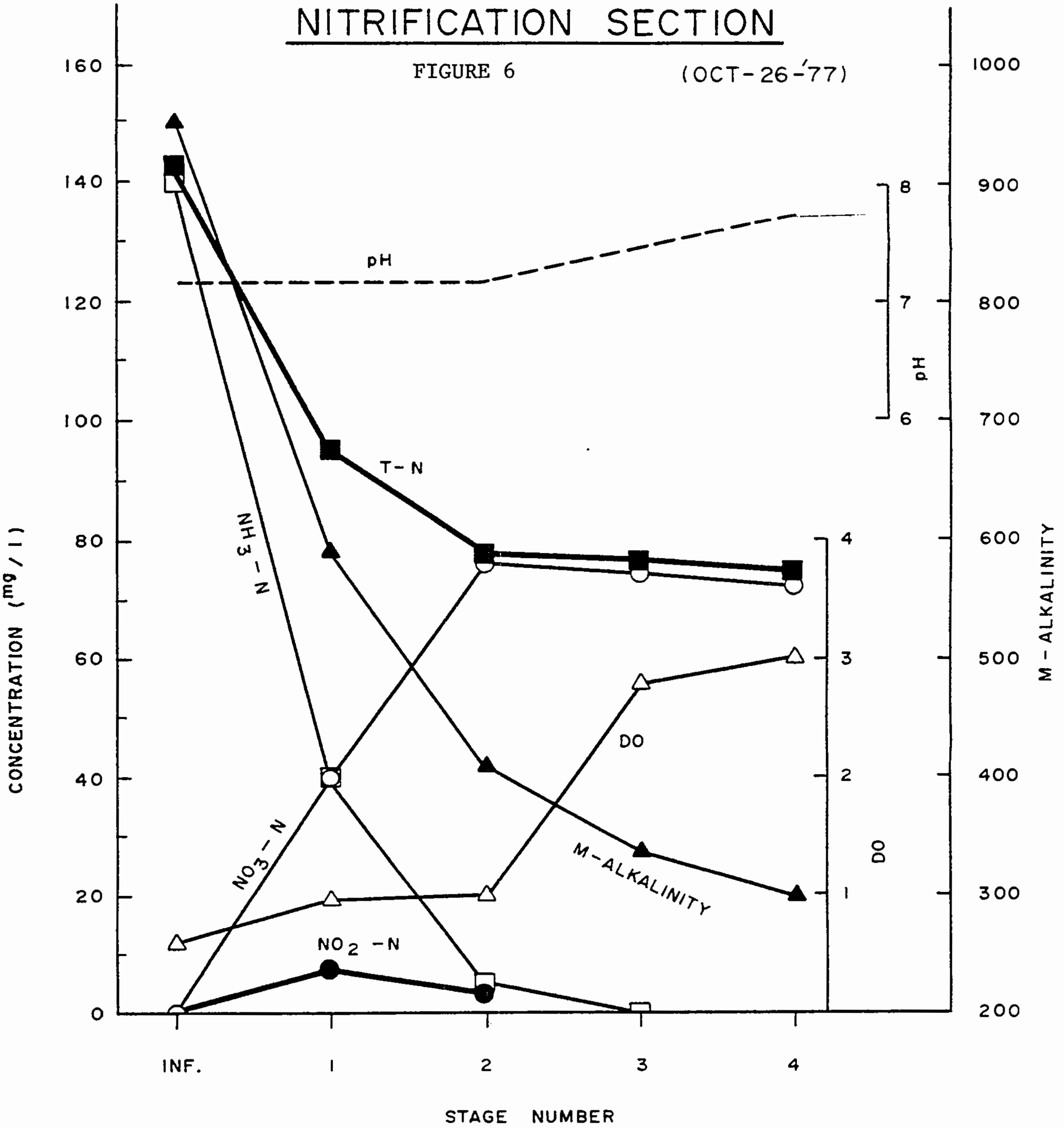

NITRIFICATION SECTION
FIGURE 6
(OCT-26-'77)
CONCENTRATION (mg/l)
160
140
120
100
80
60
40
20
0
INF.
1
2
3
4
STAGE NUMBER
pH
T-N
NH_3-N
NO_3-N
NO_2-N
DO
M-ALKALINITY
8
7
6
4
3
2
1
1000
900
800
700
600
500
400
300
200

FIGURE 7

STAGE DATA–MIYAZAKI CITY

DENITRIFICATION SECTION

(NOV-9-'77)

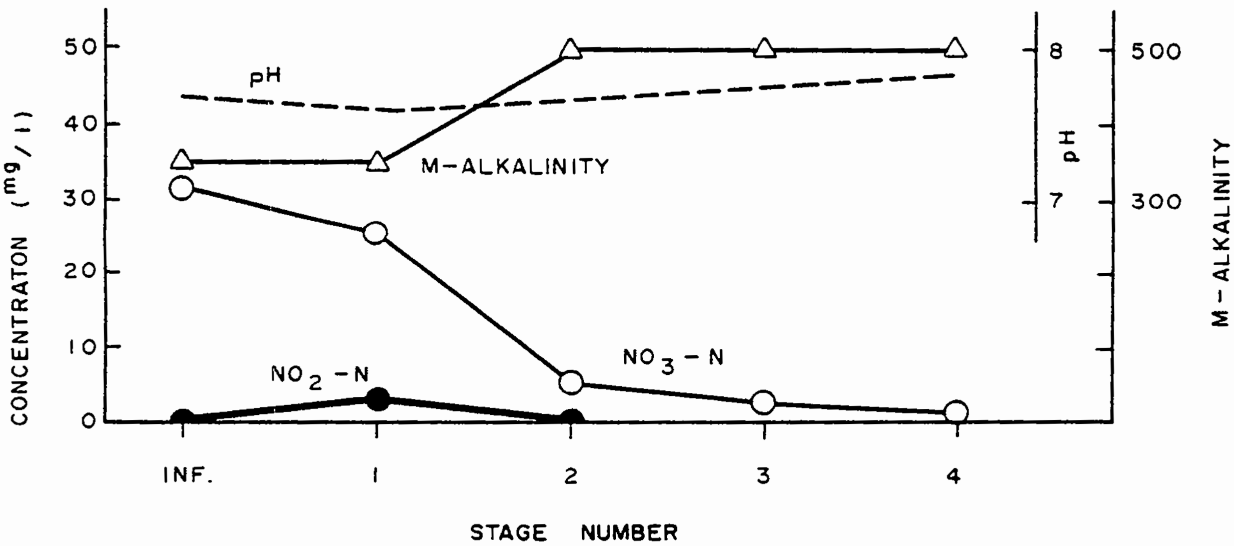
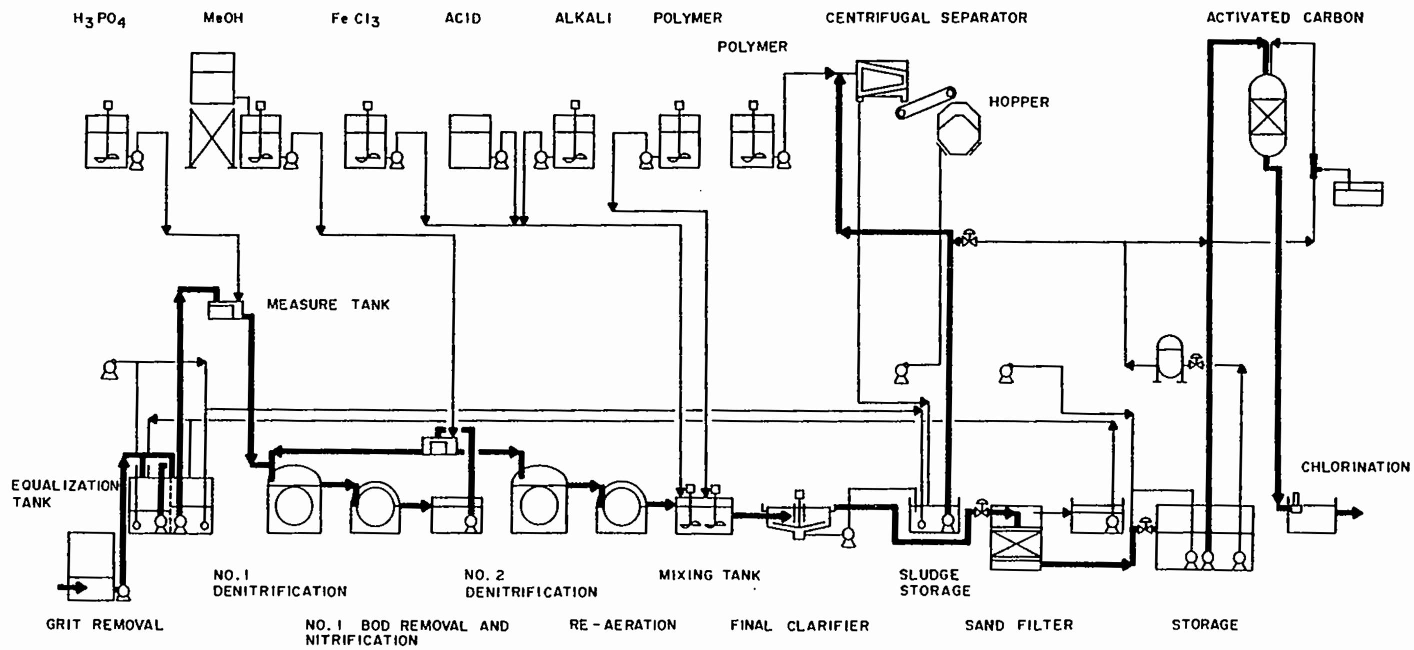


FIGURE 8

FLOW DIAGRAM OF YOKKAICHI CITY GARBAGE WASTEWATER TREATMENT PLANT



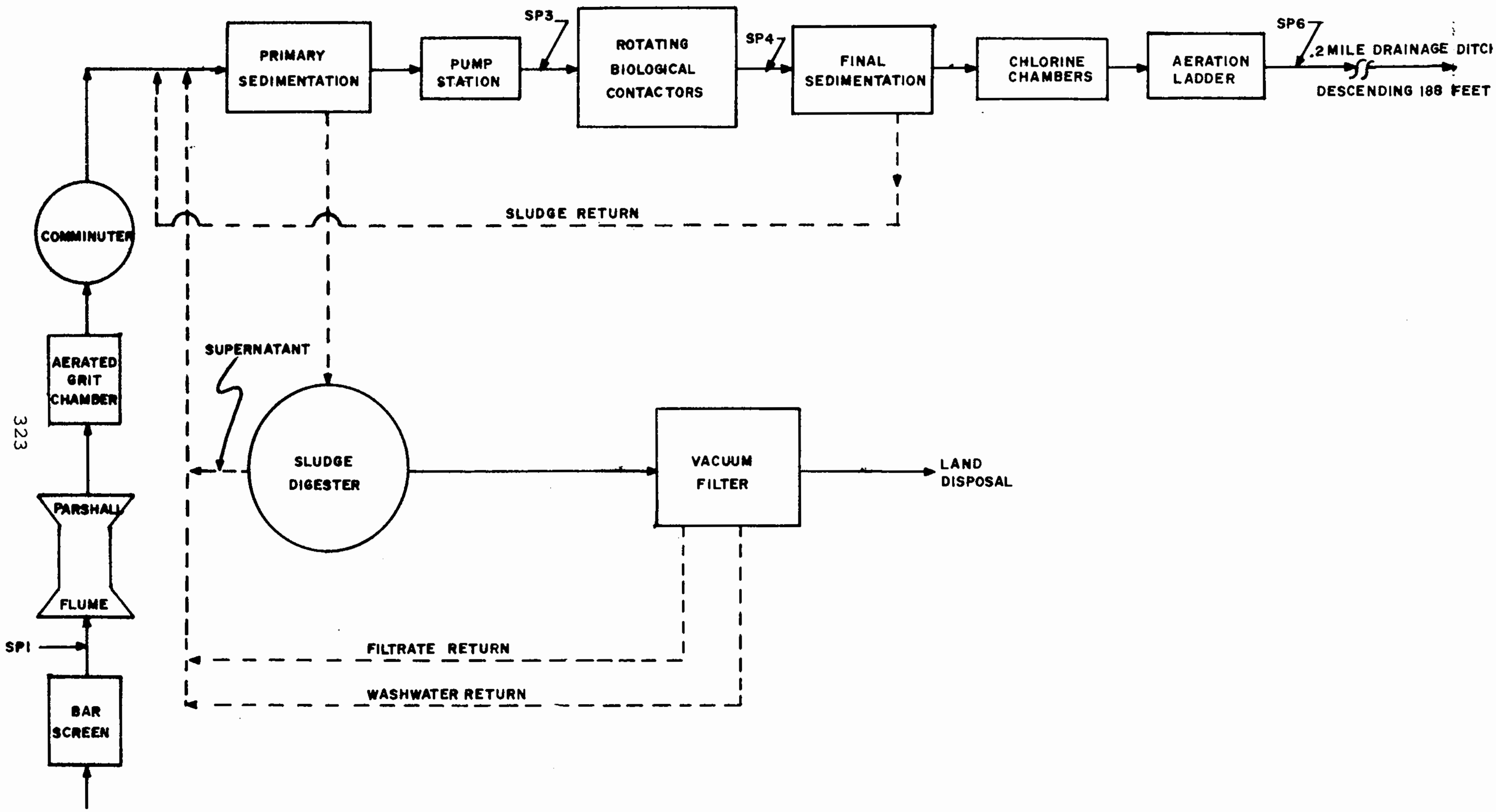


FIGURE 1. SEWAGE TREATMENT PLANT FLOW DIAGRAM AND SAMPLE POINT(SP) LOCATIONS

Table 6

DESIGN PARAMETERS - YOKKAICHI CITY

LOADING RATES

A. NO. 1 DENITRIFICATION

BOD_5:	17.4 g/m^2-D (3.57 #/1,000 ft^2-D)
Hydraulics:	140 1/m^2-D (3.45 gpd/ft^2)
Methanol:	None
NO_3-N + NO_2-N:	6.3 g/m^2-D (1.29 #/1,000 ft^2-D)
Time:	0.9 Hours

B. NO. 1 NITRIFICATION

BOD_5:	1.7 g/m^2-D (0.35 #/1,000 ft^2-D)
Hydraulics:	50 1/m^2-D (1.23 gpd/ft^2-D)
NH_3-N:	3.0 g/m^2-D (0.62 #/1,000 ft^2-D)
Time:	2.6 Hours

C. NO. 2 DENITRIFICATION

BOD_5:	13.6 g/m^2-D (2.79 #/1,000 ft^2-D)
Hydraulics:	76 1/m^2-D (1.87 gpd/ft^2-D)
Methanol:	108 kg/D (238 #/D)
NO_2-N + NO_3-N:	4.5 g/m^2-D (0.92 #/1,000 ft^2-D)

OPERATIONAL ADVANTAGES OBTAINED
BY INCORPORATING A BIO-DRUM^{TM}
IN AN ACTIVATED SLUDGE PROCESS

By

GEORGE R. FISETTE, P.E.
Product Manager

Ralph B. Carter Company
192 Atlantic Street
Hackensack, N. J.

Introduction

In the continual search for better methods of wastewater treatment, several processes have been developed in recent years that are vast improvements over the basic trickling filter and activated sludge process. These new methods have included pure oxygen activated sludge, static tube aerators, rotating biological contactors and various types of improved media for trickling filters. The most recent addition is a process that obtains a synergistic benefit from combining the activated sludge process with a rotating biological contactor (RBC).

This process concept was developed in Denmark during 1972. There are a number of Bio-Drum^{TM} systems operating in Denmark, Southeast Asia, Saudi Arabia, Canada, and the United States.

The Bio-Drum device consists of a steel cylindrical cage 8 feet in diameter by 8 feet long, a media within the cage consisting of polyethylene balls, a drive system, and mounting arms with bearings to enable multiple drums to be connected together on a single drive system. The cage consists of a structural metal framework around which is attached a perforated screening used to retain the balls. The end plates of the cage are designed to move inward to compress the balls in order to hold them in a rigid, locked position. The balls are molded of high density polyethylene to which carbon black has been added for ultraviolet protection. The drive system consists of a two speed motor attached to a gear reducer which transmits power to a chain and sprocket assembly. Multiple drums can be connected together by means of couplings and solid-lube, pillow block bearings. The bearings selected never need to be lubricated and are designed to operate in a water environment. The drums and drive assembly are designed to go together in one, two or four drum assemblies. In this manner, a single drive motor can rotate up to four drums on multiple axes. FIGURE 1 is a photograph of a Bio-Drum.

In addition to the polyethylene balls within the drum, there are installed around the perifery of the drum polyethylene containers known as "waterlifts". These waterlifts are part of the pumping and diffuser capability of the Bio-Drum.

While the utilization of fixed film media with the activated sludge process is not a new concept[1, 2], the Carter Activated Biofilm Method™ (CABM™) is the first process that effectively combines them in a single basin. FIGURE 2 is a schematic of the CABM process, see also TABLE 2, a preliminary design for a 2.5 MGD Bio-Drum CABM system. This rotating biological contactor process makes use of sludge recycle and a high strength suspended solids culture. The basic process concept is to obtain a very high rate biological removal process utilizing an extremely large bacteria inventory in order to keep detention time and basin size to a minimum. The typical hydraulic detention time is on the order of one to two hours. This results in a basin 1/4 to 1/10 the size of that for conventional activated sludge systems. It is the large inventory of bacteria obtained from both the suspended culture (MLSS) and the biofilm on the polyethylene balls that results in an overall design utilizing a minimum of equipment, land area, and basin volume. Typical MLSS concentrations in the Bio-Drum basin are on the order of 5,000-10,000 mg/l. Combined with the bacteria slime on the Bio-Drum balls, a total effective suspended solids concentration of 15,000-25,000 mg/l is obtained. Since the energy input of the Bio-Drum is into a very small basin, one obtains a process with a high degree of efficiency since the drive motor of a dual Bio-Drum is only 5 horsepower.

The Bio-Drum, utilizing the CABM process, is a high rate, short detention type process that is extremely stable to shocks, peaks, toxic compounds, and is extremely efficient[1, 2, 3, 4] due to the combination of fixed film and suspended growth media. The low horsepower, typically 10-15 horsepower per MGD, provides for a low cost operating system. In addition, the simple design of the Bio-Drum further minimizes routine maintenance and operating cost. The Bio-Drum due to its floating nature, greatly simplifies both construction cost and installation cost. Indeed, it can be easily retrofitted to existing basins and lagoons with minimal changes to existing structures. A further and highly significant advantage of a Bio-Drum system is its ability to provide for both BOD and ammonia removals in a single basin in a single stage process. This ability to obtain ammonia removals in a single stage system further reduces the amount of capital equipment, operating processes, and land required to meet present and future effluent standards.

MECHANICAL DETAILS

As previously described, the Bio-Drum consists of three major components--the cage, the polyethylene balls, and the drive assembly. The drum (cage) consists of an angle-iron sub-assembly that is designed to support the weight of the completed drum. To this sub-assembly the covering of expanded metal (screening) is attached. The two circular end plates for the drum are designed on a tracking system to move in or out approximately 6 inches each. These movable end plates are advanced by means of bolts at numerous pressure points around the cross-sectional area. The structural metal is first wire brushed then the completed drum and covering is given a finished coat of a rust inhibitive paint. The ping-pong sized (38 mm) balls are manufactured out of high density polyethylene to which carbon black has been added for UV (ultraviolet) protection. These balls are produced by an injection technique known as blow-molding. This results in a ball with minimum seam and no perforations. During manufacture, each individual ball is tested to be sure there are no holes or leaks in any individual ball. Approximately 1/4 of a million are included in each 8 foot diameter by 8 foot long cage. In addition to the balls, polyethylene containers of approximately one liter each are attached to all of the periferial structural members around the outside edge of the cage. These waterlifts are attached to the structural angle iron by means of plastic ties. FIGURE 3 shows a waterlift in use. In order to show detail in the photograph the lift is inside a small box to keep the balls back. The box does not exist in a working drum. The center axis of the cage consists of a thick walled 8 inch diameter pipe. Flanges are then welded to each end of the pipe for attaching the drum to the drive component or companion drum assemblies.

The same basic drive assembly is used to power one, two, or four connected drums. The drive assembly consists of a two-speed motor, a gear reducer, and chain and sprocket sub-assembly. The 2-speed motor is either 3, 5, or 7 1/2 Hp as necessitated by the number of drums, and runs at 900/1800 RPM. The motor is normally operated in 900 RPM condition with the higher speed being saved for temporary process conditions. The gear reducer is a Koellmann gear reducer with a 43 to 1 reduction ratio. This gear reducer operates in a flooded hydraulic condition using standard greases and lubricating oils. It comes with a sight tube for continual monitoring of the oil level within the gear reducer. The chain and sprocket sub-assembly consists of a 5" drive sprocket mounted on the gear reducer and a 39" master sprocket mounted on the companion flange of one of the drums. The chain is a heavy duty roller style chain with a master link. The supporting structure for the Bio-Drum consists of a torsion-tube yoke which attaches to the basin walls or other permanent structure. The drum axis is attached to the support arms by the means of solid-lube pillow blocks. These bearings are designed to operate in a flooded and wet environment. The bearings do not need any lubrication, and as they wear the bearings inner surfaces can be rotated to increase the life of an individual bearing. Multiple drums are connected together by a coupling which allows for some independent movement of the drums relative to each other. All drums are connected to the common drive system. FIGURE 4 shows a single drum with its mounting yoke and drive system.

When the Bio-Drum assembly is installed and operating, the manufacturer supplies an initial startup service that consists of verifying the proper wiring and rotational direction of each drum assembly and provides instructions on the simple adjustment of the end plates to maintain the balls in the state of proper compression. It is to be expected during the first several months that the end plates would have to be adjusted several times as the balls work into their final permanent locations. After this initial shakedown period, no further service or attention need be given to the balls or drum structure itself.

The only routine maintenance that is required will be for checking and lubrication of the gear reducer. It is advisable to have on hand as spare parts several master links and several feet of chain in the eventuality that a chain breaks or is damaged. Routine maintenance to the chain and sprocket assembly would consist of every 6 to 12 months inspecting the chain and sprockets for wear. It is recommended that at least every two or three years the drive sprocket and chain be automatically replaced as a preventive maintenance procedure.

Under normal and anticipated operating conditions and environment, it will not be necessary to provide a building or covers for the Bio-Drum assembly. However, it may be advantageous for process and operator convenience to have covers over the Bio-Drum assembly area. Due to the packed nature of the Bio-Drum and the fact that the major volume and bacteria film inventory is within the drum itself, sun, rain, and wind have a very minor effect on the growth of the biofilm. It must be remembered that the waterlifts are purposely designed to pour water over the balls and biofilm when they are exposed to the atmosphere. Therefore, rain will have less effect than the falling water built into the design of the drum. In cold environments, because of the short detention time within the bio-basin, there is a very small temperature change experienced in this area. As long as the incoming waste is sufficiently above the freezing point, the small amounts of ice that may form around the edge of the basin are not detrimental to the mechanical operation of the Bio-Drum. In addition, operating experience and past history show that since there is no splash, spray or foam with a Bio-Drum system there is very small tendency for ice to accumulate on any of the super-structure. However, it must be remembered that cold water temperatures are detrimental to the biological process and, in fact, result in a significant slowing (retardation) of the biological process. It may be for this purpose then, advantageous to provide covers to retain heat within the biological system.

PROCESS DETAILS

The 2-speed drive system is incorporated into the Bio-Drum for when a biological process is subjected to upsets. Whether these upsets be brought about by the normal durinal cyclic loading or by BOD peaks from industrial wastes or by infiltration, the higher second speed is available to increase the overall performance of the biological system. By increasing the rotational speed of the drum, it is possible to obtain greater oxygen transfer, up to 4 times more, greater mixing, and more rapid intermixing of the incoming BOD with the fixed film bacteria. FIGURE 5 shows the effect of recycle upon the operating performance and BOD removal of a rotating biological contactor, specifically the Bio-Drum. To date most rotating biological contactors have operated in a plug flow mode without the benefit of solids recycle. As a matter of fact, previous testing and research done by both manufacturers and independent consultants had determined that a RBC system received no benefit from the recycle of suspended solids[5,6]. Those RBC systems presently using sludge recycle all require air diffusers and accessory blowers. However, due to the mechanical aerator capability of the Bio-Drum, the Bio-Drum does not need supplemental diffusers and does obtain a benefit from the recycle of suspended solids. Indeed, the benefit seems to be as a result of a synergistic combination of the two types of biological processes, fixed film and suspended growth. FIGURE 5 shows that with recycle considerably more pounds of BOD may be applied to a given cubic foot of media or conversely, for a fixed BOD waste considerably less capital equipment is required to obtain the same degree of removal.

While the Bio-Drum is capable of sustaining a suspended culture (MLSS) of 2,000 to 10,000 mg/l, economics dictate that the concentration of MLSS be about 7,000 mg/l. However, if there are external restraints; such as existing tank size, clarifier efficiency, etc; then the CABM process can be designed to operate at lower MLSS levels.

The quantity of bacteria on the Bio-Drum media can be calculated by noting the change in the flotation level of the drum as the film accumulates, its bouyancy, and measurements of the film thickness and moisture content. If all of the biofilm were removed from the Bio-Drum and placed in suspension in the basin an effective MLSS of 15,000 to 25,000 mg/l would be obtained.

A further benefit of the combination of fixed film and suspended growth is the ability of the Bio-Drum with the CABM process to obtain removals of both BOD and ammonia in the same single step basin. It is theorized that this ability, not obtainable with other conventional RBC equipment, results from the potentially older sludge age of the bacteria in the biofilm attached to the drum itself. The recycle suspended solids are considered to consist mostly of carbonaceous bacteria used in the removal of BOD. The ammonia removing bacteria *(NITROBACTER and NITROSOMUS)* are considered to be attached to the bio-ball media. Because of the slower growth rate of these nitrifying bacteria, they have a requirement of a considerably older sludge age than the more rapid growing carbonaceous bacteria. A logical explanation then, for the ability to remove ammonia at the same time as BOD requires that there be a separate inventory of nitrifying bacteria, independent of the recycle stream which would typically have a relatively young sludge age.

It has been alluded that the Bio-Drum operates as a mechanical aerator. Several unusual features have been incorporated, or are a natural part of the design of the Bio-Drum. It is these features and design benefits that allow the Bio-Drum to operate as a mechanical aerator. The most obvious feature is the waterlifts which are attached to the external periphery of the cage. There are approximately 144 of these lifts in each drum. These lifts are capable of providing a direct water pumping rate of 100 gallons per minute per drum at the normal rotational speed (high speed is 200 gpm). This means that a single Bio-Drum is capable of providing a complete hydraulic turnover of the water within its domain in less than 1 1/2 hours by direct pumpage only. This does not take into consideration that movement of water brought about by rotation of the drum structure itself; nor intrained flow. Thus, a single drum is pumping by direct water movement approximately 150,000 gpd. As this pumped water cascades down over the media during the atmospheric portion of each cycle, oxygen is transferred from the atmosphere into this falling, cascading sheet of water. Of course, the normal oxygen transfer as experienced by rotating biological contactors takes place during the atmospheric portion of the rotation from the air directly to the bacterial slime adhered to the media. The oxygen transfer, therefore, into the wastewater is above and beyond that needed for the bacterial slime attached to the media. Secondly, as these lifts, now empty of water, are resubmerged into the wastewater, the air that is trapped within bubbles out of the lifts up through the media. In this manner, the lifts are acting as a coarse bubble diffuser. TABLE 3 details the oxygen transfer of various RBC's and diffusers. The direct air pumpage by the lifts is 13 CFM, or 19,000 cubic feet of air per day per drum. As this air flow passes up through the media, oxygen is transferred into the wastewater as the bubbles are continually broken and re-agglomerated during their travel through the rotating media. In addition to the oxygen transfer and contacting provided by the waterlifts, a considerable portion of oxygen transfer and energy input into the basin takes place as water and air flow into and out of the drum during each rotation.

The Bio-Drum is designed to rotate at 2.6 rpm, giving a peripheral speed of 1.1 ft. per second at the normal condition. The high speed condition used for special process conditions is at 5.2 rpm, giving a peripheral speed of 2.2 ft. per second.

As the drum rotates the water is continually falling out of the back side of the drum, from within the media back onto the water surface, giving a wake and a small wave. This additional agitation of the water and splashing action within the drum again results in oxygen transfer to the wastewater. Because of the many sources of oxygen transfer and the mechanical design of the Bio-Drum, the system is able to support a high concentration of mixed liquor suspended solids. The oxygen requirement for endogenous respiration and for BOD oxidation is provided by the drum for the portion of the BOD removal that takes place by the suspended culture.

Due to the extremely high concentration of bacterial solids, approaching 15,000-25,000 mg/1, a very rapid assimilation of the BOD takes place. Indeed, the biological processes are known to be completed within approximately 15 to 20 minutes. The major requirement for the longer hydraulic detention time as used within the Bio-Drum CABM design of 1 to 2 hours is brought about by the necessity to protect the design for shock loadings, for hydraulic peaks, and for the nonideality of the intermixing of the bacteria with the incoming BOD.

The basin size is primarily a function of hydraulic flow rate and such associated factors as peaking, hydraulic cycles, BOD, and temperature fluctuations. The basin conventionally is designed with a depth only slightly greater than 1/2 the diameter of the drum, (i.e., 4.5-5 feet water depth). The clearance between the bottom of the drum and the basin floor is kept as small as possible within the restraints of the design. The basin is normally designed in a fairly square to slightly rectangular pattern in order to optimize the completely mixed nature of the process. Conventional wisdom dictates that for most waste treatment plants, a parallel scheme be incorporated whereby two parallel biological basins are provided, each with a capacity slightly greater than half the hydraulic design.

An important consideration to the proper operation of the CABM system is the ability of the Bio-Drum to keep the mixed liquor suspended solids in uniform suspension. At the concentration levels normally encountered of 5,000-8,000 mg/1, it is necessary that sufficient energy be put into the wastewater to prevent separation and deposition of the solids. Even though the total operating horsepower of a Bio-Drum system is quite low, the applied horsepower, i.e., the energy imparted directly into wastewater within the basin, is quite high. Under normal design considerations, the energy applied directly to the wastewater by the rotating drum system is in excess of 2.5 hp./k ft^3. In addition, the higher speed capability of the drive system allows this value to be raised in those circumstances where a faster settling solids or a higher density solids is obtained. Furthermore, the clearance between the bottom of the drum and the basin floor is kept to a minimum in order to maximize the scouring velocity along the bottom of the basin to resuspend any solids that have a tendency to settle towards the bottom of the basin. The high direct pumping capacity of the drum itself of approximately 150,000 GPD, greatly contributes to the ability of the total Bio-Drum CABM system to maintain uniform solid suspension. In a basin with an average hydraulic detention time of 1 hour, the capability of the Bio-Drum by direct pumping alone to turn over or recirculate the entire basin volume in approximately 1.5 hours provides additional evidence of the level of agitation within the Bio-Drum basin.

A second and equally important constraint is that the Bio-Drum exhibits no tendency for internal clogging by excessive bacterial growth, sloughing off of excess growth, or developing anaerobic areas within the internal volume of the drum itself. Two features of the Bio-Drum mechanical design were intentionally incorporated to alleviate this problem. However, the proof of the drum's ability to perform without clogging and without going anaerobic is best obtained from evidence of operating units. The units that have been installed and operating for 3 to 4 years have shown no tendency to clog as evidenced by the <u>lack</u> of foul odors, or a decrease in the performance rating of the drum system with time. If the internal area of a drum were to clog or go anaerobic, or otherwise decrease the performance capability of the Bio-Drum, then a deterioration should be noted in the quality of the product produced by the treatment plant. To date there has been no evidence of a decrease in performance, rather the system appears to increase in removal ability with time. The rotational speed of the drum, combined with the small interstitual spaces, means that the water velocity around the balls and through the spaces must be very high. The size of the balls was purposely selected in order to arrive at the minimum practical void space of 35%. In this manner, the Bio-Drum design maximizes the surface area available for bacterial film to attach and maximizes the velocity of the water through the open spaces within the packing. Combined with these high internal water velocities, is the scrubbing effect obtained from the rising air bubbles released by the waterlifts. These rising air bubbles provide an expanding action or scrubbing action that, in conjunction with the high water velocity, maintains a uniform thickness of biofilm on each ball. These released/removed bacteria particles are removed as discreet and small particles within the flowing water stream as the drum rotates. In this manner, large chunks of matter as typically slough off a trickling filter have been prevented. Furthermore, during the atmospheric portion of each rotational cycle, the bacterial film is further rinsed with a gentle irrigation of water released from the waterlifts. This gentle flow further tends to remove any loosely attached and light particles as it flows down over the balls during the rotational journey. It should also be realized that any plugging tendency which is exhibited becomes self-defeating. In that as a particular area would tend to plug, the interstitual velocity in that area would increase and the excess materials would be rapidly removed due to the higher velocity of water within that section of the drum.

Since it is conceivable to use Bio-Drum in a treatment plant that does not have a primary clarifier, consideration must be given to preventing clogging or plugging of the drum screen by large foreign objects. Such things as rags and plastic bags, etc., must be removed from the drum or prevented from reaching the drum in order to prevent an aesthetic problem, a potential odor problem, and some interferences with the hydraulic flows within a drum. Therefore, it is recommended that in those applications without a primary clarifier, that a bar screen of some design and a grinder should be utilized in order to reduce the size of the extraneous objects and their potential for adhering to the drum screening. In this manner, since the drum screen itself is very smooth and presents a small profile for string-like objects to hang on, the tendency of the drum to have string or cloth attached to its outside surface should be minimized.

PROCESS DESIGN

As with all process equipment, a method of calculation or determining the size and quantity of equipment necessary to process a given load is necessary. While the theoretical equations have been derived for both the activated sludge system and for biofilm devices, at present there are no known equations describing the combined system. Therefore, the Bio-Drum and CABM system are designed on a semi-empirical basis as follows.

Based upon both pilot and full-scale testing and operating plants, a relationship has been determined which related the applied BOD, the cubic foot of drum, and the desired quality of effluent. The design parameter used to determine the quantity of drums required is a volumetric loading factor, F/V (lbs. BOD/day/ft.^{3}). This loading factor is used in the manner similar to the design parameter used for sizing trickling filters and plastic media-type filters. In conjuction with the drum loading parameter the F/M ratio (lbs. BOD/day/lb. MLSS or MLVSS as preferred), is used in an iterative (trial and error) scheme to arrive at the final design. A balance must be arrived at between the F/M ratio, the F/V ratio, and the basin size.

FIGURE 6 shows the interrelationships between BOD removals, F/V, and F/M. At a given F/V loading, the F/M must be decreased in order to improve performance. For a given BOD waste, F/M can be decreased by raising the MLSS concentration or increasing the basin volume.

The objective is to maximize the total solids inventory while maintaining the F/M concentration within a practical range that can be maintained by the Bio-Drum. This means that the typical F/M value obtained is directly related to the basin size and the concentration allowed, which ranges from 5,000 to 18,000 mg/l. The basin size is primarily governed by flow considerations. Some allowance in basin size must be considered for unusual BOD or ammonia requirements. However, the normal basin has a capacity of 1 hour hydraulic detention time. The upper range of the basin size is approximately 2 hours hydraulic detention time. Once the basin size has been established, then the MLSS concentration allowed can be determined and the F/M ratio calculated at this point. The F/M ratio so calculated must then be related to the F/V ratio to verify that the F/M is within the operating range at the F/V loading selected. TABLE 2 is a preliminary design calculation for a 2.5 MGD municipal waste.

Of prime consideration in the operation of the CABM process is the design and operation conditions of the clarifier and recycle flows. It is extremely advantageous that the recycle flow be kept to as small a value as possible in order to minimize the clarifier surface area required for proper sedimentation and solids separation. Therefore, the clarifier design selected must have a highly efficient separation capability and a high degree of thickening activity. It is imperative that the solids underflow be at the maximum percent solids concentration obtainable. With proper design of the clarifier and thickenings action, the recycle flow would be in the range of 10-100% of the inlet flow. Obviously, it is preferred to minimize the recycle flow in the range of 10-30%. This high concentration is obtainable even though the MLSS concentration is quite high, because of the improved settling characteristics of the sludge that results from the combined biofilm activated sludge system.[4]

In addition to these major factors affecting the design and process, there are several minor parameters which must also be considered in arriving at a polished design. In municipal designs, and even in some industrial applications considerations must be given to hydraulic peaks. These may be the conventional daily cyclic flows as evidenced by most municipal plants, or they may be process related flow peaks dependent upon the manufacturing processes of a specific plant, wherein washwater of batch-type processes are dumped at discreet time intervals. Furthermore, there are really two types of hydraulic peaks. In one case, the BOD concentration remains constant; in other words, the total load of BOD incoming to the plant increases in direct proportion to the hydraulic flow. In the other case, the pounds loading of BOD incoming to the plant remains essentially constant and we basically have dilution of the normal flow by increased hydraulic input. In evaluating the effect of these hydraulic flows on a Bio-Drum design, if the duration of the peak is of the order 1/2 to 1 1/2 hours, little or no changes are necessary in the basic design for a Bio-Drum system. On the other hand, if we are looking at hydraulic surges that may run 2 hours or more, then consideration must be given in either case to the effect of these flows on the Bio-Drum process. If the increased BOD loading during hydraulic peaks represents a significant portion of the day, then the plant will have to be oversized to allow for proper removal of these increased loadings. In the case of dilution or infiltration problems, the major constraint will be on the design and operation of the clarifier and recycle streams, to prevent the washout of biological solids. However, in either case the fact that both a fixed film culture and a suspended culture are available for treatment means that the effects of these hydraulic peaks may be rapidly neutralized.

If the plant is experiencing an upsurge in BOD, this can be offset and performance maintained by increasing the amount of recycle and changing the drum to its higher speed of operation. The system is now capable of providing additional oxygen and suspending additional solids to treat this surge in BOD. In the condition where one obtains a hydraulic surge without corresponding BOD increase, the optimal alternative is to increase sludge recycle only slightly and maintain a close watch on the clarifier performance such that excess solids do not wash over the top and to prevent an excessive buildup of sludge and a sludge blanket. It is also possible to obtain surges in BOD without corresponding changes in the hydraulic flow. Uner this condition, it is only necessary to increase the solids recycle and increase the drum speed to provide the additional activity necessary for removal of the increase in BOD. Again, consideration must be given in the normal design to the magnitude of this increase relative to the average design for which the plan was originally designed.

A somewhat different type of peaking problem is that experienced in small municipalities and industries which operate on a 5 day or 6 day cycle, namely the lack of flow on weekends or for other extended periods. With a conventional rotating biological contactor there is an extreme tendency over these low periods for the rotating contactor to have such a high degree of evaporation that it essentially dehydrates the entire basin over the course of one or two days. The prevention of this occurring, therefore, requires these type of plants to be designed with a hydraulic recycle strictly to maintain sufficient hydraulic flow to keep the bacterial cultures wet and active. In the Bio-Drum and CABM process, however, the recycle flow is already incorporated within the design and due to the high suspended solids inventory there is always sufficient bacterial available to act as the food source for the fixed film culture. In other words, the process can go into an endogenous type of respiration or cannibalistic activity during these periods of starvation.

A more severe problem associated with peaks and spasmodic flows is the unfortunate occurance of a toxic dump or poison entering the system. Under these conditions, the advantages of a combined fixed film and suspened growth system shows its superiority to the operation of either type of singular system. If the system were a pure fixed film culture system, the toxic load would quite probably wipe out the entire plant and severely hamper removal and process efficiency. On the other hand, with a strictly activated sludge type culture it is possible that the sludge would never acclimate to the constituants of the toxic input and, therefore, each and every time a toxic cyclic occured the activated sludge plant would also suffer in performance capability. However, in the combined Bio-Drum CABM system the biofilm can adjust over time to the toxic material and provide a reserve bacterial inventory capable of treating the toxic component.

As mentioned previously, temperature, per se, does not affect the mechanical reliability of the Bio-Drum design. However, the lowering of temperature during the winter does severely hamper the biological processes. Therefore, the Bio-Drum system must be designed with consideration for this effect. If one wishes to maintain the same degree of removal in the winter as in the summer without the use of heat or covers to retain heat, then the Bio-Drum system must be oversized in order to produce proper removals during cold weather conditions. Because of the packed type nature of the Bio-Drum and the fact that the water surface and film surface on the media are not directly exposed to the atmosphere; the evaporation and heat loss from a Bio-Drum system is minimized in comparison to other types of biological processes. This means that in temperate climates where only occasional cold weather is experienced the Bio-Drum basin will not tend to drop its temperature below the freezing point as is common with other biological contactor devices.

COMPARISON TO OTHER PROCESSES

Inspection of TABLE 4 shows that the CABM process is quite similar in design perimeters to both the high rate activated sludge process and the contact stabilization sludge process. Comparison of the Bio-Drum to fixed film type processes, listed in TABLE 5, shows that on an organic loading per unit volume of media the Bio-Drum is, in fact, considerably higher than conventional trickling filter and RBC type devices. However, if one looks at the hydraulic load in gallons per day per square foot of media, the Bio-Drum device is loaded considerably less than conventional trickling filter designs and well below the hydraulic loading for conventional or improved processes. Just as loading perimeters and loading values are specific for individual processes and different from all others; the Bio-Drum and CABM have specific design and loading perimeters that relate only to the Bio-Drum design. However, if one puts the design perimeter of all processes on a comparable basis wherein the loading is measured in pounds of active bacteria, which is nearly impossible with fixed film devices or per pound of suspended solids (MLSS) then it can be seen that the Bio-Drum is really no different than any other biological process.

Another measurement or comparison of the various processes that can be evaluated is the energy requirement, i.e. horsepower, utilized in the removal of BOD and in the operation of the various types of processes. With a few conservative assumptions, the horsepower required to treat one million gallons per day of raw waste can be estimated for the various generic types of equipment; mechanical aerators, diffused aerators, submerged turbine aerators, rotating biological contactors, and Bio-Drum. TABLE 6, lists some of the values of horsepower required to treat one MGD of waste. The 15 horsepower shown for the Bio-Drum is considerably lower than the more conventional and expected numbers of 40 to 50 horsepower for various mechanical systems. However, these high numbers are really an indication of the inefficiency of the processes in that the greater the horsepower required to treat an equivalent amount of BOD the more expensive the annual operating cost is for an individual plant. Furthermore, a comparison of the applied horsepower, defined as that horsepower transmitted into the volume of water under treatment within a given basin shows that the Bio-Drum is within the range of "conventional" treatment processes. TABLE 6 shows these values of horsepower per thousand cubic feet of basin water under treatment. Part of the reasoning for the inefficiencies of the more conventional mechanical processes is brought about by the size of the basins involved under treatment which require large quantities of horsepower to maintain proper biological mixing and contacting.

SUMMARY

The difference between the Carter Activated Biofilm Method™ and other biological processes can be summarized in a few words; the same, but with IMPROVEMENTS. However, the Bio-Drum™ differs from other biological treatment equipment in being a more efficient device for removing BOD. The features and benefits of the Bio-Drum and CABM process are summarized in TABLE 1.

TABLE 1

BIO-DRUM

MECHANICAL

Feature	Benefit
floats	reduced installation costs, no supports required
compact 8' x 8' size	easy to install
low stress on bearings	long life
low stress on drive	long life
simplified design, few moving parts	low maintenance costs

CARTER ACTIVATED BIOFILM METHOD

PROCESS

high organic loading	less capital equipment
high BOD removal/unit	low capital costs
short detention time	small basin & land required
low Hp	low operating costs
BOD & NH_3 in single basin	less unit processes, lowest capital costs
high bacteria solids inventory	very stable process-resist upsets
built in compensation for BOD, flow, and toxic variations	easy to operate

TABLE 2

PRELIMINARY DESIGN BRIEF

CARTER BIO-DRUM™ and ACTIVATED BIOFILM METHOD

Project Location: Anytown, USA — Date: 12/12/79

Consulting Engineer: RBC Consultants, Inc.
Someplace, USA — Prepared by: G. R. Fisette, PE

Attn: Joe Consultant, PE

BASIC DATA

(1)	a) Waste	Municipal	
	b) Process	Activated Biofilm	
(2)	Design, Flow	2.5	MGD
(3)	BOD Influent to Bio-Drum Basin	4170	lbs BOD/day
(4)	BOD Effluent	625	lbs BOD/day
(5)	Removal	85	%
(6)	NH_3 Influent to Bio-Drum basin	500	lbs $N\text{-}NH_3$/day
(7)	NH_3 Effluent	n/a	lbs $N\text{-}NH_3$/day
(8)	NH_3 Removal	n/a	%
(9)	SS Influent	4170	lbs SS/day
(10)	pH range	6-8	pH
(11)	Dissolved oxygen level to be maintained in the Bio-Drum basin	1.5	mg/l
(12)	Temperature of Waste in Cio-Drum basin a) Winter	15	°C
	b) Summer	25	°C

BIO-DRUMS

(13) Media loading rate (F/V) required — 0.67 lbs BOD/d/ft^3

(14) Hydraulic detention time required — 1.0 hrs.

(15) Volume of Bio-Drums required= — 6224 ft^3

$$\left(\frac{\text{item \#3}}{\text{item \#13}}\right) = \left(\frac{4170}{0.67}\right)$$

(16) Number of Bio-Drums required= — 16 units

$$= \left(\frac{\text{item \#15}}{390}\right) = \left(\frac{6224}{390}\right) =$$

(17) Basin volume occupied by Bio-Drums=

=(Item #16) (127) = (16) (127) = — 2000 ft^3

(18) Total Basin volume =

a) $\left(\frac{\text{item \#2} \times 10^6}{180}\right) \left(\text{item \#14}\right) + (\text{item \#17})$

$\left(\frac{2.5 \times 10^6}{180}\right)$ (1.0) +(2000) = — 16,000 ft^3

b) (volume, ft^3) (7.48) = (16,00) (7.48) = — 120,000 gal

(19) Minimum area required at 5 feet water depth =

$$\left(\frac{\text{item \#18a}}{5}\right) = \left(\frac{16,000}{5}\right) =$$ — 3200 ft^2

(20) Minimum area required for Bio-Drum units =

(item #16) (144) = (16) (144) = — 2300 ft^2

(21) Basin layout for dual assemblies+

a) Width = $\left(\frac{\text{item \#19 or 20}}{2}\right)^{\frac{1}{2}}$ = — 50 ft

b) Length = (width, ft) (2) = — 64 ft

(22) Power required = (item #16) (2.5) = (16) (2.5) = — 40 BHp

MLSS

(23) Solids loading rate (F/M) required 0.55 lbs BOD/d/#MLS

(24) Pounds of MLSS in basin =

$$= \frac{(\text{item } \#3\ )}{(\text{item } \#23)} = \frac{(4170)}{(0.55)} = 7580 \text{ lbs MLSS}$$

(25) Concentration of MLSS in basin =

$$\frac{(\text{item } \#24 \times 10^6)}{(\ (8.34)\ (\text{Item } \#18b))} = \left(\frac{7580 \times 10^6}{(8.34)\ (120{,}000)}\right) \quad 7575 \text{ mg/l}$$

(26) Effective mass of MLSS in system =

(item #16) (600# solids/drum) + (item #24) =

(16) (600) + (7580) = 17,200 lbs MLSS

(27) Effective concentration of MLSS in system =

$$\frac{(\text{item } \#26)\ (10^6)}{(8.34)\ (\text{item } \#18b)} = \frac{(17{,}200)\ (\ 10^6\ )}{(8.34)\ (120{,}000)} = \quad 17{,}000 \text{ mg/l}$$

(28) Estimated Sludge Age of recycle solids =

$$\left(\frac{\text{lbs MLSS in system}}{\text{lbs MLSS wasted/day}}\right) = \left(\frac{\text{item } \#24 + \text{clarifier solids}}{\text{item } \#3 * \#\ \text{MLSS}/\#\text{BOD/d}}\right) =$$

$$\left(\frac{7580 + 10{,}600}{4170 * 0.75}\right) = \quad 5.8 \text{ days}$$

CLARIFIER

(29) Approximate clarifier surface area =

$$\left(\frac{\text{item } \#2}{\text{overflow rate, GPD/ft}^2}\right) = \left(\frac{2.5 * 10^6}{500}\right) = \quad 5000 \text{ ft}^2$$

TABLE 3

OXYGEN TRANSFER CAPABILITIES OF
BIO-DRUM™ AND OTHER AERATION DEVICES[16]

Apparatus	K_La^{20}	SOTR	No
Single 8' ϕ x 8' Bio-Drum			
Normal speed @ 1.1 fps	2.88	1.4	0.70
Dual 8' ϕ x 8' Bio-Drum			
Normal speed @ 1.1 fps	3.33	2.8	0.85
Peak speed @ 2.2 fps	12.7	10.8	1.5
Pilot unit, 24"ϕ x 18" Bio-Drum[4]			
Low speed @ 0.5 fps	4.27	0.014	0.16
Normal speed @ 1.1 fps	8.40	0.028	0.25
Mid speed @ 1.6 fps	14.2	0.047	0.33
Peak speed @ 2.1 fps	22.7	0.075	0.48
High speed @ 2.6 fps	37.0	0.12	0.71
Rotating Biological Contactor (biodisc)[17]			
12' ϕ standard media-mechanical drive @ 1.0 fps	1.18	0.17	----
12' ϕ standard media-air drive @ 1.0 fps	4.10	0.58	----
12' ϕ high density media-mechanical drive @ 1.0 fps	1.13	0.16	----
12' ϕ high density media-air drive @ 1.0 fps	3.55	0.50	----
Coarse-bubble diffusers[16, 18]	---	0.8-2.0	0.8-1.6

K_La^{20} = Mass transfer coefficient, hr^{-1}

SOTR = Standard mass transfer rate, lbs O_2/hr

No = Standard mass transfer efficiency, lbs O_2/Hp-hr

TABLE 4

TYPICAL ACTIVATED SLUDGE DESIGN LOADINGS[14]

Process	O_2 Uptake	DT	MLSS	F/M	F/V	Depth	F/V'	Removal	SRT	O_2 Required	Recycle	Sludge Produced
	mg O_2/hr g MLSS	hrs	mg/l	lbs BOD/d lb MLSS	lbs BOD/d/kft^3 basin	ft	lbs BOD/d/kft^3 media	%	d^{-1}	lb O_2/lb BOD	% flow	lb/lb BOD/d
Conventional	7-15	4-8	1500-4000	0.15-0.4	20-60	10-20	-	90-95	4-8	0.8-1.1	30-100	0.4-0.6
High Rate	15-25	2-4	3000-5000	0.4-1.0	70-180	10-20	-	85-90	2-4	0.7-0.9	30-100	0.5-0.7
Step	-	3-6	-	0.2-0.4	40-60	10-20	-	-	-	-	-	-
Contact-Stab	10-30	1-6	2000-4000 6-10,000	0.15-0.5	30-70	10-20	-	80-95	3-10	0.8-1.1	25-100	-
Extended	3-8	16-24	2000-6000	< 0.05	10-15	10-20	-	90+	≥30	1.4-1.6	100-300	0.15-0.3
Lagoon (aerated)	-	24-240	< 150	-	-	6-15	-	80+	-	1.0-1.4	none	-
Pure O_2	-	1-3	3000-5000	0.25-1.0	100-250	15-25	-	90+	-	-	-	-
Complete Mix	-	4-6	3000	0.2-0.6	50-120	15-25	-	90+	-	-	-	-
Modified	20-40	0.5-2	500-1500	1.5-3.0	90-180	-	-	60-75	< 1	0.4-0.6	10-30	0.8-1.2
CABM™	-	0.5-2	6-10,000 (15-20,000)	0.1-1	-	5	0.6-1.0	80-95	variable	-	10-150	-
RBC[5, 6, 11, 12]	-	1-3	150 (10-20,000)	∞	-	6	0.5-0.2	80-95	0	-	0-100	-
ABF[11, 13, 14]	-	2-4	2-5,000	0.2-0.9	50-225	5-22	0.1-0.35	80-90	?	0.7-1.0	50-200 H_2O & 30-100 sludge	0.55-0.75

TABLE 5

TYPICAL FIXED FILM PROCESS DESIGN LOADINGS

Process	Depth	Removal	Organic Loading		Hydraulic Loading		DT
	ft	%	lbs BOD/d per kft^3	acre ft	MGD	$\frac{GPD}{ft^2}$	hrs
Convention	6-8	80-85	5-25	200-1000	1-4	25-90	--
Intermediate	6-8	50-70	15-30	700-1400	4-10	90-230	--
High Rate	3-8	65-80	25-300	1000-1300	10-20	230-900	--
Plastic media	$\leq$ 40	65-85	$\leq$ 300	?	15-90	350-2100	--
Roughing filter	3-20	40-65	$\geq$ 100	?	60-180	1400-4200	--
ABF[11, 13, 14]	10-40	80-95	100-300	-	-	-	2-4
RBC[5, 6, 11, 12]	6	80-95	70-150	-	-	1-4	1-3
CABM™	6	80-95	500-800	-	-	8-20	1/2 - 2

TABLE 6

COMPARISON OF POWER REQUIREMENTS

Basis: 1 MGD of 180 mg/l BOD

	Transfer Rate lbs O_2/Hp-hr (14)	Loading lbs BOD/d/kft^3	Hp/MGD	Hp/k#BOD/d	Mixing Hp/kft^3
low speed surface	1.7	--	40	27	1.2
Turbine-sprayer	1.4	--	48	32	1.4
Diffused air, medium bubbles	1.4	--	48	32	1.4
Pure oxygen, cryogenic	2.7	--	25	17	3.4
RBC	--	110	38	25	4.7
Bio-Drum	0.85	700	15	10	2.7
ABF (9, 14)	?	200	21	17	?

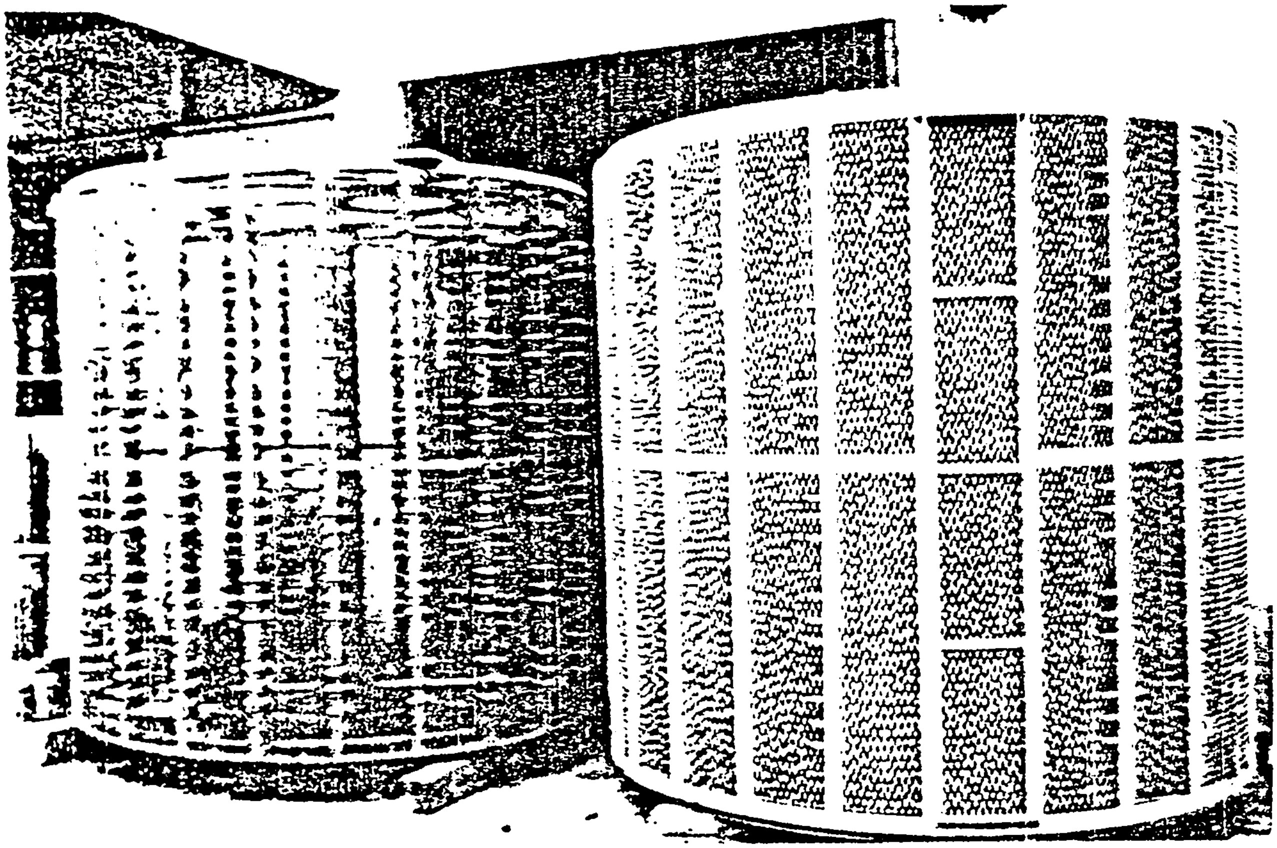


FIGURE 1

2.5 MGD BIO-DRUM™ INSTALLATION USING THE ACTIVATED BIOFILM METHOD™

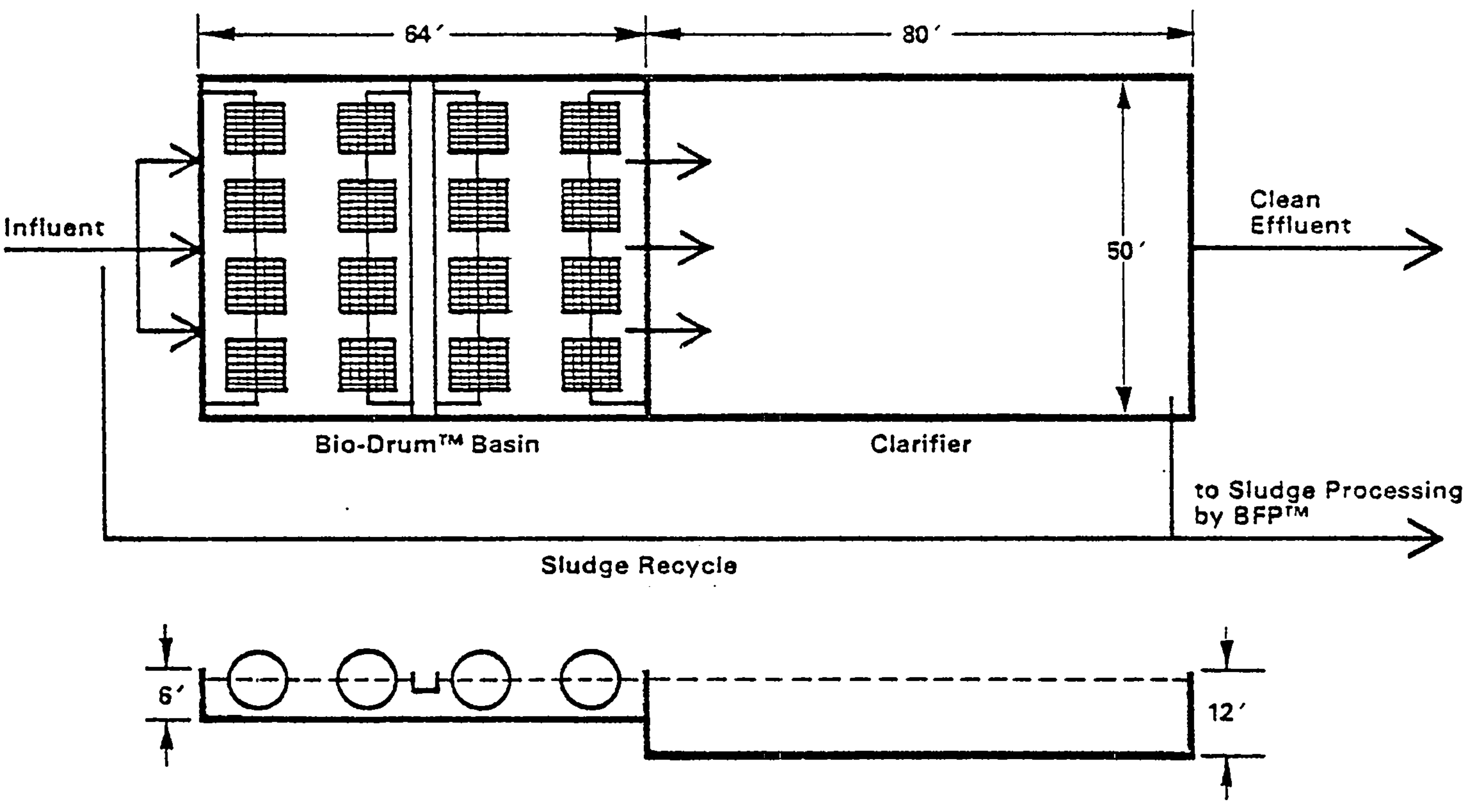


FIGURE 2

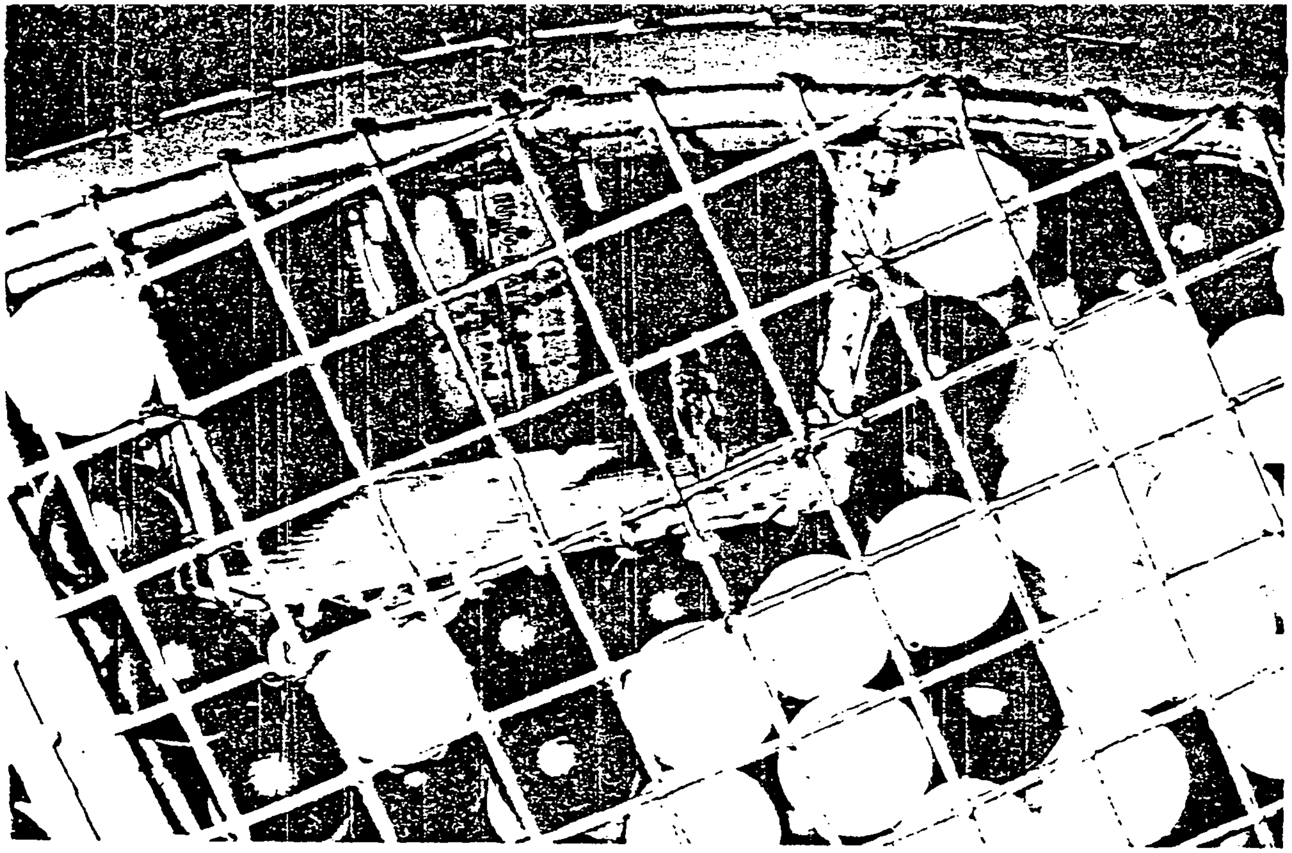


FIGURE 3

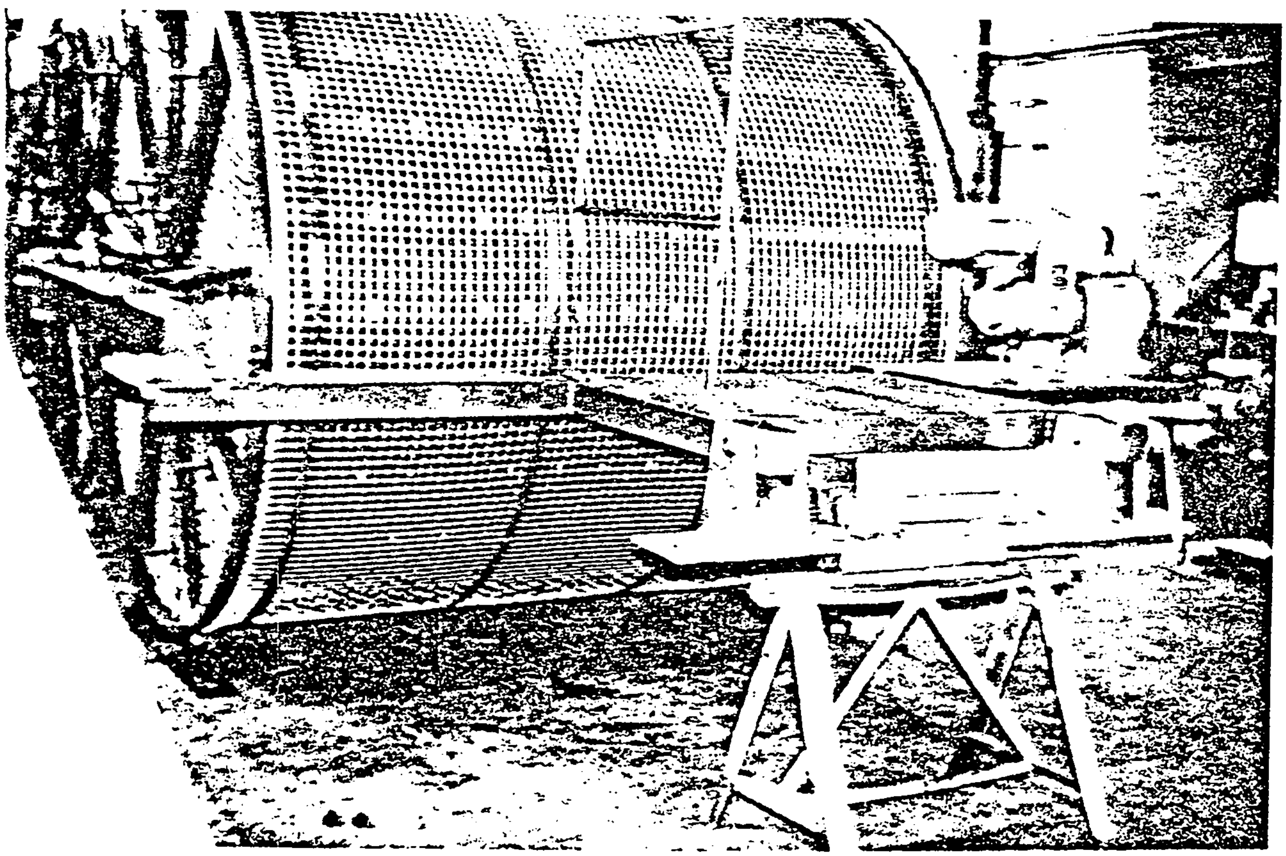


FIGURE 4

FIGURE 5

Effect of Sludge Recycle upon Performance of Rotating Biological Contactors

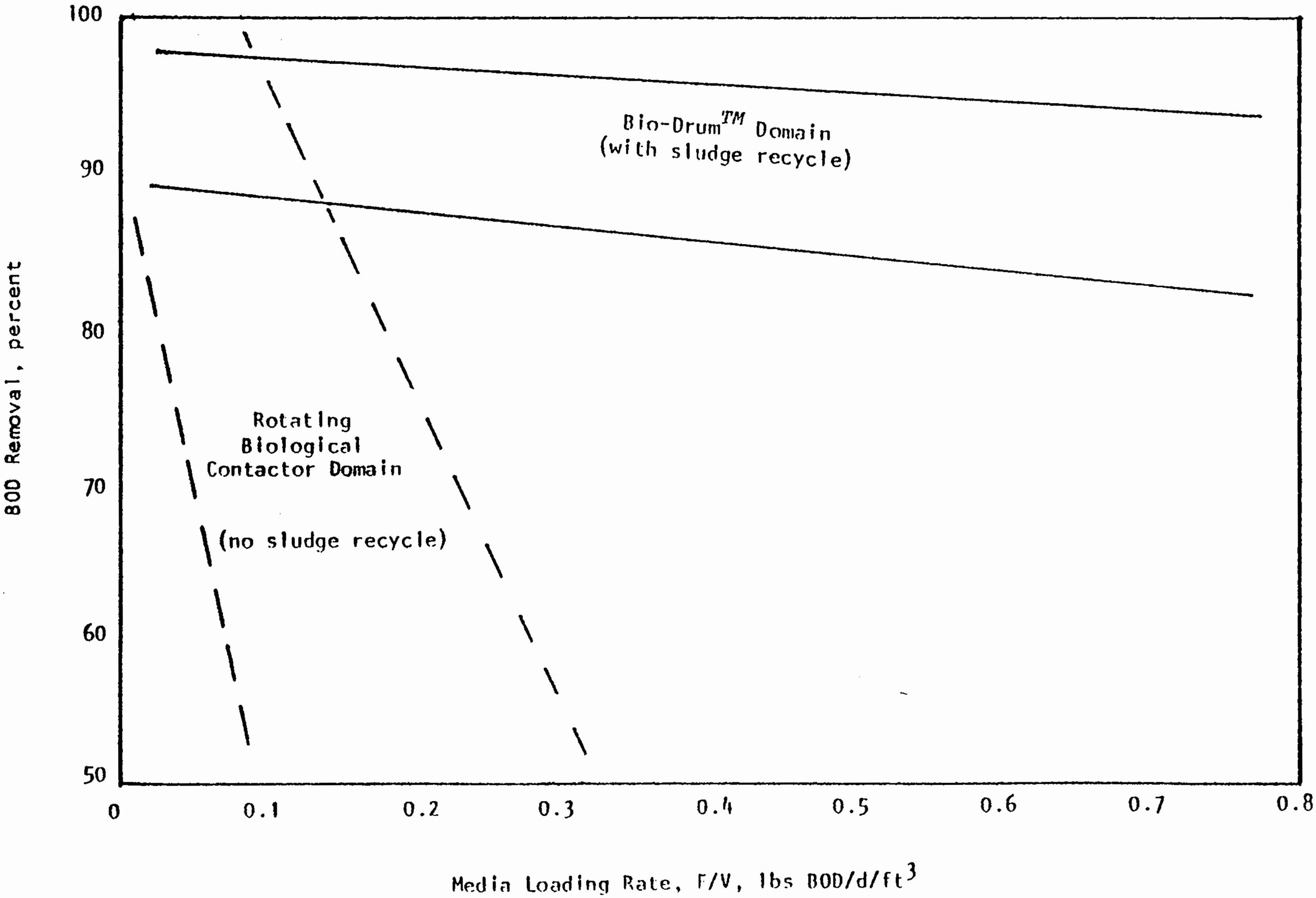


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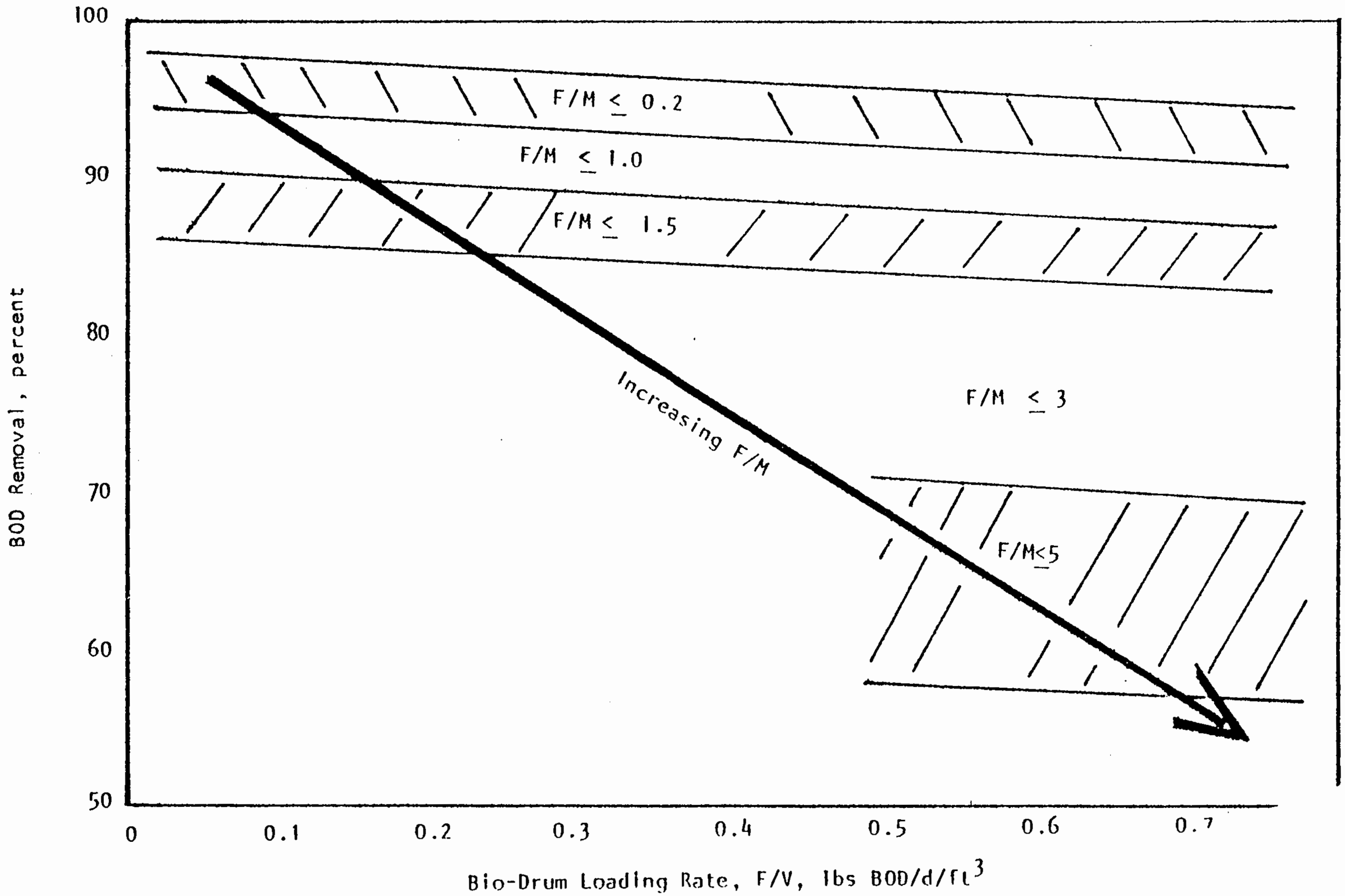
FIGURE 6

Performance of the Carter Bio-Drum^{TM} & CABM^{TM}

For BOD Removal



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EVALUATION OF A ROTATING BIOLOGICAL DISC
IN A SEWAGE TREATMENT PROCESS
IN PACKAGE PLANT APPLICATIONS

BY

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Mike Vesio
Purestream Industries, Inc.

The Rotating Biological Disc Process, sometimes referred to as the RBC, RBD, RBS, Bio-Shaft, Bio-surf, Aero-surf, Enviro-surf, etc. hereafter will be referred to, as the RBC Process. This process caused a very significant impact in the wastewater treatment market. This statement is obvious when one considers the excellent program we are participating in at the First National Symposium on Rotating Biological Contactor Technology. However, the simplicity of process flow control, minimal maintenance and operator attention, economical installed costs, low energy consumption and according to many, the panacea for all process problems, has given the RBC Process, usage for many applications. The Federal Construction Grant Program contributing needed construction monies and a tremendous marketing effort by disc manufacturers also spirited the advance and the industry acceptance of the RBC Process.

Conservatively, there are over 5,000 RBC installations in the United States and abroad. In the next ten years, this figure will easily double. What caused this phenomenal growth of a biological waste treatment process that dates back over 50 years? What resurrected this almost forgotten biological wastewater treatment process? First and foremost, immediately after World War II Germany, with high population densities located in small towns and villages surrounded by hilly and mountainous regions, needed an inexpensive, dependable, low energy consumption and minimal operator attention wastewater treatment system to produce a high quality wastewater effluent. Regulatory agencies were faced as we are today, with pollution of good water short rivers and streams. The University of Stuttgart, Germany, evaluated many processes to meet these water pollution control objectives. Among the wastewater treatment processes investigated was the

Rotating Biological Discs. Or, commonly then referred to in Germany as the Immersion Drip Filter. The use of the relatively new material, plastics, substituting for the wood and metal biological support surfaces, described in the Weigand Patent (1) and the Maltby "Biological Wheel" Patent (2), accomplished the needed economies of the RBC Process. A European disc manufacturer continued the usage of the RBC Process for relatively small sized installations. The virtues of an inexpensive, dependable, low energy consumption and minimal operator attention wastewater treatment system to produce high water quality effluents made the RBC Process ideal for low flow applications. The conservativeness of the European disc manufacturer in process application and structural design contributed heavily to the credence and reliabilities of the RBC Process.

Independently, in the mid 60's, I headed a group at Allis Chalmers Co. in Milwaukee that developed on its own a RBC Process. The European work at that time was unknown to us. Later, after our own developments and eventual communications with the European disc manufacturer, we were convinced the RBC Process had commercial application in the United States. The wastewater treatment industry at this time, was not the high technological, several billion dollar a year sales, regulated industry as we know today. Generally, the sanitary engineer used a particular wastewater treatment process he had experienced, continued to use that same process over and over for every application. Rule of thumb designs, personal experiences and time proven wastewater treatment processes were the order of the day. Some educators and a few pioneers in the industry tried to advance the technology of wastewater treatment. Unfortunately, since a waste treatment project carried the individual sanitary engineer's P. E. stamp and the engineer's reputation was on the line, very few new technologies were tried.

The Federal Water Pollution Control Administration (FWPCA) program of complete funding for research of new wastewater treatment technologies and upwards of 90% for demonstration projects of these new technologies was the breakthrough the RBC Process required in the United States. This enabled basic RBC work and several full-scale demonstration projects. Some of these projects proved an immediate success, demonstrating the RBC values of high performance, simple operation, low maintenance and low power consumption. In the mid 70's the Federal General Accounting Office (GAO) instructed all applicants for federal construction grant funds to submit cost evaluations for all applicable wastewater treatment processes including the RBC for their projects. This made consulting engineers take a realistic appraisal of the RBC Process.

The many advantages of the RBC Process found its way into numerous applications. These applications can be categorized as follows:

A. Small, medium and large municipal sewage treatment plants.

B. Commercial applications such as subdivisions, resorts, shopping centers, trailer courts, and campgrounds.

C. Complete treatment of industrial process wastewater.

D. Pre-treatment of industrial process wastewater.

This latter application for pre-treatment of industrial process wastewaters will become more prevalent as the State and municipal regulatory agencies enforce the reporting waste discharges to municipal sewage systems. The reporting of waste material with pollution concentrations larger than domestic wastewater requires additional municipal treatment capacity. The municipality will charge an extra sewage discharge fee to these industrial dischargers. Reduction of the industrial waters pollution concentration to the levels of domestic wastewater, and the resulting savings of these extra municipality discharge fees, can in most instances, amortize the cost of wastewater treatment equipment very rapidly.

The Federal Construction Grant Program and associated cost effective evaluations gave the RBC Process many opportunities for different effluent discharge standards in the municipal market. The RBC Process is used for carbonaceous BOD reduction, nitrification, following overloaded existing processes, paralleling overloaded existing processes, and more recently, the RBC Process was used in a de-nitrification application. Unfortunately, this very large municipal market with federal funds and federal regulations with so-called "non-restrictive specifications," enable some disc manufacturers to take short cuts in design and applications. Insufficient in-house testing and the inabilities to use the required process disc design parameters resulted in many problems. Horror stories of complete failures of installations are commonplace. One of the largest installation failures included 96 RBC shafts. There are several RBC installations that have been and are still being completely replaced for the third time. New installation failures are being reported daily. The disc competitive situation in the municipal marketplace will still result in future horror stories. The original purpose of the European RBC objectives are being ignored. The RBC originally developed as an inexpensive, dependable, operator insensitive waste treatment process, are causing consulting engineers to take another look before selecting the RBC Process.

The Bio-Shaft Co. which I founded, has rigidly kept the European RBC objectives in mind. Excellent process performance without structural disc or shaft problems have been our trademark.

However, you can't put bread on the table in the federally funded municipal market with this philosophy. The competitive situation does not give Tait/Bio-Shafts credit for stating "We are the only disc manufacturers not to have a shaft or disc failure."

Looking for a market area where Tait/Bio-Shafts can market our dependable RBC product, we went back to the basics. The RBC Process was initially developed for the small size installations. Many successful small size installations are in operation. Pretorius (3) said it quite well:

> "Treatment works for small communities should be relatively cheap, reliable, and easy to operate and be maintained by unskilled labor, but should produce the same high standard of effluent as that from a larger well-controlled plant. The rotating disc unit (RDU) seems to fulfill these prerequisites."

The RBC equipment is only a portion of the complete wastewater treatmentsystem. Primary treatment, secondary clarifier, disinfection and sometimes, tertiary treatment will complete a system. While we can have a simple dependable RBC biological reactor, all would be for naught if the rest of the wastewater treatment system equipment does not support the simple and dependable philosophy. We looked for a market that needed a wastewater treatment system where dependability, simplicity and high performance are the first and foremost objectives. Where is there a market that can support the development and expense of a disc manufacturer?

By coincidence, a very large market in need of these features exists. The pre-engineered complete wastewater treatment package system, more commonly referred to as the commercial market requires the RBC Process features. Dependability, simplicity and high performance. The required higher effluent quality standards and the enforced reporting of all point discharge sources make the RBC Process features desirable. In the past, the extended air and contact stabilization processes were supplied for these commercial applications. With proper, consistent and knowledgable operations, the past effluent water quality operators for small installations, enforcement of reporting standards and higher water quality would appear to see a decreasing use of these processes. The need will still exist for the pre-engineered complete wastewater treatment package system because of many small low applications.

Tait/Bio-Shafts, Inc. and its sister division, Purestream Industries, Inc., combined its expertise to offer the RBC Process in a complete pre-engineered wastewater treatment system. Figure No. 1 shows the flow scheme which is used for a complete wastewater treatment application. As can be seen, the flow scheme is surge or flow equalization, primary clarification, RBC reactor, final settling and disinfection. Sludge from final settling and primary settling is directed to the aerobic digestor. Figure No. 2 is an elevation view with cutaway of the system. Please note the surge and aerobic digestor is aerated. Both primary and secondary sludge is air lifted to the aerobic digestor. Figure 2A is an artist conception of the assembled system. Each equipment component of the system is a modular design, completely pre-engineered. All of the components selected have the same dependable history as the RBC. These components are not only dependable, but are operator insensitive, low installed and operating costs, with the ability to consistently produce high effluent water quality. Figure No. 3 shows the selection tables for the 6 ft. 8 inch diameter RBC system. Figure No. 4 shows selection tables for the 11 ft. diameter RBC system. Please note the design flow selections encompass the range of 7,000 to 250,000 GPD. Other selection tables for various system BOD removals are included in Purestream Industries brochure TBRI7500. These brochures are available in the Exhibitor Hall. We also have available for your review a scaled-down model of this complete pre-packaged RBC system.

The RBC reactor has been conservative sized for the commercial market. In the past, commercial applications equipment has been sized with a minimal amount of process information. Therefore, anticipating the same lack of complete information, the RBC has been sized to insure obtaining the required high effluent water qualities. Tait/Bio-Shaft used the staged wastewater treatment process design for BOD removal. Each stage is sized to remove 50% of the BOD level that the individual stage sees. A hydraulic loading of 8 GPD per sq. ft. per stage can easily accomplish this performance. The use of the staging concept will give the consulting engineer another valuable tool in which to select the required number of stages for a specific design output.

The use of modular system components can enable the engineer or regulatory agencies to substitute or interchange specific system components design parameters to meet the individual installation requirements. Figure No. 5 shows a basic modular component for the primary modular. This includes the surge compartment, aerobic digestor, and primary clarification. By enlarging or decreasing the length dimension, incremental capacities of each individual component can be obtained.

Figure No. 6 is the RBC reactor tank in a stage configuration. Removal of internal baffles and manipulation of the length dimension will again produce incremental system capacity changes. Figure No. 7 is the final clarification tank with chlorine contact compartment. Again, length changes can accommodate various component sizing.

The use of modular system component techniques enables the engineer to use the specific modular component for other process applications. Inserting the RBC reactor component ahead of an existing biological process will rough treat an overloaded secondary treatment system. In these applications, the bio-reactor modular alone can sufficiently reduce the BOD directed to the existing biological process to the point where the overall system will be more effective. Inserting the RBC reactor component behind an existing secondary treatment system can be used to polish or upgrade the overall system treatment efficiency. Once again, the modular concept of a bio-reactor makes an existing wastewater treatment plant upgrade or overload a simple solution.

Using the biological reactor with a new clarifier can be used very advantageously for an industrial pre-treatment installation. The treated water will be discharged to a municipal sewer system and as mentioned before, substantial savings to the industrial user can be experienced.

Applications for the small community or complete pre-engineered complete wastewater packaged system in effect will give this market area all the advantages of a larger built-in-place municipal system. It also has an inherent advantage that all components will be supplied by one company. Purestream Industries using the Tait/Bio-Shaft Rotating Biological Disc is the only company to supply a pre-engineered, complete wastewater packaged RBC system in this country.

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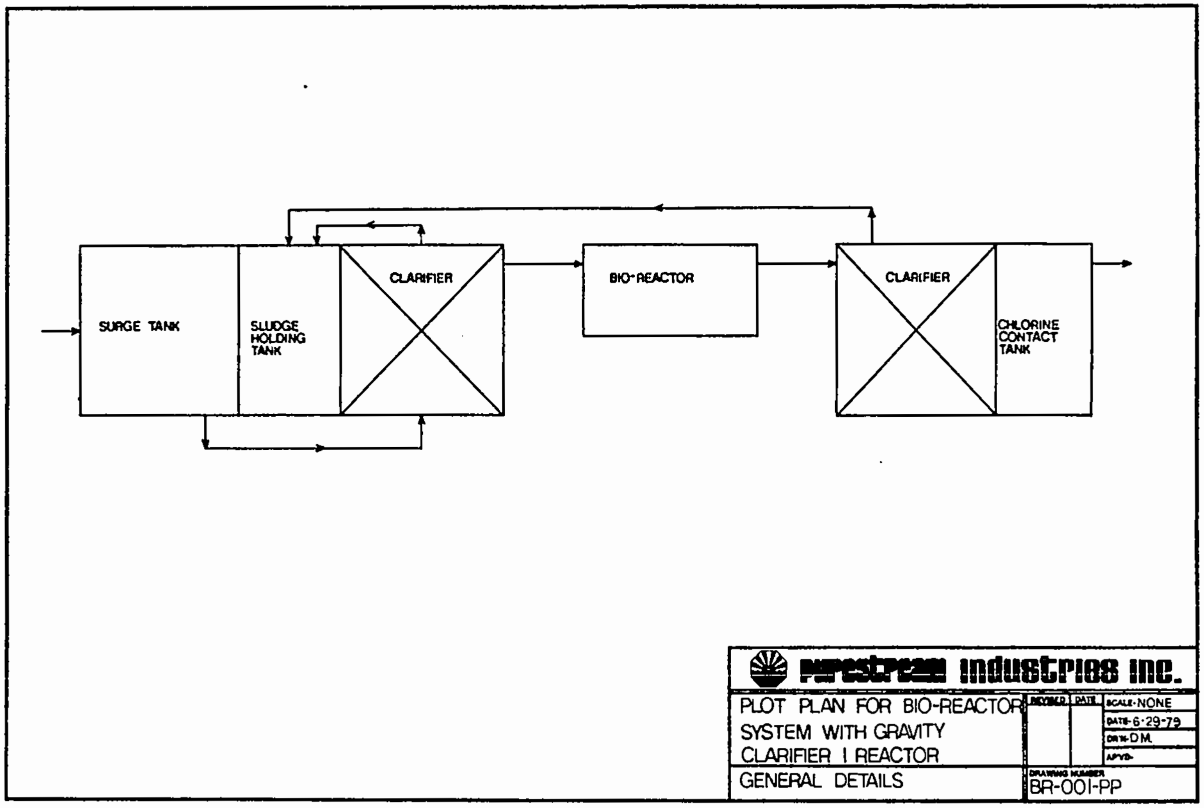

SURGE TANK
SLUDGE HOLDING TANK
CLARIFIER
BIO-REACTOR
CLARIFIER
CHLORINE CONTACT TANK
industries inc.
PLOT PLAN FOR BIO-REACTOR SYSTEM WITH GRAVITY CLARIFIER I REACTOR
GENERAL DETAILS
SCALE-NONE
DATE-6-29-79
DRN-D.M.
APVD-
DRAWING NUMBER
BR-001-PP

PBR-041

Fig. 1

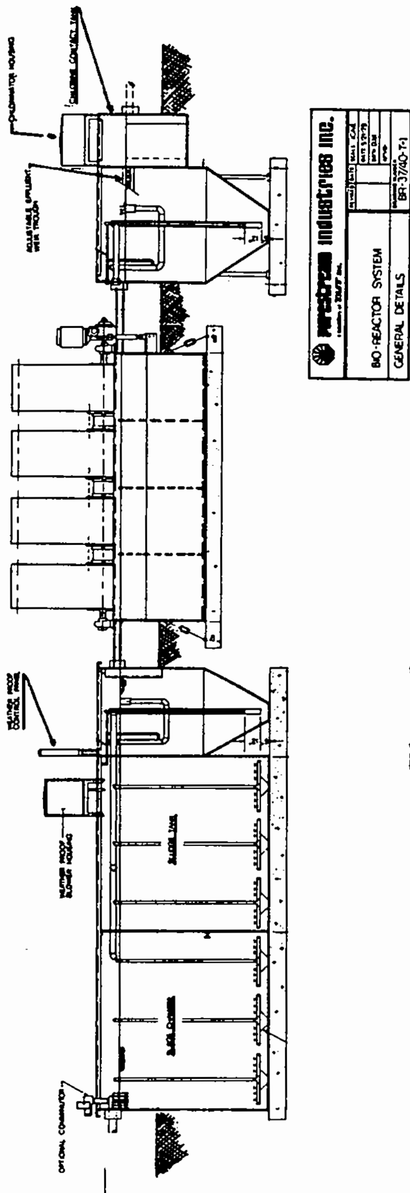


Fig. 2

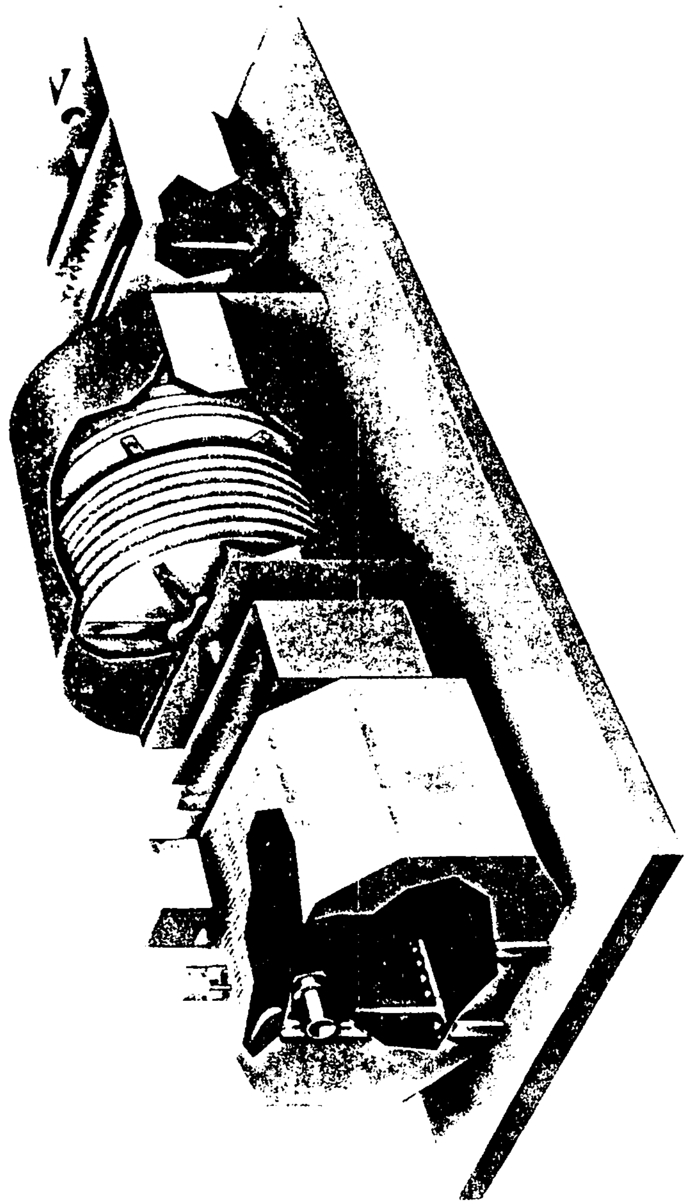


FIGURE 3

Selection tables for secondary treatment

95.6% Removal — **6 Ft.-8 In. Dia. Reactor**

FLOW	MODEL	NO. OF REACTORS	SURGE VOL.	PRIMARY VOL.	DIGESTER VOL.(F³)	SECONDARY CLAR. VOL.	CCT VOL.	TOTAL HP
7,000	PBR 6D5	1	700	670	140	670	146	3¾
8,000	PBR 6D6	1	800	670	160	670	177	3¾
9,000	PBR 6D7	1	900	750	180	750	188	3¾
10,000	PBR 6D8	1	1,000	835	200	835	208	3¾
12,500	PBR 6D9	1	1,250	1,160	250	1,160	260	3¾
15,000	PBR 6D11	1	1,500	1,250	300	1,250	312	4
17,500	PBR 6D13	1	1,750	1,500	350	1,500	365	4½
20,000	PBR 6D15	1	2,000	1,670	400	1,670	416	4½
22,500	PBR 6D17	1	2,250	1,875	450	1,875	470	4½
25,000	PBR 6D19	1	2,500	2,170	500	2,170	521	4½
30,000	PBR 6D23	1	3,000	2,500	600	2,500	625	5
35,000	PBR 6D27	1	3,500	3,000	700	3,000	729	8
40,000	PBR 6D31	1	4,000	3,340	800	3,340	835	8
45,000	PBR 6D17	2	4,500	3,750	900	3,750	940	8
50,000	PBR 6D19	2	5,000	4,170	1,000	4,170	1,050	8
60,000	PBR 6D23	2	6,000	5,000	1,200	5,000	1,250	9
70,000	PBR 6D27	2	7,000	5,830	1,400	5,830	1,460	13½
80,000	PBR 6D31	2	8,000	6,670	1,600	6,670	1,670	13½

91% Removal — **6 Ft.-8 In. Dia. Reactor**

FLOW	MODEL	NO. OF REACTORS	SURGE VOL.	PRIMARY VOL.	DIGESTER VOL.(F³)	SECONDARY CLAR. VOL.	CCT VOL.	TOTAL HP
7,000	PBR 6C5	1	700	670	140	670	146	3⅓
8,000	PBR 6C6	1	800	670	160	670	177	3⅓
9,000	PBR 6C7	1	900	750	180	750	188	3½
10,000	PBR 6C8	1	1,000	835	200	835	208	3½
12,500	PBR 6C9	1	1,250	1,160	250	1,160	260	3½
15,000	PBR 6C11	1	1,500	1,250	300	1,250	312	3½
17,500	PBR 6C13	1	1,750	1,500	350	1,500	365	3½
20,000	PBR 6C15	1	2,000	1,670	400	1,670	416	3¾
22,500	PBR 6C17	1	2,250	1,875	450	1,875	470	3¾
25,000	PBR 6C19	1	2,500	2,170	500	2,170	521	3¾
30,000	PBR 6C23	1	3,000	2,500	600	2,500	625	3¾
35,000	PBR 6C27	1	3,500	3,000	700	3,000	729	6
40,000	PBR 6C31	1	4,000	3,340	800	3,340	835	6
45,000	PBR 6C35	1	4,500	3,750	900	3,750	940	6½
50,000	PBR 6C39	1	5,000	4,170	1,000	4,170	1,050	6½
60,000	PBR 6C23	2	6,000	5,000	1,200	5,000	1,250	6½
70,000	PBR 6C27	2	7,000	5,830	1,400	5,830	1,460	9½
80,000	PBR 6C31	2	8,000	6,670	1,600	6,670	1,670	9½

NOTE: In applications requiring nitrification, CONSULT FACTORY.

FIGURE 4

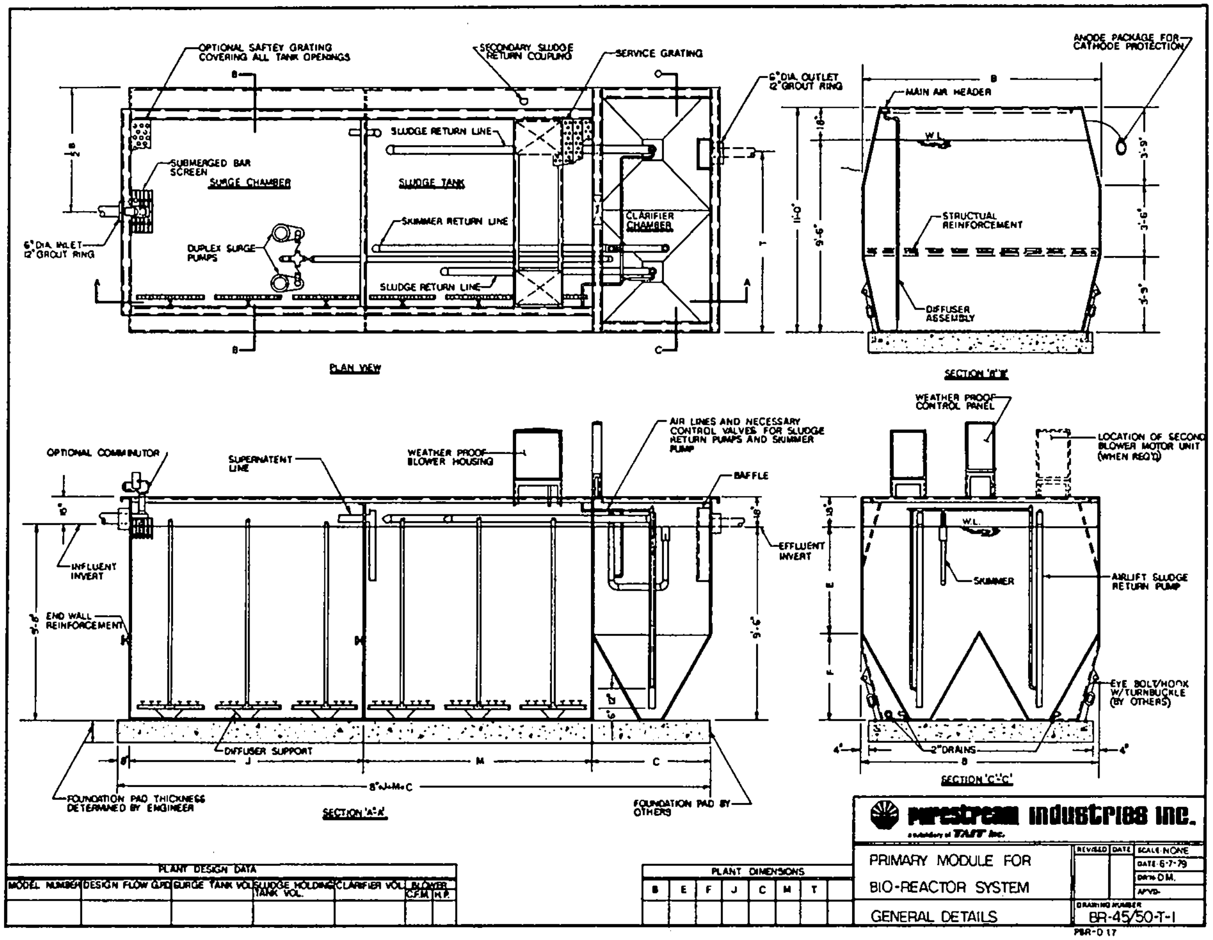
Selection tables for secondary treatment

95.6% Removal — **11 Ft. Dia. Reactor**

FLOW	MODEL	NO. OF REACTORS	SURGE VOL.	PRIMARY VOL.	DIGESTER VOL.(F³)	SECONDARY CLAR. VOL.	CCT VOL.	TOTAL HP
25,000	PBR 11D7	1	2,500	2,170	500	2,170	521	3¾
30,000	PBR 11D8	1	3,000	2,500	600	2,500	625	3¾
35,000	PBR 11D10	1	3,500	3,000	700	3,000	729	5¾
40,000	PBR 11D11	1	4,000	3,340	800	3,340	835	6
45,000	PBR 11D12	1	4,500	3,750	900	3,750	940	6½
50,000	PBR 11D14	1	5,000	4,170	1,000	4,170	1,050	6½
60,000	PBR 11D17	1	6,000	5,000	1,200	5,000	1,250	6½
70,000	PBR 11D20	1	7,000	5,830	1,400	5,830	1,460	6½
80,000	PBR 11D22	1	8,000	6,670	1,600	6,670	1,670	9
90,000	PBR 11D25	1	9,000	7,500	1,800	7,500	1,880	9
100,000	PBR 11D28	1	10,000	8,340	2,000	8,340	2,090	9
125,000	PBR 11D36	1	12,500	10,845	2,500	10,845	2,605	9
150,000	PBR 11D22	2	15,000	12,500	3,000	12,500	3,125	10½
175,000	PBR 11D24	2	17,500	15,000	3,500	15,000	3,650	13
200,000	PBR 11D28	2	20,000	16,700	4,000	16,700	4,170	13
225,000	PBR 11D32	2	22,500	38,333	4,500	38,333	4,690	18
250,000	PBR 11D36	2	25,000	41,666	5,000	41,666	5,210	18

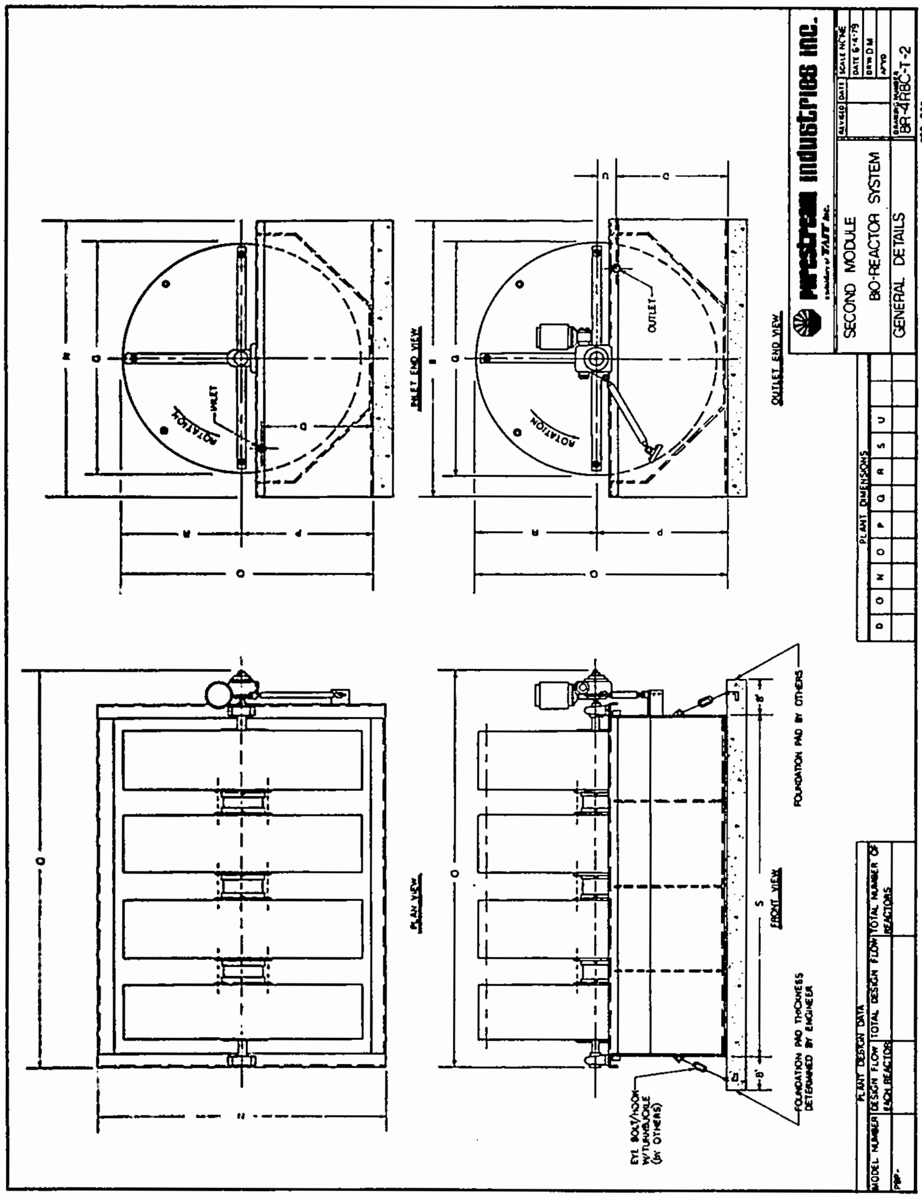
91% Removal — **11 Ft. Dia. Reactor**

FLOW	MODEL	NO. OF REACTORS	SURGE VOL.	PRIMARY VOL.	DIGESTER VOL.(F³)	SECONDARY CLAR. VOL.	CCT VOL.	TOTAL HP
25,000	PBR 11C7	1	2,500	2,170	500	2,170	521	3¾
30,000	PBR 11C8	1	3,000	2,500	600	2,500	625	3¾
35,000	PBR 11C10	1	3,500	3,000	700	3,000	729	5¾
40,000	PBR 11C11	1	4,000	3,340	800	3,340	835	5¾
45,000	PBR 11C12	1	4,500	3,750	900	3,750	940	5¾
50,000	PBR 11C14	1	5,000	4,170	1,000	4,170	1,050	6
60,000	PBR 11C17	1	6,000	5,000	1,200	5,000	1,250	6½
70,000	PBR 11C20	1	7,000	5,830	1,400	5,830	1,460	6½
80,000	PBR 11C22	1	8,000	6,670	1,600	6,670	1,670	9
90,000	PBR 11C25	1	9,000	7,500	1,800	7,500	1,880	9
100,000	PBR 11C28	1	10,000	8,340	2,000	8,340	2,090	9
125,000	PBR 11C36	1	12,500	10,845	2,500	10,845	2,605	9
150,000	PBR 11C43	1	15,000	12,500	3,000	12,500	3,125	9
175,000	PBR 11C50	1	17,500	15,000	3,500	15,000	3,650	12
200,000	PBR 11C28	2	20,000	16,700	4,000	16,700	4,170	13
225,000	PBR 11C32	2	22,500	38,333	4,500	38,333	4,690	16
250,000	PBR 11C36	2	25,000	41,666	5,000	41,666	5,210	18



OPTIONAL SAFTEY GRATING COVERING ALL TANK OPENINGS
SECONDARY SLUDGE RETURN COUPLING
SERVICE GRATING
6" DIA. OUTLET 12" GROUT RING
ANODE PACKAGE FOR CATHODE PROTECTION
MAIN AIR HEADER
SUBMERGED BAR SCREEN
SURGE CHAMBER
SLUDGE TANK
SLUDGE RETURN LINE
SKIMMER RETURN LINE
CLARIFIER CHAMBER
DUPLEX SURGE PUMPS
6" DIA. INLET 12" GROUT RING
SLUDGE RETURN LINE
PLAN VIEW
W.L.
STRUCTUAL REINFORCEMENT
DIFFUSER ASSEMBLY
SECTION 'B'-'B'
WEATHER PROOF CONTROL PANEL
LOCATION OF SECOND BLOWER MOTOR UNIT (WHEN REQ'D)
OPTIONAL COMMINUTOR
SUPERNATENT LINE
WEATHER PROOF BLOWER HOUSING
AIR LINES AND NECESSARY CONTROL VALVES FOR SLUDGE RETURN PUMPS AND SKIMMER PUMP
BAFFLE
EFFLUENT INVERT
INFLUENT INVERT
END WALL REINFORCEMENT
SKIMMER
AIRLIFT SLUDGE RETURN PUMP
EYE BOLT/HOOK W/TURNBUCKLE (BY OTHERS)
DIFFUSER SUPPORT
2" DRAINS
8"+J+M+C
FOUNDATION PAD THICKNESS DETERMINED BY ENGINEER
SECTION 'A'-'A'
FOUNDATION PAD BY OTHERS
SECTION 'C'-'C'
purestream industries inc.
a subsidiary of TAFT Inc.
PRIMARY MODULE FOR
BIO-REACTOR SYSTEM
GENERAL DETAILS
REVISED
DATE
SCALE: NONE
DATE 6-7-79
DRN: D.M.
APVD:
DRAWING NUMBER
BR-45/50-T-1
PSR-0 17
PLANT DESIGN DATA
MODEL NUMBER
DESIGN FLOW G.P.D.
SURGE TANK VOL.
SLUDGE HOLDING TANK VOL.
CLARIFIER VOL.
BLOWER
C.F.M.
H.P.
PLANT DIMENSIONS
B
E
F
J
C
M
T

Fig. 5



purestream industries inc.
SECOND MODULE
BIO-REACTOR SYSTEM
GENERAL DETAILS
BR-4RBC-T-2
PLAN VIEW
FRONT VIEW
INLET END VIEW
OUTLET END VIEW
ROTATION
INLET
OUTLET
FOUNDATION PAD BY OTHERS
FOUNDATION PAD THICKNESS DETERMINED BY ENGINEER
EYE BOLT/HOOK W/TURNBUCKLE (BY OTHERS)
PLANT DESIGN DATA
PLANT DIMENSIONS

Fig. 6

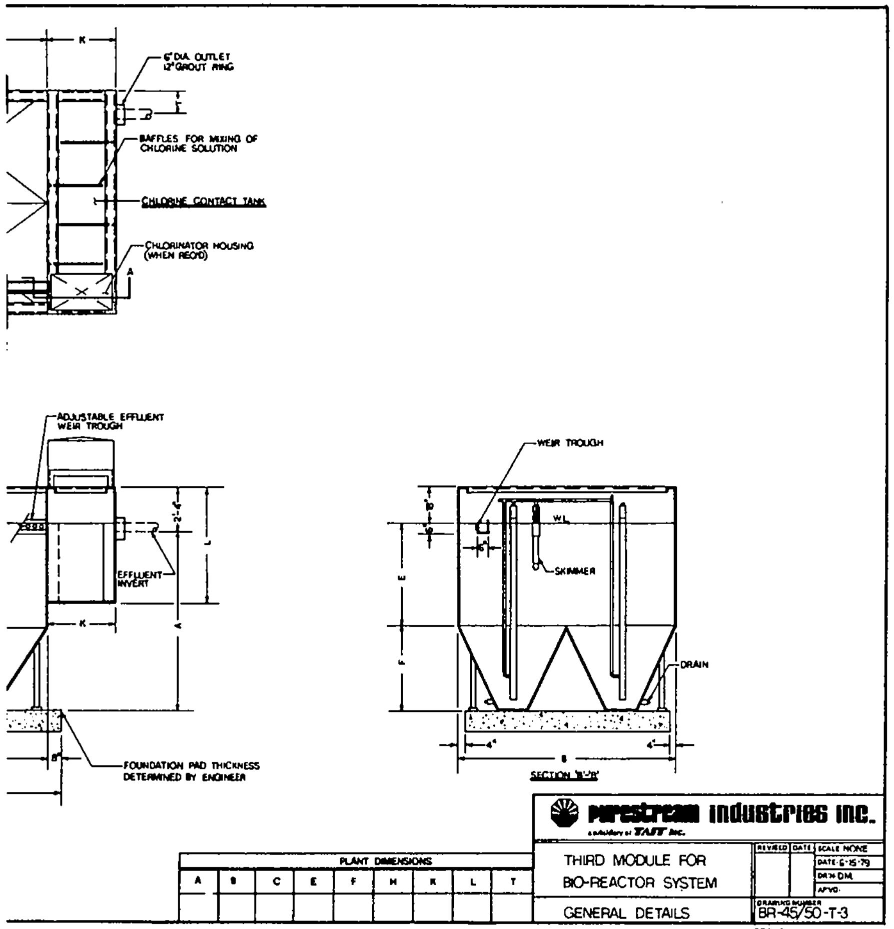

6" DIA. OUTLET
12" GROUT RING
BAFFLES FOR MIXING OF
CHLORINE SOLUTION
CHLORINE CONTACT TANK
CHLORINATOR HOUSING
(WHEN REQ'D)
ADJUSTABLE EFFLUENT
WEIR TROUGH
EFFLUENT
INVERT
FOUNDATION PAD THICKNESS
DETERMINED BY ENGINEER
WEIR TROUGH
W.L.
SKIMMER
DRAIN
SECTION 'B'-'B'
PLANT DIMENSIONS
A
B
C
E
F
H
K
L
T
purestream industries inc.
THIRD MODULE FOR
BIO-REACTOR SYSTEM
GENERAL DETAILS
SCALE NONE
DATE 6-15-79
BR-45/50-T-3
PBR-035

Fig. 7

PART IV: BIOKINETIC STUDIES

DYNAMICS OF MICROBIAL FILM PROCESSES

by

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Abstract

Microbial film formation at a surface is the net result of several physical, chemical and microbial processes including the following:

1. Transport of dissolved and particulate matter from the bulk fluid to the surface.
2. Firm microbial cell attachment to the surface.
3. Microbial transformations (growth, reproduction, etc.) at the surface resulting in production of organic matter.
4. Partial detachment of the deposit due primarily to fluid shear stress.

The properties and structure of the resulting biofilm accumulation reflect the relative rates of the above processes and influence the observed rate coefficients (e.g., coefficients for attachment rate and microbial substrate removal rate) and transport coefficients (e.g., diffusion coefficient, thermal conductivity and viscosity of biofilms). The transport coefficients are useful in analyzing the effects of biofilm development on fluid frictional resistance and heat transfer resistance.

This paper presents a framework for analyzing the interrelated processes contributing to biofouling. Available rate and composition data are presented so that the relative process rates can be compared.

INTRODUCTION

Bacteria stick firmly, and often with specificity, to almost any surface submerged in an aqueous environment. The bacteria attach by means of a matrix of polymers, primarily polysaccharides, that extend from the cell surface and form a mass of tangled fibers, termed a glycocalyx.[3] The adhesion mediated by the glycocalyx determines particular locations of bacteria in many aquatic environments. The cells grow and reporduce at the surface increasing the mass of cells and their associated material, i.e., the biofilm.

Biofilm processes may be beneficial as exemplified by fixed-film wastewater treatment processes (e.g., trickling filters and rotating biological contactors). In addition, biofilms frequently play a major role in stream purification processes. In fact, microbial activity in natural waters is concentrated at the interfaces.[36,29,3] However, biofilms can be quite troublesome in certain engineering systems. For example, biofilms in water conduits can cause energy losses resulting from increased fluid frictional resistance and increased heat transfer resistance. Table 1 lists the effects and relevance of biofilm processes to various sectors of our society.

Prior research from our group has been directed to problems of fouling (primarily sponsored by NSF and Electric Power Research Institute). Fouling refers to the formation of inorganic and/or organic deposits on surfaces. In cooling systems, these deposits form on condenser tube walls increasing fluid frictional resistance and heat transfer resistance. Four types of fouling, alone or in combination, may occur:

1. crystalline fouling caused by precipitation of $CaCO_3$, CaSOy or silicates

2. corrosion fouling resulting from formation of insulating layers of metal oxides on the tubes

3. fouling due to adherence of particulate matter on tube surfaces

4. biofouling resulting from attachment and growth of microorganisms.

Our work has been concerned with biofouling.

The most common method for controlling fouling biofilm development and maintaining heat exchange performance is periodic chlorination. Chlorine, added to cooling water, serves either to kill the microorganisms or to hydrolyze the extracellular polymers which hold the biofilm together. The chlorine dosage and application schedule is typically determined by (1) observation of condenser performance as indicated by plant steam back-pressure, or (2) operator experience.

Recently, concern over residual toxicity from hypochlorous acid or its reaction products has resulted in federal regulations which limit the allowable concentrations of free available chlorine in cooling water discharges. At the present time, there is no sound basis for assessing the impact of the regulations. Our previous investigations stemmed from the apparent need for a better basic understanding of biofilm development and biofilm destruction so that the impact of these new regulations on power generation could be evaluated.

The nature of the problem and the widespread concern regarding biofilm processes (evidenced in Table 1) led us to more fundamental studies in the dynamics of biofilm processes and its relevance in areas besides biofouling. Results of these and other studies are described in succeeding sections of this document. We propose to extend the work by further study of the fundamental processes which contribute to biofilm formation and destruction including the following:

Table 1. Effect and relevance of biofilms on various rate processes.

Effects	Specific Process	Concerns
Heat transfer reduction	Biofilm formation on condenser tubes and cooling tower fill material. Energy losses.	Power industry Chemical process industry U.S. Navy Solar energy systems
Increase in fluid frictional resistance	Biofilm formation in water and sewage conduits as well as condenser and heat exchange tubes. Causes increased power consumption for pumped systems or reduced capacity in gravity systems.	Municipal water supply and sewage collection Power industry Chemical process industry Solar energy systems
	Biofilm formation on ship hulls. Causes increase in fuel consumption.	U.S. Navy Shipping industry
Mass transfer and chemical transformations	Accelerated corrosion due to processes in the lower layers of the biofilm. Material deterioration in metal condenser tubes, sewage conduits, and cooling tower fill material.	Power industry U.S. Navy Municipal water supply and sewage collection Chemical process industry
	Biofilm formation on remote sensors, submarine periscopes, sight glasses, etc. Reduces effectiveness.	U.S. Navy Water quality data collection
	Detachment of microorganisms from biofilms in cooling towers. Releases pathogenic organisms (e.g., *Legionella*) in aerosols.	Public health
	Biofilm formation and detachment in drinking water distribution systems. Changes water quality in distribution system.	Water supply industry Public health
	Biofilm formation on teeth. Dental plaque and cavities.	Dental health
	Attachment of bacteria to animal cells. Diseases of lungs, intestines, and urinary tract.	Human health

Table 1. (continued)

Effects	Specific Process	Concerns
	Extraction of and oxidation of organic and inorganic compounds from water and wastewater. Reduction in "pollutant" load. For example, rotating biological contacters, biologically-aided carbon adsorption, and benthal stream activity.	Wastewater treatment Water treatment Stream analysis
	Biofilm formation in industrial production processes. Reduced product quality.	Pulp and paper industry
	Immobilized microorganisms or community of microorganisms for conducting specific chemical transformations.	Chemical process industry

1. transport rate of microbial cells from the bulk liquid to the surface.

2. interfacial phenomena resulting in adsorption/attachment of microbial cells to the surface.

3. reactions within the biofilm which increase the mass of the accumulation (e.g., microbial growth, exopolysaccharide production).

4. strength of the biofilm deposit as related to partial biofilm removal in a fluid shear field.

PROCESS ANALYSIS

Process analysis refers to the application of scientific methods to the recognition and definition of problems and the development of procedures for their solution. This generally requires 1) mathematical specification of the problem for the given physical situation, 2) development of a mathematical model, and 3) synthesis and systematic presentation of results to ensure full understanding. The process denotes an actual series of operations or treatment of materials as contrasted with the model, which is a mathematical description of the process (1).

Fundamentals (after Churchill (2))

The fundamental relationships which underlie process analysis are the equations for conservation of mass, momentum and energy which are a result of the laws of thermodynamics and Newton's laws of motion. The conservation equations are generally expressed in terms of intensive factors (independent of system mass) such as composition, velocity and temperature. The conservation equations also introduce physical properties such as thermodynamic properties, stoichiometric coefficients, transport rates and chemical reaction rates.

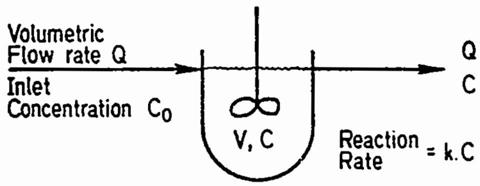
The thermodynamic properties, such as density, heat capacity and chemical equilibrium constants can be estimated with reasonable confidence from mechanistic models and can be measured with reasonable accuracy. The transport coefficients, such as viscosity, thermal conductivity and diffusivity, and the chemical reaction rate coefficients can rarely be predicted and are difficult to measure. Even the definition of these latter quantities is somewhat arbitrary.

The physical, chemical and biological transformations of interest in biofouling are completed in a certain period of time. With respect to biofouling, a specified change may signal the shutdown of operations and the beginning of cleaning operations. The time required for this specified change is inversely proportional to the rate at which the process occurs. Thus, rate is the most important quantity in process analysis.

The equations for conservation of mass, energy, momentum and chemical species equate the rate of accumulation to the net rate of input by flow and the net rate of input by various rate processes such as chemical reaction, diffusion, radiation, convection and viscous dissipation. The <u>process rates</u> are fundamental quantities in that they can be generalized and correlated simply with factors such as temperature, pressure, composition, velocity and diameter which describe the environment. The <u>rate of accumulation</u> and the <u>net rate of input by flow</u> are herein called <u>rates of change</u>. These rates of change are observed or measured quantities which may be the result of several process rates. They cannot be correlated simply or generalized. It is essential that rates of change not be confused with process rates (Fig. 1).

The procedures involved in process analysis are: 1) the mathematical description of the rate of change, i.e., the rate of accumulation in batch operations and the net rate of input by flow in continuous operations, 2) the experimental measurement of the rate of change and the determination

RATE PROCESS ANALYSIS



MATERIAL BALANCE ON COMPONENT C

$$\frac{d(vc)}{dt} = Q\,(C_0 - C) - k.C$$

Rate of Accumulation	Net Rate of Input by Flow	Output by Reaction
RATES OF CHANGE		PROCESS RATE

Figure 1

PROCESS RATES can be correlated with temperature, composition and geometry of process rate from the experimental measurements, 3) the correlation and generalization of the process data, and 4) the use of the rate data in process calculations, including conditions outside the range of the experimental conditions.

Introduction to Microbial Fouling Processes

Microbial fouling is the net result of several physical, chemical and biological processes including the following:

- organic adsorption at the wetted surface
- transport of the microorganisms to the wetted surface
- microorganism attachment to the surface
- metabolism and growth of attached microorganisms
- detachment or reentrainment of biofilm by fluid shear stress

Organic adsorption. Microorganisms select their habitats on the basis of many factors, including the nature of the wetted surface (material of construction and surface roughness).

Figure 2 illustrates an initially "clean" surface exposed to turbulent flow of a fluid containing dispersed microorganisms, nutrients, and organic macromolecules. Adsorption of an organic monolayer occurs within minutes of exposure as shown in Fig. 3. Investigations have shown that materials with diverse surface properties (e.g., wettability, surface tension, electrophoretic mobility) are rapidly conditioned by adsorbing organics when exposed to natural waters with low organic concentrations.

Transport of microbial particles to the surface. Figure 4 indicates the physical transport of bacterial particles from the bulk fluid to the surface covered by an organic film. Within a turbulent flow regime, particles suspended within the fluid are transported to the solid surface by two mechanisms: molecular diffusion and turbulent eddy transport. Theory indicates that the flux of particles to the surface increases with

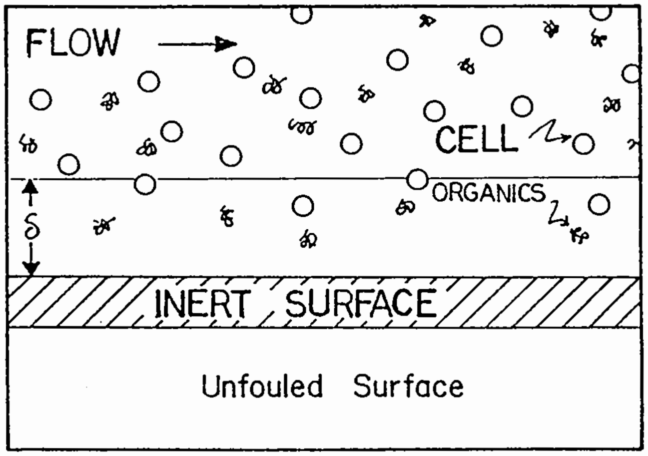


Fig. 2. Initially clean surface exposed to a turbulent flow of fluid containing microorganisms and associated material.

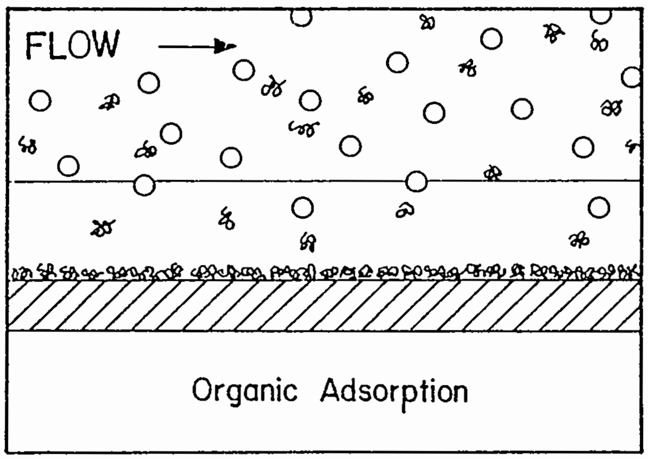


Fig. 3. Adsorption of organic material from the bulk fluid.

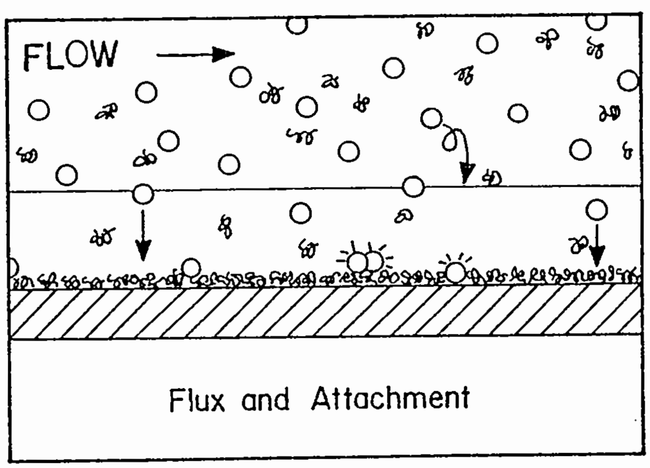


Fig. 4. Flux and attachment of microbial cells to the surface from the bulk fluid.

increasing fluid velocity and particle concentration. However, particle flux is also strongly dependent on the physical properties of the particles (e.g., size, shape, density) and may be influenced by other hydrodynamic processes near the attachment surface.

Microorganism attachment to the surface. Research suggests the existence of a two-stage attachment process: reversible adhesion followed by an irreversible attachment. Reversible attachment refers to an initially weak adhesion of a bacterium to a surface. Organisms still exhibit Brownian motion and are readily removed by mild rinsing. Conversely, irreversible attachment is usually aided by the production of extracellular polymers and attached cells can not be removed easily.

Many microbial cell attachment studies have been conducted at relatively low fluid shear rates or under quiescent conditions. Rates of accumulation determined from these studies are potentially mass transfer-limited and may not be applicable to condenser biofouling where fluid shear rates are quite high (equivalent to average fluid velocities of 6 ft/sec in a 3/4' I.D. tube).

Metabolism and growth of attached microorganisms. Attached and dispersed microorganisms assimilate nutrients, synthesize new biomass and produce extracellular polymers. Biomass production on the surface is the net result of cell division and extracellular product formation, as shown in Fig. 5.

Biofilm growth has been described by a wide variety of rate expressions whose rate constants are functions of pH, temperature, limiting nutrient concentration, nutrient type, terminal electron acceptor, and organism concentration.

Postulated rate expressions for nutrient depletion by a fixed biofilm

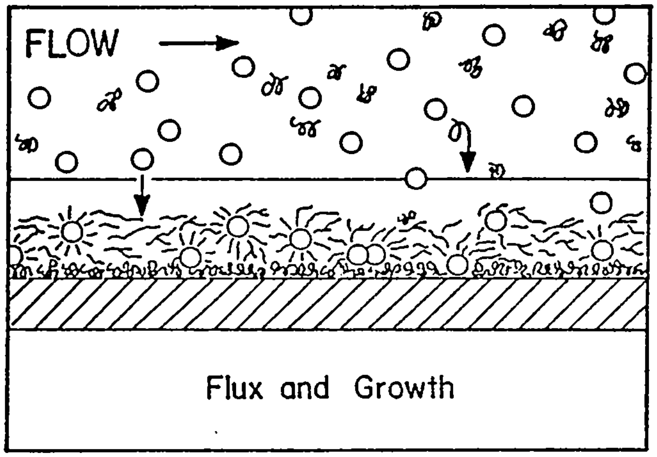


Fig. 5. Continued flux of microbial cells to the surface with simultaneous growth processes occurring.

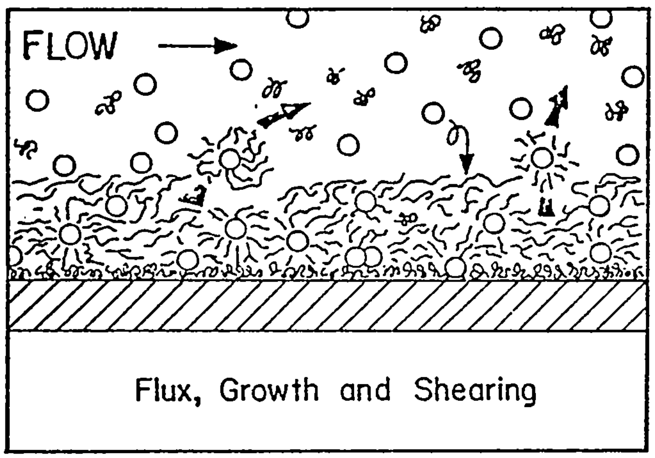


Fig. 6. Continued flux of microbial cells to the surface and simultaneous growth processes opposed by attachment or reentrainment of biomass due to fluid shear.

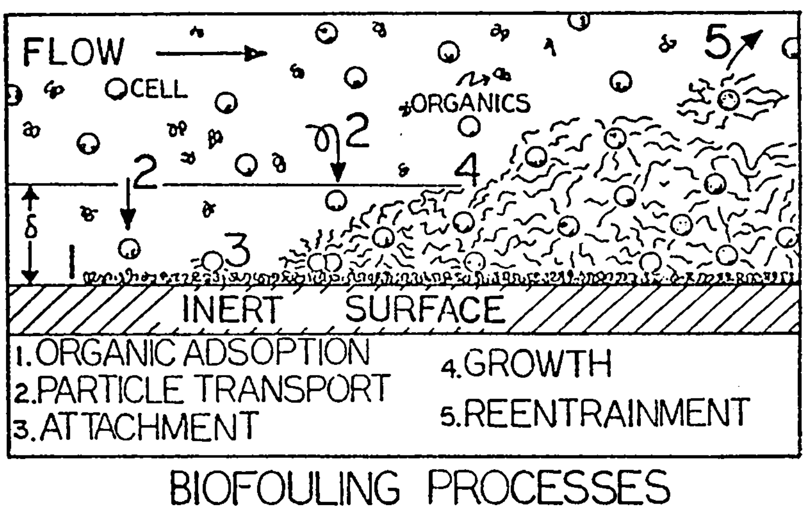


Fig. 7. Summary diagram of biofouling processes.

are numerous, but all agree that nutrient depletion rates are first order in biofilm mass for thin films and that diffusion rates in the biofilm can often control the overall removal rate of nutrients.

Reentrainment of biofilm by fluid shear. At any point in the development of a biofilm, portions of biofilm peel away from the inert surface and are reentrained in the fluid flow (Fig. 6). Reentrainment is a continuous process of biofilm removal and is highly dependent on hydrodynamic conditions. Sloughing, on the other hand, appears to be a random, massive removal of biofilm attributed to oxygen/nutrient depletion deep within biofilms. Sloughing is more frequently witnessed with thicker, dense films especially in laminar flow. More work is needed to quantify either effect.

In summary biofouling is the net result of all these processes occurring simultaneously (Fig. 7). However, at specific times during biofilm development, certain processes may contribute more than others.

PROPERTIES AND COMPOSITION OF BIOFILMS

Microorganisms, primarily bacteria, adhere to surfaces ranging from the human tooth and intestine to the metal surface of condenser tubes exposed to turbulent flow of water. The microorganisms "stick" by means of extracellular polymer fibers, fabricated and oriented by the cell, that extend from the cell surface to form a tangled matrix termed a "glycocalyx" by Costerton *et al.*, (3). The fibers may conserve and concentrate extracellular enzymes necessary for preparing substrate molecules for ingestion, especially high molecular weight or particulate substrate which is available in natural waters.

The biofilm surface is highly adsorptive, partially due to its polyelectrolyte nature, and can collect significant quantities of silt, clay and other detritus in natural waters.

Physical, chemical and biological properties of biofilms are dependent on the environment to which the attachment surface is exposed. The physical and chemical microenvironment combine to select the prevalent microorganisms which, in turn, modify the environment of the surface. As colonization proceeds and a biofilm develops, the microenvironment is altered and bulk biofilm properties change. Biofilm properties and changes that occur during biofilm development are critical to determining the effect of biofilms on fluid and heat transport in turbulent flow systems but have been largely ignored in past studies.

Physical Properties

The most fundamental biofilm properties are volume (thickness) and mass. In turbulent flow systems, wet biofilm thickness (Th) seldom exceeds 1000 μm (4). The biofilm mass can be determined from the wet biofilm thickness if the biofilm dry mass density (ρ_{Th}) is known. ρ_{Th} reflects the attached dry mass per unit wet biofilm volume and measured values in turbulent flow systems range from 10 - 50 mg/cm^3. ρ_{Th} increases with increasing turbulence (4) and increasing substrate loading (5,4) as indicated in Figs. 8 and 9. The increase in ρ_{Th} with increasing turbulence may be caused by one of the following phenomena:

1. selective attachment of only certain microbial species from the available population
2. microorganism response to environmental stress
3. fluid pressure forces "squeeze" loosely bound water from the biofilm.

The relatively low biofilm mass densities compare well with observed water content of biofilm (6,7,8).

The transport properties of biofilm are of critical importance in quantifying effects of biofilms on mass, heat and momentum transfer.

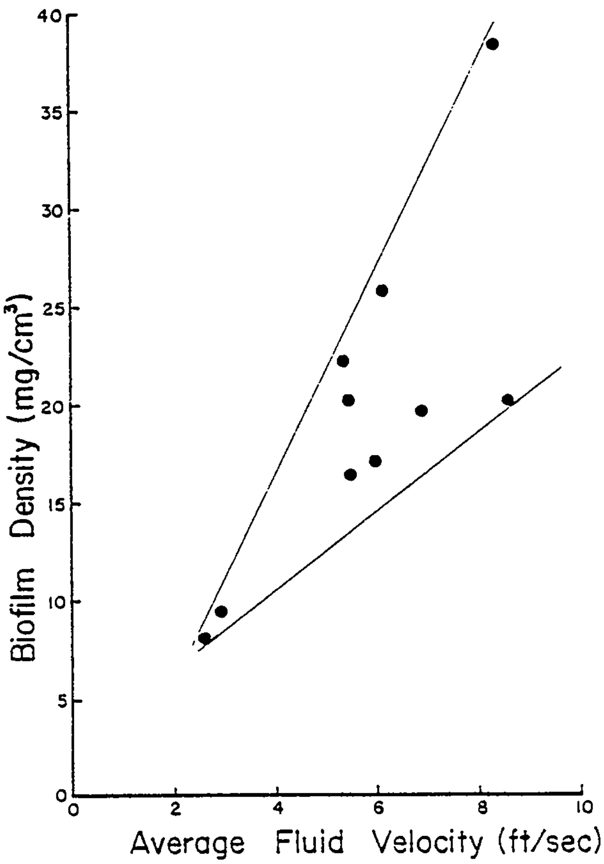


Fig. 8. Effect of fluid velocity on biofilm density for glucose loading rate < 25 mg m^{-2} min^{-1} in a 1.27 cm I.D. tube [4].

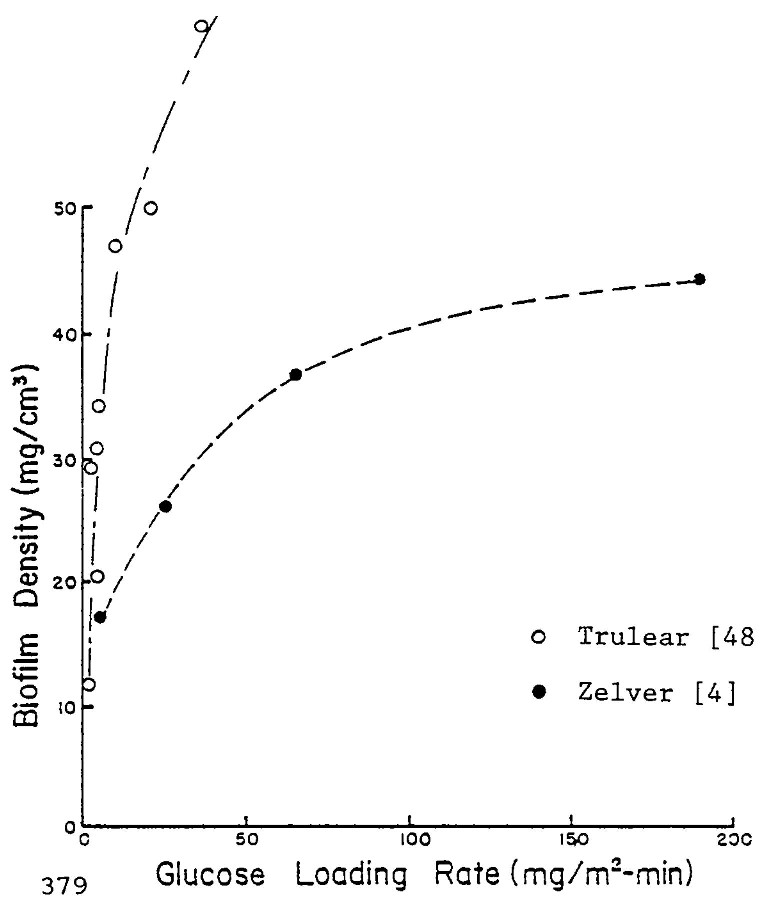


Fig. 9. Effect of glucose loading rate on biofilm density.

Table 2. Experimental diffusion coefficient measurements from the literature [9].

Reactant	Diffusivity $10^{-5}cm^2s^{-1}$	D_{floc}/D_{H_2O} x 100%	Biomass Type	Growth System	Procedure	Reference
Oxygen	1.5	70	Bacterial Slime	Rotating Tube	Reaction Products Analysis	[12]
Oxygen	0.21	8	Fungal Slime *Zooglea ramigera*	Fluidized Reactor	Nonlinear Curve Fit	[13]
Glucose	0.048	8	*Zooglea ramigera*	Fluidized Reactor	Nonlinear Curve Fit	[14]
Glucose	0.06-0.6	10-100	Mixed Culture	Fluidized Reactor	Two Chamber	[15]
Oxygen Ammonia Nitrate	2.2 1.3 1.4	90 80 90	Nitrifier Culture	Fluidized Reactor	Two Chamber	[16]
Oxygen*	0.4-2.0	20-100	Mixed Culture	Fluidized Reactor	Two Chamber	[17]
Glucose*	0.06-0.21	10-30				

* Tests conducted under a variety of experimental conditions.

Table 3. Viscoelastic properties of biofilm developed at 40°C at a fluid shear stress of 3.3 $N\,m^{-2}$. Glucose was growth-limiting and was applied at 6.2 $mg\,m^{-2}\,min^{-1}$ [7].

Elastic (storage) Modulus	59.5 $N\ m^{-2}$
Viscous (loss) Modulus	118 $N\ m^{-2}$

Table 4. Thermal conductivity of biofilm and other selected materials relevant to biofouling of heat exchangers.

Material	Thermal Conductivity ($W\ m^{-1}\ {}^{\circ}K^{-1}$)	Temperature (°C)	Reference
Biofilm	0.68 ± 0.27	28.3 ± 0.3	[6]
	0.71 ± 0.39	26.7 ± 0.3	
	0.57 ± 0.10	28.3 ± 0.3	
Water	0.61	26.7	[18]
	0.62	28.3	
Carbon Steel	51.92	0–100	[19]
Steel	46.86	18	[20]
Stainless Steel (type 316)	16.30	0–100	[19]
Aluminúm 5052	138.46	20	[19]
	205.85	100	[20]
Cupronickel 10% 706	44.71	0–100	[19]
Copper	384.	18	[20]
Titanium (commercial pure)	16.44	0–100	[19]
Glass	0.6 – 0.9	---	[18]

Diffusion coefficients for various compounds through microbial aggregates have been reported in the literature (9), mostly for floc particles (Table 2). Matson and Characklis (9) report variation in the diffusion coefficient for glucose and oxygen with growth rate and carbon-to-nitrogen ratio. In biofilms, the diffusion coefficient is most probably related to biofilm density. *In situ* rheological measurements indicate that the biofilm is viscoelastic with a relatively high viscous modulus as indicated in Table 3 (7). Reported biofilm thermal conductivities are presented in Table 4. As expected from reported water content, biofilm thermal conductivity is not significantly different from water.

Chemical Properties

Inorganic composition of biofilms undoubtedly varies with the chemical composition of the bulk water and probably affects the physical and biological structure of the film. Calcium, magnesium and iron affect intermolecular bonding of biofilm polymers which are partially responsible for the structural integrity of the deposit. For example, EDTA is effective, although costly, in detaching biofilm (7). In heat exchangers, corrosion products and inert suspended solids can adsorb to the biofilm matrix and influence its chemical composition. Table 5 reports the range of inorganic composition observed in selected biofilms.

The organic composition of the biofilm is strongly related to the energy and carbon sources available for metabolism. Classical papers (10,11) have demonstrated the effect of environment and microbial growth rate on the composition of the cells and their extracellular products. For example, nitrogen limitation can result in production of copious quantities of microbial extracellular polysaccharides. Table 6 presents data on the composition of biofilms developed in the field and in the laboratory. In terms of ma-

Table 5. Chemical properties of biofilms obtained from fouled surfaces experiencing excessive frictional losses (after Characklis [8]).

	REFERENCE				
	[21]	[22]	[22]	[23]	[7]
Water	87	85.6	90	95	96
Volatile Fraction	2.5	2.7	1.9	2.4	3.2
Fixed Fraction	10.5	11.7	8.1	2.6	0.8
Si (as percent fixed fraction)		7.0	11.8	12.5	
Fe		18.5	7.9	1.4	
Al		7.5		3.9	
Ca		1.0	5.6		
Mg		2.5		3.2	
Mn		59.5	56.3	4.9	

Table 6. Chemical composition of biofilms obtained in the field and laboratory emphasizing the primary constituents (C,N,P).

	% DRY WEIGHT						
Source	C	N	P	Fixed Solids	C/N	C/P	Reference
Biofilm - power plant condenser	6.4 - 13.8	0.5-3.0	--	--	2-27	---	[24]
Biofilm - laboratory reactor	42.8	10.0	--	--	4.3	---	[25]
Biofilm - laboratory reactor	19.0	9.2	1.8	20	2.1	10.5	[7]
E. coli	50.0	14.0	3.0	--	3.6	16.7	[26]

cromolecular composition, Bryers (27) has measured protein-to-polysaccharide mass ratios ranging from 0 to 10 (polysaccharide concentration in terms of glucose and protein concentration based on casein) with increasing biofilm accumulation. Other chemical analyses of biofilm have been reported by Bryers and Characklis (28).

Biological Properties

The organisms which colonize the attachment surface will strongly influence biofilm development rate and biofilm chemical and physical properties. However, organism-organism and organism-environment interactions undoubtedly shift population distributions during biofilm accumulation. Several investigators have observed succession during biofouling (29,30).

The first visible sign of microbial activity on a surface is usually small "colonies" of cells distributed randomly on the surface. As biofilm development continues, the colonies grow together forming a relatively uniform biofilm. The viable cell numbers are relatively low in relation to the biofilm volume (10^4 - 10^8 cm^{-3} biofilm) occupying only from 1-10% of the biofilm (7). Costerton (31) and Jones _et al._, (32), present photomicrographs which corroborate these data in natural and laboratory systems.

In many cases, filamentous forms emerge as the biofilm develops further. _Hyphomicrobium_, _Sphaerotilus_ and _Beggiatoa_ are frequently identified. The filamentous forms may gain an ecological advantage as the biofilm develops since their cells can extend into the flow to obtain needed nutrients or oxygen which may be depleted in the deeper portions.

RATE PROCESSES CONTRIBUTING TO BIOFILM DEVELOPMENT

Transport from the Bulk Fluid to the Wall

The transport and deposition of entrained particles from a turbulent stream onto the walls of a pipe have been investigated by Friedlander and Johnstone (33) for large particles (>1μm in diameter) and by Lin (34) for small particles (soluble components). Beal (35) has described a unified approach to cover a size range including these extremes which agrees reasonably well with published data. In microbial fouling, the particles of interest range in size from approximately 0.5-20 μm and are "in between" the size ranges that have occupied many researchers' interests. Therefore, Beal's model will be used for this discussion.

Beal begins with an equation describing particle flux, N, as follows:

$$N = (D + D_e) \frac{dC}{dy} \qquad (1)$$

where N = particle flux $(L^{-2}t^{-1})$

D = diffusion coefficient (L^2t^{-1})

D_e = eddy diffusion coefficient (L^2t^{-1})

C = particle number concentration (L^{-3})

y = distance from the wall (L)

The diffusion coefficients for various particles of interest are presented in Table 7. Eddy diffusion coefficients are dependent on turbulent intensity and therefore vary across the pipe section for a given particle. Eddy diffusion coefficients are characteristically orders of magnitude greater than diffusion coefficients (i.e., $D_e >> D$). Beal integrated Eq.(1) to obtain N_w, the flux at the wall:

$$N_w = K_D C_{avg} \qquad (2)$$

Table 7. Calculated diffusion coefficients for particles of interest.

	Molecular Weight ($g\ mol^{-1}$)	Particle Diameter (μm)	Diffusion Coefficient x 10^{10} ($cm^2\ sec^{-1}$)	Temperature (°C)
Tobacco mosaic virus	31,400,000	-	58.8 [1]	20
Myoglobin	17,000	-	5320 [1]	20
Glucose	180	-	97300 [1] 67300 [2]	25
Particle[3]		10	4.3	20
		5	8.6	20
		3	14.2	20
		3	18.5	30
		1	42.8	20
		0.1	430	20
		0.01	4300	20

[1] Calculated from Wilke-Chang equation.

[2] Measured Value [8]

[3] Calculated from Stokes-Einstein equation.

where

$$K_D = \frac{K v_r P}{K + v_r p} \tag{3}$$

K_D = deposition coefficient (Lt^{-1})

v_r = radial velocity of the particle (Lt^{-1})

K = transfer coefficient (Lt^{-1})

p = "sticking" efficiency of the particles (dimensionless)

The radial velocity of the particle, v_r, can be considered the result of the following: (1) motility generated by internal energy (e.g., chemotactic response by a bacterial cell), v_M, (2) sedimentation, v_s, (3) Brownian motion, v_B, and (4) fluid motion, v_f. An analysis of the relative contribution of the components indicates that v_B is dominant for particles up to about 1 μm in turbulent flow. Above 1 μm, fluid motion (in the turbulent flow regime) is most important.

Two simplified forms of Eq. (3) are useful. If $K \gg pv_r$, the only particle concentration gradient is very close to the wall and

$$N_w = pv_r C_{avg} \qquad (K \gg pv_r) \tag{4}$$

If $pv_r \gg K$, transport through the bulk fluid in the pipe controls wall flux and

$$N_w = KC_{avg} \qquad (pv_r \gg K) \tag{5}$$

For microbial fouling, $pv_r \gg K$, based on calculations assuming the following (Table 6):

particle density	=	1.07 g/cm^3
particle diameter	=	0.01 - 10 μm
avg. fluid flow velocity	=	50-350 cm/sec (in a 1.27 cm ID tube)
sticking efficiency	=	1.0

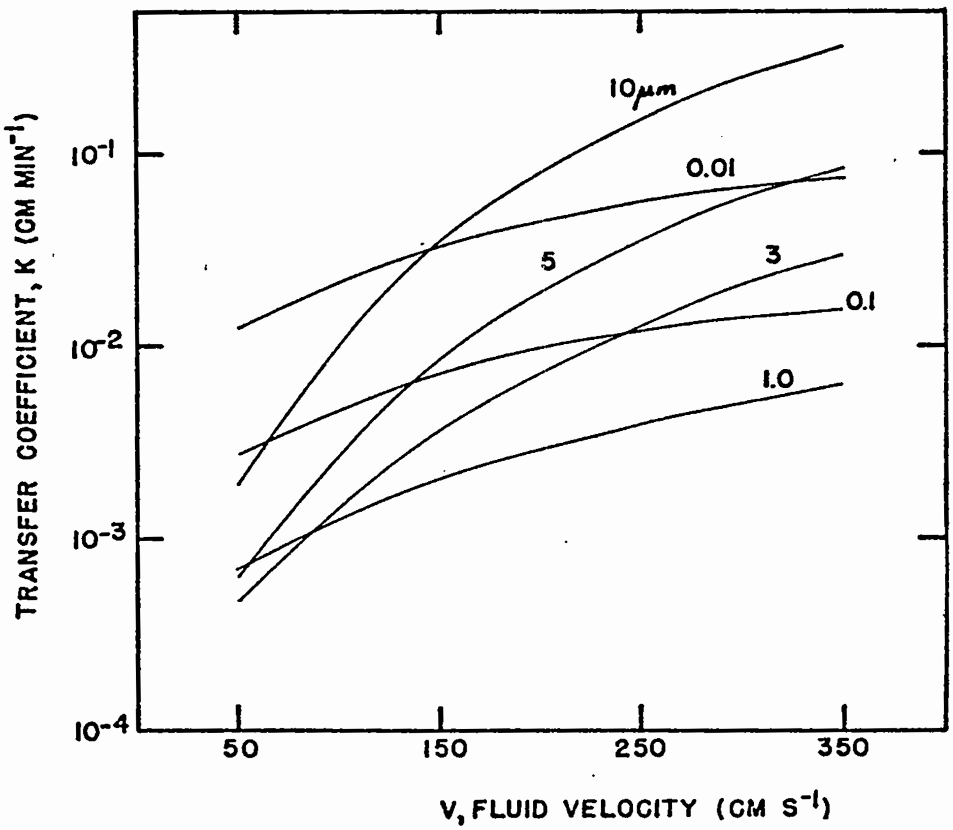


Fig. 10. The influence of fluid velocity (1.27 cm I.D. tube) and particle diameter (particle density = 1.07 g cm^{-3}) on the transfer coefficient, K, calculated from Beal [35].

Figure 10 illustrates the effect of fluid velocity and particle diameter on K. Bryers (27) results suggest that sticking efficiency is considerably less than 1.0 (approximately 0.01) and calculations reported in Fig. 10 must be used cautiously.

Accumulation of organic material at the wall in the early stages of biofouling (assuming growth is negligible) is a result of two fluxes: (1) microbial cells and (2) organic macromolecules. So

$$(N_w)_{total} = (N_w)_M + (N_w)_X \tag{6}$$

where

$$(N_w) = K_M C_{Mavg} \tag{7}$$

$= \text{flux of macromolecules } (L^{-2}t^{-1})$

$$(N_w)_X = K_X C_{Xavg} \tag{8}$$

$= \text{flux of microbial cells } (L^{-2}t^{-1})$

$$(N_w)_{total} = K_M C_{Mavg} + K_X C_{Xavg} \tag{9}$$

where

K_M = transfer coefficient for macromolecules (Lt^{-1})

K_X = transfer coefficient for microbial cells (Lt^{-1})

C_{Mavg} = average macromolecule number concentration in the bulk fluid (L^{-3})

C_{Xavg} = average microbial cell number concentration in the bulk fluid (L^{-3})

In calculating flux, the units of measurement become critical in interpreting the data with respect to biofouling. Assuming a microbial cell diameter of 3 μm and a cell density of 1.07 g/cm^3, the mass flux of macromolecules and microbial cells are equal when

$$\frac{K_M C_{Mavg} M_w}{K_X C_{Xavg}} = 10^{12} \tag{10}$$

where M_w = molecular weight of macromolecule ($gmol^{-1}$)

An illustrative example is presented in Table 8 which suggests that deposition of macromolecules is negligible in terms of mass flux.

Table 8. A comparison of particle flux and mass flux for a hypothetical water source flowing at 150 cm $^{-1}$ in a circular tube (1.27 I.D.). Macromolecular particle flux is much greater than microbial particle flux. However, the reverse is true for mass flux.

	Macromolecule	Microbial Cell
Carbon Content (%)	40	50
Particle Mass Concentration in Bulk Fluid ($\mu g\ cm^{-3}$)	2	7.6 [1]
Molecular Weight ($g\ mol^{-1}$)	50,000	--
Particle Number Concentration in Bulk Fluid (cm^{-3})	6×10^{13}	1×10^{6}
Transfer Coefficient ($cm\ s^{-1}$) (from Table 7)	6×10^{-4}	1×10^{-4}
Particle Flux ($cm^{-2}\ s^{-1}$)	3.6×10^{10}	100
Mass Flux ($g\ cm^{-2}\ s^{-1}$)	5×10^{-23}	1.5×10^{-9}

[1]Assuming cell diameter = 3 μm and cell density = 1.07 g/cm^3

Molecular Fouling

Figure 2 illustrates an initially "clean" surface exposed to turbulent flow of a fluid containing dispersed microorganisms, nutrients, and organic macromolecules. Within a short period (minutes) of exposure, adsorption of measurable quantities of the organic molecules occurs (Fig. 3). These molecules are usually polysaccharides of glycoproteins. Loeb and Neihof (37) and DePalma et al. (38) have measured rates of adsorption in seawater, and Bryers (27) has observed adsorption in a laboratory system. Rates and extent of adsorption of these investigations are presented in Table 9. Maximum accumulation from molecular fouling is less than 0.1 μm. The rate of molecular fouling is much greater than rates of microbial fouling which is generally reported in terms of days or weeks. Consequently, molecular fouling can be considered instantaneous. However, based on "thickness" measurements, molecular fouling can have no significant effect on fluid flow or heat transfer. Nevertheless, the surface properties resulting from adsorption of an organic film may affect the sequence of microbial events which follow.

Costerton et al. (3) have discussed the pronounced specificity of some bacteria that attack only a particular animal host tissue and suggest that specificity may be explained by the specificity of the host-tissue glycocalyx. It remains to be seen whether a surface, wetted by the adsorption of organic molecules indigenous to that environment, will be initially colonized by a specific microbial cell.

Brash and Samak (39) present experimental evidence that significant turnover occurs in molecular fouling films (proteinaceous) on polyethylene. This may be occurring in microbial systems even when relatively thick films have developed.

Table 9. Maximum rate and extent of molecular fouling.

Maximum Rate (nm/min)	Maximum Accumulation (nm)	Maximum Accumulation (μg COD/cm^2)	Surface	Reference
0.15 - 0.45	30 - 80		Pt[1]	[37]
0.004	7.1		Ge[2]	[38]
0.044	77.3		Ti[2]	
0.01 [5]	13.5 [5]	1.5	glass[3]	[27]
0.22 [5]	22.5 [5]	2.5	glass[4]	

[1]Immersed in quiescent Chesapeake Bay water (3-4° C) containing 2.3 mg carbon/l, salinity between 9-16 °/oo and pH between 7.9 - 8.2

[2]Gulf of Mexico water (22° C) flowing past the surface as a fluid shear stress of 7.1 N/m^2. Salinity was 34 °/oo. Carbon concentration not reported.

[3]Medium consisted of a sterile 1:1 w/w of trypticase soy broth - glucose mixture (34° C; pH 8). The glass surfaces were immersed in tubes placed in a mechanical shaker. Carbon concentration was approximately 80 mg carbon/l.

[4]Medium was effluent (30° C; pH 8) from a chemostat (10-20 mg/l COD, 3 mg/l polysaccharide) with no primary substrate remaining. Microorganisms were present (approximately 10^6 cells/ml) but no cells attached during the period of interest. Fluid shear stress was 3.8 N/m^2.

[5]Estimated from measurements of chemical oxygen demand (COD) adsorbed per unit area. Assumed COD of protein is 0.855 mg COD/mg protein and protein density is 1.3g protein/cm^3.

Attachment of Microorganisms

Previous research (40,41) suggests the existence of a two-stage attachment process: (1) reversible adhesion followed by, (2) an irreversible attachment. Reversible attachment refers to an initially weak adhesion of a cell which can still exhibit Brownian motion but is readily removed by mild rinsing. Conversely, irreversible attachment is a permanent bonding to the surface, usually aided by the production of extracellular polymers. Cells attached in this way can only be removed by rather severe mechanical stress. Marshall (29) and Corpe (42) have implicated polysaccharides and glycoproteins in irreversible attachment.

Most of the research on cell attachment has been conducted at very low fluid shear stress or in quiescent conditions. There is yet to be a demonstration of reversible adhesion in turbulent flow.

Fletcher (43) presents an excellent analysis of the rate of attachment of a marine pseudomonad. The model and results are very similar to those observed in molecular adsorption from solutions onto surfaces. In both cases, the process rate is controlled by the concentration in the bulk solution. The Langmuir adsorption isotherm model was used by Fletcher and assumes the following:

1. adsorption is limited to a monocellular layer
2. adsorbed components are restricted to specific adsorption sites
3. the heat of adsorption is independent of surface coverage.

Fletcher observed no cell desorption and modified the Langmuir model accordingly.

According to Fletcher's development, the rate of attachment is proportional to the cell concentration in the bulk fluid and the probability of a cell contacting a vacant adsorption site:

$$R = k\ X_B\ (1 - \theta) \tag{11}$$

where R = rate of cell attachment $(L^{-2}t^{-1})$

X_B = cell number concentration in the bulk fluid (L^{-3})

θ = fraction of surface covered by cells (dimensionless)

k = rate constant (Lt^{-1})

The number of cells adsorbed is proportional to the fraction of surface covered so:

$$X_A = k'\theta \tag{12}$$

where X_A = number of cells adsorbed per unit area (L^{-2})

k' = number of cells required for total coverage (L^{-2})

Then, by combining Eqs. (11) and (12):

$$R = k\ X_B\ (1 - \frac{X_A}{k'})$$

k' can be obtained from "adsorption isotherms" and is reported to be from 2-3 x 10^7 cells/cm^2. k is estimated at 1-8 x 10^5 cm/min from Fletcher's data.

Fletcher indicates more rapid attachment with log phase cells and slower rates with cells from the stationary phase and death phase, respectively. More cells attach at higher temperatures (ranging from 3 - 20°C).

The nature of the attachment surface is an important factor affecting attachment in heat exchangers. Wettability or critical surface tension, is the property used most frequently to describe surface characteristics in microbial attachment studies (44,45). In seawater, cell attachment increases with increasing critical surface tension of the surface (including glass, copper, polyethylene, teflon) with the exception of the copper surface on which fewer cells attached (44). The copper may inhibit cell attachment by inhibiting a metabolic process necessary for attachment. Even so, there are many examples of biofouling on cupronickel condenser surfaces (46).

Microbial Reaction Processes

Table 10 is a matrix representation of the fundamental microbial reaction processes. The rows of the matrix define the stoichiometry of the process. For example, the growth process (Row 1) can be represented by the following stoichiometric equation where the ν's are stoichiometric coefficients:

$$-\nu_1 s - \nu_2 z - \nu_3 e + \nu_4 x + \nu_5 p + \nu_6 a = 0 \qquad (14)$$

The columns of the matrix define the "observed" rate as opposed to the process rate. For example, observed substrate removal rate (Col. 1) can be represented by the sum of process rates as follows:

$$-q_s = \alpha_1 \mu + \alpha_2 m + \alpha_3 k_p - \alpha_4 k_L \qquad (15)$$

The α's are "observed" stoichiometric ratios. Table 10 only begins to describe the complexity of microbial metabolism. Nevertheless, this amount of "structure" is useful in modelling the rate processes affecting biofilm development.

Trulear and Characklis (5) have observed substrate removal rate, q_s, and net biomass production rate, μ_n, in an experimental biofilm reactor. It was convenient to express the substrate removal rate as:

$$q_s M_A = q_s \rho A Th \qquad (16)$$

where q_s = specific substrate removal rate (t^{-1})

M_A = total biofilm mass (M)

ρ = biofilm density (ML^{-3})

Th = biofilm thickness (L)

A = wetted surface area (L^2)

Table 10. A matrix representation for the fundamental microbial rate processes.

PROCESS RATE → STOICHIOMETRY		REACTANTS			PRODUCTS				
FUNDAMENTAL PROCESS		Substrate	Nutrient	Electron Acceptor	Biomass		Product		Metabolite
Process	Rate	s	z	e	x_T	x_d	p_e	p_i	a
Growth	μ	–	–	–	+		+	(+)	+
Maintenance									
exogenous	m	–		–			+		+
endogenous	k_e		+	–	–	(+)	+	–	+
Product Formation	k_p	–	–	–			+	+	+
Death									
loss of viability	k_d				–	+			
lysis	k_L	(+)			–	(+)	+		
OBSERVED RATE		q_s	q_z	q_e	μ_n		q_p		q_a

q = specific production or removal rate (t^{-1})

μ_n = net specific growth rate or specific biomass production rate (t^{-1})

x_T = total biomass concentration (ML^{-3})

x_d = inert solids concentration (ML^{-3})

p_e = extracellular microbial product concentration (ML^{-3})

p_i = intracellular microbial product concentration (ML^{-3})

s = substrate concentration (ML^{-3})

z = nutrient concentration (ML^{-3})

e = electron acceptor concentration (ML^{-3})

The substrate removal rate, defined in this way, increases in proportion to biofilm thickenss up to a critical thickness beyond which removal rate remains constant (Fig. 11). The critical thickness is observed to increase with influent substrate concentration (S_i) or, more fundamentally, surface loading rate. This behavior is confirmed by other investigators (47,25,4) and is attributed to nutrient diffusional limitations within the biofilm. Once the biofilm thickness exceeds the depth of substrate or oxygen penetration into the biofilm (Fig. 12), the removal rate is unaffected by further biofilm accumulation.

Observed substrate removal rate cannot be used to distinguish between growth, maintenace, product formation, and death. It seems clear from other data (27) that product formation (primarily polysaccharide) is significant in the early stages of biofilm formation. Maintenance requirements become important as the film gets thicker and substrate does not entirely penetrate the biofilm. These other process rates have not been measured and are critical for determining stoichiometric coefficients and predicitng biofilm development rates.

The substrate removal rate is also dependent on fluid velocity (Fig. 13). At low fluid velocities, a relatively thick mass transfer boundary layer (δ) can cause a liquid phase diffusional resistance which decreases substrate concentration at the liquid-biofilm interface and thereby decreases substrate removal rate (Fig. 14).

Detachment of Biofilm

As the biofilm grows thicker, the fluid shear stress at the biofilm interface generally increases. Also, as biofilms grow thicker, the potential for substrate, oxygen or nutrient limitation in the deeper portions is great. These limitations may weaken the biofilm matrix and cause de-

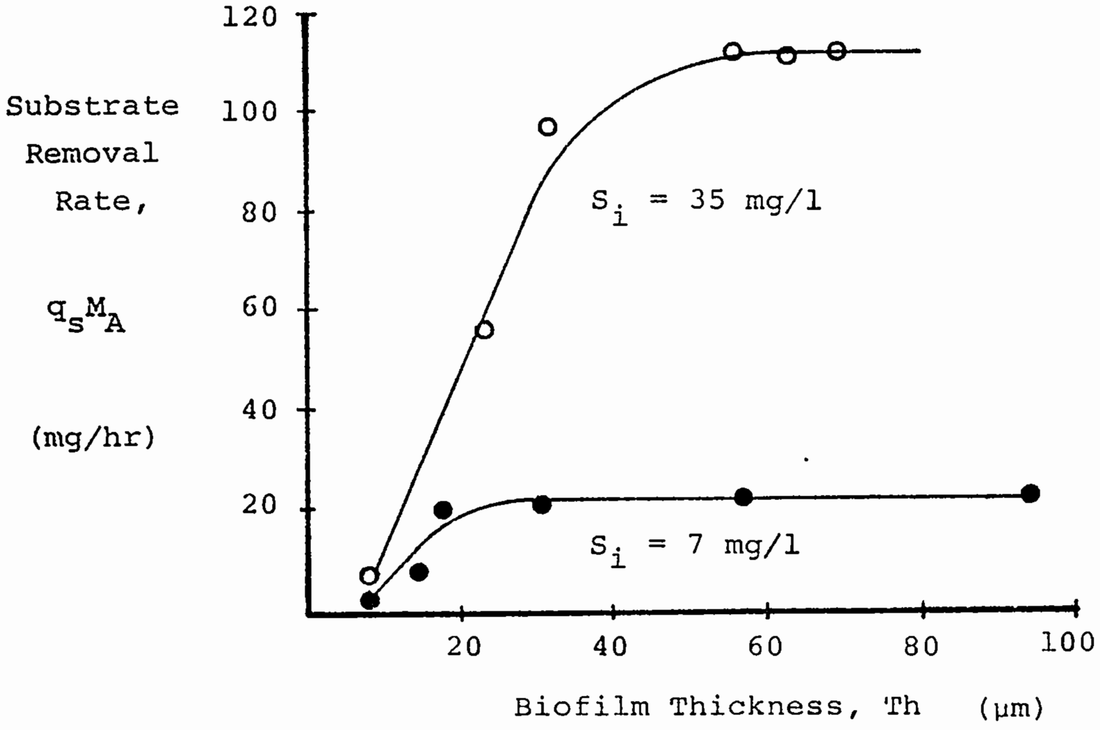


Fig. 11. The influence of biofilm thickness and substrate loading on substrate removal rate [5].

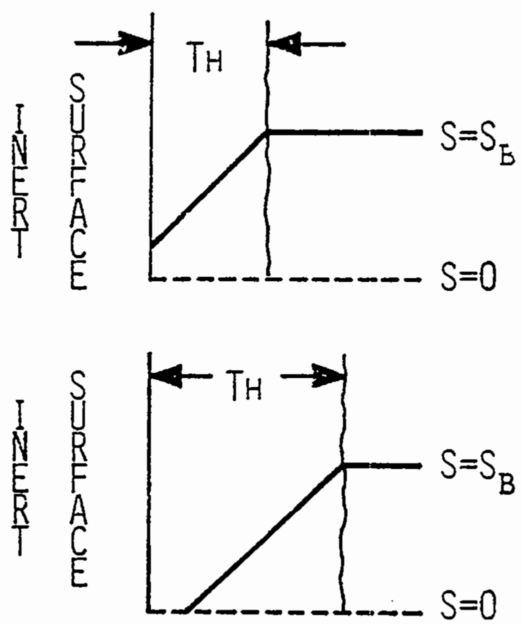


Fig. 12. Diffusional resistances in biofilms result in a constant substrate removal rate after the biofilm reaches a critical thickness beyond which substrate cannot penetrate.

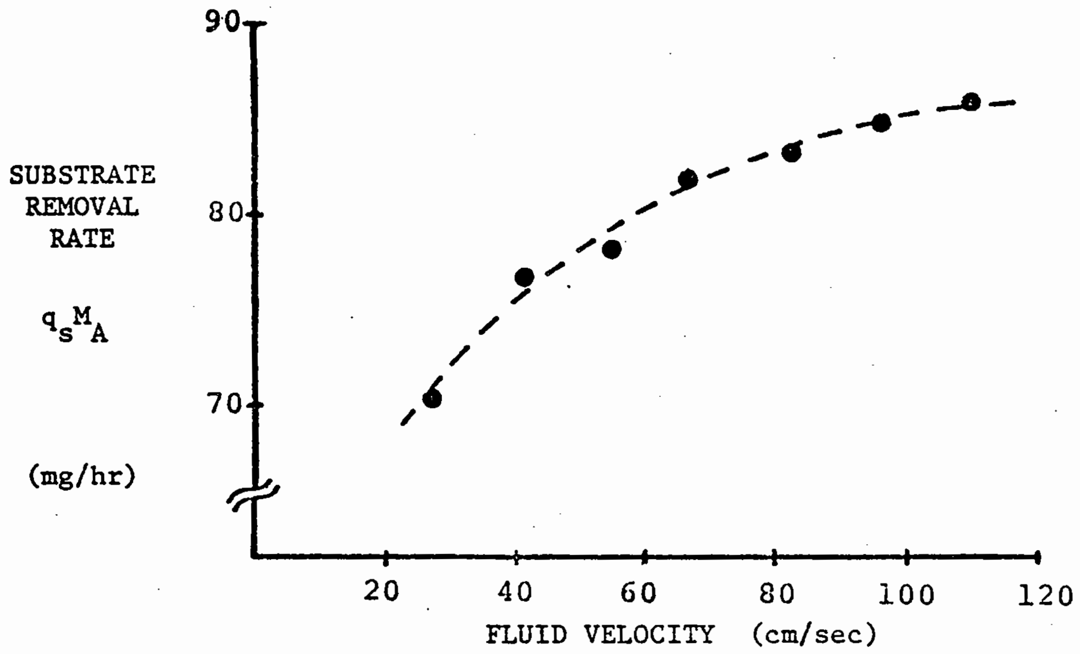


Fig. 13. Influençe of fluid velocity on substrate removal rate [5].

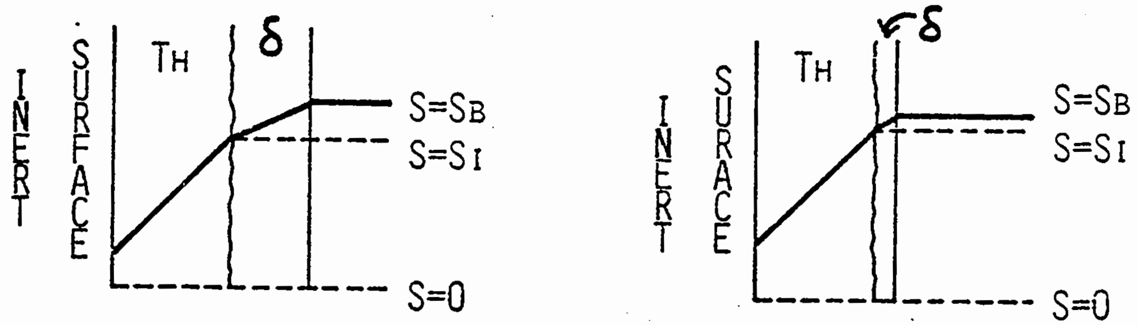


Fig. 14. Increasing fluid velocity past the biofilm results in a smaller viscous sublayer (δ) which increases mass transfer rate through the liquid phase.

tachment. Trulear and Characklis (5) report that the biofilm detachment rate increases with increasing substrate removal rate, probably because thicker biofilms result (Fig. 15).

Overall Rate of Biofilm Development

The development of a biofilm is adequately described by a sigmoidal-shaped curve (Fig. 16). The slope of this curve at a particular time is the net biofilm "development" rate, R_p, and is plotted vs time in Fig. 17a for two experiments at different substrate loadings (5). The rate increases to a maximum value corresponding to the sigmoidal inflection point and then decreases to zero. Trulear (48) has measured maximum biofilm development rates during an experiment which range from 8.3 - 66 x 10^{-5} mg biofilm cm^{-2} min^{-1} for glucose concentrations of 6 and 130 mg l^{-1}, respectively. Since the biofilm detachment rate, R_D, is proportional to biofilm thickness, a higher detachment rate is observed in the high substrate loading experiment (Fig. 17b). Thus,

$$R_D + R_p = \mu_n \qquad (17)$$

where μ_n is the net biomass production rate. At steady state $\mu_n = R_D$ since thickness remains constant.

The effect of fluid velocity on the plateau (or maximum) biofilm thickness is illustrated in Fig. 18 for various substrate loadings. An increase in fluid velocity increases biofilm detachment rate which minimizes the plateau biofilm thickness. However, at low substrate loadings, fluid velocity seems to have little effect on the plateau thickness.

SUMMARY

Microbial film formation has been discussed in terms of the more fundamental physical, chemical and biological processes which contribute to the biomass accumulation at a surface. The discussion suggests that

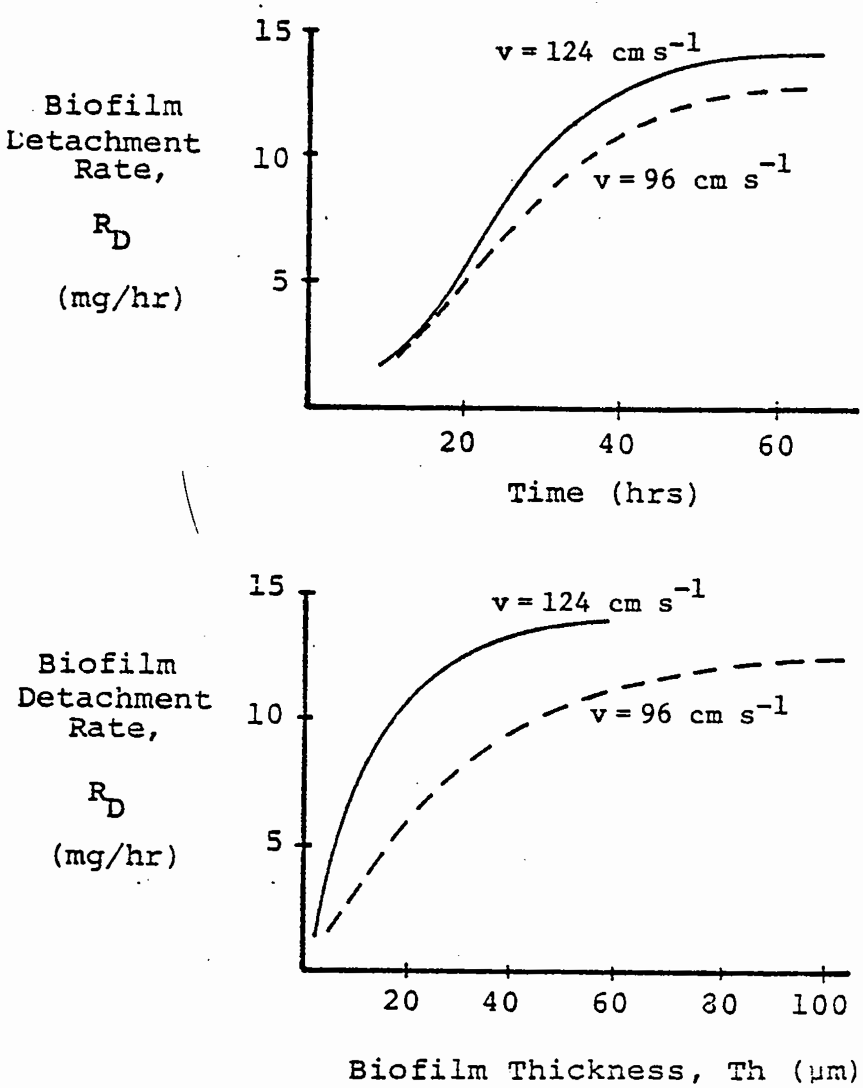


Fig. 15. The influence of biofilm thickness and fluid velocity on biofilm detachment rate [5].

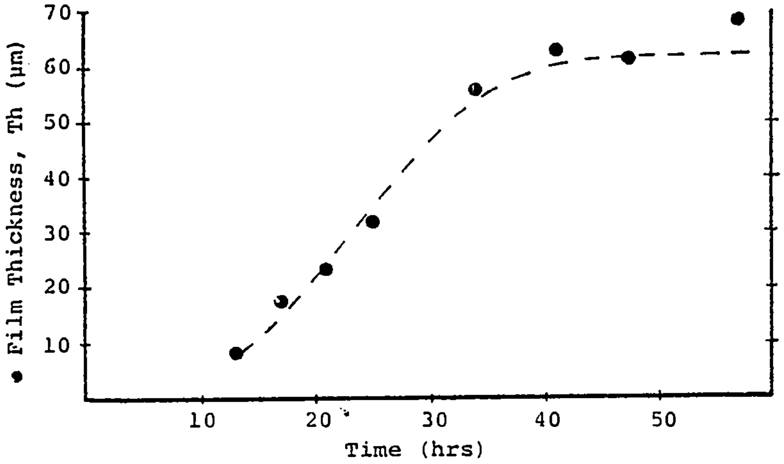


Fig. 16. The progression of biofilm development [5].

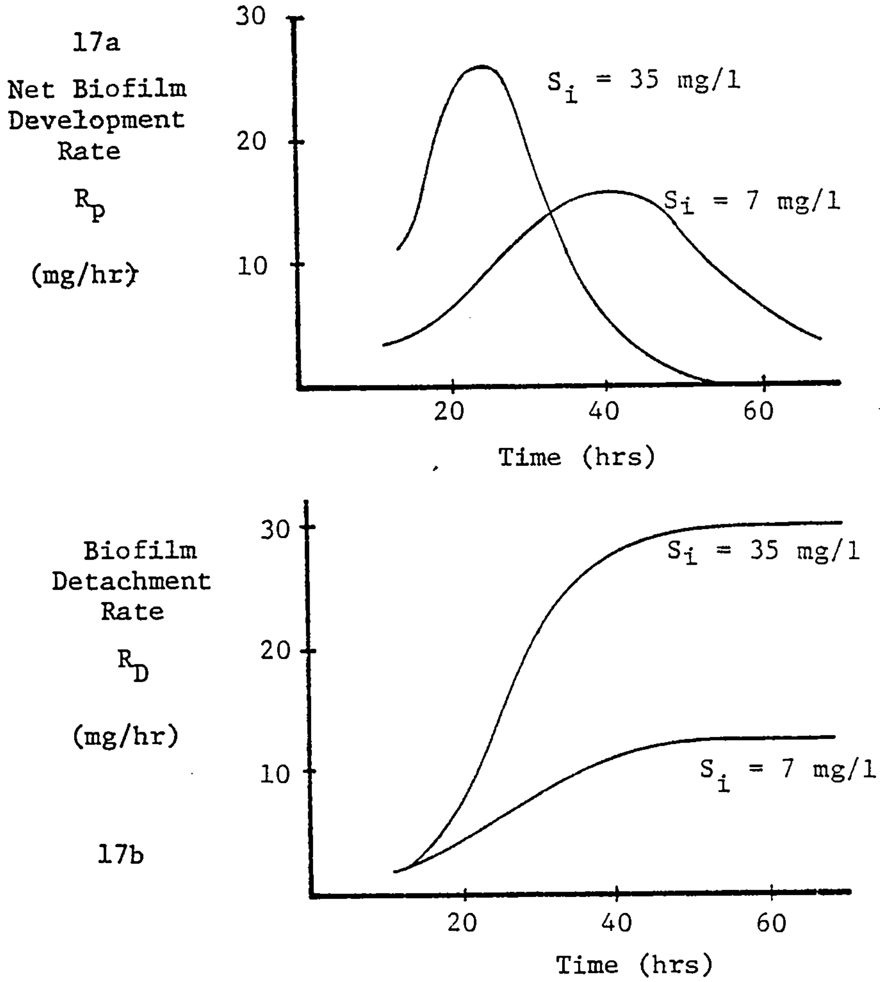


Fig. 17. Changes in biofilm development rate, R_p, and biofilm detachment rate, R_D, at different substrate loading rates. The sum, $R_D + R_p$ is the net biomass production rate, μ_n, for the system [5].

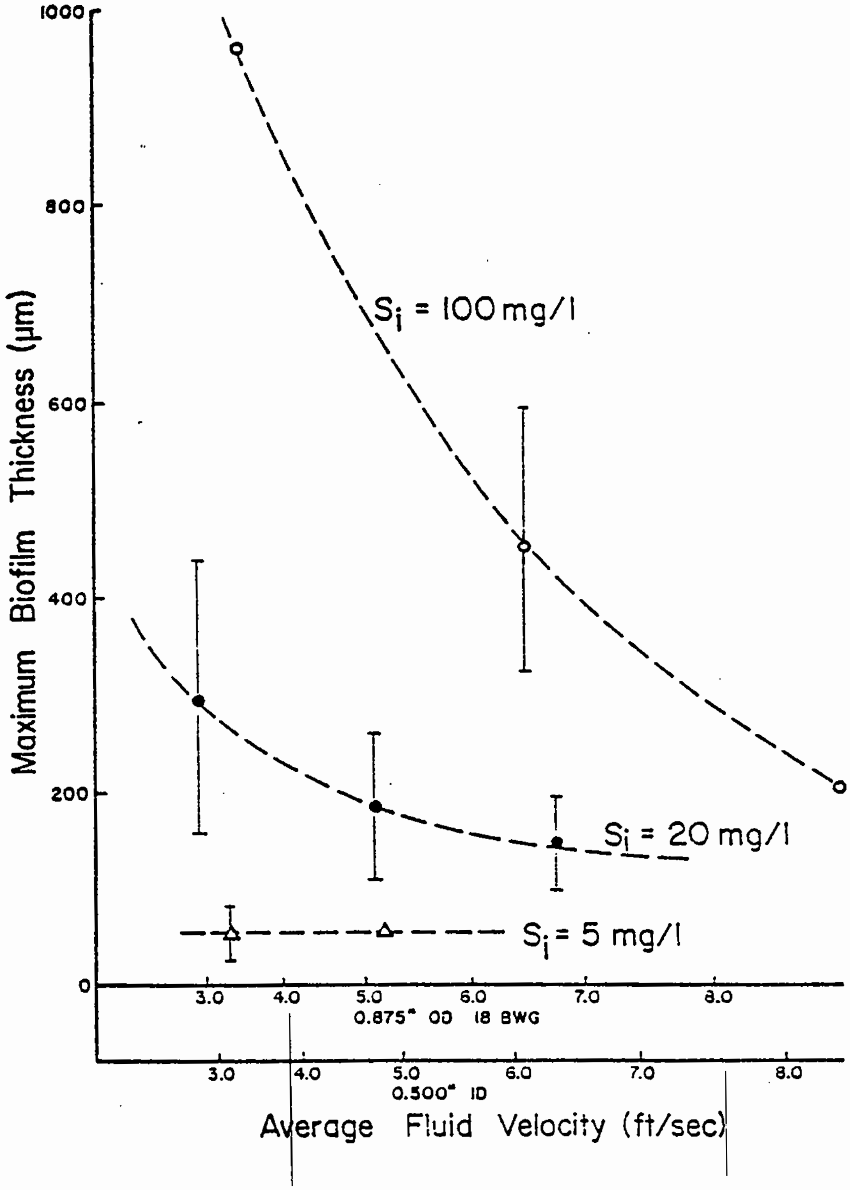


Fig. 18. The influence of fluid velocity and substrate loading on the maximum (or plateau) thickness attained in a 1.27 cm I.D. tube [7].

more attention must be directed at the following topics:

1. More information is needed on the physical, chemical and biological properties and structure of biofilms as a function of the interfacial environment.

2. Mathematical models for process rates as a function of bulk concentrations, surface characteristics, and biofilm composition are needed to ascertain the rate-controlling process in a given environment.

3. The models must be tested by experiments under controlled conditions.

The results of such programs will lend insight necessary for scientists and engineers to design appropriate systems which utilize reactive fixed-film surfaces.

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EFFECTS OF ORGANIC LOADING AND MEAN SOLIDS RETENTION TIME ON NITRIFICATION IN RBC SYSTEMS

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Oxidation of ammonia-nitrogen in rotating biological contactor systems can be economically achieved with domestic and industrial wastewaters using biofilm growths containing nitrifying bacteria. The design of such wastewater nitrification systems is controlled by the kinetics of growth of the autotrophic bacteria, Nitrosomonas and Nitrobacter, which grow at rates lower than those of heterotrophic bacterial populations routinely used in removing carbonaceous organic matter. In addition, autotrophic nitrifying populations are sensitive to such wastewater characteristics as temperature, pH and dissolved oxygen concentration. Therefore, in systems developed to concurrently remove carbonaceous organic matter and oxidize ammonia nitrogen, the growth characteristics of nitrifying bacteria establish the minimum specific growth rate for use in the design of the process system.

Considerable data and experimental relationships[1-9] are available which quantitatively express the effects of wastewater characteristics and process operational parameters on the growth of pure and enriched culture suspensions of *Nitrosomonas* and *Nitrobacter*. The objective of the research herein was therefore to experimentally determine if these fundamental relationships for nitrifying bacteria could be applied directly to attached films in RBC systems. It was, therefore, necessary to develop procedures for evaluating growth rates of nitrifying bacteria in mixed heterogenous cultures. Mean solids retention time, a parameter routinely utilized in the evaluation of net growth rates in suspended growth systems, was utilized to relate RBC reactor performance to the growth characteristics of attached biofilms. The primary objective was to determine if growth and ammonia removal rates for attached nitrifying bacteria could be predicted with relationships established for growth rates of nitrifying bacteria in pure and mixed culture suspensions. The ultimate objective of the research initiated with this project is to determine if these growth rate relationships can be used in design, operation and evaluation of RBC systems.

EXPERIMENTAL PROCEDURES

Experimental Reactor. A laboratory-scale RBC system was used to evaluate the effect of organic loading and mean solids retention time, SRT, on nitrification. The continuous-flow RBC reactor used in the study is presented schematically in Figure 1. As indicated in Figure 2, mixed liquor from the RBC reactor was continuously pumped through a glass-walled heat exchanger and discharged to the surface of a filtration sieve (U. S. Sieve No. 80; sieve opening = 0.177 mm). Mixed liquor passed through the sieve by gravity into the RBC reactor leaving only sloughed biomass on the sieve surface. The mixed liquor recycle flow rate was maintained at 1 ℓ/min, a rate sufficient to circulate the total volume of the RBC reactor approximately one time every 6 min. The use of the heat exchanger, in conjunction with a reactor cover which minimized heat and evaporative losses, was sufficient to maintain mixed liquor temperatures between 21 and 25°C. The gravity-flow sieve effectively retained sloughed biomass and mixed liquor suspended solids concentrations were maintained within limits typically experienced in full scale systems.[10]

The physical characteristics of the RBC reactor are indicated in Table 1.

TABLE 1
Physical Characteristics of Laboratory-scale RBC Reactor

Characteristic Parameter	Value
Number of discs	12
Disc diameter	25.4 cm
Disc thickness	0.3 cm
Disc spacing	2.5 cm
Disc submergence	7.6 cm
Total wetted area	1.0 m^2
Rotational velocity	0.5 rps
Tip velocity	40 cm/s

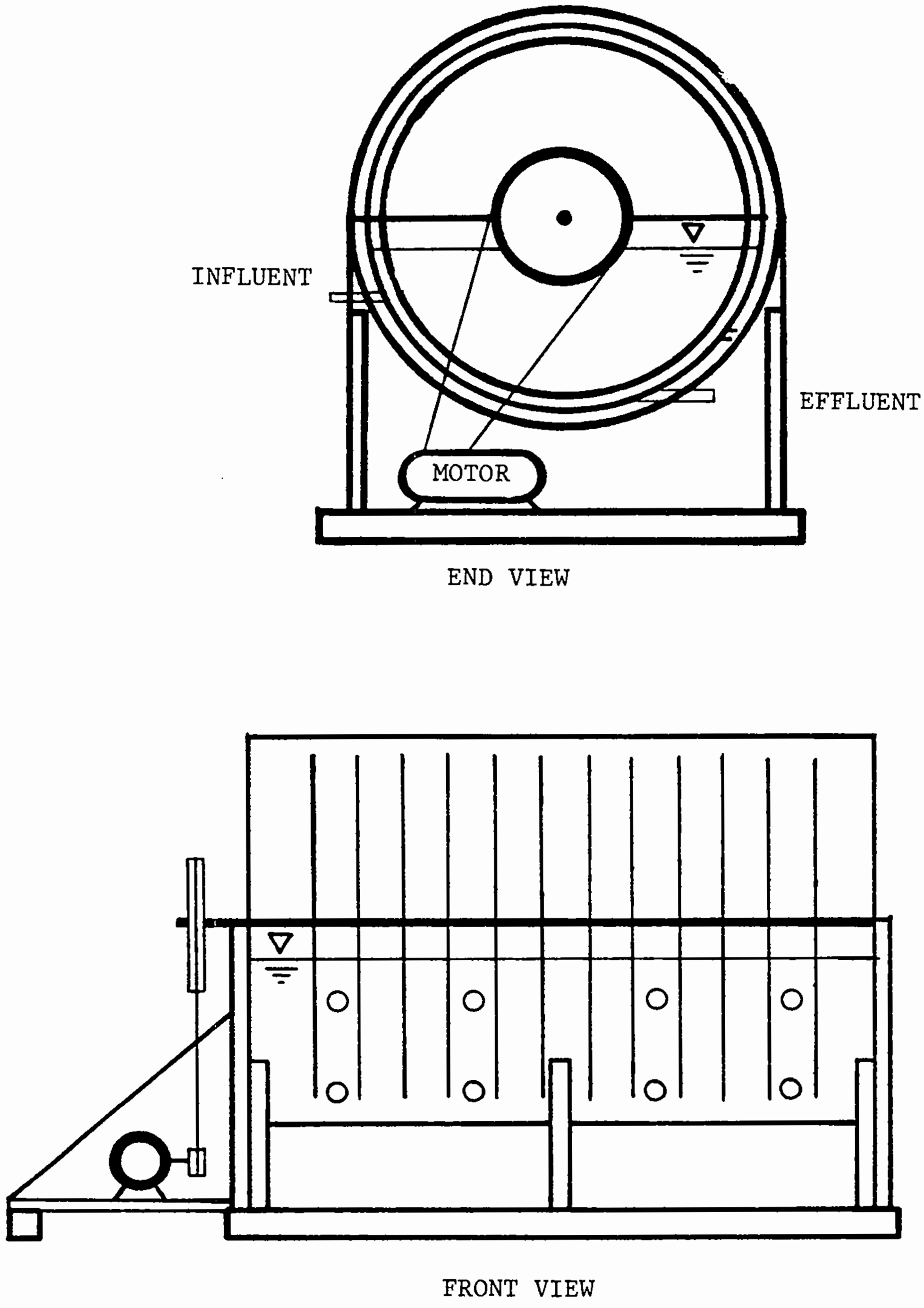

Figure 1. Schematic Diagram of End and Front View of Laboratory-scale RBC Reactor

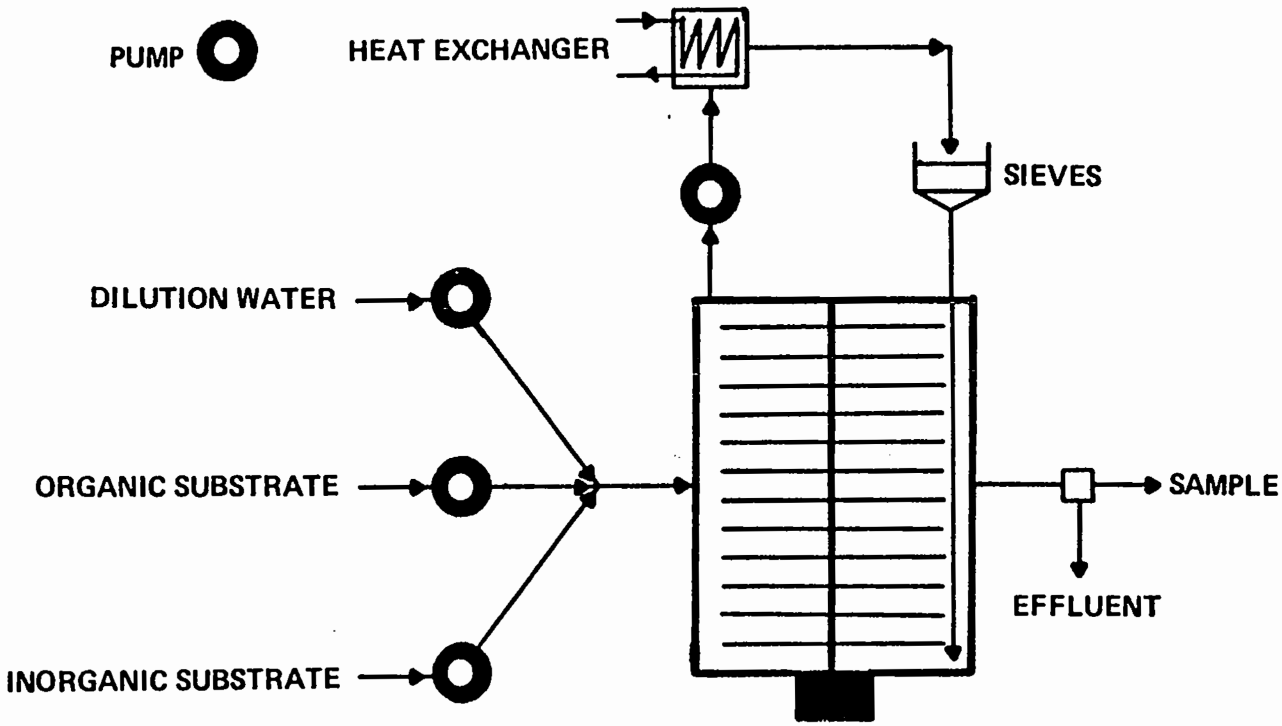

Figure 2. Schematic Diagram of RBC Reactor System

These parameters were held at constant values throughout all experimental runs. Disc rotational velocity was set at 0.5 rps to provide an adequate supply of dissolved oxygen throughout the experimental study. An examination of gas transfer characteristics of the reactor system indicated that the overall oxygen mass transfer coefficient was approximately 1.1×10^{-3} cm/s (20°C), which resulted in a maximum oxygen transfer rate of 8.7 g O_2/d.

Wastewater Composition. A synthetic wastewater was utilized to simulate the soluble portion of a domestic wastewater. The synthetic wastewater was introduced into the reactor in three separate flows, i.e. an inorganic substrate flow, an organic substrate flow and a dilution water flow. The inorganic and organic substrates were each applied at a rate of 5 ml/min, while dilution water was applied at 40 ml/min for a total flow of 50 ml/min and a hydraulic retention time of 2.0 hr. The composition of the inorganic and organic substrates is presented in Table 2. The COD concentration of the composite influent wastewater was

TABLE 2
Composition of Synthetic Wastewater

Inorganic Substrate		Organic Substrate	
Compound	Mass Fraction	Compound	Mass Fraction
$(NH_4)_2SO_4$	12.15	Acetic Acid	1.7
NH_4Cl	0.22	Benzoic Acid	1.4
		Butyric Acid	1.7
KH_2PO_4	1.12	Citric Acid	4.6
K_2HPO_4	2.87	Formic Acid	3.2
$Na_2HPO \cdot 7H_2O$	4.40	Lactic Acid	3.2
		Propionic Acid	1.8
$CaCl_2$	0.66	Valeric Acid	1.7
$MgSO_4$	0.66		
$NaHCO_3$	22.15	Arabinose	10.3
Na_2CO_3	27.95	Galactose	10.3
Na_2SO_4	26.37	Sucrose	26.0
		Xylose	10.3
$MnSO_4 \cdot H_2O$	0.07		
$ZnCl_2$	0.13	Phenol	0.2
$CuSO_4 \cdot 5H_2O$	0.04		
$CoCl_2 \cdot 6H_2O$	0.05	Humics (Tea)	22.6
$Na_2B_4O_7 \cdot 10H_2O$	1.16		
		$(FeCl_3 \cdot 6H_2O)$	(1.0)

varied from 49 to 190 mg/l while influent ammonia concentrations ranged from 15.6 to 20.5 mg NH_4^+-N/ℓ. The COD/N ratio of the influent wastewater, therefore, ranged from 2.5 to 10.5. Phosphorus and other essential nutrients were provided in sufficient quantities so as to not limit the growth of attached microbial populations.

Operation of RBC Reactor. A primary objective of the study was to determine if classical data[1-9] for nitrification growth and substrate removal rates could be applied to attached films in RBC reactor systems. To examine net microbial growth rates in the RBC system, mean solids retention time, SRT, values were controlled during a series of eight experimental runs operated over a range of organic loading rates. Since only organic loading rates were varied and nitrification growth rates were to be examined, an experimental technique was developed to simultaneously control and monitor attached film growth rates. Several procedures were used to routinely monitor the rate of accumulation of (1) sloughed biomass and (2) attached biomass.

Biomass which sloughed periodically from disc surfaces was collected and monitored daily as mixed liquor suspended solids and effluent suspended solids. Mixed liquor suspended solids were those collected on a sieve in the recycle system and those actually in suspension within the RBC reactor. Effluent suspended solids were collected with an effluent composite sampling system.

In addition to natural sloughing of attached biomass, a portion of the biomass attached to disc surfaces was mechanically removed on a regular basis to more effectively control SRT values for the attached biofilm. The twelve discs in the experimental reactor were subdivided into 4 groups of 3 discs each, with disc sides numbered from 1 to 6. The biomass attached to similarly numbered disc sides (i.e., one side of each of four discs) was mechanically removed at intervals of 6, 12 or 18 days. All disc surfaces were sequentially scraped one time every 6, 12 or 18 days. For example, at a scraping interval of 6 days, all biomass on disc sides numbered 1 was removed on day 1, all biomass on disc sides numbered 2 was removed on day 2 and so forth until all sides had been scraped once in a 6 day period. The mass quantity of biomass removed from each disc was monitored every 1, 2 or 3 days, i.e. for scraping intervals of 6, 12 or 18 days, respectively. The scraping of biomass was continued until a steady state response was achieved for each reactor.

The controlled scraping and recovery of attached biomass, and subsequent analysis of biomass contained in the mixed liquor and effluent, provided sufficient data for determination of the net rate at which biomass accumulated in the RBC system. To calculate SRT values for the RBC system, a measure of the total quantity of attached biomass was required. Periodically, the total quantity of attached biomass was estimated through removal of a fixed portion of biomass from each of the 6 disc sides in a single set of discs. Experimental data collected in this manner indicated that attached biomass was linearly distributed as a function of time of biomass accumulation. Therefore, controlled scraping of disc surfaces and recovery of all biomass naturally sloughed within the RBC reactor allowed for the examination of biomass growth rates as measured with SRT values. Further details of the scraping procedures and biomass monitoring techniques are presented by Cruz[11] and Pope[12].

Analytical Procedures. Analytical methods and procedures presented in Standard Methods[13] were followed in the analysis of wastewater and biomass properties. A micro-COD[11-13] procedure was used to monitor all effluent COD concentrations. The coefficient of variation for replicate effluent samples containing 10 to 50 mg COD/ℓ was 2 to 5%, indicating excellent analytical precision. Ammonia nitrogen was determined using an ammonia specific-ion electrode (Orion Research, Cambridge, MA) and a standard addition technique.[11,12] Nitrite- and nitrate-nitrogen concentrations were determined with diazotization and chromotrophic acid procedures,[13] respectively. Total kjeldahl nitrogen of biomass solids and filtered effluent samples was determined with an automated method (Industrial Method 28-69A, Technicon Corp., New York). Suspended solids measurements for mixed liquor, effluent and attached biofilm samples were determined with Gooch crucibles containing glass fiber filter mats.[13]

RESULTS

A total of nine experimental runs were performed sequentially in three phases over a period of 11 months. Phase A included four runs in which the effect of attached biofilm SRT on nitrification efficiency was examined. Organic and nitrogen loading rates were maintained at relatively constant values of 4-4.6 g COD/m^2 • d (0.82-0.94 lb COD/1000 ft^2 • d) and 1.11-1.45 g N/m^2 • d (0.23-0.3 lb N/1000 ft^2 • d), respectively.

Phase B included one experimental run in which the effect of hydrolysis of organically-bound nitrogen was examined. Glycine was the sole source of nitrogen during this run and was supplied at a rate of 1.62 g N/m^2 • d (0.33 lb N/1000 ft^2 • d). Organic loading for this run was increased to 5.9 g COD/m^2 • d (1.21 lb COD/1000 ft^2 • d) as a result of the use of glycine as a nitrogen source.

Four experimental runs were included in Phase C to examine the effect of increased organic loading on nitrification efficiency. Organic loading was sequentially increased from 6.4 to 13.7 g COD/m^2• d) (1.31 to 2.8 lb COD/1000 ft^2 • d) while nitrogen loading rates remained at 1.3 to 1.44 g N/m^2 • d (0.26 to 0.29 lb N/1000 ft^2 • d). Experimental conditions for the nine experimental runs are summarized in Table 3.

SRT Values. The calculation of SRT was accomplished using values for total attached and suspended biomass in the RBC system, M_T, and the average rate of accumulation (i.e. wastage) of biomass, r_w, in the reactor system. SRT was then calculated using Equation 1.

$$SRT = \frac{M_T}{r_w} \tag{1}$$

The rate of wastage of biomass, r_w, was equal to the summation of the rate of accumulation of biomass in the mixed liquor and rate of discharge

TABLE 3
Experimental Operating Conditions for Phases A, B and C

Parameter	Experimental Run								
	A1	A2	A3	A4	B1	C1	C2	C3	C4
Scraping Interval, d	None	6	12	18	18	12	12	12	12
Influent DO, mg/l	3.1	3.4	2.4	2.6	2.2	2.2	1.5	3.1	3.3
Influent COD, mg/l	49.0	54.0	59.0	62.0	84	90	136	172	190
Influent Nitrogen,* mg/l	19.6	17.5	19.8	15.6	23.1	19.2	20.5	20.2	18
Hydraulic Retention Time, h	2.1	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydraulic Loading, gpd/ft^2	1.67	1.74	1.79	1.74	1.72	1.74	1.72	1.72	1.77
Organic Loading, lb COD/1000 ft^2•d	0.84	0.82	0.94	0.90	1.21	1.31	1.95	2.46	2.8
Nitrogen Loading, lb N/1000 ft^2•d	0.27	0.25	0.30	0.23	0.33	0.28	0.29	0.29	0.26

*Nitrogen as NH_4^+-N for all runs except B1 which was as Organic-N.

of biomass in the reactor effluent. The rate of accumulation of mixed liquor biomass was equal to that retained on the recirculation sieve and that removed directly from the mixed liquor in the reactor. Both mixed liquor fractions were removed for analysis on a 1, 2, or 3 day cycle, depending on the scraping interval used, and expressed herein as average concentrations of mixed liquor suspended solids.

Measurement of total attached biomass could not routinely be achieved without significant disruption of process performance. However, the net accumulation of total biomass attached to disc surfaces over 6, 12 and 18 day scraping intervals was monitored throughout each experimental run and these data were used to calculate total system biomass. By periodically scraping a portion of the attached biomass from the six disc surfaces in a set of four discs, the distribution of attached biomass on disc surfaces was established for each run.

Data presented in Figure 3 indicate the distribution of attached biomass solids during experimental runs A4 and B1. The organic loading rates during run B1 were higher than those for A4 and, as expected, total attached biomass was higher for run B1. Furthermore, the data in Figure 3, and those for other runs, indicate that the quantity of attached biomass on disc surfaces was a linear function of the time of accumulation, i.e. the time since biomass was last scraped from a disc surface. The rate of growth of attached biomass was therefore uniform throughout the reactor system and total quantities of attached biomass on disc surfaces varied primarily as a result of disc scraping techniques. Furthermore, the quantity of biomass scraped from similarly numbered disc surfaces, in accord with the scraping intervals used, was indicative of the total quantity of attached biomass on other disc surfaces. Total attached biomass, M_T, was then estimated with solids data for biomass scraped from disc surfaces and the associated elapsed time intervals for which biomass had accumulated on other disc surfaces in the reactor system.

Average values of the individual suspended and attached solids fractions for the eight experimental runs during which controlled wastage of attached biomass was practiced are presented in Table 4. Mixed liquor suspended solids concentrations (calculated as an average value using data for biomass solids collected on the recirculation sieve) ranged from 33 to 487 mg/l and were generally consistent with values reported by Antonie[10] (i.e. 49 to 275 mg/l). Effluent suspended solids concentrations were maintained at low concentrations, i.e. 3.7 to 9.8 mg/l and reflected the high efficiency with which mixed liquor solids were removed from suspension in the recirculation system. This furthermore indicated that the majority of the biological uptake of influent carbonaceous and nitrogenous oxygen demand was achieved with attached, and not suspended, biomass.

SRT values for the eight runs varied from 1.3 to 3.6d reflecting the rapid growth of the attached biomass. These rapid growth rates were dictated by intentional biomass wastage, however, natural sloughing of biomass accounted for a significant portion of overall biomass wastage rates, as reflected in the disparity between actual SRT values and the scraping intervals of 6, 12 and 18 days.

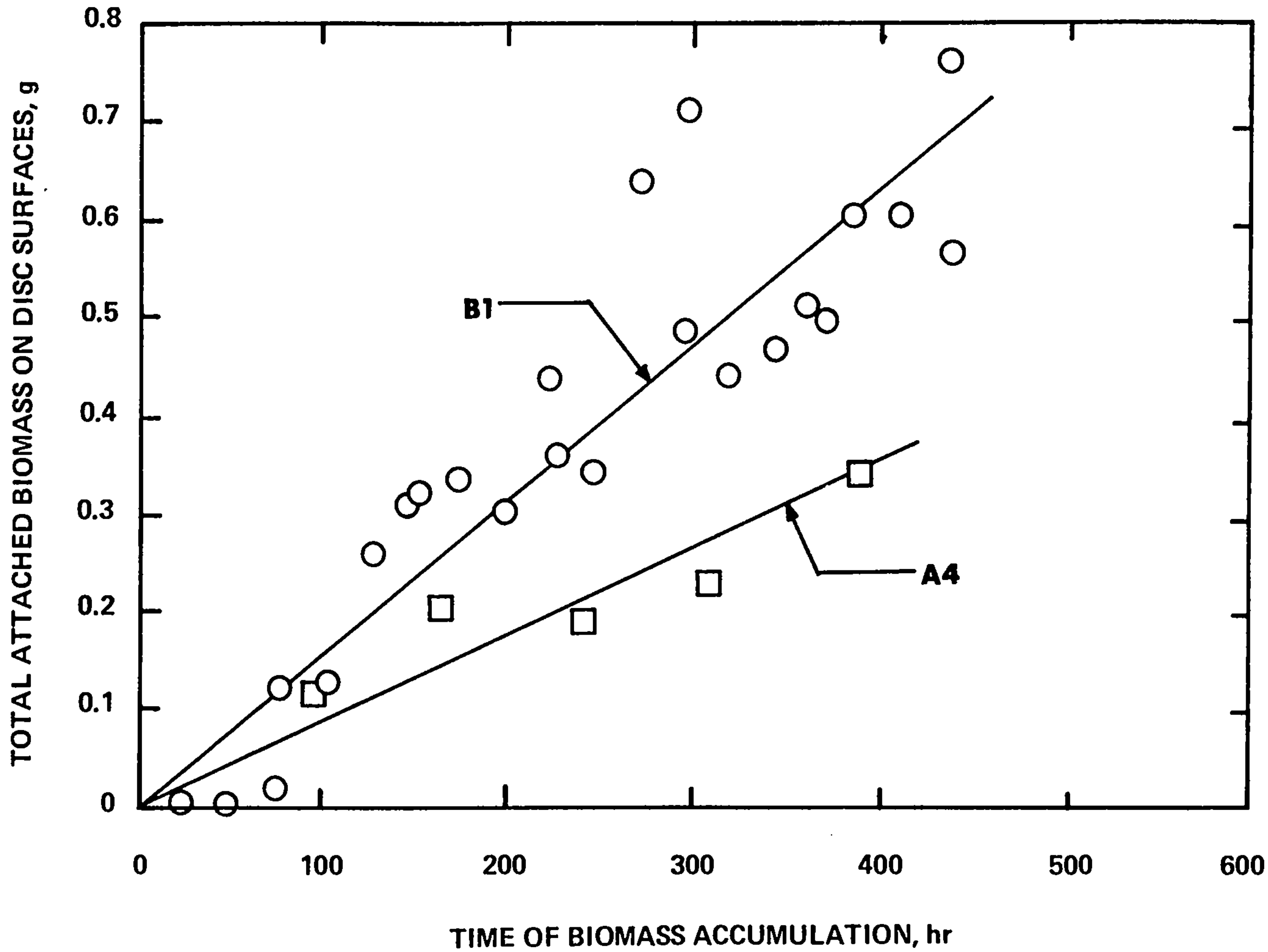

Figure 3. Distribution of Attached Biomass in the RBC System for Experimental Runs A4 and B1

TABLE 4
Suspended and Attached Solids and Mean SRT Data for the RBC System

Experimental Run	Mixed Liquor Suspended Solids (mg/l)	Effluent Suspended Solids (mg/l)	M_T Total RBC Biomass (g)	r_w Rate of Wastage of Biomass (g/d)	Mean SRT $\frac{M_T}{r_w}$ (d)
A2	33	5.2	1.63	1.26	1.3
A3	212	4.3	5.64	2.4	2.4
A4	232	5.0	4.77	1.43	3.3
B1	290	9.8	7.23	2.03	3.6
C1	198	5.1	3.04	1.66	1.8
C2	363	7.3	6.25	3.18	2.0
C3*	330	3.7	5.27	2.66	2.0
C4*	487	7.7	7.96	4.04	2.0

*Calculations are for highly transitory periods

Nitrification Efficiency. During experimental phase A, organic and nitrogen loading rates were held at constant levels. The average COD/N ratio of the influent wastewater was 3.1, indicating a low organic loading rate (i.e. influent COD = 49-62 mg/ℓ). SRT was varied from 1.3 to 3.3d for the three runs (A2-A4) in which controlled wastage of attached biomass was performed. During the first experimental run (A1), an attached biofilm was allowed to develop without controlled wastage of attached biomass. While it was not measured, the SRT value for this run was much higher than the highest value (SRT = 3.3d for A4) achieved during runs with controlled biomass wastage.

Effluent nitrogen data for phase A were presented elsewhere[14] and are summarized in Table 5. Effluent ammonia and nitrate concentrations for runs A1, A3 and A4 indicated that nitrification was achieved at the 91 to 99% level, i.e. virtually complete nitrification. Nitrification was not achieved during run A2, the run with the lowest SRT value (i.e. SRT = 13d), due to the washout of nitrifying bacteria from the attached film.

TABLE 5
Steady State Ammonia - Nitrogen Concentrations in Influent and Effluent Wastewaters During Phase A

Run	Influent NH_4^+-N (mg/ℓ)	Effluent NH_4^+-N (mg/ℓ)	Effluent NO_3^--N (mg/ℓ)	Effluent NO_2^--N (mg/ℓ)	Percent Nitrification*
A1	19.6	0.1	15.8	<0.01	99
A2	17.5	14.4	0.2	<0.01	1
A3	19.8	1.0	18.7	0.80	91
A4	15.6	0.1	14.0	<0.01	99

* Percent of effluent soluble nitrogen attributable to NO_3^--N

Since effective nitrification was achieved at low SRT values using ammonia as the sole source of nitrogen, an experimental run was performed during phase B to examine the effect of hydrolysis of organically-bound nitrogen on nitrification efficiency. The RBC system was operated during run B1 at the same scraping interval as the previous experimental run (A4) to minimize the time to achieve steady state conditions. The low SRT value associated with this run was used to determine if the rate of hydrolysis of organic-nitrogen was a rate-limiting step with respect to nitrification. As indicated in Table 6, influent organic-N, i.e. glycine, was hydrolysed and liberated ammonia was oxidized at a 93% efficiency level. Hydrolysis of organic nitrogen was, therefore, not a rate limiting step, even at a high growth rate, i.e. a SRT value of 3.6d.

Four experimental runs were conducted during phase C to examine the effects of increased influent organic matter concentrations on nitrification. Influent COD was increased in step intervals, as indicated in Figures 4 and 5, from an average COD concentration of 90 mg/ℓ to 190 mg/ℓ and approached that

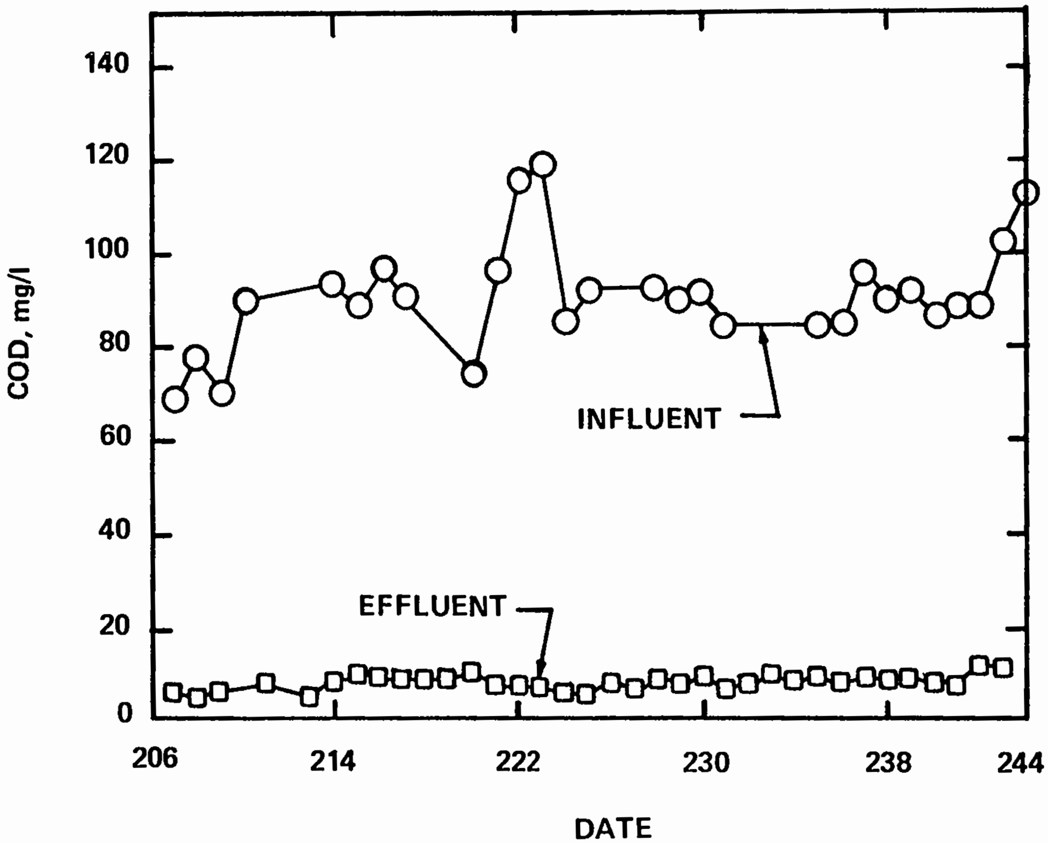

Figure 4. Influent and Effluent COD Concentrations for Experimental Run C1

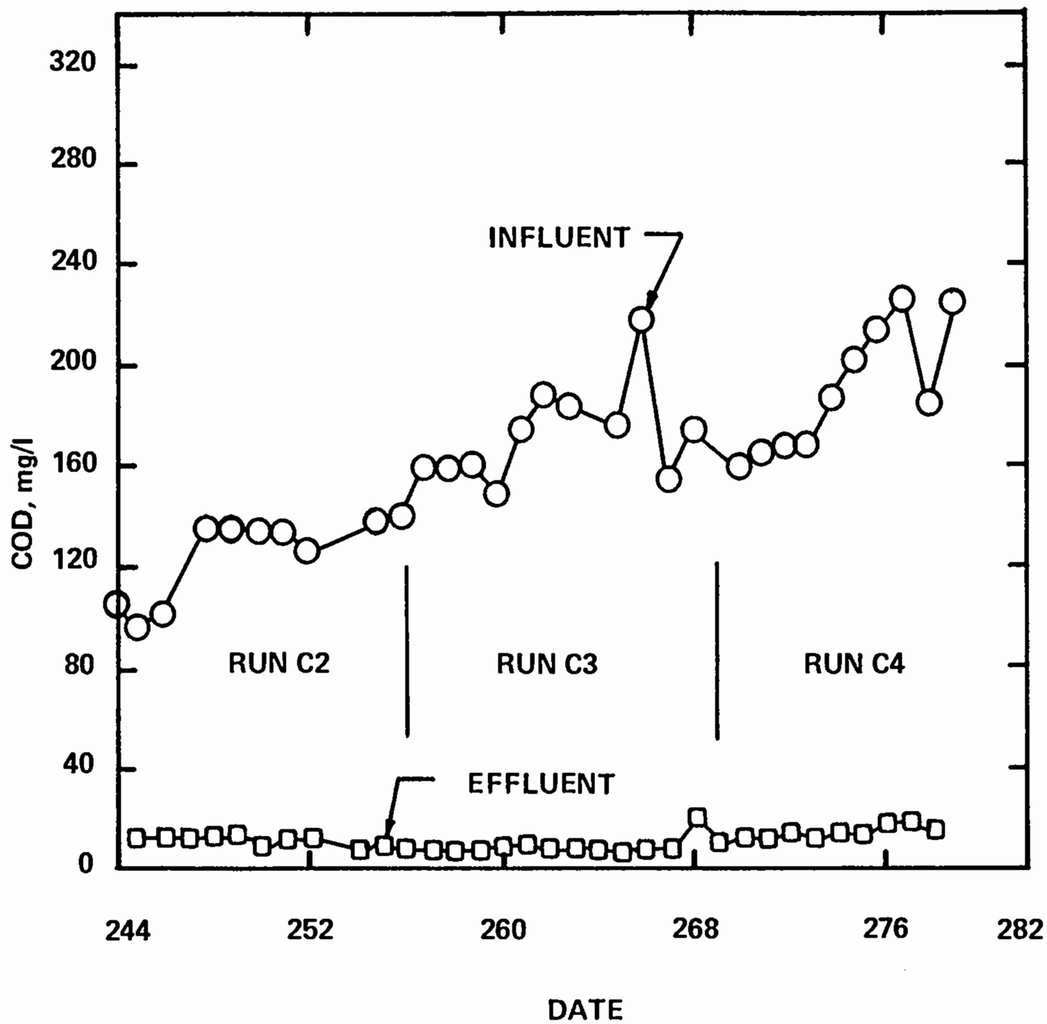

Figure 5. Influent and Effluent COD Concentrations for Experimental Runs C2-C4

TABLE 6

Steady State Ammonia- and Organic-Nitrogen Concentrations in Influent and Effluent Wastewaters for Phases B and C

Run	Influent NH_4^+-N (mg/ℓ)	Influent Org-N (mg/ℓ)	Effluent NH_4^+-N (mg/ℓ)	Effluent Soluble Org-N (mg/ℓ)	Effluent NO_3^--N* (mg/ℓ)	Percent Nitrification**
B1	--	23.1	0.3	0.5	14.8	93
C1	19.2	--	0.5	--	16.7	98
C2	20.5	--	0.1	--	16.0	99
C3	20.2	--	0.2	--	15.9	98
C4	18	--	0.8	--	12.3***	95

* NO_2^--N concentration was less than detectable limit (<0.01 mg N/ℓ)

** Percent of effluent soluble nitrogen attributable to NO_3^--N

*** Effluent nitrate-N indicated was not a steady-state value

equivalent to a low strength domestic wastewater during run C4. Effluent ammonia-nitrogen concentrations remained low throughout the four runs as indicated in Figures 6 and 7.

High levels of nitrification were achieved during all runs ranging from 95 to 98%, as indicated in Table 6. Therefore, organic loading up to 13.7 gCOD/$m^2 \cdot d$ (2.8 lb/1000 $ft^2 \cdot d$) had no negative impact on nitrifying bacteria in attached biofilms. Factors contributing to this favorable response included mixed liquor dissolved oxygen concentrations which averaged 5.0 mg/ℓ. Mixed liquor pH values, in addition, were stable at 7.3 and temperature averaged 23°C. Further study is required to examine attached film nitrification, especially at low mixed liquor dissolved oxygen concentrations.

Nitrogen Mass Balance. Formation of nitrate-nitrogen by nitrifying bacteria may result in denitrification in anoxic portions of an attached biofilm. To determine if denitrification occurred in the RBC system and to examine the extent to which nitrogen was removed by inclusion into microbial cell mass, a detailed nitrogen balance was conducted for the nine experimental runs.

The mass flows of nitrogen in influent and effluent wastewaters and in the biomass removed from mixed liquor and disc surfaces were examined for each experimental run and are presented in Table 7. Quantitative data were obtained for all nitrogen fractions except the wasted biomass nitrogen for run A1. Within the limits of cumulative analytical capabilities, the majority of influent nitrogen was detected in the effluent wastewater and wasted biomass. For runs A2 through C4, the overall nitrogen balance averaged 103% of influent nitrogen. Therefore, no measurable denitrification occurred within the RBC system. In addition, wasted biomass nitrogen data indicated that biomass nitrogen content varied from 8.7 to 13.5% nitrogen and increased with biomass SRT value.

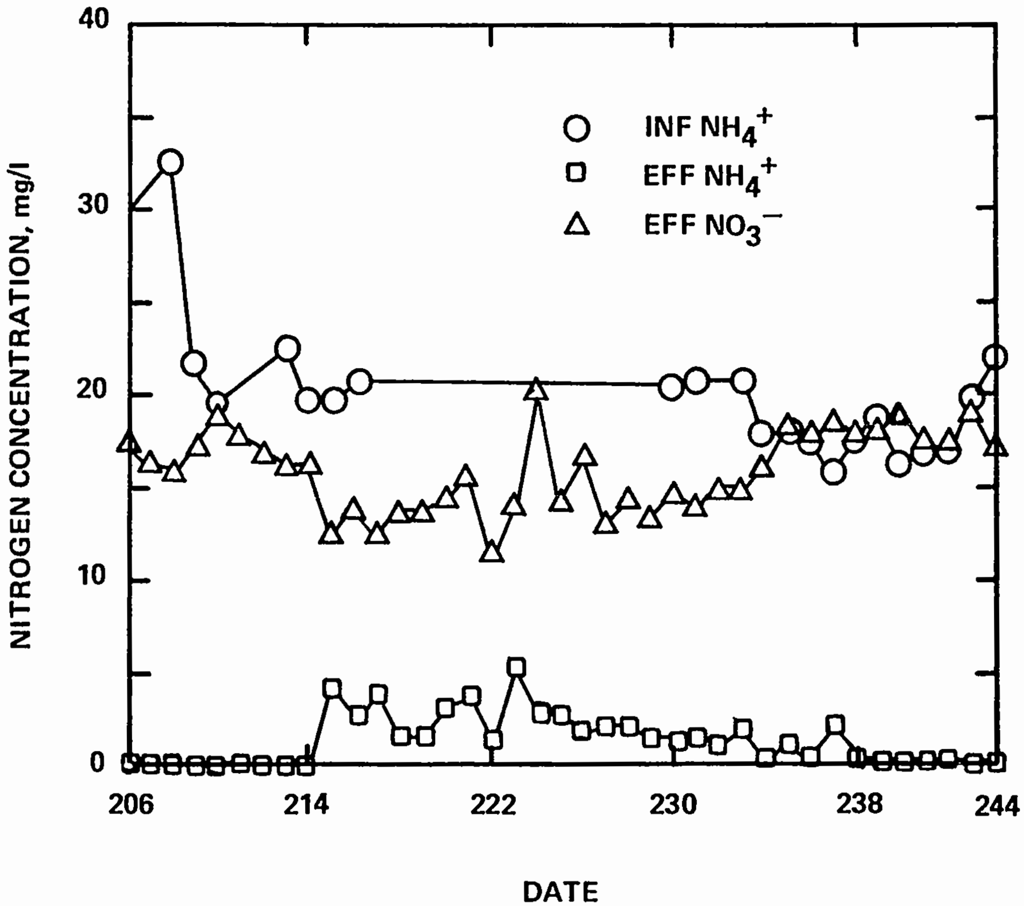

Figure 6. Influent and Effluent Soluble Nitrogen Concentrations during Experimental Run C1

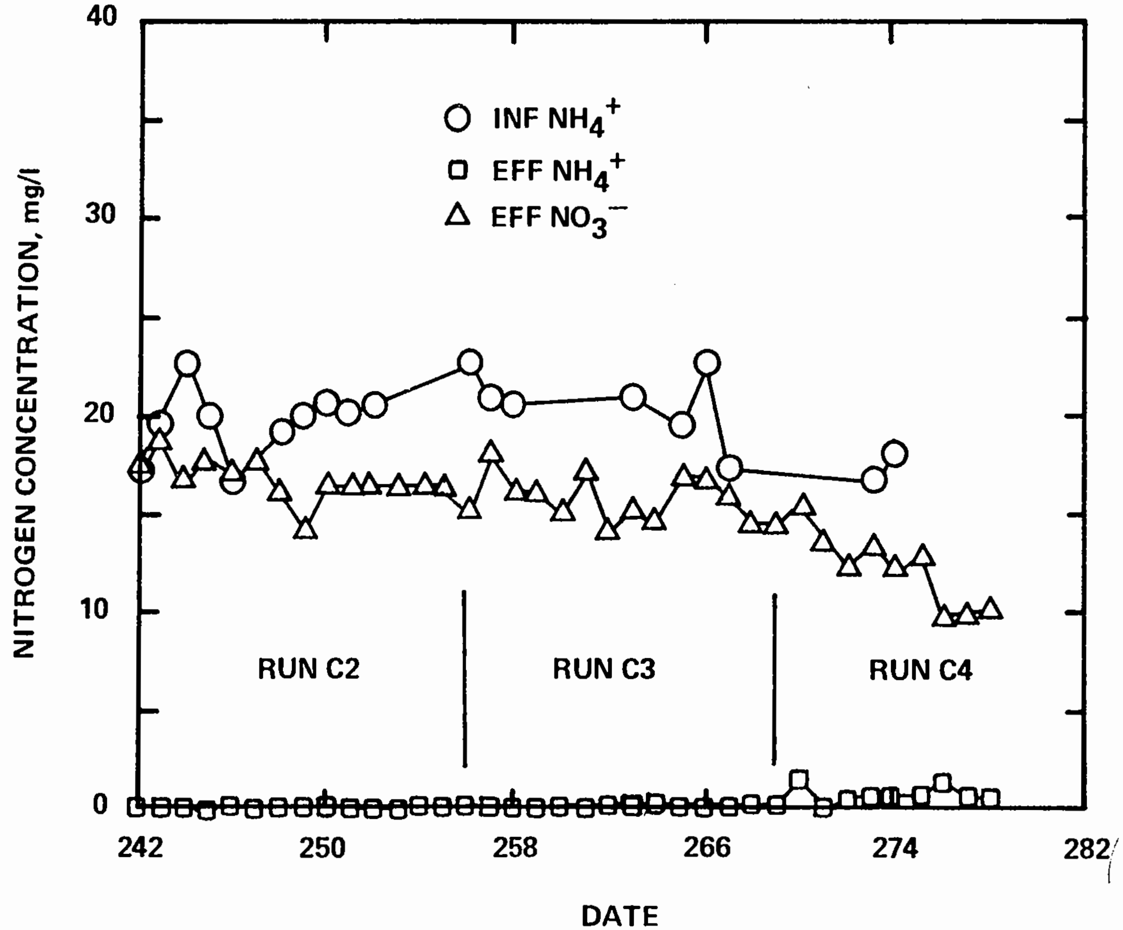

Figure 7. Influent and Effluent Soluble Nitrogen Concentrations during Experimental Runs C2, C3 and C4

TABLE 7
Nitrogen Balance for RBC System

	A1	A2	A3	A4	B1	C1	C2	C3	C4
INFLUENT NITROGEN*									
NH_4^+-N	1.33	1.24	1.45	1.11	--	1.36	1.44	1.41	1.30
Organic-N	--	--	--	--	1.62	--	--	--	--
EFFLUENT NITROGEN*									
NH_4^+-N	0.01	0.97	0.07	0.01	0.02	0.03	0.01	0.02	0.05
NO_2^--N	0	0	0.06	0	0	0	0	0	0
NO_3^--N	1.08	0.01	1.35	1.02	1.05	1.25	1.12	1.11	0.88
Organic-N	--	--	--	--	0.06	--	--	--	--
WASTED BIOMASS NITROGEN*									
Organic-N	--	0.12	0.20	0.19	0.30	0.19	0.39	0.30	0.48
NITROGEN BALANCE (% of influent-N)	82	89	116	110	88	108	106	101	108

* Nitrogen expressed as $(g\text{-}N/m^2 \cdot d)$

Organic Removal Efficiency. The removal of influent COD was excellent for all experimental runs as indicated in Figure 8. Effluent organic quality improved slightly with increased SRT values, as expected[15]. At organic loading rates as high as 13.6 $gCOD/m^2 \cdot d$ (2.8 $lbCOD/1000\ ft^2 \cdot d$) COD removal efficiency was 92% and compared very favorably with pilot-scale systems loaded at similar levels[10].

DISCUSSION

Examination of SRT values in Table 4 and nitrification data in Tables 5 and 6 indicated that nitrification efficiency was related to SRT. As presented in Figure 9, nitrification did not occur below a SRT value of 1.8 days. However, at and above SRT values of 1.8 days, nitrification was virtually complete and the data indicated a response typical of the growth of nitrifying bacteria. When examining growth relationships for these bacteria, a Monod-like hyperbolic relationship[16] is used, i.e.

$$\mu = \frac{\hat{\mu} S}{K_s + S} \tag{2}$$

where μ = net specific growth rate constant, $\hat{\mu}$ = maximum net specific growth rate, S = concentration of limiting substrate and K_s = half-velocity constant. Values of K_s for nitrifying bacteria range from 0.18 to 1.0 mg N/ℓ[3] indicating that ammonia and nitrite oxidation reactions proceed at maximum rates as near-zero order reactions, at substrate nitrogen concentrations of 1.5 mgN/ℓ[1-5]. Nitrifying bacteria then typically grow very rapidly at or near critical wash-out growth rates (i.e. $\hat{\mu}$) while continuing to remove ammonia- and nitrite-nitrogen to sub-mg/ℓ levels[1-5].

The response of the attached biofilms in the experimental RBC system was consistent with that of nitrifying bacteria. Washout of nitrifying populations from the RBC system occurred abruptly between SRT values of 1.3 (Run A2) and 1.8 (Run C1) days. This furthermore indicated that the maximum net specific growth rate constant, $\hat{\mu}$, for the nitrifying population was between 0.56 and 0.83 d^{-1}, in accord with the relationship SRT = $(1/\mu)$[15].

Of the nitrifying bacteria, Nitrosomonas is the slowest growing bacterium[1,4,8] and, therefore, the oxidation of ammonia to nitrite is the rate-controlling reaction. Examination of reported values of $\hat{\mu}$ for Nitrosomonas, as presented in Table 8, indicated that the range of $\hat{\mu}$ values obtained in this experimental study were consistent with pure and mixed culture data for Nitrosomonas, i.e. $\hat{\mu}$ = 0.17 - 1.08 d^{-1}.

While determining the precise $\hat{\mu}$ value for the nitrifying population in the experimental system was not possible, the response of the attached biofilm population was virtually identical with that predicted with data for nitrifying bacteria. Therefore, the use of classical nitrification data in predicting and modelling the response of RBC systems is justified. This conclusion is further substantiated in Figure 10, in which experimental nitrification data from numerous wastewater treatment studies[19,21-26] are

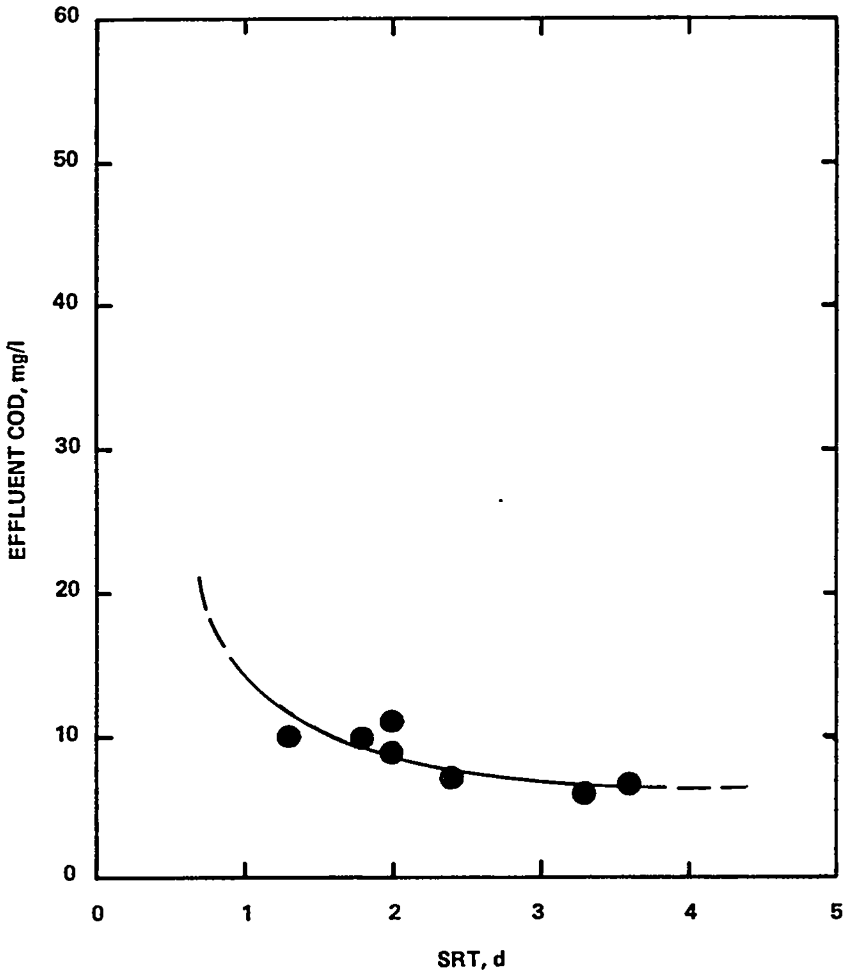

Figure 8. Effluent Soluble COD for RBC Reactor System

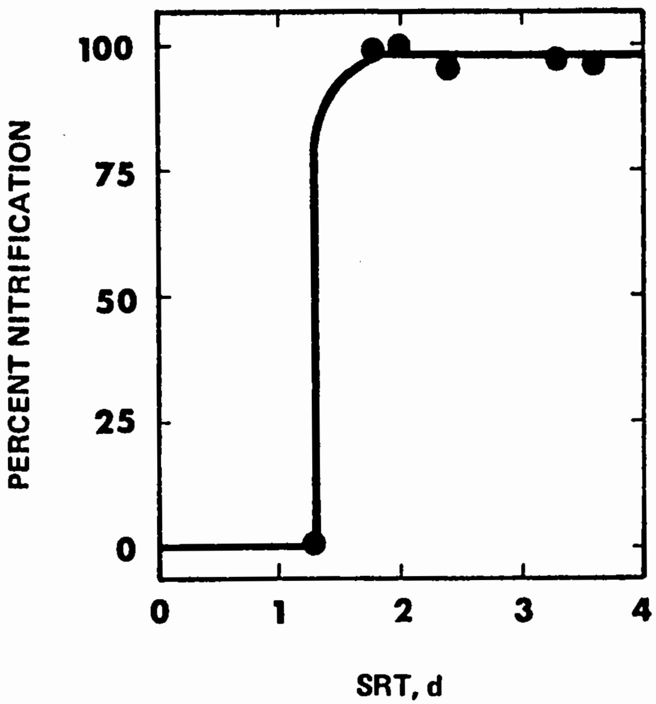

Figure 9. Nitrification Efficiencey for Attached Biofilms as a Function of Biofilm SRT (14)

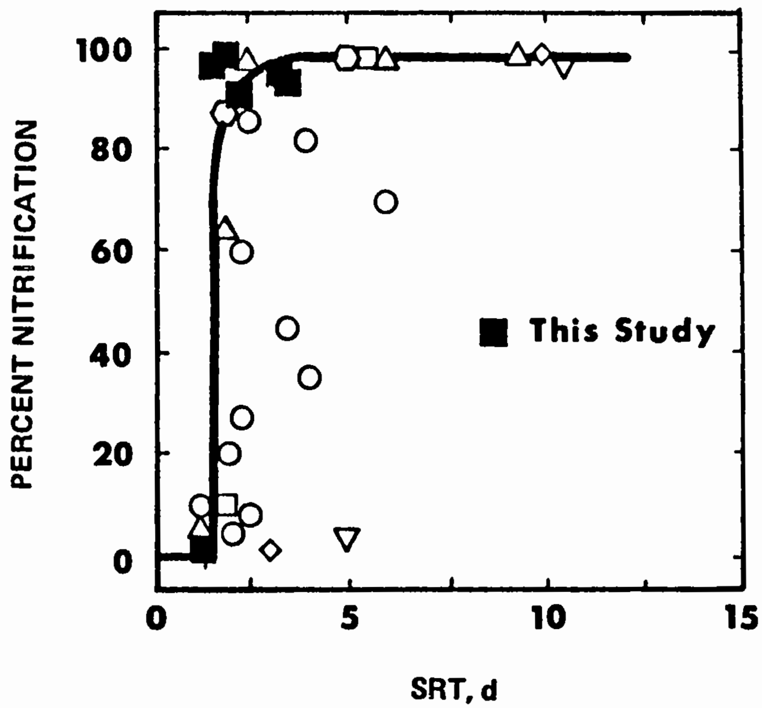

Figure 10. Nitrification Efficiency as a Function of SRT from Numerous Wastewater Studies[19,21-26](14)

presented with those from this study. The response of the RBC system in this study was then consistent with that for numerous wastewater treatment studies, providing further justification for the use of classical growth and substrate removal relationships[1-9] in the design and evaluation of RBC systems.

TABLE 8
Values of $\hat{\mu}$ for *Nitrosomonas* in Pure and Mixed Cultures

Temperature ^{o}C	$\hat{\mu}$ d^{-1}	Reference
21	0.85	8
25	0.88	17
25	0.55	15
21	0.85	18
23	0.37	19
25	0.17	20
23	1.08	3
20	0.94	6

With regard to modelling RBC systems for concurrent carbonaceous BOD removal and nitrification, Mueller *et al*.[27] presented a comprehensive mathematical model of an RBC system. The model was calibrated and verified with actual operational data from full-scale systems. A critical component of the RBC model was the use of the classical Monod-like growth relationship[16] to predict growth and substrate removal rates for nitrifying bacteria. Although the predicted values of the growth rate constants, $\hat{\mu}$ and K_s, for nitrifying bacteria were significantly lower than reported values[27], results of the study reported herein strongly indicate that such modelling approaches should be vigorously pursued.

SUMMARY AND CONCLUSIONS

Nitrification can be concurrently achieved with removal of carbonaceous organic matter in single-stage RBC systems at high biofilm growth rates. In addition, mean solids retention time, SRT, is a critical variable with respect to the retention of nitrifying bacteria in attached biofilms. Reported values for the growth constants, $\hat{\mu}$ and K_s, can be used to predict critical SRT values at which washout of attached nitrifying populations will occur, as well as establish effluent ammonia levels from RBC systems.

The hydrolysis of organically-bound nitrogen, when provided as simple amino acids, i.e. glycine, is not a rate limiting reaction and does not impede nitrification, even at high biofilm growth rates. Organic loading rates up to 13.6 $gCOD/m^2 \cdot d$ did not impede nitrification at attached biofilm SRT values slightly higher than those resulting in washout. The continuous availability of dissolved oxygen within the RBC system, however, provided condtions for near optimal growth of nitrifying bacteria. Further studies must be pursued under oxygen limiting conditions to more effectively evaluate the use of Monod kinetic relationships[16] in the design of RBC systems. The close agreement of the results of this experimental study with kinetic data and relationships for Nitrosomonas and Nitrobacter cultures confirmed the validity of the use of these relationships in modelling, designing and operating RBC systems.

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ROLE OF SUSPENDED SOLIDS IN THE KINETICS OF RBC SYSTEMS

BY

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INTRODUCTION

The rotating biological contactor is receiving increased interest as a biological wastewater treatment process. As a wastewater treatment process, the kinetics of the biological population are very important in the design and/or operation of the process. While some researchers have speculated that the RBC process may possess some of the behavioral traits of both the fixed film and suspended culture processes, the majority, if not all, of the design methods developed have discounted the effects of the mixed liquor suspended solids present in the system. The biofilm concept has been commonly used for kinetic description of the RBC process.

Kornegay and Andrews (1) developed a kinetic model to describe fixed reactors. In their mass balance they included the suspended solids. However, they made the assumption that RBC would be operating under "wash out" conditions and the suspended solids would play no part in substrate removal. Therefore, their actual model does not include suspended solids. They also found that there was an active film thickness. The substrate utilization reached a steady state value after the biological film reached a thickness of 70 μ. Hoehn and Ray (2) fully supported the active film thickness theory, however, they found that the active film thickness was 200 μ instead of 70 μ.

In general, the kinetic models do not take into account the actual biomass. They consider the area that the biofilm covers and base their models on this area. This study was undertaken to study whether or not the suspended solids play a role in the RBC process. It was also of interest to determine whether or not the consideration of actual biomass would provide a means of comparing the RBC process with the activated sludge process.

MATERIALS AND METHODS

The model RBC unit used in this study consisted of a plexiglass tank divided into four stages with four polyethylene discs in each stage. Each disc was approximately 1/8 inch thick and 6 inches in diameter. This resulted in a total disc surface area of 6.28 square feet or 1.57 square feet per stage. The volume of the liquid in the reactor was 5.1 liters. This provided a forty percent submergence of the discs. The hydraulic flow rates to the RBC were maintained through the use of a constant head tank which received a continuous flow of tap water. The flow from the constant head tank was regulated by a valve combined with a flow meter on the tank outlet line. Water from the constant head tank fed by gravity into a wet well, where it was mixed with the concentrated synthetic waste to achieve the desired organic concentration. The synthetic waste was pumped to the wet well using a Cole-Parmer Masterflex pump. From the wet well, mixture flowed by gravity into the first stage of the RBC. The rotational speed of the discs was maintained at 10 rpm.

The synthetic waste used in this study contained glucose as the sole carbon source. All required nutrients were added in excess so that carbon was the limiting growth factor. The COD of the influent wastewater was maintained at 300 mg/l.

The RBC was initially seeded with effluent from the primary clarifier of the Stillwater, Oklahoma, Wastewater Treatment Plant. The RBC was allowed to operate as a batch unit for three days and then operated as a continuous flow reactor. The RBC was operated in this manner for two weeks to allow the development of a biological growth. Analyses were initiated after two weeks. COD and suspended solids were run daily until a steady state condition was established. After steady state had been achieved, samples were collected on two consecutive days. These were averaged and recorded as the results of that phase of the study. In addition to COD and suspended solids measurements, samples were also collected for determining the residual COD and for conducting a batch growth study.

The residual COD determination consisted of taking one liter of the RBC effluent and aerating it as a batch reactor for one week. COD analysis was conducted at various times to ascertain the residual COD. Batch growth studies using a shaker were conducted to determine the maximum specific growth rate.

Studies were conducted at five different flow rates. These flow rates were 36 ℓ/day, 71 ℓ/day, 143 ℓ/day, 178 ℓ/day and 250 ℓ/day. These resulted in hydraulic loadings of 1.5 gpd/ft^2, 3.0 gpd/ft^2, 6.0 gpd/ft^2, 7.5 gpd/ft^2 and 10.5 gpd/ft^2.

RESULTS

The ΔCOD remaining at each stage for the five flow rates studied is shown in Figure 1. ΔCOD represents the total amount of organic matter available as substrate to the microorganisms. It is determined by substrating the residual COD from the observed COD. Figure 1 shows that the ΔCOD removed by stages follows zero order kinetics for flow rates of 36 ℓ/day (1.5 gpd/ft^2) and 71 ℓ/day (3.0 gpd/ft^2). All other flow rates showed kinetics approximating first order.

Figure 2 shows the mixed liquor suspended solids concentrations obtained at each stage for the flow rates studied. The suspended solids increased with the first stages and then decreased in the latter stages. Also, the suspended solids concentrations decreased as the flow rates increased. Suspended solids concentrations of 2000mg/l were reached for flow rates of 36 and 71 ℓ/day, whereas, a suspended solids concentrations of 820 mg/l was the maximum achieved at a flow rate of 250 ℓ/day.

Figure 3 shows a comparison of the dilution rate and the maximum specific growth rate that was obtained by batch studies. The dilution rate was greater than the maximum specific growth rate for all growth rates studied. However, the dilution rate and growth rate were close for the 36 ℓ/day and 71 ℓ/day flow rates. It must be recognized that the batch procedure used for obtaining growth rates may not give the true growth rates that are occuring in the continuous flow reactor. Therefore, it is possible that the true growth rate in the RBC reactor was greater than the dilution rate. If this is true, then the suspended solids would be effective in removing substrate at these flow rates. Also, it appears that suspended solids are retained in each stage rather than being held for only the detention time. This would provide a smaller dilution rate for the solids. The retention of solids has been observed at low hydraulic loadings for all RBC studies conducted in this laboratory. The maximum specific growth rates for the 143 ℓ/day, 178 ℓ/day and 250 ℓ/day flow rates were much lower than the dilution rate and the possibility that the suspended solids would be effective in removing substrate is very small. Therefore, it appears that the suspended solids may remove substrate at low dilution rates (high detention times) but not be responsible for substrate removal at higher dilution rates (low detention times).

Two parameters were used to evaluate the role of the suspended solids. These were specific substrate utilization and specific substrate utilization rate. The specific substrate utilization is given as

$$\frac{S_i - S_e}{X_m} \qquad \text{(suspended solids not included)}$$

or

$$\frac{S_i - S_e}{X_m + X_s V} \qquad \text{(suspended solids included)}$$

where

S_i = influent COD, mg/ℓ

S_e = effluent ΔCOD, mg/ℓ

X_m = mass of microrganisms on discs, mg

X_s = suspended solids in reactor, mg/ℓ

V = liquid volume of RBC reactor, ℓ

The specific substrate utilization rate is given as

$$\frac{(S_i - S_e)\ F}{X_m}$$

or

$$\frac{(S_i - S_e)\ F}{X_m + X_s F}$$

where

F = flow rate, ℓ/day

The mass of microorganisms on the rotating discs were calculated by using the active film theory. It is evident that an accurate estimate of the amount of biological solids present on the discs and actively participating in the substrate removal is not easy to make. The mass of microorganisms was calculated by multiplying the disc surface area times the density of the microorganisms, times the active film thickness. The density of the microorganisms was taken as 95 mg/cm^3 and the active film thickness was taken as 200 µ. As mentioned earlier, Kornegay and Andrews (1) reported an active film thickness of 70 µ and Hoehn and Ray (2) reported an active film thickness of 200 µ. Famularo, Mueller, and Mulligan (3) used their model to calculate an active film thickness of 120 µ. This provides a rather large range of active film thicknesses to choose from. In the study being reported, an active film thickness of 200 µ was selected because this value gave results that appeared to be more reasonable than those obtained from other film thickness. The biomass per stage would be

$$1.57\ ft^2/stage \times 929\ cm^2/ft^2 \times 95\ mg/cm^3 \times 0.02\ cm = 2771\ mg/stage$$

and the substrate utilization may be calculated for various stages and flow rates.

Specific substrate utilization calculated by using only the active film biomass is shown as a function of the effluent ΔCOD in Figure 4. It is clearly seen that two different relationships exists. One for the 36 ℓ/day and 71 ℓ/day flow rates and another relationship for the 148 ℓ/day, 178 ℓ/day and 250 ℓ/day flow rates.

Specific substrate utilization can also be calculated by taking into account the suspended solids for the 36 ℓ/day and 71 ℓ/day flow rates. The

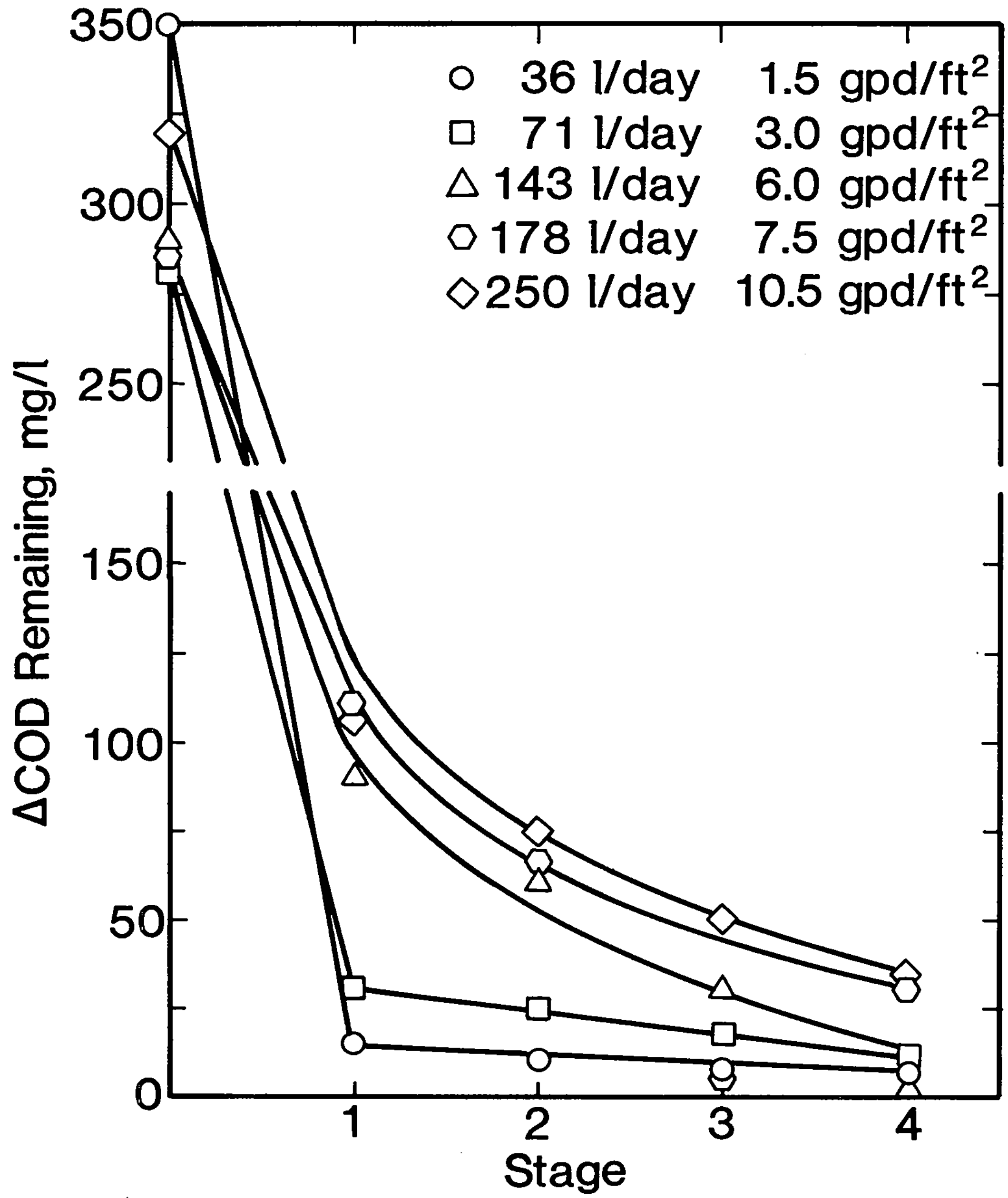

Figure 1. ΔCOD Remaining vs. RBC Stage.

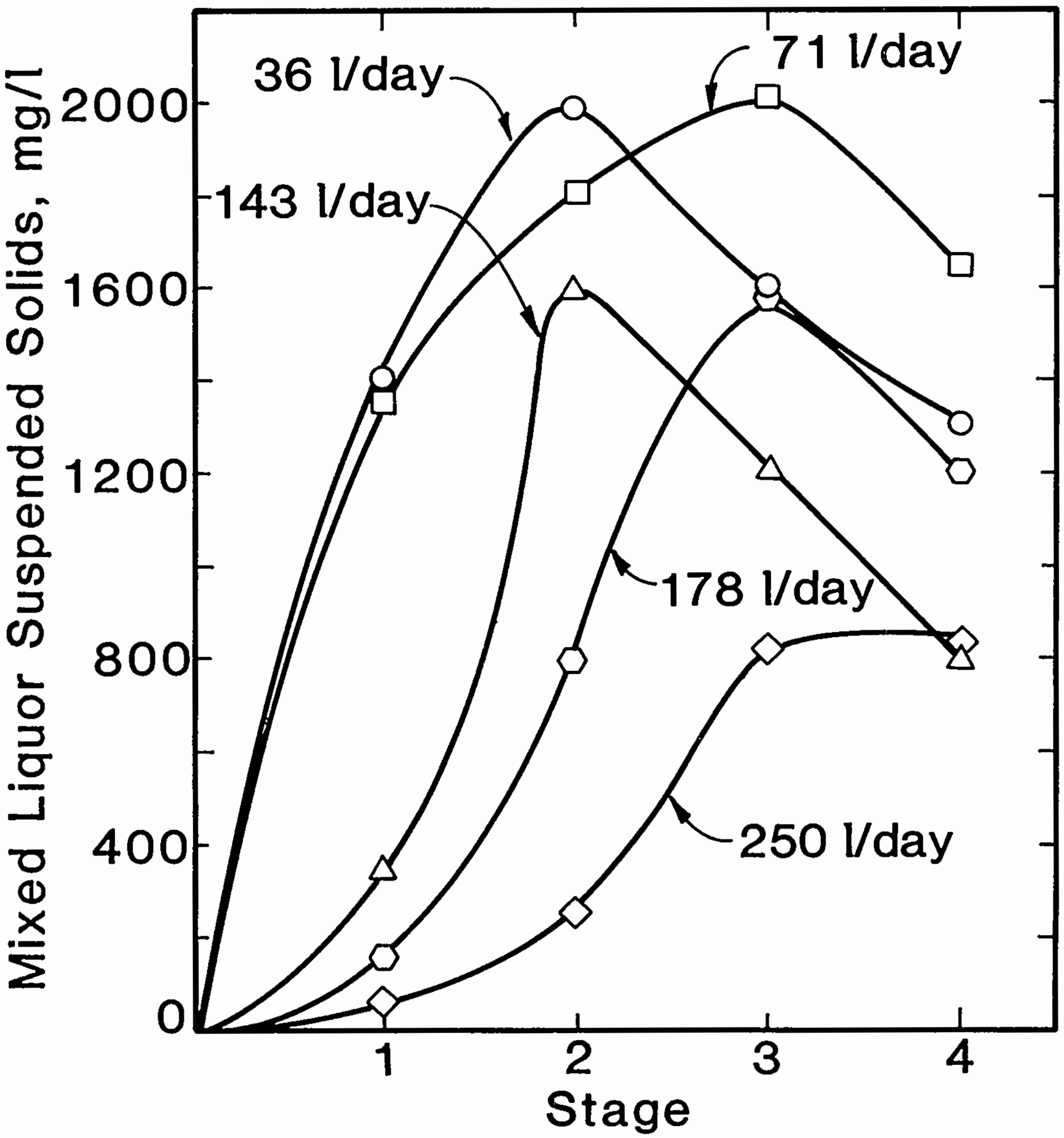

Figure 2. Suspended Solids vs. RBC Stage.

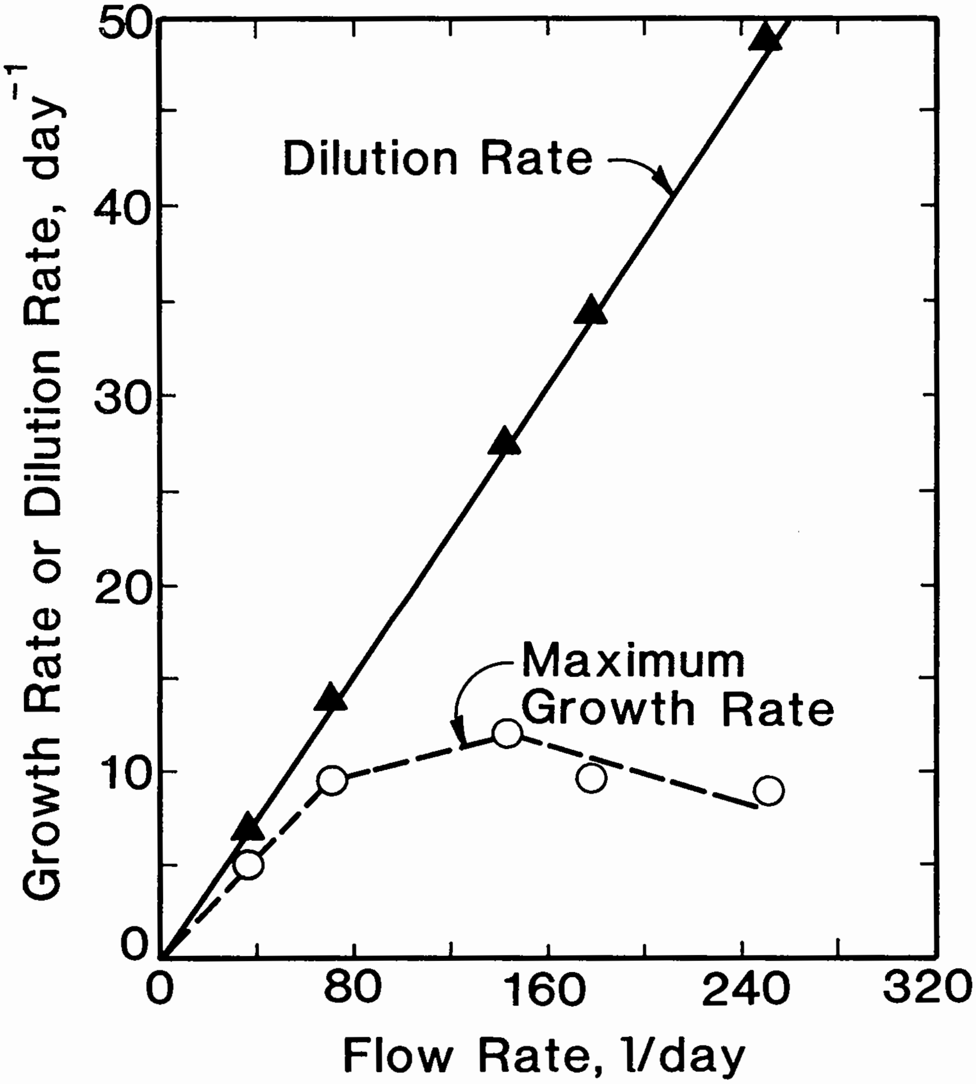

Figure 3. Maximum Growth Rate and Dilution Rate vs. Flow Rate.

suspended solids were not included for the higher flow rates. This relationship is shown in Figure 5. It is seen that all data fit one curve. This is in contrast to Figure 4 in which there are two different curves.

Another parameter that is often used to describe activated sludge is the specific substrate utilization rate. This parameter includes the hydraulic flow rate. Figure 6 shows the relationship between the specific substrate utilization rate (active biofilm solids only) and the effluent ΔCOD. The data fits one curve with the relationship

$$\frac{(S_i - S_e)\ F}{X_m} = \frac{39\ S_e}{232 + S_e}$$

When the suspended solids are included with the 36 ℓ/day and 71 ℓ/day flow rates, a straight line relationship is obtained (Figure 7). The mathematical relationship is

$$\frac{(S_i - S_e)\ F}{X_m + X_s\ V} = 0.13\ S_e$$

Regression analyses were conducted with the data presented in Figures 6 and 7 and these relationships gave the best fit. The correlation coefficient for the data presented in Figure 6 was 0.93 and the correlation coefficient for the data presented in Figure 7 was 0.95.

The food to microorganism ratio has been used to evaluate activated sludge processes for a number of years with much success. Bentley and Kincannon (4) have also used the food to microorganism ratio to evaluate biological towers. This ratio can also be used to evaluate the performance of an RBC. The active film biomass and suspended solids at flow rates of 36 ℓ/day and 71 ℓ/day were used to calculate the food to microorganism ratio. Figure 8 shows the treatment efficiency obtained at various ratios. This Figure also suggests that the suspended solids should be considered as part of the active biomass at higher detention times. It is also seen that removal rates greater than 95 percent were achieved at food to microorganism ratios below 1.0.

Figure 9 shows the treatment efficiencies obtained for various hydraulic detention times. The only flow rates that produced detention times capable of achieving treatment efficiencies of 95 percent or greater were 36 ℓ/day and 71 ℓ/day. These flow rates gave a hydraulic loading of 1.5 gpd/ft^2 and 3.0 gpd/ft^2.

DISCUSSION

The results of this study show that there are conditions where the suspended solids are active in removing substrate in an RBC. The general concept is that the suspended solids are washed out of the reactor and are not active in substrate removal. This certainly is the case when the dilution rate exceeds the maximum specific growth rate. However at low hydraulic loadings, the dilution rate may be less than the maximum specific growth rate.

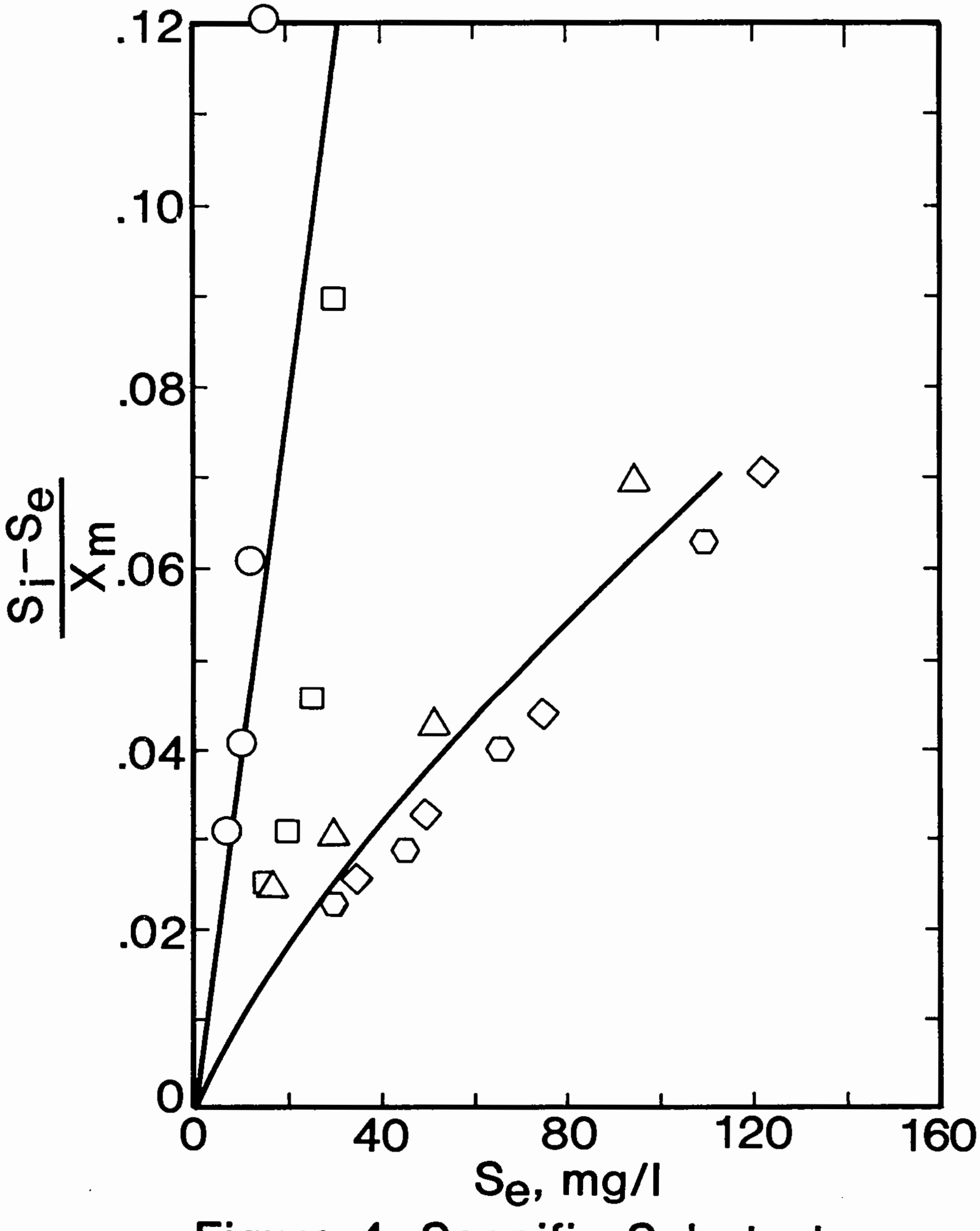

Figure 4. Specific Substrate Utilization Without Suspended Solids vs. Effluent ΔCOD.

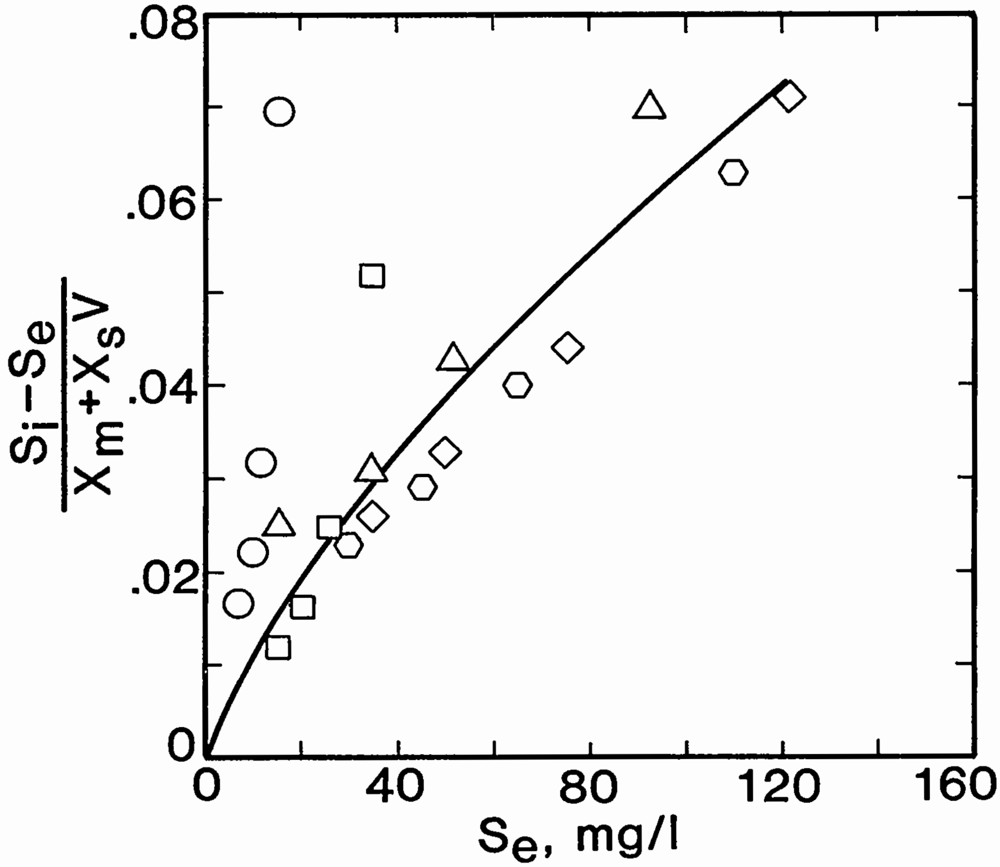

Figure 5. Specific Substration Utilization Including Suspended Solids vs. Effluent ΔCOD.

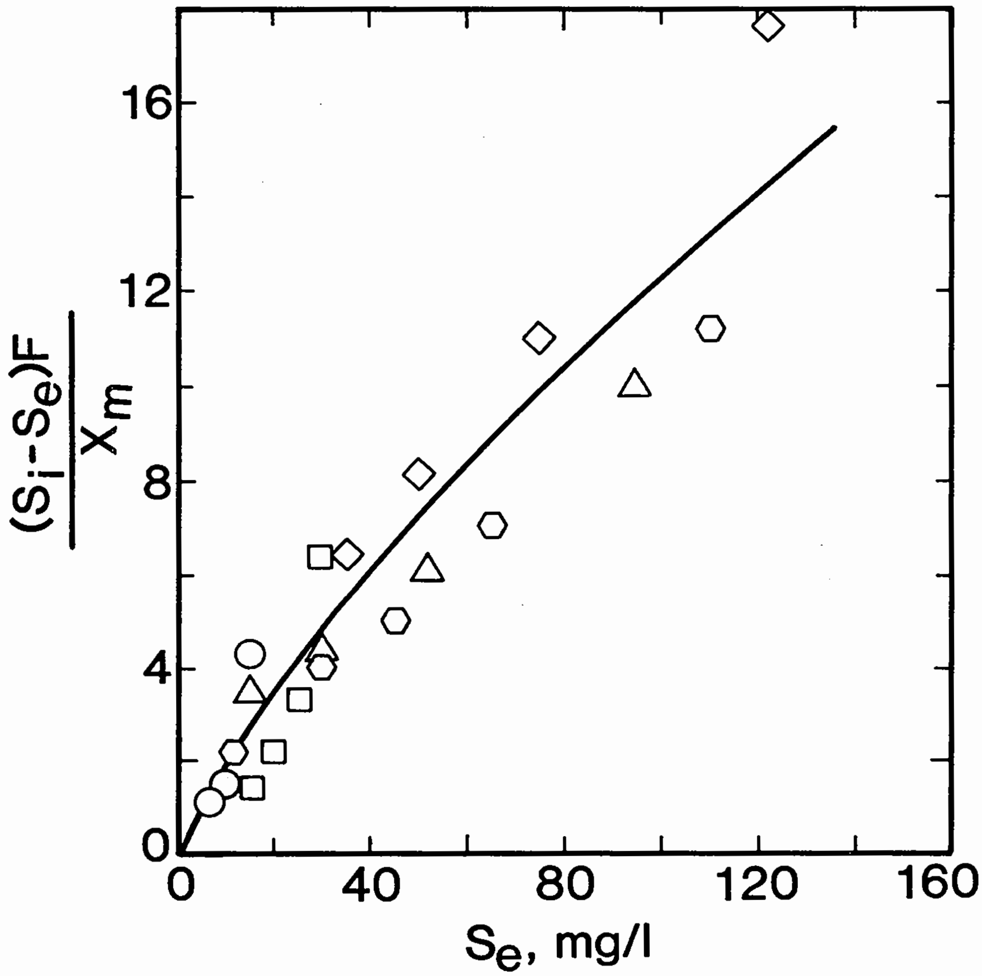

Figure 6. Specific Substrate Utilization Rate Without Suspended Solids vs. Effluent ΔCOD.

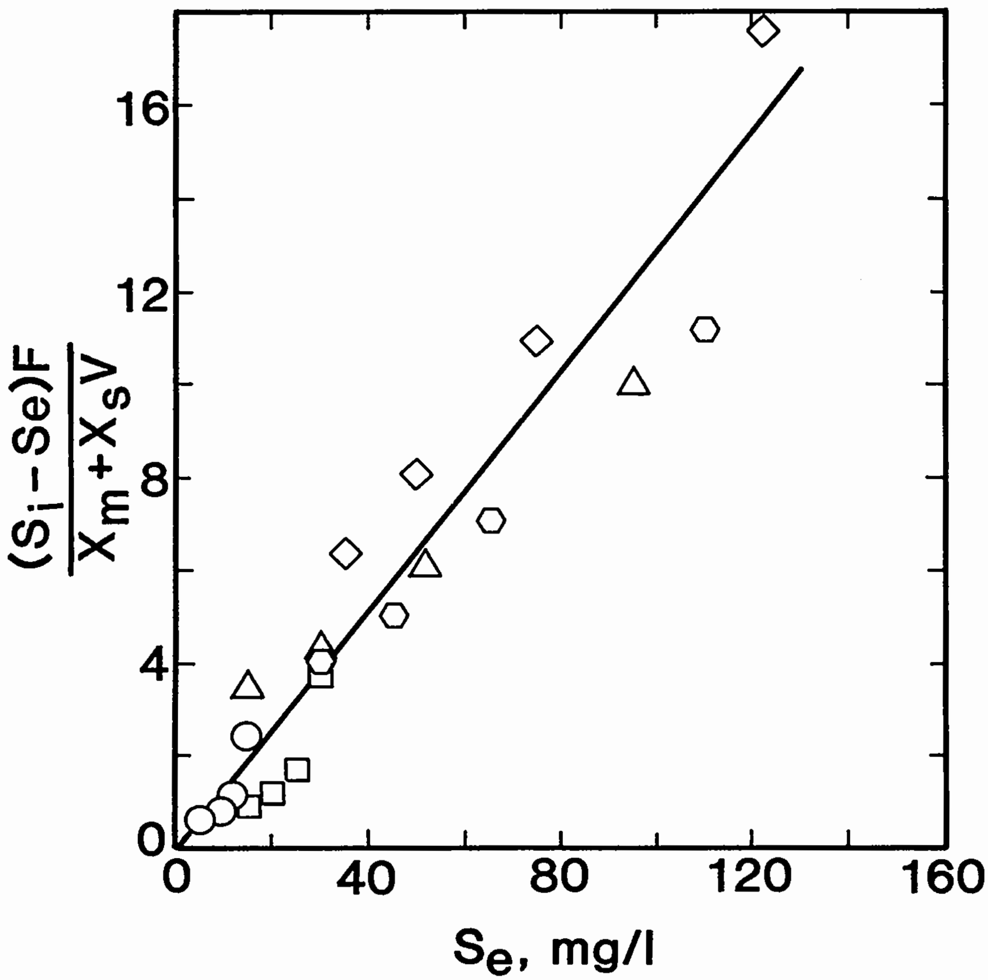

Figure 7. Specific Substrate Utilization Rate Using Suspended Solids vs. Effluent ΔCOD.

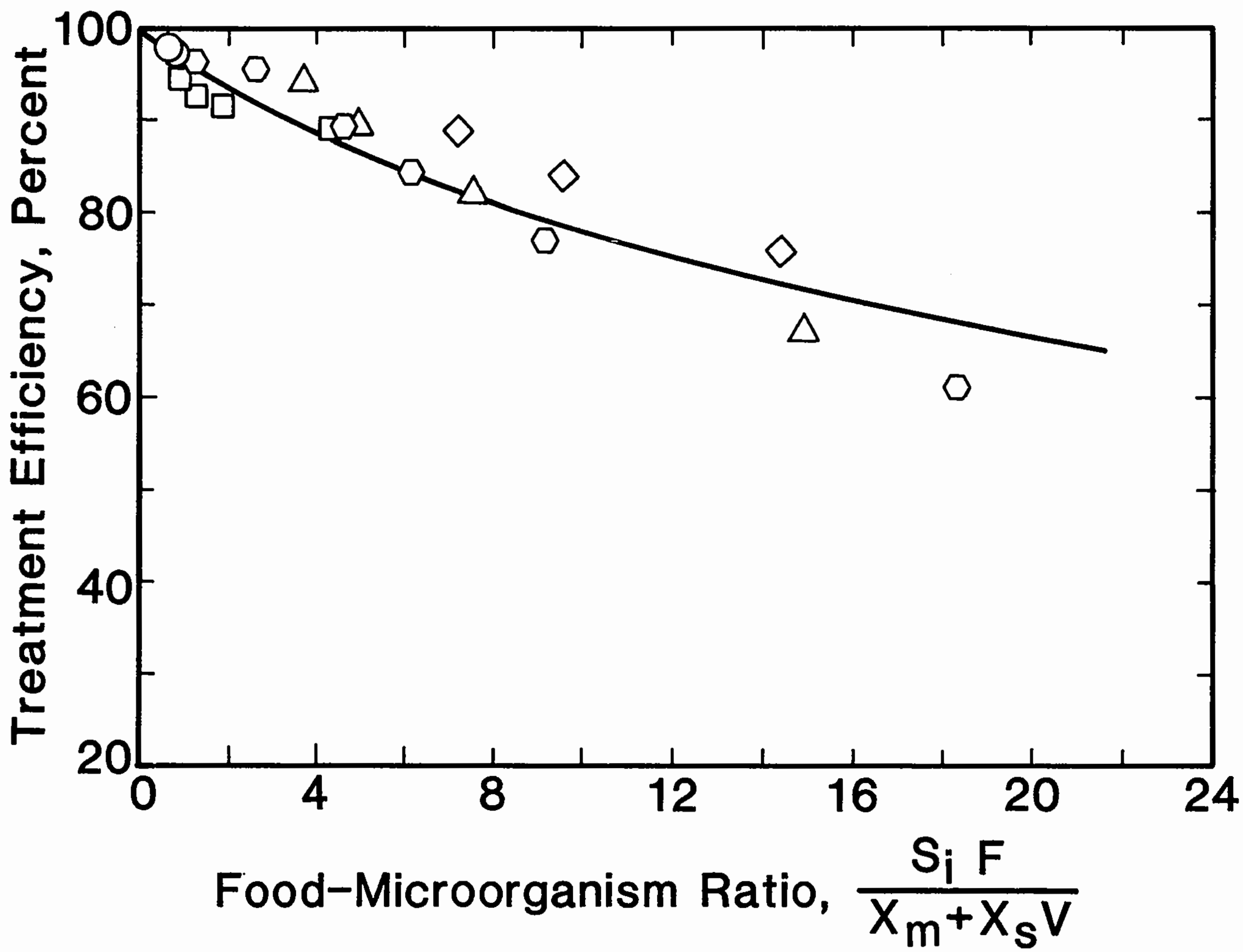

Figure 8. Treatment Efficiency vs. Food–Micro-organism Ratio.

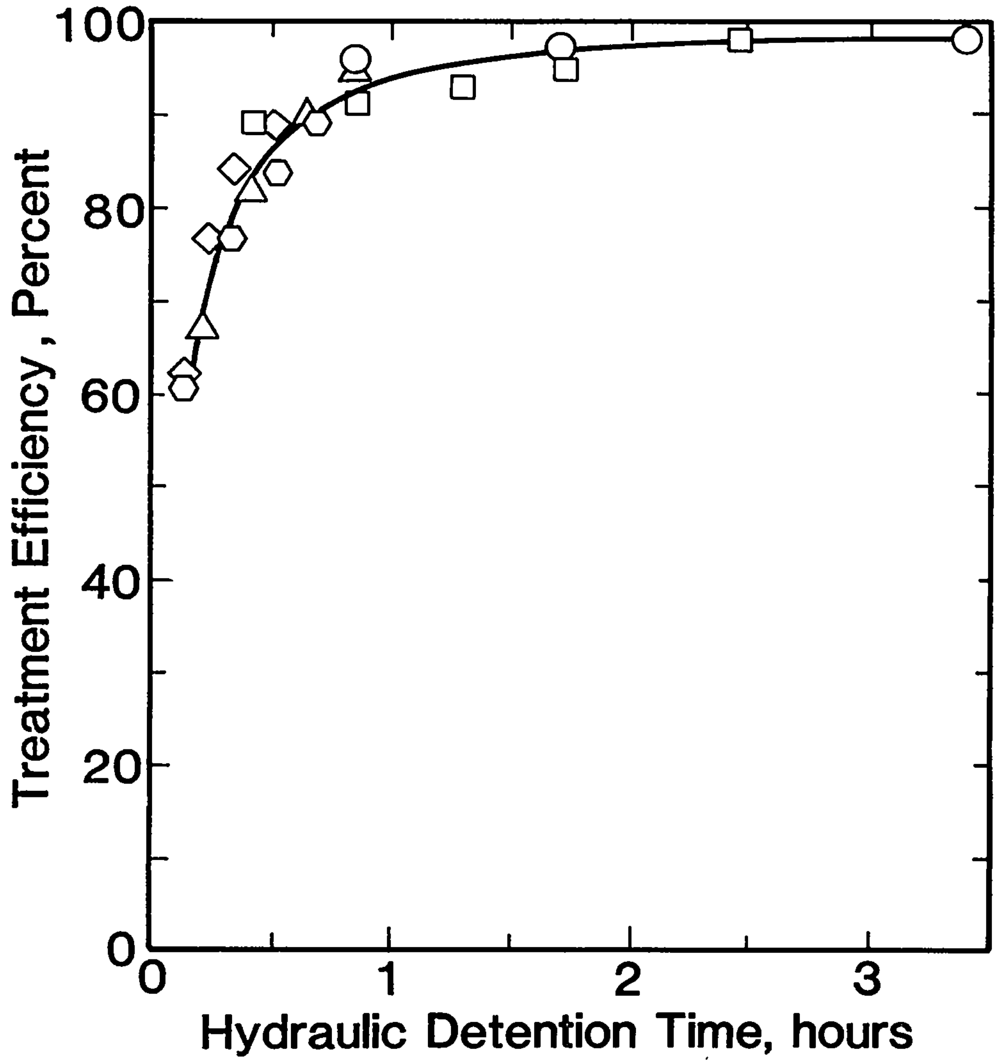

Figure 9. Treatment Efficiency vs. Hydraulic Detention Time.

Under these conditions, the suspended biological solids would be active in utilizing substrate. It appears that this is what happened in this study. At hydraulic loadings of 1.5 and 3.0 gpd/ft^2, the batch growth studies gave maximum specific growth rates close to the dilution rates. Higher hydraulic loadings resulted in the dilution rate being much greater than the maximum specific growth rate.

After making the assumption that the biological suspended solids were only active in substrate removal at hydraulic loadings of 1.5 and 3.0 gpd/ft^2, comparisons were made between parameters using suspended solids and those not using suspended solids. A single relationship between specific substrate utilization and effluent ΔCOD was obtained only when the suspended solids were included in calculating the specific substrate utilization.

A Monod like relationship between the specific substrate utilization rate and effluent ΔCOD was observed when the suspended solids were not included. However, a first order relationship was observed when the suspended solids were included.

These two comparisons show that the suspended solids do exert an effect on the kinetics of the RBC process. This is especially true at low effluent substrate requirements. The normal way to achieve a low effluent substrate is by using a low hydraulic loading. This produces a detention that allows the suspended solids to not "wash out". This then allows the suspended solids to be active in removing substrate. Since present design models do not consider suspended solids, most designs have a built-in safety factor. This may be an advantage of not considering suspended solids when designing an RBC process. However, suspended solids consideration could provide a more economical design.

It was also an interest of this study to determine whether or not parameters using biomass could be used to describe the RBC process. Such parameters would allow a more direct comparison between the RBC process and the activated sludge process. It was found that food to microorganism ratios could be calculated using an active biomass and suspended solids that compare well with food to microorganism ratios calculated for activated sludge. RBC food to microorganism ratios of one or less gave treatment efficiencies of 95 percent or better.

A reaction rate constant that compares with Eckenfelder's activated sludge reaction rate constant was also determined in this study. Figure 7 shows this relationship. The slope of the straight lines is 0.13 l/mg/ℓ-day. This reaction rate compares well with Eckenfelder's constant that has been reported for easily biodegraded organic wastewaters.

This study has shown that suspended solids can play an important role in the kinetics of RBCs and that they should not be completely disregarded. The actual importance of the suspended solids depends upon a particular situation.

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THE KINETICS OF A ROTATING BIOLOGICAL CONTACTOR TREATING DOMESTIC SEWAGE

by

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INTRODUCTION

Background

The Rotating Biological Contactor (RBC) has been employed for biological treatment of municipal and industrial wastewater for several years. The process has been shown to be efficient and cost effective in various types of applications. Traditionally, the design of RBC systems has been based primarily on empirical relationships and design curves developed from pilot plant studies. This empirical design approach has ignored the basic concepts of biological substrate removal kinetics. Although several recent investigations have developed various models to describe the performance of the RBC process, there is limited data available describing the kinetic constants associated with substrate removal. Also, there is very little available information concerning the effect of temperature on the kinetic constants associated with RBC substrate removal.

Objectives

The general objective of this study is to develop the kinetic constants describing carbonaceous substrate removal in the RBC process treating domestic sewage and to determine the effect of temperature on these constants. The kinetic constants developed will then be employed to develop a rational approach to RBC systems design for treatment of municipal wastewater.

To accomplish the above general objective, the following specific objectives will be achieved.

1. Develop a Monod type [*Monod* 1942], steady state model which describes carbonaceous substrate removal in the RBC process.

2. Determine the values of the kinetic constants (maximum specific growth rate, $\hat{\mu}$; half saturation constant, K_s; decay coefficient, k_d; and the yield coefficient, Y) for the above model as a function of temperature.

3. Develop, using laboratory scale RBC units treating domestic sewage, the data base required to achieve the above specific objectives.

4. Employ the kinetic constants produced to develop a rational RBC design procedure for domestic wastewater.

Scope

This paper details preliminary results obtained from the initial phase of the study. The results presented in this paper are limited to only one temperature (15°C) and the analysis is preliminary in nature. The study is currently underway at Utah State University, Logan, Utah, and will be completed in the near future. The results of the entire study will be presented at a later date.

PREVIOUS INVESTIGATIONS

Initial attempts to model the performance of rotating biological contactors (RBC) were empirical in nature and mainly employed regression analysis [*Hartman* 1965, *Jost* 1969, *Antonie and Welch* 1969, and *Weng and Molof* 1974]. Their efforts generally ignored temperature effects and were not directly related to microbial substrate removal. Substrate kinetic removal equations have been developed by several investigators [*Grieves* 1972, *Hansford et al.* 1976, *Benjes* 1978, *Kornegay and Andrews* 1968, *Kornegay* 1972, *Kornegay* 1975, and *Clark et al.* 1978]. In general these models employed either saturation kinetics [*Monod* 1942] or first order kinetics to describe substrate removal. Usually these equations were limited to a single stage system or treated a multi-stage system as a single unit.

Kornegay and Andrews [1968] investigated the kinetics of fixed film biological reactors at 25°C using a rotating drum and glucose as the substrate. Under controlled flowrates, glucose concentration, and attached film thickness, they found active biomass thickness to be 70μ, the half saturation constant (K_s) 121 mg/ℓ (glucose), and the maximum specific growth rate ($\hat{\mu}$) 0.28 hr^{-1} (6.7 day^{-1}). *Grieves* [1972] developed a theoretical dynamic and steady state model for the RBC using kinetic constants from the literature. He verified his model by conducting dynamic tests at 20°C using glucose as the substrate.

Clark et al. [1978] investigated the kinetics of BOD removal under varying wastewater flows and concentrations, using primary effluent with soluble BOD_5 of 32-88 mg/ℓ. Working with a four-stage RBC unit, they obtained a yield coefficient (Y) of 0.96 (based on soluble BOD_5), a half saturation constant (K_s) from 431 mg/ℓ (first stage) to 18 mg/ℓ (fourth stage), and maximum growth rates ($\hat{\mu}$) of 4.4 day^{-1} (first stage) to 0.3 day^{-1} (fourth stage).

Recent investigations have employed either mass transfer models or have combined mass transfer concepts with substrate removal kinetics to describe RBC performance [*Schroeder* 1976, *Friedman et al.* 1976, and *Famularo et al.* 1978]. These equations have generally been applied in oxygen limited substrate removal situations.

MATERIALS AND METHODS

Four, four stage, 38 cm diameter, laboratory scale rotating biological contactor (RBC) units furnished by the Environmental Systems Division, George A. Hormel and Company, Coon Rapids, Minnesota, will be employed to develop the data necessary to determine the values of the kinetic constants. These units will receive settled domestic sewage from the Hyrum City Wastewater Treatment Plant, Hyrum, Utah. The sewage is collected during a 30 minute period three times per week and is transported to the Utah State University campus and stored at approximately 2°C until fed to the RBC units. The sewage is collected at the same time of day to ensure uniform composition and strength.

The units will be operated at four different organic loading rates at four different temperatures. This paper details the preliminary results of the 15°C experimental design. The hydraulic loading rate will be held constant and is the same for each unit. The experimental apparatus is shown schematically in Figure 1.

Tables 1 and 2 indicate the hydraulic loading rate, the organic loading rate, and the liquid temperature of each unit during the 15°C experimental phase. The organic loading rates ranged from 5.9 g COD/m^2/day in Unit A to 22.0 g COD/m^2/day in Unit D. The various organic loading rates are achieved by diluting the raw sewage with dechlorinated tap water. The liquid temperature in the RBC units ranged from 14.1 to 16.6 °C. This temperature variation was a result of heat loss through the stages rather than temperature differences between similar stages.

The influent to the system and the effluent from each stage was monitored using 24-hour composite samples (20 minute intervals) and occasionally grab samples every two days during steady state conditions

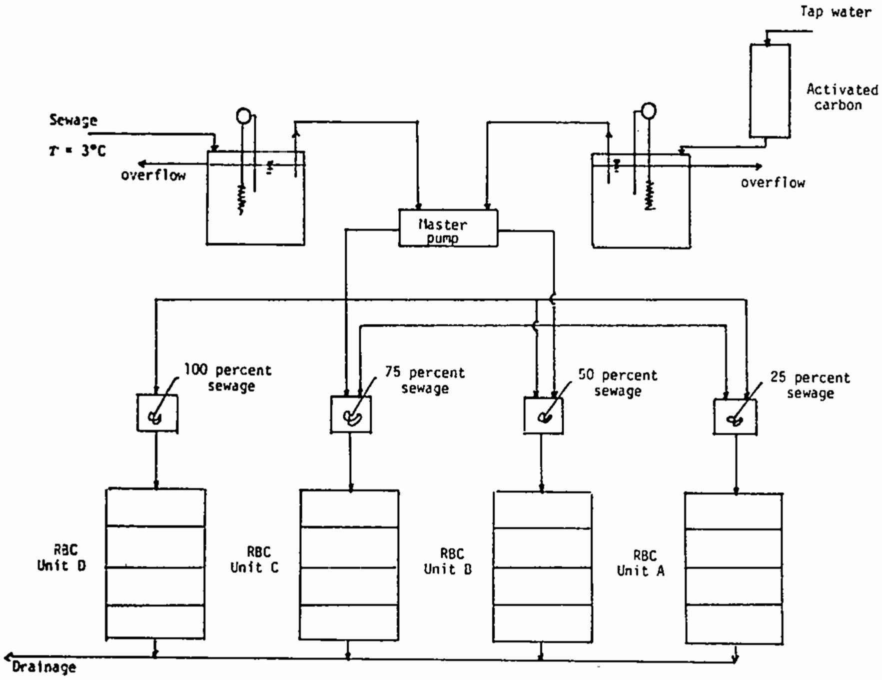

Figure 1. Schematic of experimental apparatus employed to develop data base for developing kinetic constants.

Table 1. Average hydraulic and organic loading rates in the four laboratory scale rotating biological contactor units.

	Unit			
Parameter	A	B	C	D
Average hydraulic loading rate (ℓ/day)	274±12	285±6	283±14	291±9
Average organic loading rate (grams of COD/ m^2/day)	5.9	11.5	15.5	22.0

Table 2. Average liquid temperature (°C) in each stage of each laboratory scale rotating biological contactor unit during data collection phase.

Stage Number	Unit				Average Temperature (°C)
	A	B	C	D	
1	16.6±0.4	16.0±0.3	16.0±0.4	16.4±0.3	16.3
2	15.4±0.5	15.4±0.4	15.0±0.4	15.7±0.2	15.4
3	14.8±0.3	14.4±0.5	14.4±0.5	15.1±0.2	14.7
4	14.8±0.3	14.1±0.5	14.1±0.4	14.7±0.3	14.4

for chemical oxygen demand (COD), suspended solids, volatile suspended solids, ammonia-nitrogen, nitrite-nitrogen, nitrate-nitrogen and total Kjeldahl nitrogen. In addition, in-situ measurements of flow, temperature, pH and dissolved oxygen were conducted. The ampule technique [*Oceanographics* 1978] was employed for COD analysis while all other analyses were conducted according to Standard Methods [*APHA* 1975].

At the conclusion of the data collection phase, the entire solids from each disc in each stage was removed from the disc and analyzed for total and volatile solids. Thus, the total biomass of the system was determired.

Kinetic Constants

The model employed in the study is based on *Monod* [1942] substrate removal kinetics and assumes steady state conditions.

The substrate removal in a single stage Rotating Biological Contactor (RBC) can be equated to the growth of microorganisms as shown in equation 1.

$$Q(S_0-S_e) = \frac{1}{Y}\frac{dX}{dt} \qquad (1)$$

where

Q = flowrate

S_0 = influent substrate concentration

S_e = effluent substrate concentration

Y = yield coefficient = $\frac{\text{mass of biomass produced}}{\text{mass of substrate removed}}$

$\frac{dX}{dt}$ = change in biomass per unit time

Monod [1942] indicated that the change in biomass under substrate limiting conditions can be represented by a saturation function

$$\frac{dX}{dt} = \hat{\mu}\left(\frac{S\,X}{K_s + S}\right) \qquad (2)$$

where

$\hat{\mu}$ = maximum specific growth rate

X = biomass concentration

K_s = half saturation constant (*i.e.*, substrate concentration at a growth rate equal to half the maximum specific growth rate)

S = limiting substrate concentration

Equations 1 and 2 can be combined and expressed in linear form as shown below.

$$\frac{X}{Q(S_0 - S_e)Y} = \left(\frac{K_s}{\hat{\mu}}\right)\frac{1}{S_e} + \frac{1}{\hat{\mu}} \qquad (3)$$

If the value of the yield coefficient, Y, is known, experimental data can be fitted to equation 3 and the value of the kinetic constants, $\hat{\mu}$ and K_s, can be determined.

The value of the yield coefficient, Y, can be obtained by writing a mass balance biomass in a single RBC stage as shown in equation 4. This equation assumes no change in the mass of biomass occurring once steady state conditions are achieved.

$$\text{Biomass Produced} = \begin{matrix}\text{Biomass sloughed}\\ \text{from the disc}\end{matrix} + \text{Decay} \qquad (4)$$

Equation 4 may be expressed mathematically as

$$\frac{dX}{dt} = Q(X_e) + k_d X \qquad (5)$$

where

X_e = effluent biomass concentration (generally measured as volatile suspended solids)

k_d = decay coefficient

X = total biomass (generally measured as volatile suspended solids)

Equation 5 does not distinguish between "active" and "nonactive" biomass. Combining equation 5 with equation 1 and rearranging into linear form results in equation 6.

$$\frac{Q(X_e)}{X} = \frac{YQ(S_0 - S_e)}{X} - k_d \tag{6}$$

Experimental data fitted to equation 6 will result in values for the yield coefficient, Y, and the decay coefficient, k_d. Thus, all of the desired kinetic constants can be determine from equations 3 and 6. Solving these equations with experimental data collected at various temperatures will yield kinetic constant values at these temperatures. Thus, the effect of temperature on the kinetic constants can be determined. A previous study conducted by the authors indicates that this effect may be described by an Arrhenius type relationship [*Mikula* 1979, *Mikula et al.* 1980].

RESULTS AND DISCUSSION

Process Performance

The laboratory scale rotating biological contactor (RBC) units were operated for approximately one month before steady state conditions were achieved in each stage. The attached biomass in the first stage

Table 3. Dissolved oxygen and pH of the laboratory scale rotating biological contactor (RBC) units.

Unit	Stage Number	December 1, 1979		December 24, 1979	
		pH	Dissolved Oxygen (Mg/1)	pH	Dissolved Oxygen (Mg/1)
A	1	7.80	5.0		
	2	7.85	6.7		
	3	7.90	7.4		
	4	8.00	7.8		
B	1	7.70	3.9		
	2	7.80	5.3		
	3	7.85	6.8		
	4	7.90	7.5		
C	1	7.70	3.1	7.70	4.0
	2	7.80	4.6	7.65	5.1
	3	7.85	5.8	7.60	4.3
	4	7.85	7.6	7.60	5.2
D	1	7.65	2.5	7.80	2.8
	2	7.80	3.6	7.75	3.2
	3	7.85	4.5	7.65	3.4
	4	7.85	10.2*	7.50	3.9

* Suspected analytical error

Table 4. Total volatile solids biomass present in each stage of the laboratory scale rotating biological contactors.

Stage Number	Volatile Solids in Each Stage (grams/m^2)			
	Unit A	Unit B	Unit C	Unit D
1	27.8	36.4	47.5	46.7
2	6.6	29.6	36.0	38.8
3	4.1	10.9	21.5	23.2
4	2.1	6.2	13.8	16.8

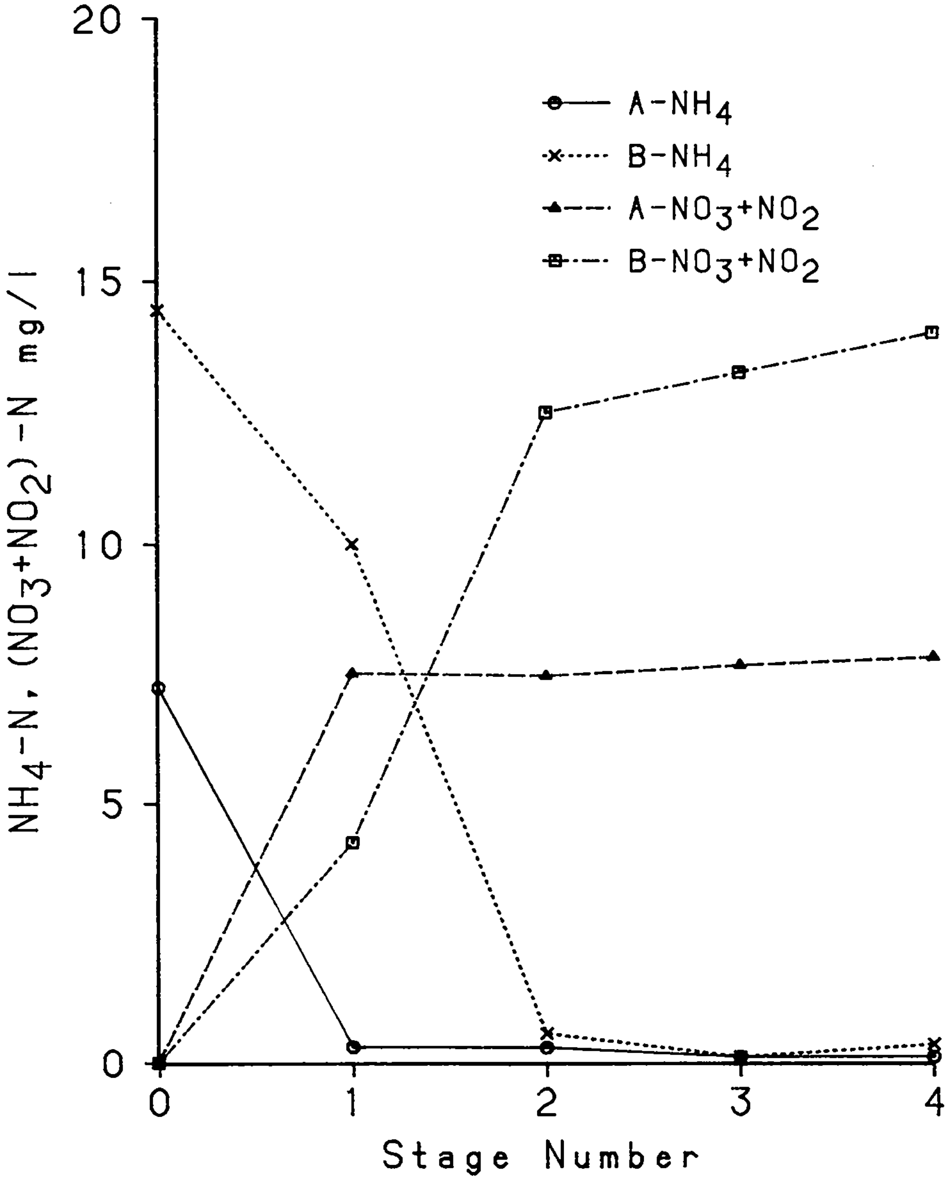

Figure 2. Nitrogen response of laboratory scale rotating biological contactors, Units A and B.

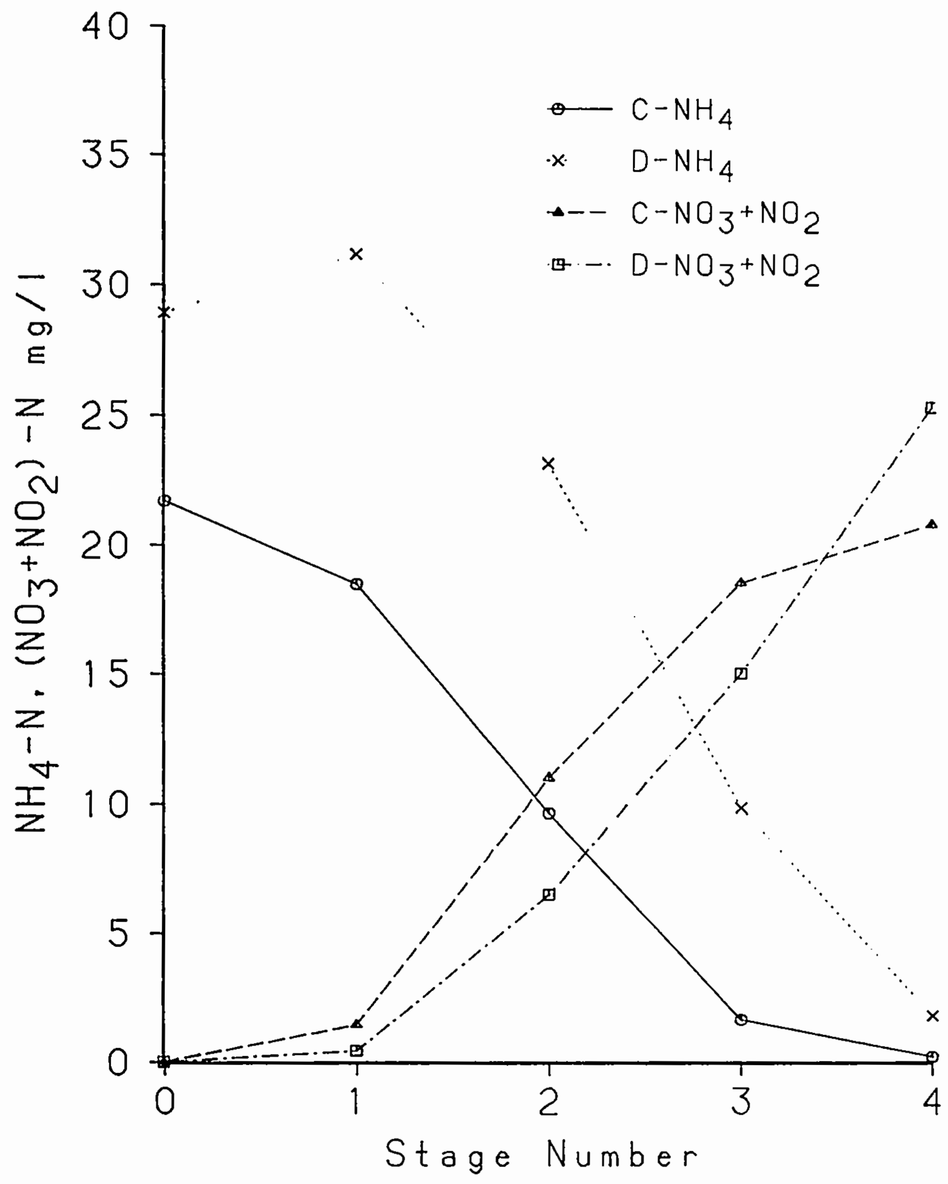

Figure 3. Nitrogen response of laboratory scale rotating biological contactors, Units C and D.

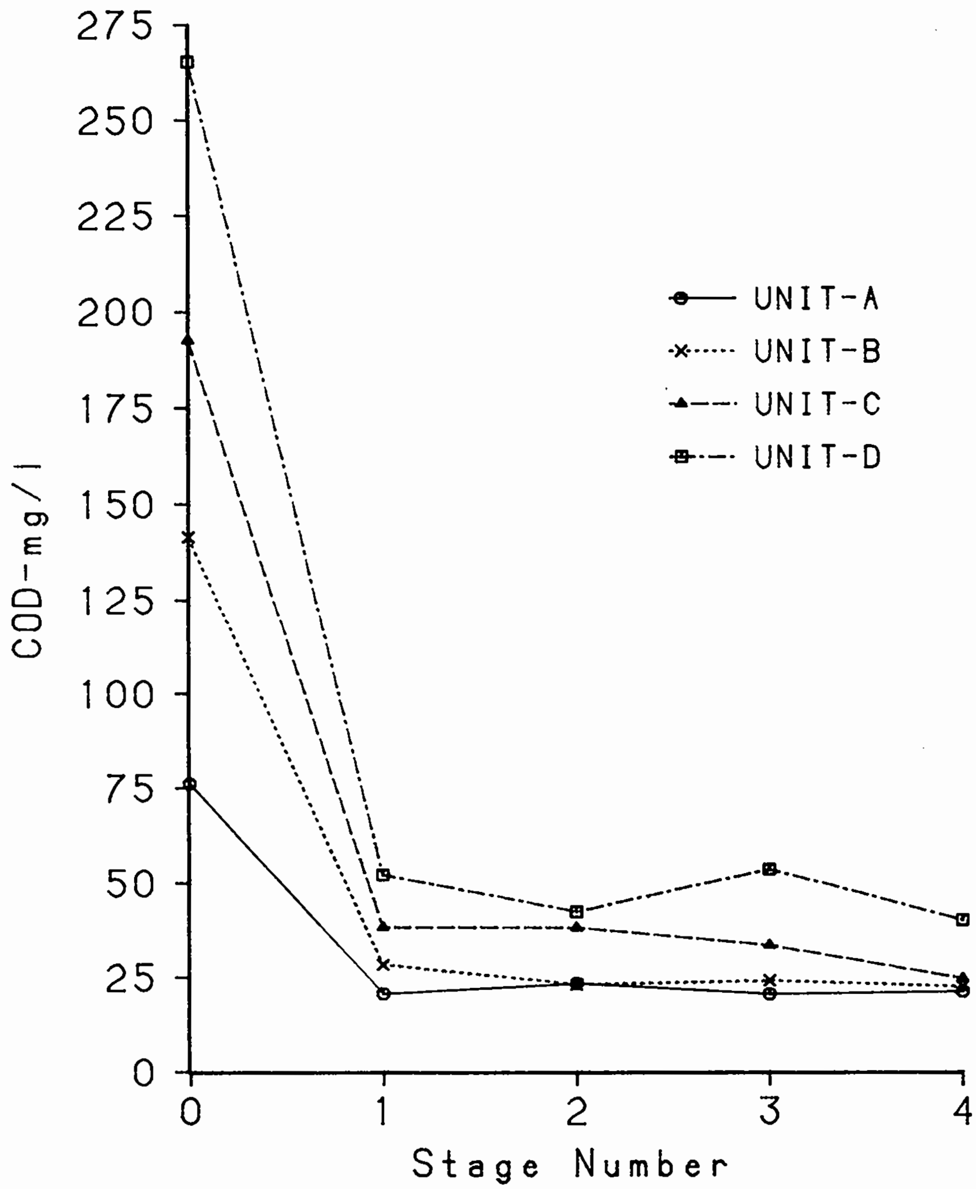

Figure 4. Carbonaceous substrate removal in the laboratory scale rotating biological contactor units. Note: (1) 0 = Total Chemical Oxygen Demand (COD) of the influent to the system. (2) Effluent values for stages 1, 2, 3, and 4 are in terms of soluble COD.

Kinetic Constant Determination

The steady state data employed with equations 3 and 6 to determine the carbonaceous substrate removal kinetic constants are reported in Table 5. The values for the influent substrate concentration, S_0, are measured in terms of total chemical oxygen demand (COD). While the effluent substrate concentrations are measured in terms of soluble COD. This approach assumes that all the influent particulate organic material is solubilized within the reactor and is available for uptake by the microorganisms.

The values for the total biomass, X, present in each reactor are measured in terms of volatile solids and the entire depth of biomass is assumed to be active. This assumption was made because measurements for the fraction of active biomass were not conducted.

The data obtained from Unit A were not included in the development of the kinetic constants due to the high degree of nitrification occurring within this unit. Initial attempts to include the Unit A data in the analysis resulted in unrealistic and inconsistent results. Future data analysis at other temperatures will attempt to account for the nitrification phenomenon.

The results of the linear regression for equation 6 to determine the values for the yield coefficient, Y, and the decay coefficient, k_d, are reported in Table 6. The yield coefficients varied from 0.81 in the fourth stage to 1.44 in the second stage. These values are based on COD removal. These values are similar to those found by *Clark et al.* [1978] who reported yield coefficients of 0.68 to 1.11 based on soluble biochemical oxygen demand (BOD_5).

The values for the decay coefficient, k_d, range from 0 days^{-1} in the third stage to 0.44 days^{-1} in the second stage. These values appear to be relatively high and are probably highly influenced by the incorporation of the total volatile biomass in the analysis rather than only the "active" biomass.

The values for the other kinetic constants developed from equation 3 are reported in Table 7. The regression plots of equation 3 for each stage are shown in Figures 5 to 8. The values for the maximum specific growth rate, $\hat{\mu}$, appear to decrease with stage, except for the second stage. The half saturation values, K_s, also decrease significantly in the fourth stage. The values of $\hat{\mu}$ range from 1.47 day^{-1} to 2.92 day^{-1}. *Kornegay and Andrews* [1968] reported a $\hat{\mu}$ of 6.7 days^{-1} at 25°C and *Clark et al.* [1978] reported a $\hat{\mu}$ of 4.4 days^{-1} in the first stage of an RBC unit and 0.3 days^{-1} in the fourth stage of an RBC unit. The relatively small values of $\hat{\mu}$ reported in this present study are probably due to the low temperature (15°C). *Reynolds* [1975] reported that the value of $\hat{\mu}$ for algae is temperature related and is smaller at lower temperatures.

The values for the half saturation constant, K_s, range from 6.0 mg COD/ℓ is the fourth stage to 67.4 mg COD/ℓ in the second stage.

Table 5. Steady state data employed to develop kinetic constants for carbonaceous substrate removal at 15°C.

Unit	Stage Number	Influent Total COD[†]		Effluent Total COD[§]		Effluent Soluble COD		Effluent Volatile Suspended Solids	
		No. of Samples	Average Concentration (mg/ℓ)	No. of Samples	Average Concentration (mg/ℓ)	No. of Samples	Average Concentration (mg/ℓ)	No. of Samples	Average Concentration (mg/ℓ)
A	1	8	76.1	2	41.9	4	20.7	4	24.0
	2			4	53.3	3	23.4	5	27.7
	3			2	35.0	2	20.6	5	25.3
	4					2	21.3	5	20.7
B	1	9	141.4	2	80.0	2	28.3	5	52.3
	2			3	47.4	3	23.1	5	37.8
	3			2	55.9	2	24.1	5	30.2
	4			2	41.5	3	22.5	5	33.7
C	1	11	192.6	5	123.6	6	38.3	7	85.4
	2			4	101.9	5	38.2	7	58.5
	3			5	96.7	5	33.5	7	72.0
	4			4	77.1	4	24.6	7	54.8
D	1	12	265.2	5	168.5	6	52.1	7	141.3
	2			5	131.7	7	42.4	7	122.5
	3			5	136.2	6	53.6	7	99.1
	4			4	107.7	6	40.2	7	73.5

† Influent to first stage

§ Influent to next stage

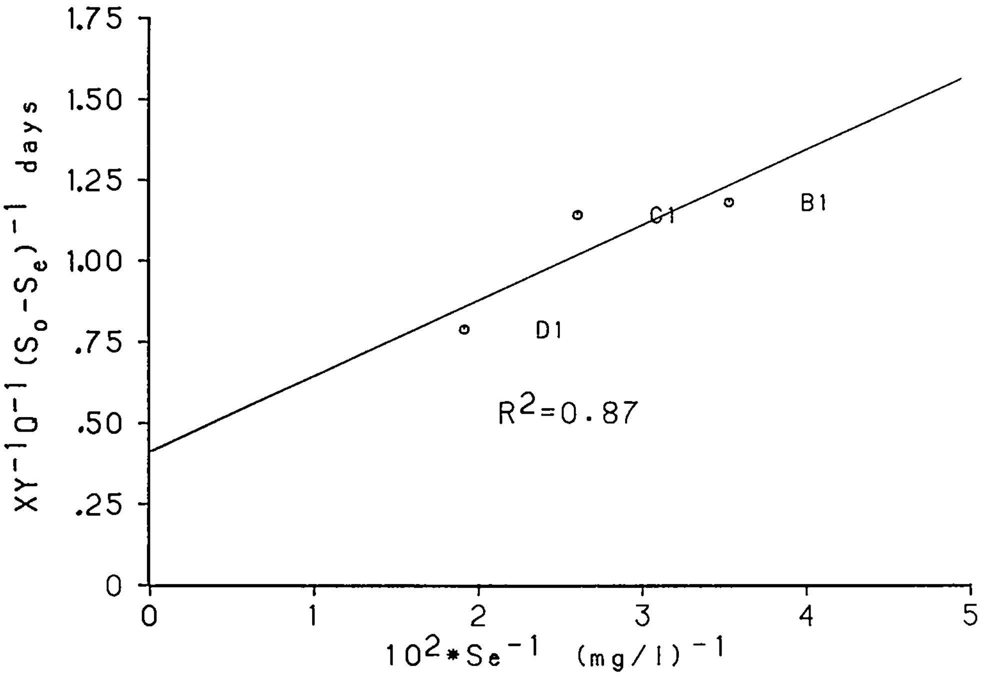

Figure 5. Regression analysis for determination of the maximum specific growth rate, $\hat{\mu}$, and half saturation constant, K_S, in the first stage.

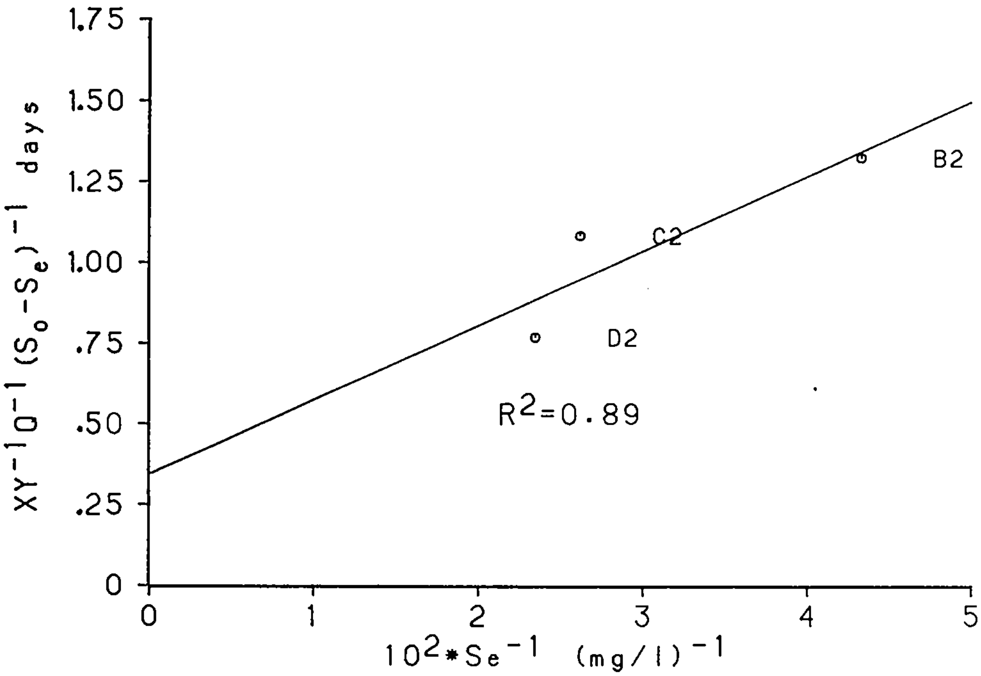

Figure 6. Regression analysis for determination of the maximum specific growth rate, $\hat{\mu}$, and the half saturation constant, K_S, in the second stage.

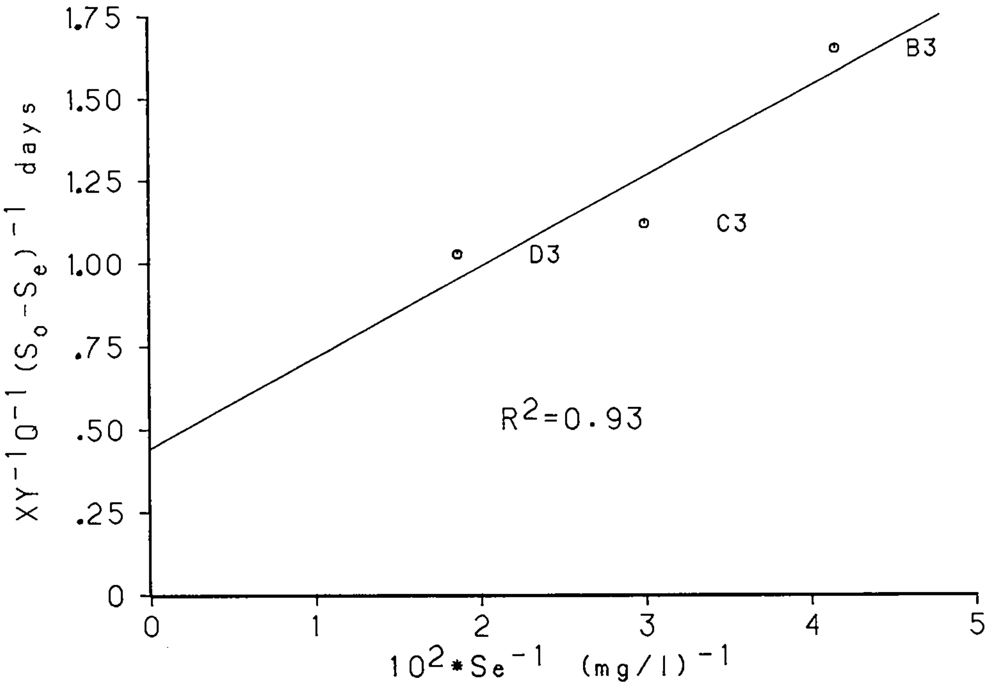

Figure 7. Regression analysis for determination of the maximum specific growth rate, $\hat{\mu}$, and the half saturation constant, K_s, in the third stage.

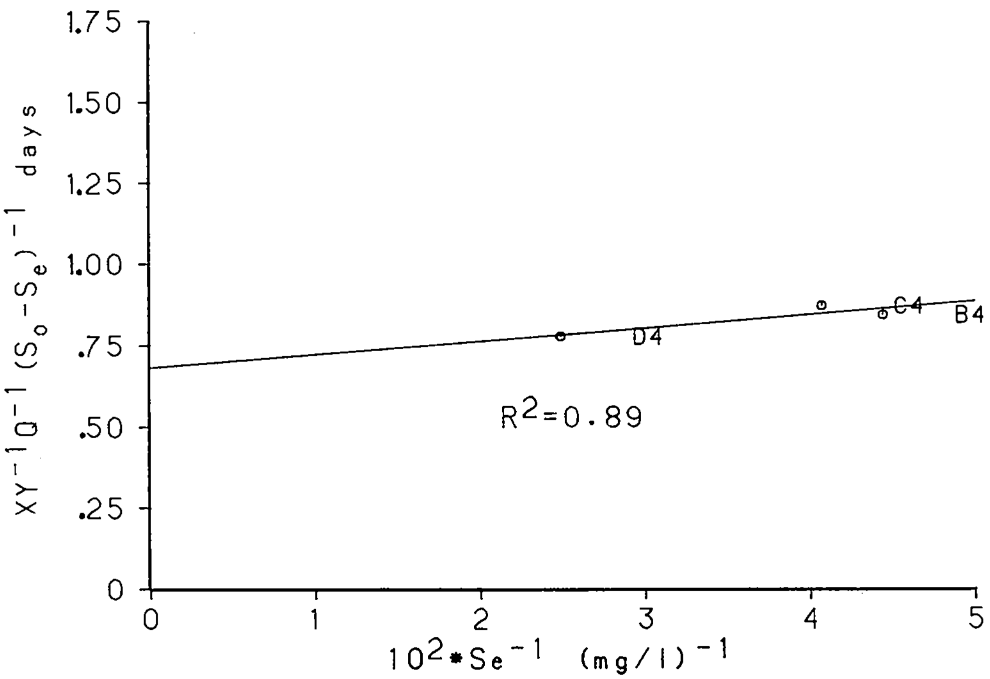

Figure 8. Regression analysis for determination of the maximum specific growth rate, $\hat{\mu}$, and the half saturation constant, K_s, in the fourth stage.

Table 6. Results of analysis for determination of the yield coefficient, Y, and the decay coefficient, k_d.

Parameter	Stage Number			
	1	2	3	4
Yield coefficient, Y	1.00	1.44	1.04	0.81
Decay coefficient, k_d, (days^{-1})	0.42	0.44	0	0.07
Regression coefficient, R^2	0.991	0.990	0.880	*

* The values for the fourth stage are based on Units C and D only and, therefore R^2 is meaningless.

Table 7. Values for the maximum specific growth rate, $\hat{\mu}$, and half saturation constant, K_s, at 15°C.

Parameter	Stage Number			
	1	2	3	4
Maximum specific growth rate, $\hat{\mu}$, (days^{-1})	2.43	2.92	2.26	1.47
Half saturation constant, K_s, (mg COD/ℓ)	56.6	67.4	61.5	6.0
Regression coefficient, R^2	0.87	0.89	0.93	0.89

The K_s values for the final three stages are approximately the same. The fourth stage K_s value is substantially less than the other three values. This could be due to the occurrence of nitrification in the fourth stage. The low fourth stage K_s value also indicates that the liquid entering the fourth stage is less biodegradable. This is expected as the easily biodegradable organic compounds are assimilated in the earlier stages of treatment. These K_s values are slightly lower than those reported by *Clark et al.* [1978] who reported K_s values ranging from 8 mg BOD_5/ℓ to 431 mg BOD_5/ℓ. However, the K_s values in this study are similar to those reported by *Mikula* [1979], who reported K_s values ranging from 10 to 186 mg COD/ℓ.

SUMMARY AND CONCLUSIONS

Little information is available concerning the kinetics of carbonaceous substrate removal in the rotating biological contactor (RBC). The general objective of this study was to develop the kinetic constants describing carbonaceous substrate removal in the RBC process treating domestic sewage and to determine the effect of temperature on these constants. This paper details preliminary results of only the 15°C phase of the study. The study is currently underway and the completed study results will be reported at a later date.

The values for the kinetic constants in a Monod steady state carbonaceous substrate removal equation have been developed for 15°C using data collected from four laboratory scale four stage RBC units treating domestic sewage. The values for the yield coefficient, Y, ranged from 0.81 to 1.44, while the values for the decay coefficient, k_d, ranged from 0 to 0.44 days^{-1}. The maximum specific growth rate, $\hat{\mu}$, varied from 1.47 to 2.92 days^{-1} while the half saturation constant, K_s, ranged from 6.0 to 67.4 mg COD/ℓ.

Additional study will verify the values of these kinetic constants and also indicate the effect of temperature on substrate removal. These results will be presented at a later date.

ACKNOWLEDGMENTS

Equipment and financial assistance for this study were provided by the Environmental Systems Division, George A. Hormel and Company, Coon Rapids, Minnesota.

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APPENDIX

Nomenclature

a = Particulate COD to VS

BOD = Biochemical oxygen demand

COD = Chemical oxygen demand

DO = Dissolved oxygen

k_d = Endogeneous respiration

K_s = Half saturation constant

Q = Flow rate

S = Substrate concentration

S_0 = Influent substrate concentration

S_e = Effluent substrate concentration

t = Time

V = Stage volume

VS = Volatile solids

X = Biomass amount

Y = Yield coefficient

$\hat{\mu}$ = Maximum specific growth

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FIXED FILM NITRIFICATION SURFACE REACTION KINETICS AND ITS APPLICATION IN RBC SYSTEMS

By

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INTRODUCTION

The waste removal process in a biological fixed film, as in the trickling filter, or in the rotating biological contactor (RBC), depends on mass transfer from the wastewater to the slime layer, followed by metabolism of the waste by microorganisms. The carbonaceous substrate removal in the fixed film process has been found to be a substrate diffusion-limited, heterogeneous model by numerous studies (Ames, et al., 1962, Atkinson, et al., 1967, Gulerich, et al., 1968, Maier, 1969, and Ross, 1970). The nitrification process, however, may differ from the carbonaceous substrate removal process because the autotrophic nitrifiers have a much lower growth rate than the heterotrophs and also the ammonia diffusivity in water is several times higher than the carbonaceous substrate diffusivity such as glucose (Perry, 1963). Therefore, the surface reaction model in nitrification process should be investigated, so that the correct model can be applied in the fixed film nitrification process design.

This paper includes two stuides: STUDY I used a stationary inclined plate on which a nitrifying biological film was developed for nitrification surface reaction model study; STUDY II used a bench scale rotating biological contactor to check if the findings from the stationary biological film process are also applicable to the rotating biological film process.

SURFACE REACTION MODELS IN BIOLOGICAL FIXED FILM PROCESS

Two possibilities exist for the geometry of the interface between the microorganisms and the liquid phase:

1. The Pseudo-Homogeneous Model: This model assumes the biological oxidation process takes place throughout the liquid film as if the microbial population were suspended in the liquid film, and there is no diffusional resistance to retard the rate of reaction, as shown in Figure 1(A);

2. The Heterogeneous Model: This model assumes that the biochemical reaction occurs at the interface of the liquid and microbial mass as depicted in Figure 2(A).

STUDY I: STATIONARY BIOLOGICAL FILM STUDY

1. Model Development

In order to describe the surface reaction models mentioned above in a more comprehensible form, the mathematical models should be developed. To describe the problem in mathematical terms, some assumptions were made:

a. A steady-state, in the biological sense, exists in that the thickness, composition and mass of the biological film remains invariant with respect to time;

b. The laminar flow regime is fully developed; and

c. The liquid film thickness on the inclined plate is defined by Nusselt's equation:

$$\delta = \left(\frac{3\mu Q}{\rho g \sin\theta}\right)^{1/3} \quad \text{.......................(1)}$$

in which δ = thickness of liquid film, μ = viscosity of liquid, Q = hydraluic loading, ρ = liquid density, g = acceleration due to gravity, θ = angle of the plane with the horizontal.

(1) The Pseudo-Homogeneous Model--

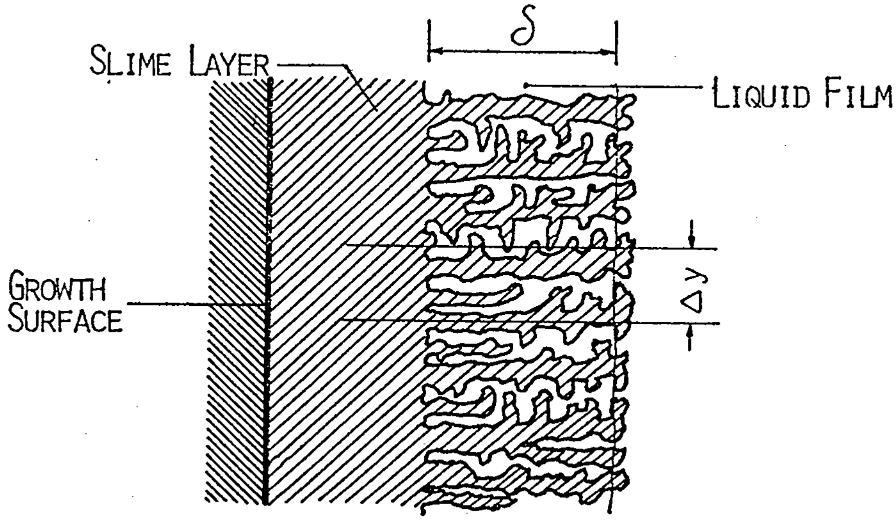
If the liquid phase diffusional resistance is neglected, a mass balance applied to a length, Δy, of unit width, as in Figure 1(B), yields:

$$Q\,dS = r\delta dy \quad \text{.......................(2)}$$

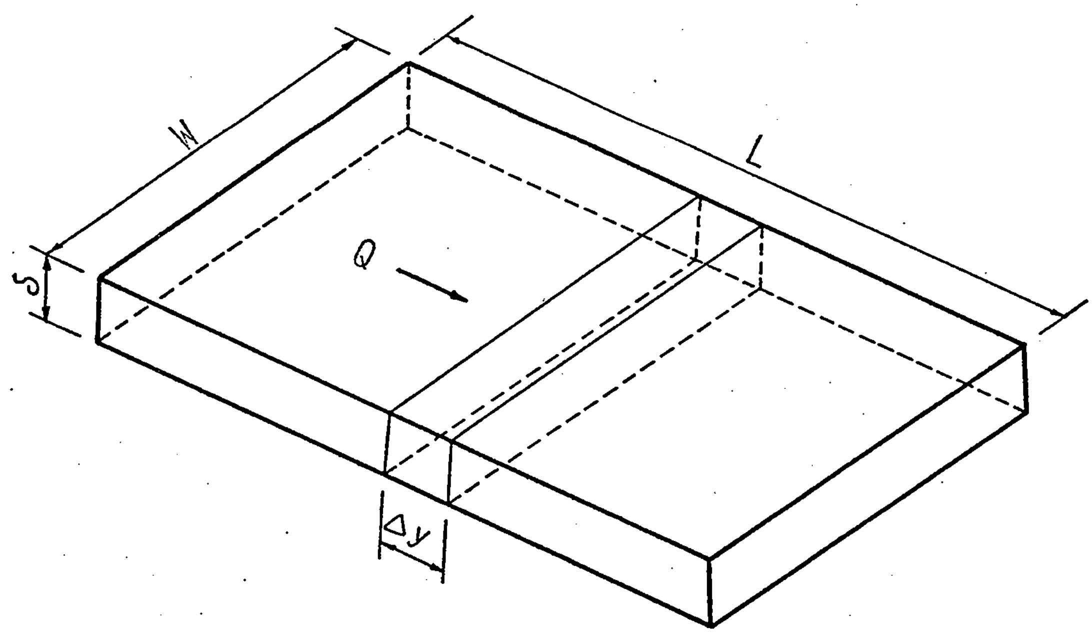
in which Q = hydraulic loading, S = substrate concentration, r = reaction rate, and δ = liquid film thickness.

The reaction rate in the nitrification process on a fixed film reactor was found to be zero-order by Huang, et al. (1974a), That is,

$$r = -k \quad \text{.......................(3)}$$

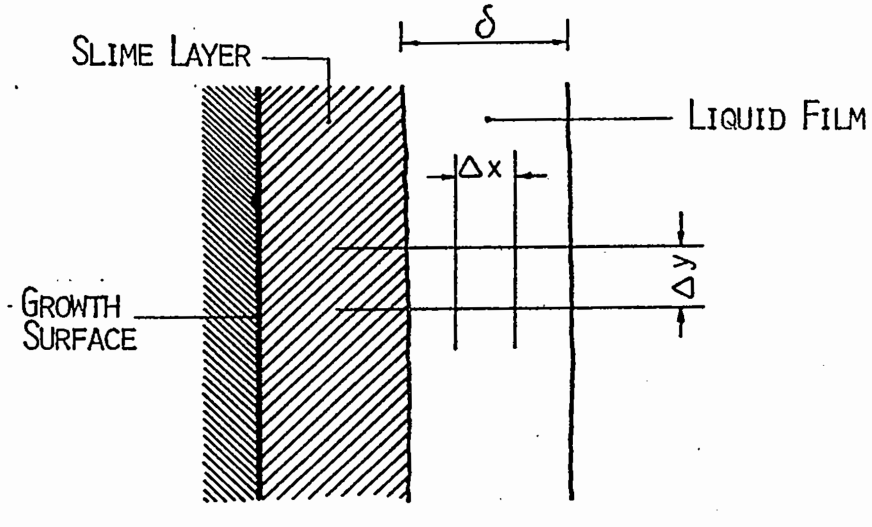


(A). POSSIBLE GEOMETRY FOR THE INTERFACE BETWEEN MICROORGANISMS AND LIQUID

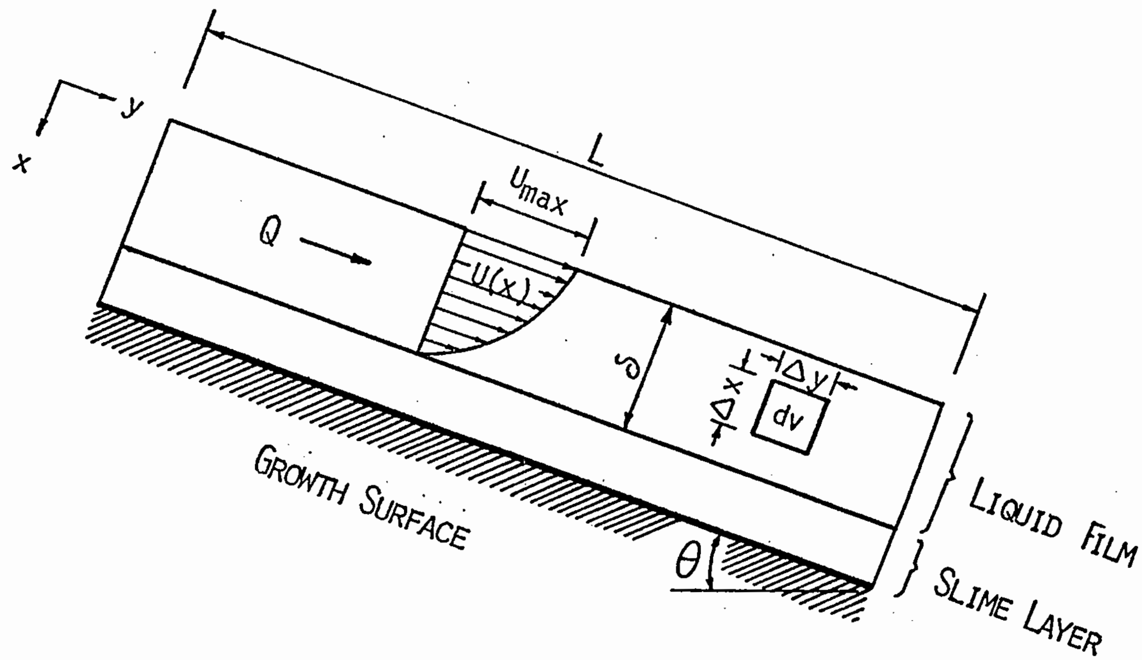


(B). SCHEMATIC DIAGRAM

FIGURE 1 - PSEUDO-HOMOGENEOUS MODEL OF THE BIOLOGICAL FILM



(A). POSSIBLE GEOMETRY FOR THE INTERFACE BETWEEN MICROORGANISMS AND LIQUID



(B). SCHEMATIC DIAGRAM

FIGURE 2 - HETEROGENEOUS MODEL OF THE BIOLOGICAL FILM

in which k = reaction rate constant, and Equation (2) becomes:

$$Q\ dS = -k\delta dy \qquad (4)$$

Integration Equation (4) and substituting Equation (1) gives:

$$S_o - S_e = (1/R_e)^{2/3} \cdot (R_v \cdot P_a)^{1/3} \qquad (5)$$

in which S_o and S_e = influent and effluent substrate concentration, respectively;

Reaction Number, $R_v = \frac{3\ k^3\ \mu}{\rho g^2\ \sin^2\theta}$; Packing Number, $P_a = \frac{L^3\ \rho^2\ g\ \sin\theta}{\mu^2}$;

Reynolds Number, $R_e = \frac{\rho Q}{\mu}$ (Fulford, 1964); and L = length of biological reactor.

For the pseudo-homogeneous film flow reactors, it has been found that recirculation has a beneficial effect on the removal of total soluble substrate (Kehrberger, et al., 1969, and Atkinson, et al., 1963).

(2) The Heterogeneous Model--

Under steady-state conditions an equilibrium situation must exist such that the rate of transfer of substrate from the main body of the liquid film to the bacterial phase will be balanced by the rate of removal and degradation by the organisms. Thus the overall rate will be determined by the relative resistances due to diffusion and removal.

The heterogeneous model can be broken into other sublevels:

(a) A reaction controlled situation- if the substrate removal rate at the reaction site is much slower than the substrate diffusional rate to the reaction site, the reaction rate will control the surface reaction model;

(b) A diffusion controlled situation- if the diffusion rate is so low that the substrate diffused into the reaction site is practically consumed immediately, the diffusional rate will control the surface reaction model;

(c) A diffusion limited situation- this is the condition which exists in between the reaction controlled situation and the diffusion controlled situation and so both the reaction rate and the diffusional rate are controlling.

The nitrification rate on a fixed film reactor was found to be a zero-order reaction (Huang, et al., 1974a). That is, there is always an excess amount of substrate at the reaction surface to allow for maximum reaction rate. The condition in which there is always an excess amount of substrate at the reaction site indicates that neither a substrate diffusion controlled situation nor a diffusion limited situation can exist in the fixed film nitrification process. For this reason, only the reaction controlled situation warrants a further investigation.

Under the situation that the substrate diffusional rate will not control the reaction rate, a differential mass balance applied to a length Δy in a unit width, as in Figure 1(B), can be described as follows:

$$Q\ dS = r\ dy \qquad (6)$$

Substituting Equation (3) into Equation (6) and integrating and substituting previously defined numbers, Equation (6) becomes:

$$S_o - S_e = (1/R_e) \cdot (R_v \cdot P_a)^{1/3} \qquad (7)$$

Equation (5) and Equation (7) can be placed in the general form:

$$S_o - S_e = K \cdot (1/R_e)^M, \text{ or}$$

$$\frac{S_o - S_{e1}}{S_o - S_{e2}} = (R_{e2}/R_{e1})^M \qquad (8)$$

in which $K = (R_v \cdot P_a)^{1/3}$, M = exponential constant, S_o = influent substrate concentration, S_{e1} and S_{e2} = effluent substrate concentration in the first hydraulic loading and the second hydraulic loading, respectively; R_{e1} and R_{e2} = Reynolds Number in the first hydraulic loading and the second hydraulic loading, respectively.

From Equation (5), the M value for a pseudo-homogeneous model is 2/3, and from Equation (7), the M value for a reaction controlled heterogeneous model is 1.

2. Method and Experimental Results

A stationary inclined plate surface, 4-in. wide and 36-in. long, provided a support surface for the slime growth. The surface of the plate was covered with a fiberglass screen which served as a structural framework for slime growth. This reactor was supported by angle frames with an inclined slope of 10^o with the horizontal. Feed was introduced into a stilling basin, where it passed over a precise mechanical overflow weir. A schematic view of the physical model is shown in Figure 3. This kind of biological film reactor has been used by many researchers for the study of biological behavior in wastewater treatment processes.

A synthetic wastewater was used for this study. A summary of the constituents of the synthetic wastewater is as follows:

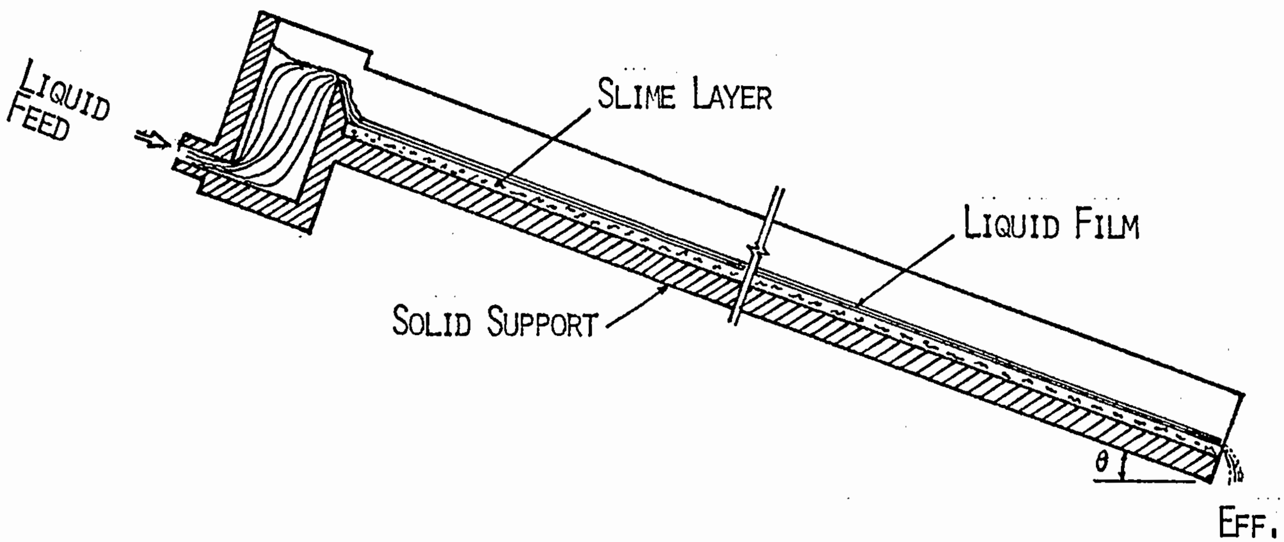


FIGURE 3 - THE BIOLOGICAL FILM REACTOR

$(NH_4)_2SO_4$	varied
N_aHCO_3	300 mg/l
$FeCl_3$	3.5 mg/l
$CaCl_2$	1 ml/l*
$MgSO_4$	1 ml/l*
Phosphate Buffer Solution	1 ml/l*
$Na_2MoO_4 \cdot 2H_2O$	0.5 mg/l
Glucose	10 mg/l
Tap water	100 ml/l
Distilled water	To make up 1 liter

*Note: Standard BOD dilution water nutrient (Standard Methods, 1975)

A boric acid-sodium hydroxide buffer solution was used to maintain a pH of 8.5 ± 0.1, for the optium pH for nitrifiers is in this range (Huang, et al., 1974b).

The analytical methods in this study followed Standard Methods (1975). Four forms of nitrogen constituents, i.e., NH_3-N, NO_2-N NO_3-N, and organic-N were analyzed. The organic nitrogen content of the synthetic water was nil, and the recovery of all other nitrogen forms was good. Since the ammonia-nitrogen oxidation controls the nitrification rate, the ammonia-nitrogen analysis was the primary measurement of nitrification efficiency.

Surface Reation Model Study--

According to Equation (8), the M values can be obtained by varying the Reynolds Numbers, i.e., by varying the hydraulic loadings. By using two sets of inclined fixed film reactors and reproducing the experiment at different times, the M values were calculated for the data obtained and plotted in Figure 4. Figure 4 reveals that the surface reaction model of nitrification in a biological film is pseudo-homogeneous.

As indicated previously, recirculation of the effluent has a beneficial effect on removal of total soluble substrate in a pseudo-homogeneous film flow reactor; therefore, a side study for recirculation effect was also performed. The results for recirculation ratios of 0.5, 1.0, 1.5, and 2.0 all showed that recirculation improved the nitrification in the fixed film reactor as shown in Figrue 5. Improving the efficiency of the system by recirculation also supports the conclusion that the correct surface reaction model for the nitrification in the fixed film process is of the pseudo-homogeneous type.

The design equation for a fixed film nitrification process can be derived by putting Equation (5) in another form:

$$S_o - S_e = ak/(Q)^{2/3} \quad \text{.......................}(9)$$

in which a = constant reflecting the characteristics of the packing media. Equation (9) indicates that the nitrification efficiency in a fixed film

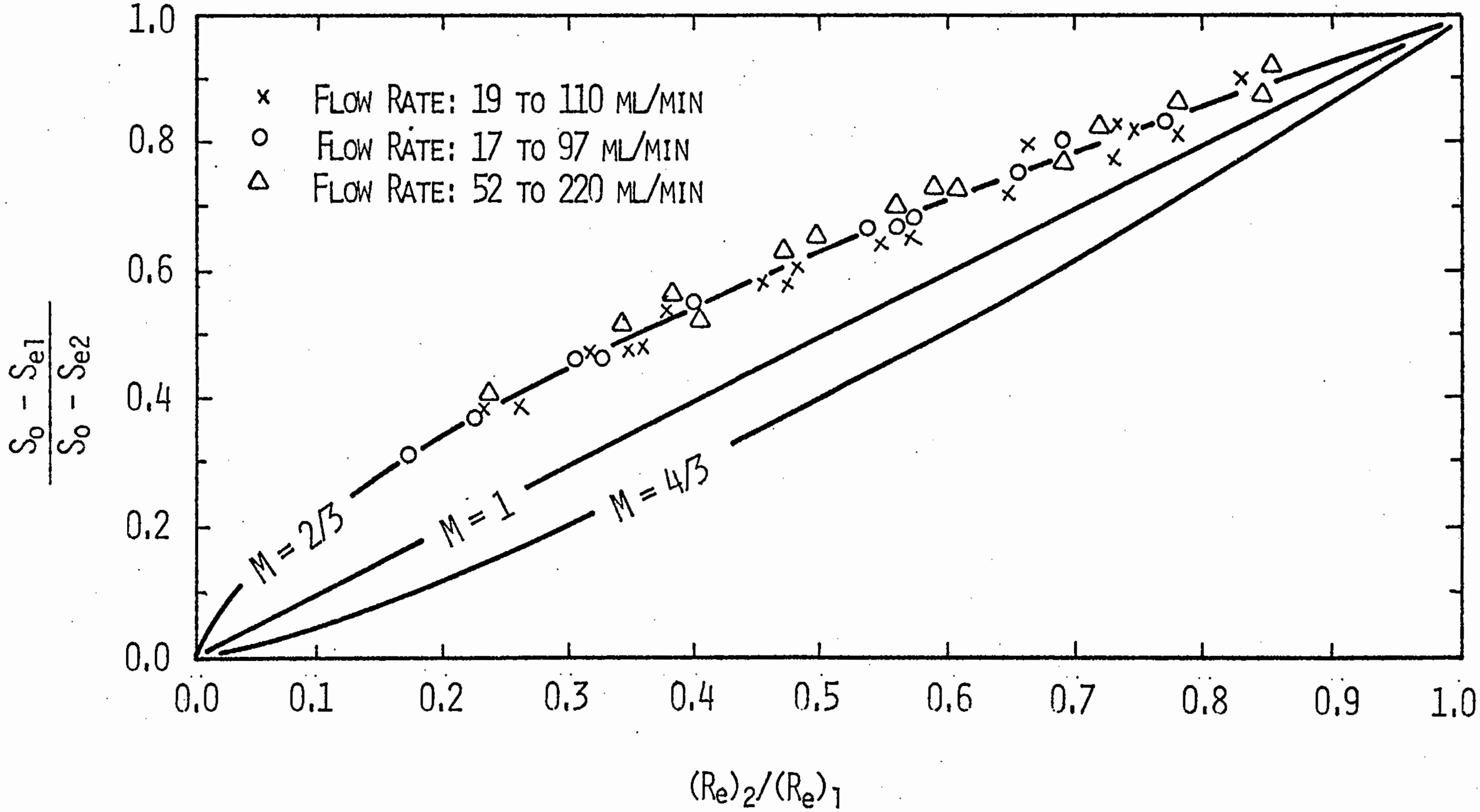


FIGURE 4 - REYNOLDS NUMBER RESPONSE STUDIES

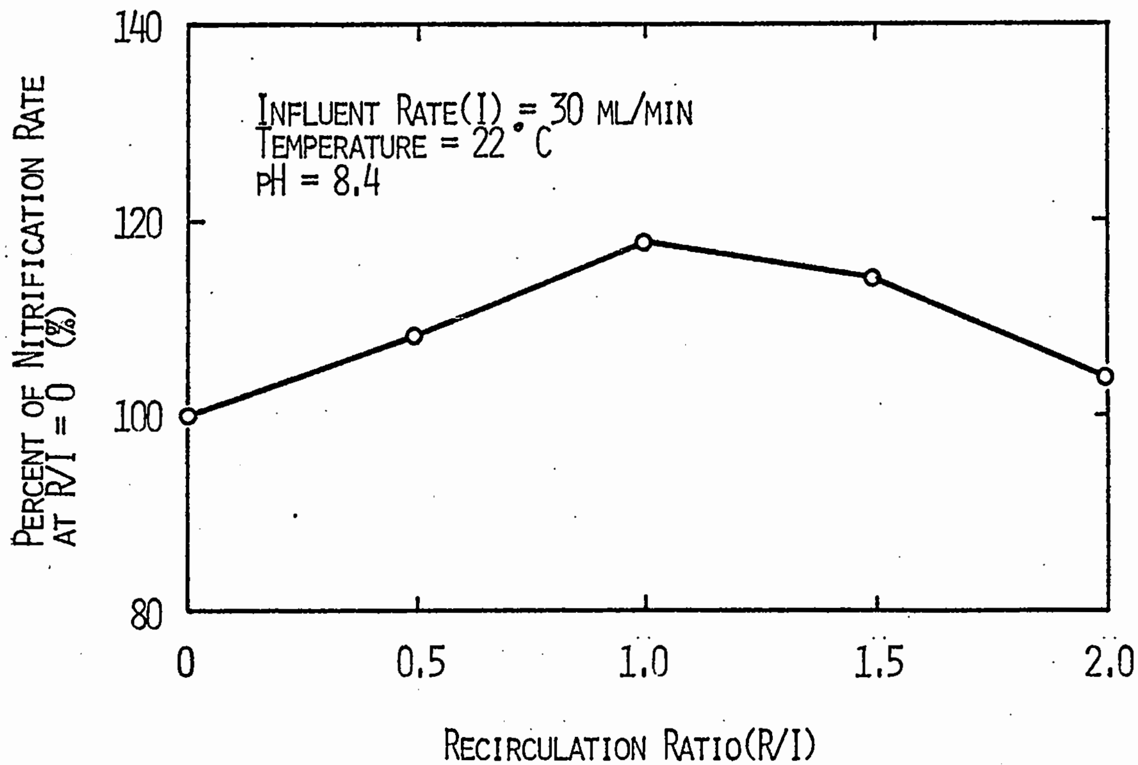


FIGURE 5 - RECIRCULATION EFFECT ON THE NITRIFICATION RATE

process depends on the characteristics of the packing media used and the reciprocal of the two-third power of the hydraulic loading. Therefore, the hydraulic loading and the packing media should be properly selected during the treatment plant design.

STUDY II: ROTATING BIOLOGICAL CONTACTOR STUDY

Conceptually, RBC units are similar to other kinds of fixed film biological treatment systems such as trickling filters. However, rotation of the RBC media provides a more positive supply of oxygen and nutrients to the bacteria than trickling filters.

In organic substrate removal studies of RBC systems, an oxygen-substrate diffusion limited situation are often assumed for organic substrate removal kinetics derivation (Famularo, et al., 1978, Hansford, et al., 1978, and Schroeder, 1976). In RBC nitrification process kinetics, however, much less work has been done. The pseudo-homogeneous model found from the stationary fixed film process, as described in STUDY I, could be applied to the RBC system.

1. Mass Balance in RBC Unit

If the pseudo-homogeneous surface reaction model is applied to an RBC system (see Figure 6(A)), the mass nitrification rate per disc face is then

$$M_Z = r\delta_N A_1 \quad \text{(10)}$$

In which M_Z = mass nitrification rate per unit time; r = reaction rate; δ_N = liquid film thickness on the slime layer at rotating speed N rpm; and A_1 = contact surface area per disc face.

Because the discs are closely spaced, a continuous nitrification rate function can be made which is analogous to a plug-flow reaction through a whole shaft of the RBC unit. Therefore,

$$V_1 \frac{dS}{dt} = M_Z \quad \text{(11)}$$

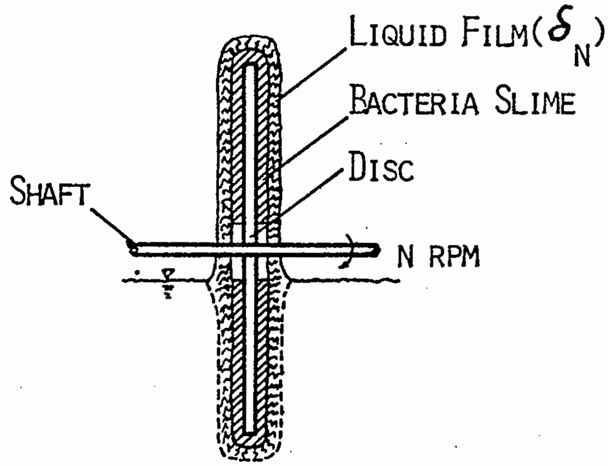
in which V_1 = liquid volume per disc face; S = NH_3-N concentration; and t = contact time.

The RBC nitrification reaction order found to follow a zero-order reaction by several studies (Murphy, et al., 1973, Torpey, et al., 1973, and Weng, et al., 1974) and by this study, which will be described later. That is,

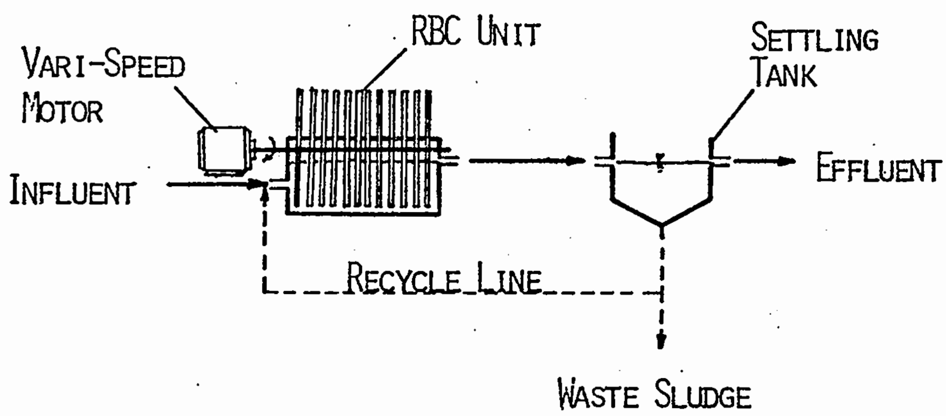
$$r = -k \quad \text{(3)}$$

The liquid film thickness, δ_N, can be defined as follows (Bintanja, et al., 1975):

$$\delta_N = K_1 \left(\frac{\mu N R}{\rho g}\right)^{1/2} \quad \text{(12)}$$



(A) SCHEMATIC CROSS-SECTIONAL VIEW OF AN RBC DISC



(B) FLOW DIAGRAM OF THE RBC TEST UNIT

FIGURE 6 - SCHEMATIC FLOW DIAGRAM OF AN RBC SYSTEM

in which δ_N = liquid film thickness at rotating speed N rpm (μm); K_1 = constant; μ = viscosity of wastewater (kg mass)/(m)(sec); N = rotating speed (rpm); R = radius of disc at the average tangential velocity point (m); ρ = density of wastewater (kg/m^3); and g = acceleration due to gravity (m/sec^2).

The constant K_1 calculated by Bintanja, et al. (1975) is 0.93.

At a certain wastewater temperature, and a certain disc size, Equation (12) can be expressed as

$$\delta_N = K_2 \cdot N^{1/2} \quad \text{.......... (13)}$$

where K_2 is a constant.

Substituting Equation (3), (10), and (13) into Equation (11), and integrating over the entire RBC unit, yields

$$V\,(S_o - S_e) = K \cdot N^{1/2}\,A\,t \quad \text{.......... (14)}$$

in which V = liquid volume in RBC unit; S_o, S_e = influent and effluent NH_3-N concentration, respectively; K = k K_2 = reaction rate constant; A = total contact surface area; and t = contact time.

Since the contact time is close to the hydraulic retention time,

$$t = V/Q \quad \text{.......... (15)}$$

where Q is the flow rate. Substituting Equation (15) into Equation (14), yields

$$\frac{Q(S_o - S_e)}{A} = K\,N^{1/2} \quad \text{.......... (16)}$$

Equation (16) indicates that NH_3-N mass removal rate per unit area is a function of the square root of the rotating speed.

2. Method and Experimental Results

A bench scale RBC unit which consists of 3-3/4 in. diameter discs was used for this study. The flow direction was parallel to the disc shaft. The schematic flow diagram is shown in Figure 6(B). Recycling was employed only during the Recirculation Study mode. The synthetic wastewater used was similar to the one used in STUDY I. The analytical procedures also followed Standard Methods (1975).

a. Reaction Rate Study--

The reaction rate study was performed by feeding the matured RBC unit with different initial NH_3-N concentrations at a constant flow rate and a constant rotating speed. The results for pH = 8.5 and pH = 7.2 are plotted in Figure 7. The test for pH = 7.2 was done in few days because the nitrifying bacteria will acclimate to a lower pH as shown in Figure 9 (Huang, et al., 1974b).

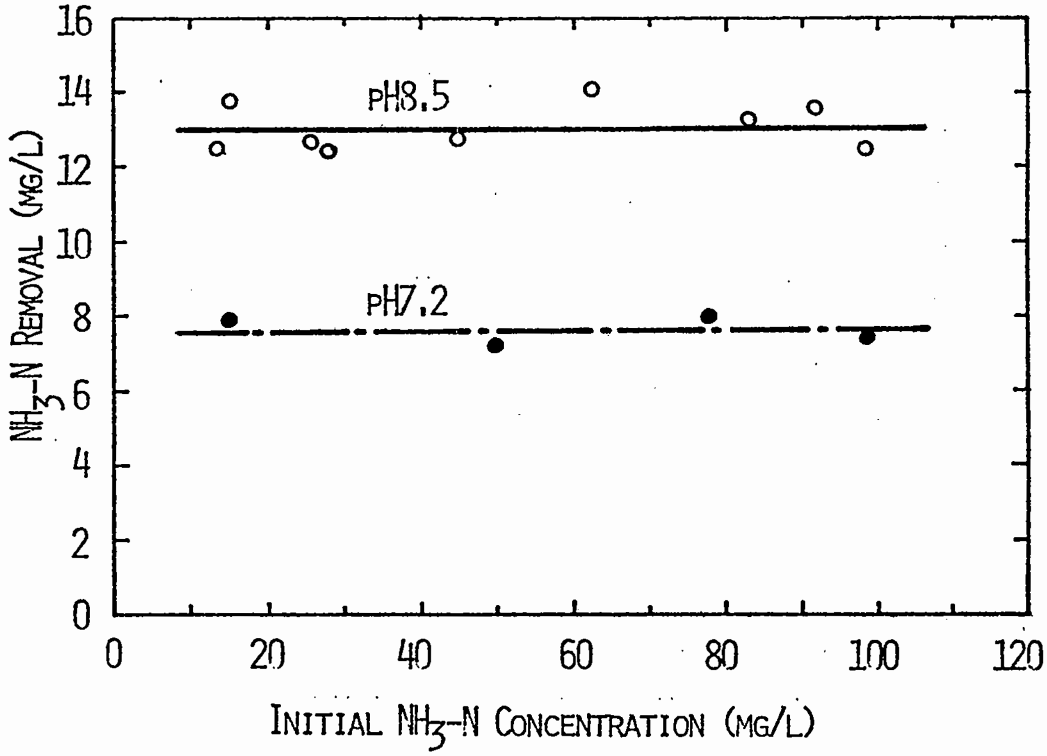


FIGURE 7- INITIAL NH_3-N CONCENTRATION VS. NH_3-N REMOVAL

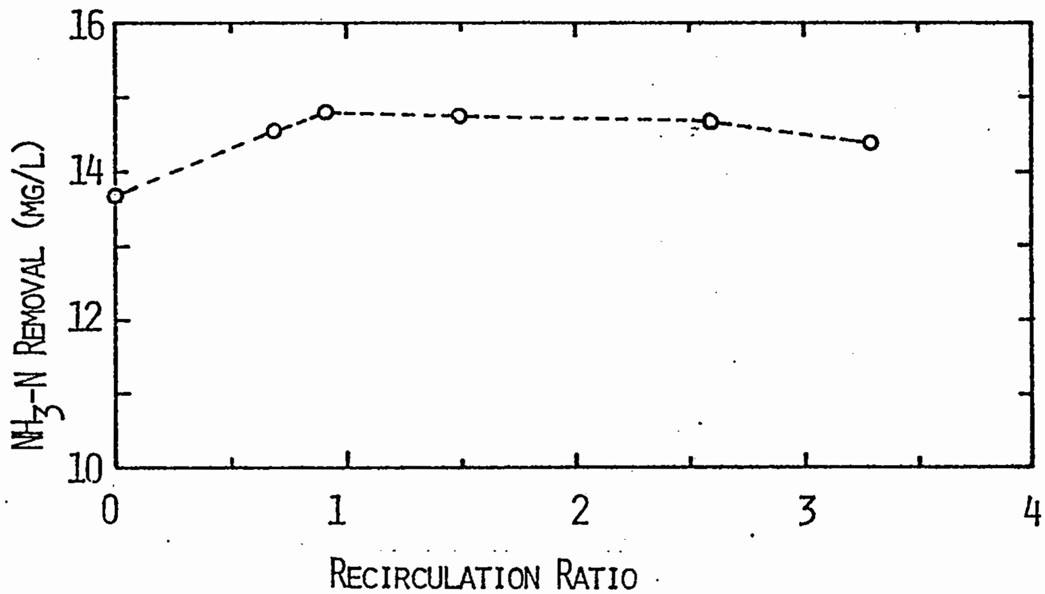


FIGURE 8 - RECIRCULATION RATIO VS. NH_3-N REMOVAL

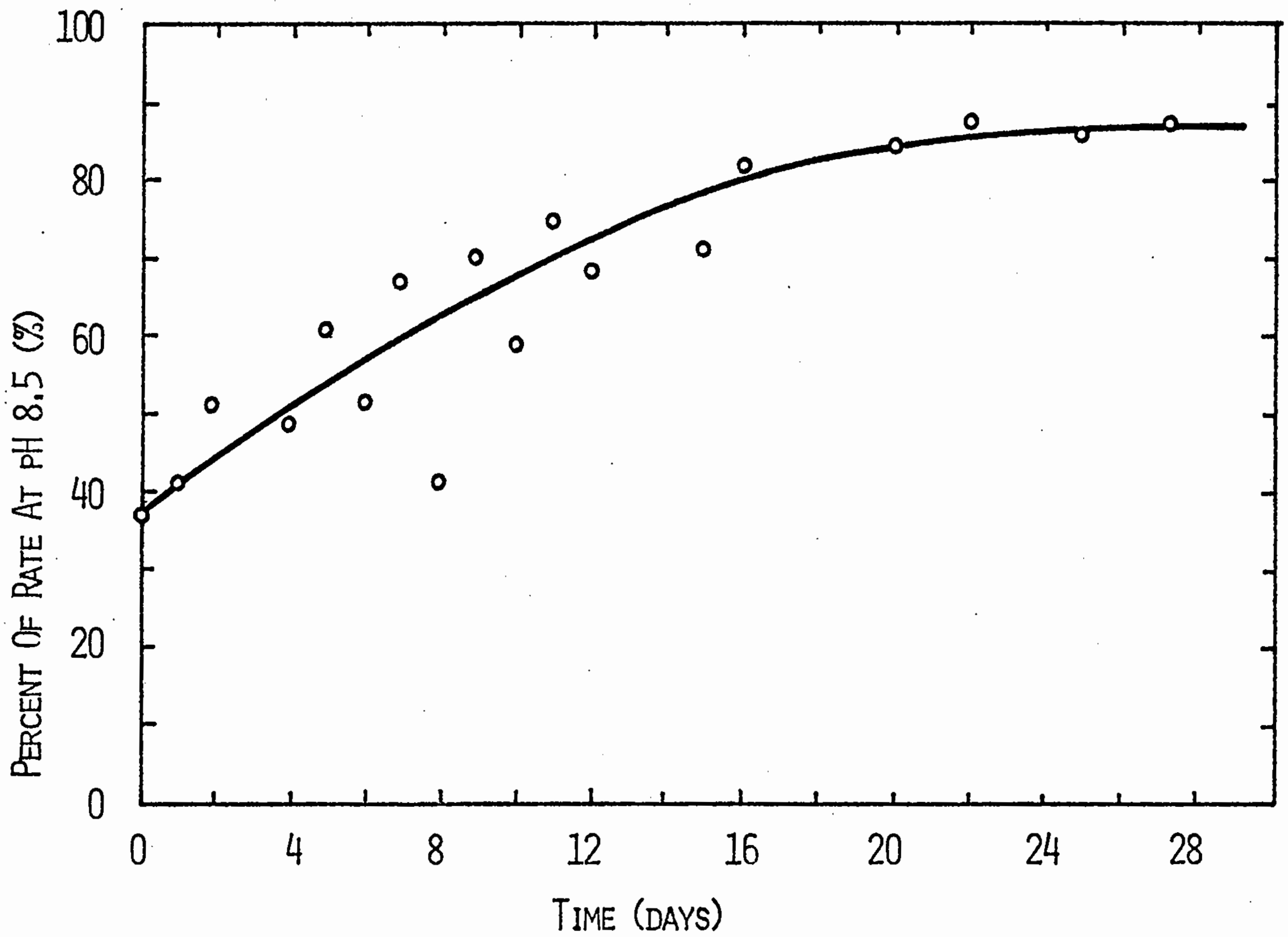


FIGURE 9 - ADAPTION TO pH 6.6 LEVEL AS A FUNCTION OF TIME AFTER LOWERING pH FROM 8.5

Figure 7 indicates that the NH_3-N removal process follows a zero-order reaction.

b. Recirculation Study--

A recirculation study was performed by recycling the underflow from the settling tank. The underflow MLSS concentration was approximately 600 mg/l. The results are shown in Figure 8. It can be seen from Figure 8 that the NH_3-N removal rate increased up to 16 percent at a recirculation ratio of 0.9.

c. Rotating Speed Study--

During this mode of study, the influent NH_3-N concentrations and hydraulic loadings were kept in four groups:

Group	Flow Rate (ml/min)	Initial NH_3-N Conc. (mg/l)
I	12 ± 2	33 ± 2
II	20 ± 2	43 ± 2 and 52 ± 3
III	30 ± 2	47 ± 3
IV	35 ± 2	54 ± 4

Rotating speeds were varied from 6.5 rpm to 128 rpm. Temperature was room temperature and wastewater pH was 8.5 ± 0.2. The results are plotted in Figure 10.

Figure 10 reveals that the NH_3-N mass removal rate per unit area increases proportional to the square root of the rotating speed up to a point (Point B in Figure 10), and then levels off. The rotating speed at Point B is approximately 75 rpm, which corresponds to a peripheral velocity of 1.23 ft/sec. This indicates that the pseudo-homogeneous model is also valid for the RBC unit used in this study with the rotating speeds from 6.5 rpm to 75 rpm.

The liquid film thickness at 75 rpm is 72 μm as calculated from Equation (12). According to the conceptive geometry of the pseudo-homogeneous model (See Figure 1(B)), a liquid film thickness of 72 μm may be close to the "effective" slime thickness. And this is probably why the NH_3-N mass removal rate levelled-off at a rotating speed of 75 rpm in this study.

DISCUSSION

According to STUDY I and STUDY II, the pseudo-homogeneous surface reaction model derived from a stationary inclined nitrifying fixed film also applies to an RBC unit. That is, the NH_3-N mass removal rate per unit area in proportional to the 0.5 power of the rotating speed up to 75 rpm, or equivalent to a peripheral velocity of 1.23 ft/sec in the bench scale RBC unit used. The NH_3-N mass removal rate per unit area then levelled off at a higher rotating speed.

Weng, et al. (1974) used a 6-in. diameter RBC unit for their

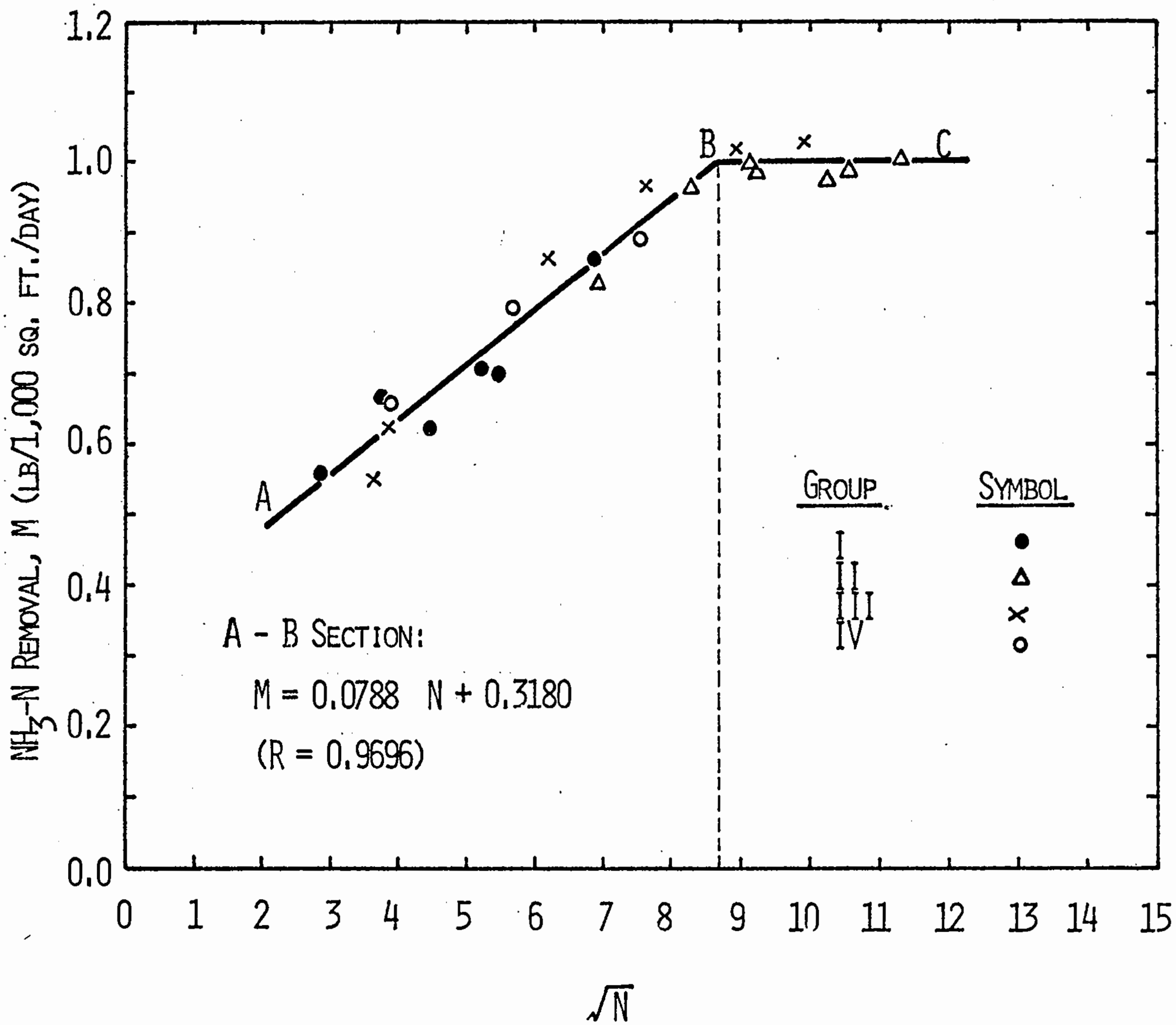


FIGURE 10 - SQUARE ROOT OF ROTATING SPEED VS. NH_3-N REMOVAL

nitrification study. The rotating speeds tested were from 10.5 rpm to 42 rpm, or a peripheral velocity of from 0.27 ft/sec to 1.10 ft/sec. Their results indicated that the NH_3-N removal is a function of the 0.53 power of the rotating speed.

The liquid film thickness on an RBC disc is a function of the square root of the rotating speed as shown in Equation (12). That is, the liquid film thickness increases if the RBC rotating speed increases. According to the conceptive geometry of a pseudo-homogeneous model as shown in Figure 1(B), increasing the liquid film thickness will also increase the NH_3-N removal until the liquid film thickness approaches to the "effective" slime thickness. Therefore, the NH_3-N removal level-off point on Figure 10 may be used to estimate the "effective" slime thickness on a nitrifying RBC unit. The RBC unit used in this study has an "effective" slime thickness of 72 μm estimated from this method. The thickness of the effective microbial film in carbonaceous substrate removal system has been reported with the range from 70 μm to 200 μm (Kornegay & Andrews, 1969, and Tomlinson & Snaddon, 1966).

According to one RBC manufacturer (Autotrol Corp., 1978), the nitrification RBC unit design should be based on hydraulic loading and the desired effluent NH_3-N concentration. The peripheral velocity is recommended at 1.0 ft/sec. This peripheral velocity corresponds to 1.6 rpm in a 12-ft diameter RBC unit. According to this study, the maximum NH_3-N removal will be achieved at a peripheral velocity of 1.23 ft/sec, or equivalent to 2.0 rpm in a 12-ft diameter RBC unit. Of course, the differences in RBC configurations and in wastewater characteristics may show a different optimum rotating speed for NH_3-N removal. Also, both the peripheral velocity and the rotating speed should be used for pilot study scaleup factors as reported by Friedman, et al. (1979). Therefore, a pilot plant test may be required for a nitrification RBC plant design. The NH_3-N mass removal rate and power consumption data at different rotating speeds should be obtained to evaluate the economic tradeoffs so that an optimum system can be designed.

CONCLUSIONS

Based on the results of these two studies, some conclusions can be drawn as follows:

1. The nitrification process follows a zero-order reaction both in stationary fixed film system and in rotating fixed film system.

2. The pseudo-homogeneous surface reaction model can be applied to both trickling filter systems and RBC systems. That is, the nitrification efficiency in a trickling filter process depends on the characteristics of the packing media used and the reciprocal of the two-third power of the hydraulic loading; in RBC nitrification system, NH_3-N removal is a function of the square root of the rotating speed up to a certain speed.

3. The active microbial film in a nitrifying RBC unit may be estimated by locating the optimum NH_3-N removal rotating speed and the corresponding liquid film thickness at that rotating speed. The RBC unit used in this study has an active (or effective) slime thickness of approximately 72 μm.

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A KINETIC MODEL FOR TREATMENT OF CHEESE PROCESSING WASTEWATER WITH A ROTATING BIOLOGICAL CONTACTOR

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INTRODUCTION

Rotating Biological Contactors (RBC) have become established as competitive, cost-effective systems for the treatment of biodegradable wastewaters. These units are being used to treat many different types of wastewaters in a multitude of treatment process configurations. Although RBC's are used in many industrial and domestic wastewater treatment schemes, there is incomplete information on the performance capabilities of the RBC. Specifically, the mechanisms of substrate removal have not been fully postulated and verified.

The general objective of this study was to develop a kinetic model which describes the performance of a rotating biological contactor treating cheese processing plant wastewater. The model employs Michaelis-Menton-Monod kinetics and steady state conditions to describe the system performance as a function of temperature and organic loading rate.

To accomplish the above general objectives the following specific objectives were achieved using a four stage, four foot diameter, pilot plant rotating biological contactor (RBC) treating cheese processing plant wastewater.

1. Determine the pilot plant performance at three separate organic loading rates (hydraulic loading rates) under steady state conditions.

2. Develop a Michaelis-Menton-Monod equation which describes the performance of the pilot plant on a stage by stage basis.

3. Determine the kinetic constants associated with the model by stage and with temperature, using the pilot plant performance data.

PREVIOUS INVESTIGATIONS

Initial attempts to model the performance of rotating biological contactors (RBC) were empirical in nature and mainly employed regression analysis [*Hartman* 1965, *Jost* 1969, *Antonie and Welch* 1969, and *Weng and Molof* 1974]. Their efforts generally ignored temperature effects and were not directly related to microbial substrate removal. Substrate kinetic removal equations have been developed by several investigators [*Grieves* 1972, *Hansford et al.* 1976, *Benjes* 1978, *Kornegay and Andrews* 1968, *Kornegay* 1972, *Kornegay* 1975, and *Clark et al.* 1978]. In general these models employed either saturation kinetics [*Monod* 1942] or first order kinetics to describe substrate removal. Usually these equations were limited to a single stage system or treated a multi-stage system as a single unit.

Recent investigations have employed either mass transfer models or have combined mass transfer concepts with substrate removal kinetics to describe RBC performance [*Schroeder* 1976, *Friedman et al.* 1976, and *Famularo et al.* 1978]. These equations have generally been applied in oxygen limited substrate removal situations.

MATERIALS AND METHODS

Wastewater Source

The Cache Valley Dairy Association Plant in Amalga, Utah, served as the site for the research project. Monterey jack, swiss, and cheddar cheese are the principal products of the plant. Approximately 430 farms contribute a total of 327,000 kg/day (720,000 lb/day) of raw milk for the cheese making process.

All of the process wastes from the plant are discharged to an aerated primary lagoon followed by several facultative lagoons. A portion of the final lagoon effluent is recycled back to the plant for use as non-contact cooling water which, after use, is returned to the wastewater treatment facility.

Experimental Apparatus

Wastewater was pumped from the headworks of the treatment facility to a 15.2 cm (6 inch) wide, pilot-scale Hydrasieve screen (Bauer Brothers Division, Combustion Engineering). The Hydrasieve was equipped with a 1.0 mm (.04 inch) screen size opening during the two lower hydraulic loading rates (HLR) in the experimental procedure. At the highest loading rate, a 1.5 mm (.06 inch) screen size opening was utilized. This system removed most of the suspended solids from the wastewater (see Figure 1).

The rotating biological contactor (RBC) (Environmental Systems Division, Geo. A. Hormel & Company) was a four stage, 1.19 m (47 inch) diameter, pilot-scale system. The disc media was type SC (sinusoidal circular) extruded polyethylene. Each stage contained 12 discs, which yielded a total system area of 145 m^2 (1680 ft^2). Flow through the stages was perpendicular to the shaft, or parallel to the disc faces. Flow to successive stages was allowed via externally mounted 5.1 cm (2 inch) pipe connections. The discs were kept at approximately 37 percent submergence by adjusting the fourth stage outlet piping.

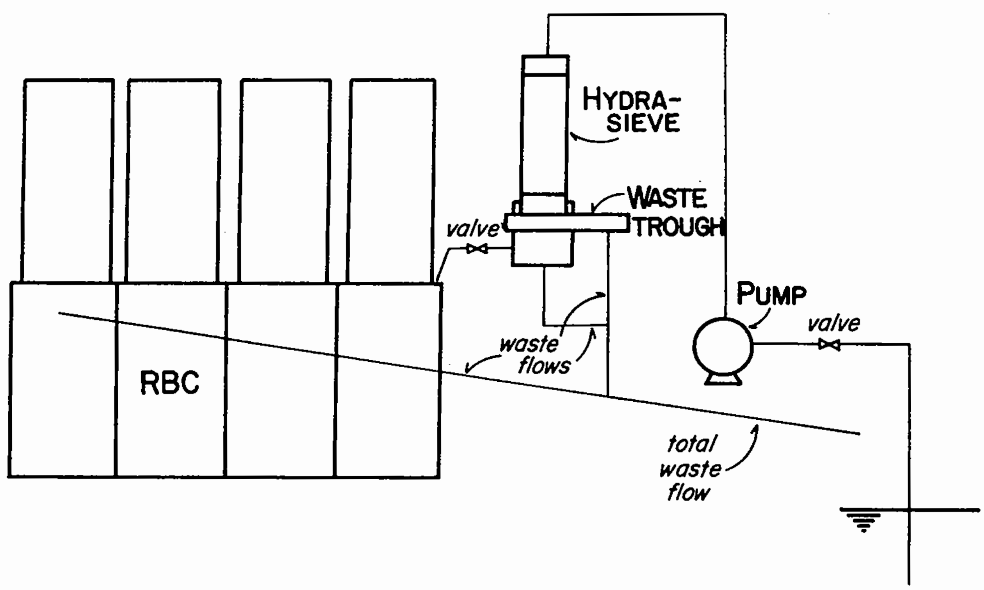

Figure 1. Schematic drawing of the experimental apparatus.

Because of the placement and hydraulics of the RBC unit, the first stage had a liquid depth of 44.4 cm (17.5 inches) while stages two through four had a liquid depth of 45.7 cm (18 inches). The dimensions of the RBC are summarized in Table 1.

The influent and effluent from each stage was monitored on a weekly basis using 24-hour composite samples with analyses being performed for suspended solids, volatile suspended solids, total chemical oxygen demand, soluble chemical oxygen demand, and nutrients. All analyses were conducted according to Standard Methods [*APHA* 1975]. In addition, in-situ measurements of flow, temperature, pH, and dissolved oxygen were conducted.

Table 1. Summary of the pilot-scale RBC dimensions.

Parameter	Value	Units
Number of stages	4	
Number of discs/stage	12	
Area/disc	3.25 (35)	$m^2(ft^2)$
Area/stage	39 (420)	$m^2(ft^2)$
Total area	156	$m^2(ft^2)$
Net first stage volume	.113 (3.99)	$m^3(ft^3)$
Remaining stage volumes (net)	.123 (4.34)	$m^3(ft^3)$
Specific surface area	119.4 (36.4)	$m^2/m^3(ft^2/ft^3)$
Rotation speed	6	rpm
Linear velocity	22.5 (73.9)	m/min (ft/min)

Experimental Procedure

Full-scale testing began in January, 1978, and was completed at the end of June, 1978. The system was operated at three different hydraulic loading rates (HLR) as shown in Table 2. Rotation speed was kept constant at 6 rpm throughout the experiment.

Table 2. Summary of the experimental hydraulic loading rates.

Loading Number	Mean daily flow rate mℓ/min (gpm)	Mean hydraulic loading rate m^3/m^2day (gpd/ft^2)	Mean daily flow ℓ (gal)	Hydraulic retention time (hrs)
1	850 (.22)	0.008 (.19)	1225 (320)	9.5
2	1250 (.33)	0.011 (.28)	1880 (475)	6.5
3	3300 (.87)	0.030 (.75)	4750 (1250)	2.4[1]

[1] Actual measured retention time. Retention times for the first two loading rates were interpolated.

Each HLR was maintained until steady-state conditions were reached. The criterion defining steady-state was a relatively constant effluent soluble COD concentration, *i.e.*, three consecutive weeks within ± 10 percent of a mean concentration. Once steady-state was reached, the system was switched to the next HLR and operated until steady-state was again reached. In this fashion, data were collected for three complete and distinct loading rates.

MODEL DEVELOPMENT

Derivation

The model development is similar to that employed by *Kornegay* [1972] and employs the following assumptions:

1. The liquid portion of each stage is completely mixed.
2. The growth of attached microorganisms is limited by the concentration of a particular nutrient.
3. Organism decay is negligible as compared with microorganism growth.
4. Substrate removal is a saturation function with respect to substrate applied, *i.e.*, Monod kinetics apply.
5. Substrate removal occurs both within the biofilm of the attached microorganisms and in the liquid portion of the RBC.
6. Substrate removal for microorganism maintenance is negligible compared with substrate removal for growth.
7. The attached microorganism growth on the entire disc area is included.
8. Substrate removal in the biofilm occurs through the entire depth of the biofilm.
9. Suspended microorganism growth contributes to substrate removal.
10. Suspended biomass consists entirely of sloughed attached biomass, and there is no growth of suspended microorganisms.
11. The stage total suspended solids concentration consists entirely of suspended biomass.

The model development in this study diverges from *Kornegay* [1972] due to differing assumptions. The first difference is in the use of the entire surface area of the discs in computing the mass of attached microorganisms. In this study, the biofilm developed uniformly across the entire surface area of the discs; therefore, the entire disc area served as a support medium for the microorganisms. *Kornegay* assumed that growth was confined to the wetted submerged segment of the discs.

The second difference is in the use of the entire depth of biofilm in computing the active mass of attached microorganisms. Here it was assumed that the entire depth of biofilm contributed to substrate removal, and thus the total mass of attached growth (dry weight mass) consisted of active microorganisms. *Kornegay* differentiated between total and active biofilm depth, and assumed that substrate removal occurred only in an active portion of the total depth. The remainder of the biofilm was assumed to be diffusion limited.

Thirdly, it is assumed that the suspended organisms contribute to substrate removal, whereas *Kornegay* assumed that substrate removal was confined solely to the active layer of the biofilm.

It is also assumed that the suspended biomass consists entirely of sloughed attached biomass, and there is no growth of suspended microorganisms. Furthermore, it is assumed that the stage total suspended solids (TSS) concentrations consist entirely of suspended biomass.

The total active mass of microorganisms, then, in each stage (X_1) equals the total attached biomass per stage plus the suspended biomass (TSS) in each stage. That is,

$$X_1 = A_w X_f + V X_s \tag{1}$$

where

A_w = total area of disc (L^2)

X_f = total attached biomass concentration (M/L^2)

V = liquid volume of the reactor (L^3)

X_s = stage TSS concentration (M/L^3)

The model was derived in a fashion analogous to suspended growth model derivations. Performing a substrate mass balance on a single stage yields:

$$\begin{pmatrix}\text{net mass rate of}\\ \text{change of substrate}\end{pmatrix} = \begin{pmatrix}\text{mass input}\\ \text{rate}\end{pmatrix} - \begin{pmatrix}\text{mass removal}\\ \text{rate}\end{pmatrix} - \begin{pmatrix}\text{mass output}\\ \text{rate}\end{pmatrix} \tag{2}$$

which may be expressed mathematically as:

$$\left[\frac{d}{dt}\ (VS_1)\right]_{\text{net}} = F_0 S_0 - \left[\frac{d}{dt}\ (VS_1)\right]_{\text{growth}} - F_1 S_1 \tag{3}$$

where

V = liquid volume of the reactor (L^3)

F_0 = influent flow rate (L^3/T)

F_1 = effluent flow rate (L^3/T)

S_0 = influent concentration of the growth-limiting substrate (M/L^3)

S_1 = effluent concentration of the growth-limiting substrate (M/L^3)

t = time (T)

The substrate mass removal rate is represented by microbial growth. Since V, the reactor liquid volume, is a constant, it can be removed from the differentiation. Therefore, equation (3) becomes:

$$\left[V\left(\frac{dS_1}{dt}\right)\right]_{\text{net}} = F_0S_0 - \left[V\left(\frac{dS_1}{dt}\right)\right]_{\text{growth}} - F_1S_1 \tag{4}$$

Employing the concepts for microorganism growth and substrate removal developed by *Monod* [1942], assuming steady state conditions and utilizing Equation 1, Equation 4 can be reduced to:

$$V\left[\left(\frac{dS_1}{dt}\right)\right]_{\text{net}} = F_0S_0 - \frac{\mu(A_wX_f + VX_s)}{Y} - F_1S_1 \tag{5}$$

where

$\hat{\mu}$ = maximum specific growth rate of the microorganisms (T^{-1})

Y = yield = $\frac{dX}{dS}$

K_s = half saturation constant (M/L^3); substrate concentration where $\mu = \frac{1}{2}\hat{\mu}$

Here K_s and μ represent the kinetic constants for the total biomass population, *i.e.*, the attached and the suspended biomass. Since the suspended biomass is assumed to represent only sloughed attached biomass, the suspended biomass specific growth rate μ is assumed to equal μ for the attached biomass. *Bintanja et al.* [1976] made the same assumption in developing a steady-state model to predict carbonaceous removals in an RBC, and achieved excellent fit of the data.

The growth yield Y is a measure of the mass of microorganisms produced per mass of substrate removed. In this study it is defined as:

$$Y_n = \frac{\text{TSS}_n}{(\Delta\text{soluble COD})_n} \tag{6}$$

where

n = stage number.

Although many authors use the volatile suspended solids (VSS) concentration to indicate the viable fraction of the mass of microorganisms produced, in this study TSS concentration was used.

Reaction rate constants show a definite temperature dependency. This is due to the fact that higher temperatures cause more energetic and more frequent collisions between molecules, and thus increase the chance of chemical reaction. This temperature dependency is usually represented by an Arrhenius-type relationship [*Bailey and Ollis* 1977]. Similarly, growth rates of bacterial cultures are proportional to temperature [*Stanier et al.* 1975]. Thus the temperature dependency of the maximum specific growth rate can be incorporated into an Arrhenius-type relationship:

$$\hat{\mu} = Ae^{-E_a/RT} \qquad (7)$$

where

A = frequency factor (T^{-1})

E_a = activation energy (cal/mol)

R = universal gas constant (1.987 cal/mol°K)

T = absolute temperature (°K)

Substituting Equation 7 into Equation 5 results in Equation 8 which describes substrate removal as a function of temperature.

$$F(S_0 - S_1) = \frac{Ae^{-E_a/RT}(A_w X_f + VX_s)}{Y}\left[\frac{S_1}{K_s + S_1}\right] \qquad (8)$$

In this study it is assumed that the system parameters in Equation (8) all change with stage. *Wu and Kao* [1976] and *Clark et al.* [1978] found that the kinetic constants for a two stage suspended growth system, and a four-stage RBC, respectively, all changed with stage. *Kornegay* [1972] assumed that the kinetic constants did not vary with stage; however, *Kornegay* did assume that substrate removal was additive with stage. The concept of summing the substrate removal contributions of each stage can be utilized to describe RBC stage performance [*Kornegay* 1972]. It follows that Equation (8) can be generalized for a multi-stage reactor as:

$$F(S_0 - S_n) = \sum_{i=1}^{n} \frac{A_i e^{-E_{a_i}/RT_i}(A_w X_{f_i} + V_i X_{s_i})}{Y_i}\left(\frac{S_i}{K_{s_i} + S_i}\right) \qquad (9)$$

where

n = total number of stages

i = number of specific stage

Since F is constant through the stages, it can be removed from the summation. Equation (9) represents steady-state substrate removal in a multi-stage reactor, and is the general model used in this study.

Application

The primary intent of the modeling effort in this study is to obtain the kinetic parameters - A, the frequency factor, E_a, the activation energy, and K_s, the half-saturation constant - for each of the four stages in the reactor. One solution technique is to input stage data into a linearized version of Equation (9), and determine the stage kinetic parameters from the resultant plots. However, since there are three unknowns, one of the parameters would have to be assumed in order to use the linearization method.

A more rigorous approach is to use a nonlinear curve-fitting solution technique to determine the optimum values for the kinetic parameters. The "NONLIN" FORTRAN program developed by *Grenney* [1973] is clearly suited for this purpose. Following the User's Manual written by *Cleave* [1978], the stage data from this study is entered into the NONLIN program along with initial estimates from A, E_a, and K_s. The NONLIN program uses an iterative finite-difference technique to converge to the optimum correlation. Thus the "best-fit" values of A, E_a, and K_s for each stage are determined.

Nonlinear equations used in the NONLIN program must be of the form $y = f(x,z, \ldots)$ where y is the dependent variable, and x, z, etc., are the independent variables. Rearranging Equation (9) so that F becomes the single dependent variable, Equation (9) becomes:

$$F = \frac{1}{S_0 - S_n} \sum_{i=1}^{n} \frac{A_i e^{-E_{a_i}/RT_i} (A_w X_{f_i} + V_i X_{s_i})}{Y_i} \left(\frac{S_i}{K_{s_i} + S_i} \right) \qquad (10)$$

Utilizing Equation (10) is not as cumbersome a task as it may appear. This is because the previous stage best-fit solution becomes a constant in solving for the successive stage best-fit. For example, for the first stage, Equation (10) is:

$$F_1 = \frac{1}{S_0 - S_1} \text{ (stage 1 term)} \qquad (11)$$

where F has been subscripted to indicate the stage in question. For simplicity, the entire term in the summation is called the "stage 1 term". For the second stage, Equation (10) becomes:

$$F_{1,2} = \frac{1}{S_1 - S_2} \text{ (stage 1 term)} + \frac{1}{S_0 - S_2} \text{ (stage 2 term)} \qquad (12)$$

From Equation (11),

$$\text{(stage 1 term)} = F_1(S_0 - S_1) \qquad (13)$$

Therefore, Equation (12) becomes:

$$F_{1,2} = F_1 \frac{(S_0 - S_1)}{S_0 - S_2} + \frac{1}{S_0 - S_2} \text{ (stage 2 term)} \qquad (14)$$

Here F_1 is the flow rate predicted from the best-fit of the first stage data. Similarly, for the third stage, Equation (10) becomes:

$$F_{1,2,3} = F_1\frac{(S_0-S_1)}{(S_0-S_3)} + F_2\frac{(S_0-S_2)}{(S_0-S_3)} + \frac{1}{(S_0-S_3)}\ (\text{stage 3 term}) \tag{15}$$

where F_2 is the predicted flow rate contribution of the second stage, and is found from equation (14):

$$F_{1,2} - F_1\,\frac{(S_0-S_1)}{(S_0-S_2)} = \frac{1}{(S_0-S_2)}\ (\text{stage 2 term}) \tag{16}$$

Therefore,

$$F_2 = F_{1,2} - F_1\,\frac{(S_0-S_1)}{(S_0-S_2)} \tag{17}$$

For the fourth stage, Equation (10) becomes:

$$F_{1,2,3,4} = F_1\frac{(S_0-S_1)}{(S_0-S_4)} + F_2\frac{(S_0-S_2)}{(S_0-S_4)} + F_3\frac{(S_0-S_3)}{(S_0-S_4)} + \frac{1}{(S_0-S_4)}\ (\text{stage 4 term}) \tag{18}$$

where F_3 is found in the manner analogous to the above solution for F_2.

In general, for $n \geq 2$, Equation (10) becomes:

$$F_{2,3,\cdots n} = -\frac{1}{(S_0-S_n)}\sum_{i=1}^{n-1} F_i(S_0-S_i) + \frac{1}{(S_0-S_n)}\ (\text{stage } n \text{ term}) \tag{19}$$

Equation (19) serves as the general form of the model incorporated into the NONLIN program for stages 2 through 4. Equation (11) serves as the simplified form for the first stage.

RESULTS AND DISCUSSION

Steady-State Data

In this study, steady-state was defined as non-time-varying effluent soluble COD concentration. The criterion defining steady-state was three consecutive weeks of effluent soluble COD concentrations with values between ± 10 percent of their mean concentration. The weeks chosen to represent steady-state for each of the three loading conditions are summarized in Table 3.

The three-week periods chosen represented the best steady-state approximations for their respective loading conditions. However, the

Table 3. Summary of the steady-state parameter values entered into the summation model (Equation 9).

Hydraulic loading rate, mℓ/min	Stage	Influent soluble COD, mg/ℓ (S_0)	Stage effluent soluble COD, mg/ℓ (S_n)	Stage temperature °K (T)	Attached biomass g/m² X_f	Suspended biomass TSS, mg/ℓ X_s	Growth yield Y	Flow rate m³/day F
850	1	357.2	98.8	295.3	74.0	115.7	.45	1.22
850		389.8	101.8	295.3	74.0	365.0	1.27	1.22
850		506.8	76.8	295.3	74.0	165.3	.38	1.22
1250		269.7	81.3	300.6	66.5	179.7	.95	1.80
1250		502.2	81.8	300.6	66.5	636.3	1.51	1.80
1250		469.0	76.9	300.6	66.5	291.1	.74	1.80
3300		1263.8	496.6	305.8	88.2	245.1	.32	4.75
3300		464.6	330.9	305.8	88.2	206.8	1.55	4.75
3300		855.0	461.7	305.8	88.2	183.8	.47	4.75
850	2	357.2	76.0	294.6	31.1	154.6	.51	1.22
850		389.8	78.7	294.6	31.1	219.9	.61	1.22
850		506.8	65.3	294.6	31.1	96.3	.21	1.22
1250		269.7	63.2	299.1	36.6	132.3	.56	1.80
1250		502.2	63.2	299.1	36.6	339.7	.71	1.80
1250		469.0	66.0	299.1	36.6	302.0	.73	1.80
3300		1263.8	453.9	304.4	69.2	447.6	.53	4.75
3300		464.6	264.0	304.4	69.2	437.5	1.66	4.75
3300		855.0	313.5	304.4	69.2	384.8	.58	4.75
850	3	357.2	64.6	293.9	16.9	196.0	.64	1.22
850		389.8	69.1	293.9	16.9	137.5	.37	1.22
850		506.8	48.0	293.9	16.9	63.5	.14	1.22
1250		269.7	79.6	297.4	24.0	79.6	.35	1.80
1250		502.2	184.3	297.4	24.0	184.3	.40	1.80
1250		469.0	106.0	297.4	24.0	106.0	.22	1.80
3300		1263.8	368.3	302.9	53.5	368.3	.36	4.75
3300		464.6	337.9	302.9	53.5	337.9	.96	4.75
3300		855.0	285.9	302.9	53.5	285.9	.47	4.75
850	4	357.2	57.0	293.2	14.7	101.0	.30	1.22
850		389.8	61.4	293.2	14.7	131.5	.35	1.22
850		506.8	55.6	293.2	14.7	482.7	1.05	1.22
1250		269.7	55.8	295.9	22.0	259.6	1.21	1.80
1250		502.2	39.0	295.9	22.0	302.0	.54	1.80
1250		469.0	46.7	295.9	22.0	238.7	.53	1.80
3300		1263.8	181.6	301.5	44.0	615.0	.34	4.75
3300		464.6	126.7	301.5	44.0	684.0	.76	4.75
3300		855.0	167.2	301.5	44.0	299.7	.16	4.75

criterion defining steady-state was not met by the 1250 and 3300 mℓ/min loading rates. For the 1250 mℓ/min HLR, the upper and lower boundaries of the three data points deviated 18.2 and -13.5 percent from the mean, respectively. For the 3300 mℓ/min HLR, the upper and lower boundaries deviated 14.6 and -20.0 percent from the mean, respectively.

The other parameters required to solve the model equation (Equation 9) were obtained from the data of these nine weeks. The actual data values of RBC influent soluble COD (S_0), stage effluent soluble COD (S_n), stage temperature (T), attached biomass (X_f), suspended biomass (X_s), growth yield (Y), and flow rate (F) were used in the model (Table 3). The value entered into the model for the total disc area per stage (A_w) was 39.0 m^2. The stage volumes (V_i) used were .113 m^3 for stage one, and .123 m^3 for stages 2, 3, and 4.

The equation for growth yield was modified, assuming that the stages cannot be treated as independent reactors in terms of growth yield. Thus the growth yield was assumed to be represented by a summation model, with the growth yield in any particular stage equal to the total system growth yield minus the particular stage growth yield. For the first stage, then, the equation was the same, *i.e.*, growth yield equalled $TSS_n/(\text{ soluble COD})_n$. For stages 2 through 4, Y was defined as:

$$Y_n = \frac{\sum_{i=1}^{n} TSS_i}{\sum_{i=1}^{n} (\Delta\text{soluble COD})_i} - \frac{\sum_{i=1}^{n-1} TSS_i}{\sum_{i=1}^{n-1} (\Delta\text{soluble COD})_i} \qquad (20)$$

where n equals stage number. The resulting values of Y predicted by Equation (20) varied from 0.16 to 1.66.

Model Results

The best-fit solutions for the stage kinetic parameters, subject to the constraints listed, are shown in Table 4. The results in Table 4 show that the frequency factor A is essentially the same for stages 1 and 2, but decreases through stages 3 and 4. The frequency factor is related to the number of chemical reactions in the biomass, and thus shows that the number of reactions is biomass concentration dependent.

The activation energy E_a is a quantitative measure of the energy input needed to cause a reaction to occur. In terms of microorganism growth, it is the heat energy required for substrate removal by the microorganisms. The results of this study show that E_a is independent of stage, since the values for all four stages are essentially constant.

Table 4. Summary of the stage kinetic parameters found using the summation model (Equation 9) with the NONLIN program.

Constraint	Stage	Frequency Factor, A day^{-1}	Activation Energy, E_a cal/mol	Half-Saturation Constant, K_s (mg/ℓ)	Maximum Specific Growth Rate, $\hat{\mu}$, day^{-1} Flow Rate, mℓ/min 850	1250	3300	R^2
None	1	1.5×10^{12}	17780	32	.109	.185	.307	.8283
	2	1.4×10^{12}	17850	186	.084	.132	.223	.8559
	3	$.46 \times 10^{12}$	17810	59	.027	.039	.068	.8404
	4	$.26 \times 10^{12}$	17790	-36	.015	.020	.034	.7636
Fourth Stage K_s fixed at 0.1 mg/ℓ	1	1.5×10^{12}	17780	32	.109	.185	.307	.8283
	2	1.4×10^{12}	17850	186	.084	.132	.223	.8559
	3	$.46 \times 10^{12}$	17810	59	.027	.039	.068	.8404
	4	1.6×10^{12}	18510	0.1	.027	.036	.064	.7404
Stages 3 and 4 lumped together as one stage	1	1.5×10^{12}	17780	32	.109	.185	.307	.8283
	2	1.4×10^{12}	17850	186	.084	.132	.223	.8559
	3-4	2.5×10^{12}	17800	10	.145	.202	.349	.7825

The mean value of E_a for all four stages is 17807.5 cal/mol (17.81 kcal/mol), and the standard deviation is 30.95.

Murphy et al. [1977] found A, the frequency factor, to be 9.45 × 10 × hr^{-1} (2.27 × 10^{12} day^{-1}), and E_a to be 13900 cal/mol in an RBC pilot-scale nitrification scheme. *Wong-Chong and Loehr* [1975], in a chemostat study of nitrification, found E_a for the ammonia oxidation step to be between 16.0 and 21.6 kcal/mol, depending on the pH in the chemostat. The values of E_a for nitrite oxidation were similarly pH dependent, and ranged from 14.0 to 39.6 kcal/mol.

The values of E_a found in this study are in the range reported by *Wong-Chong and Loehr*, and the values of A are of the same order of magnitude as those reported by *Murphy et al.* This may indicate that the activation energy E_a and the frequency factor A are independent of the type of substrate - carbonaceous versus nitrogenous - used by the microorganisms. Corresponding values of E_a and A for carbonaceous removal systems could not be found in the literature.

The values for K_s in Table 4 are listed as whole numbers because of the tremendous variability of these constants in mixed cultures. The first stage value of K_s indicates an abundance of usable organic carbon in the substrate. Thus one-half the maximum growth occurs at the low concentration of 32 mg/ℓ. The low K_s value also implies that the microbial population is well adapted to the carbonaceous substrate, *i.e.*, assimilation of the compounds in the substrate occurs quite readily.

The K_s value of 186 mg/ℓ in the second stage indicates that the second stage microbial mass has a reduced concentration of usable carbonaceous substrate in its environment, and thus maximum growth requires a much higher "second stage carbonaceous substrate" concentration. The higher K_s concentration in the second stage may also be indicative of greater species diversity, with resulting varying nutritional needs.

The value of 59 mg/ℓ for the third stage K_s indicates that the third stage biomass is more adapted to its available substrate than the second stage biomass, and thus a lower concentration of "third stage substrate" will achieve maximum growth.

The negative K_s in the fourth stage is indicative of model inadequacy beyond the third stage. This appears to be an artifact of the summation model and is not to be considered a true representation of the fourth stage kinetics. The summation model adds a decreasing increment to Equation (9) with each successive stage. Thus a point can be reached where the stage data is best-fit by subtracting it from the previous stage terms. This is what happened at the fourth stage in this study. The values of K_s found in this study show an increase from stage 1 to stage 2, followed by a decrease through stages 3 and 4. A similar trend in K_s values was found by *Clark et al.* [1978] in a four stage pilot scale RBC treating domestic wastewater. The K_s values found by *Clark* and coworkers for the four stages were

431, 546, 32, and 8 mg/ℓ, respectively. The K_s values found in the first two stages were much higher than the corresponding K_s values found in this study. This was probably due to a number of reasons, *e.g.*, different substrate, higher loading conditions, and lower temperatures.

Because Equation (9) yielded a negative K_s in the fourth stage, it was decided to verify whether or not zero-order kinetics actually governed carbonaceous substrate removal in the fourth stage. Therefore, the fourth stage data were entered into the NONLIN program with K_s arbitrarily fixed at 0.1 mg/ℓ. The resulting predictions (Table 4) for A, E_a, and $\hat{\mu}$, the maximum specific growth rate, were inconsistent with the previous stage data. Therefore, it appears that zero-order kinetics did not govern substrate removal in the fourth stage.

Also, since the fourth stage data yielded a negative K_s, the data from stages 3 and 4 were lumped together as one stage and entered in the NONLIN program to determine if these two stages could be treated as one stage. The resulting values for $\hat{\mu}$ (Table 4) were greater than for any previous stage, indicating the inappropriateness of the concept, *i.e.*, the two stages have different kinetic characteristics.

The values for $\hat{\mu}$ in Table 4 show a decrease with stage, which is expected. The highest maximum specific growth rate in the first stage is about nine times the highest maximum specific growth rate in the fourth stage. The $\hat{\mu}$ values have been calculated for each flow rate, since these flow rates occurred at different average temperatures.

The high values for R^2, the correlation coefficient, indicate that the summation model fits the data, and that carbon was in fact the limiting nutrient. The decreased R^2 for the fourth stage is attributed to the inadequacy of the model at that stage.

In comparing the values of $\hat{\mu}$ obtained in this study with literature values, a tremendous discrepancy is noted. The values of $\hat{\mu}$ obtained in this study are an order of magnitude less than published literature values. For example, *Gaudy et al.* [1967] found $\hat{\mu}$ to be 5.10 day^{-1} (BOD_5 basis) for a suspended growth reactor treating skim milk. *Wu and Kao* [1976] found a $\hat{\mu}$ of .91 day^{-1} (BOD_5 basis) in a suspended growth reactor treating yeast waste. *Clark et al.* [1978], using a four-stage pilot scale RBC, found a $\hat{\mu}$ of 4.4 day^{-1} (BOD_5 basis) in the first stage.

The discrepancy appears to be based on two reasons. First of all, although several authors have used COD as the basis for kinetic equations [*Heukelekian et al.* 1951, *Benedek and Horvath* 1967, *Gaudy and Gaudy* 1971, and *Bintanja et al.* 1976], the trend has been to apply BOD_5 as the basis. In this study, the COD removed per stage was probably much greater than the BOD_5 removed per stage because of the high COD:BOD_5 ratio. Therefore, since COD removal is in the denominator of our model equation, this could have severely reduced $\hat{\mu}$.

Secondly, the discrepancy in $\hat{\mu}$ is probably more directly related to the low hydraulic loading rates employed in this study. For example, the hydraulic loading rates in this study varied from .008 to .030 $m^3/m^2 \cdot day$ (.19 to .75 gpd/ft^2). *Clark et al.* [1978] varied HLR's from .044 to .196 $m^3/m^2 \cdot day$ (1.09 to 4.81 gpd/ft^2). This difference alone almost negates the order of magnitude difference in $\hat{\mu}$.

In this study, the hydraulics of the RBC prevented any higher increase in HLR. The .028 $m^3/m^2 \cdot day$ HLR (.75 gpd/ft^2) was the maximum safe loading rate for the chosen rotation speed that would allow each stage to operate without carryover of wastewater into the next stage.

Simulation

For the data simulation, the stage kinetic parameters obtained from the NONLIN solution technique (Table 4) were substituted back into the model. The average influent soluble COD concentration of the three data points used for each HLR was then entered into the model as S_0, along with the average values of the remaining steady state data (Table 3). Stage effluent concentrations were then solved for, and the predicted stage effluent concentration became the influent for the next stage. Thus the model was forced to fit the same data used in generating the kinetic parameters to predict the stage effluent concentrations. This procedure is not a verification of the model, but an exercise to show the model "fit."

The simulation results, using the value of 0.10 mg/ℓ for the fourth stage K_s, are shown in Figures 2-4. As it is seen, the simulations become less accurate with increasing loading rate. For the 850 mℓ/min and the 1250 mℓ/min flow rates (Figures 2 and 3), the assumption of zero-order kinetics in the fourth stage is inaccurate. This is especially noted in the 850 mℓ/min flow rate, where the fourth stage simulation deviates from the actual data. The 3300 mℓ/min HLR (Figure 4) shows a different pattern of stage soluble COD concentration. The steady-state data show a greater range in concentrations than the other two flow rates, and the fourth stage has significant effects on substrate removal. The system reached its maximum biomass production at the 3300 mℓ/min HLR, *i.e.*, saturation was reached and zero-order kinetics (with respect to substrate) governed. Therefore, the simulation shown in Figure 4 appears to accurately represent the data. The downward inflection after stage 3 does correspond to the decrease in fourth stage soluble COD concentration.

The simulations of the steady-state data, lumping stages 3 and 4 together, are shown in Figures 5-7. The resulting curves do not simulate the third and fourth stages, and this is most apparent at the lower concentrations, *i.e.*, the 850 mℓ/min flow rate (Figure 5). The two stages act differently, and lumping them together creates greater inaccuracy in the simulation.

Several possible reasons exist for the poor model simulation at the higher loading rates. They are:

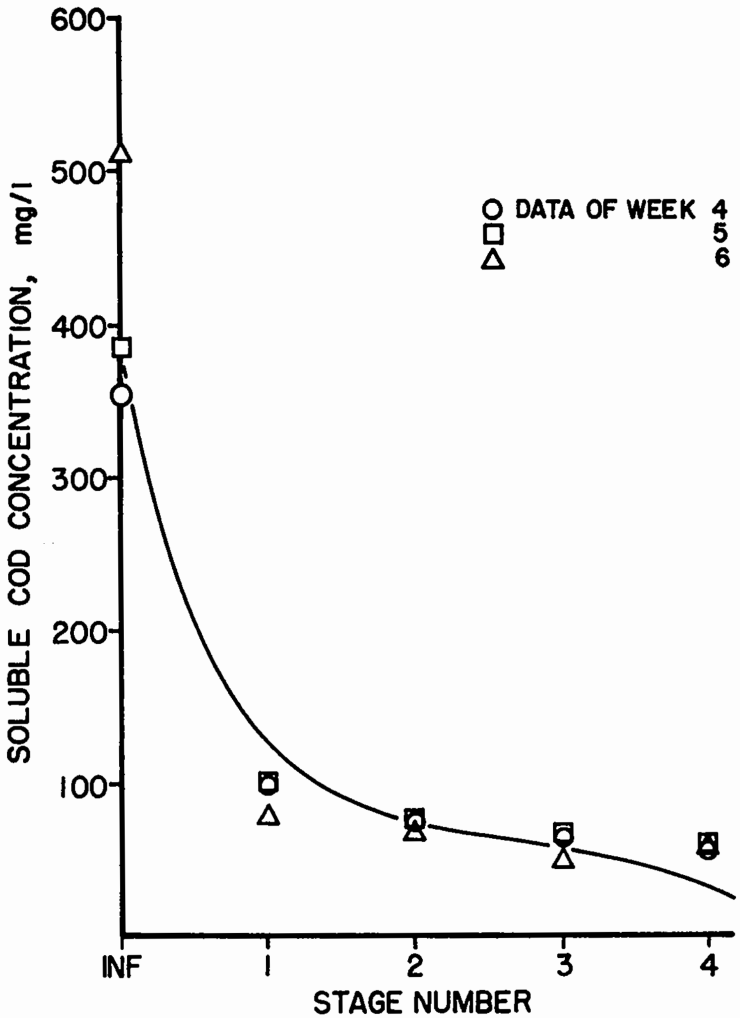

Figure 2. Simulation of the 850 mℓ/min HLR steady-state data, using 0.1 mg/ℓ as the fourth stage value of K_s.

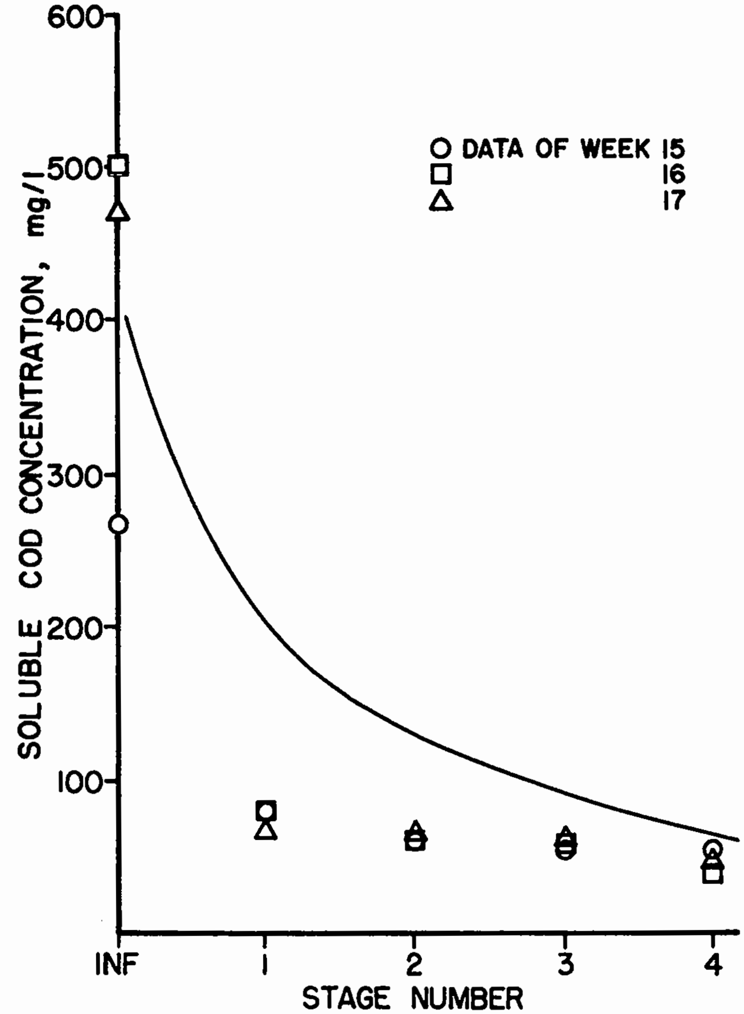

Figure 3. Simulation of the 1250 mℓ/min HLR steady-state data, using 0.1 mg/ℓ as the fourth stage value of K_s.

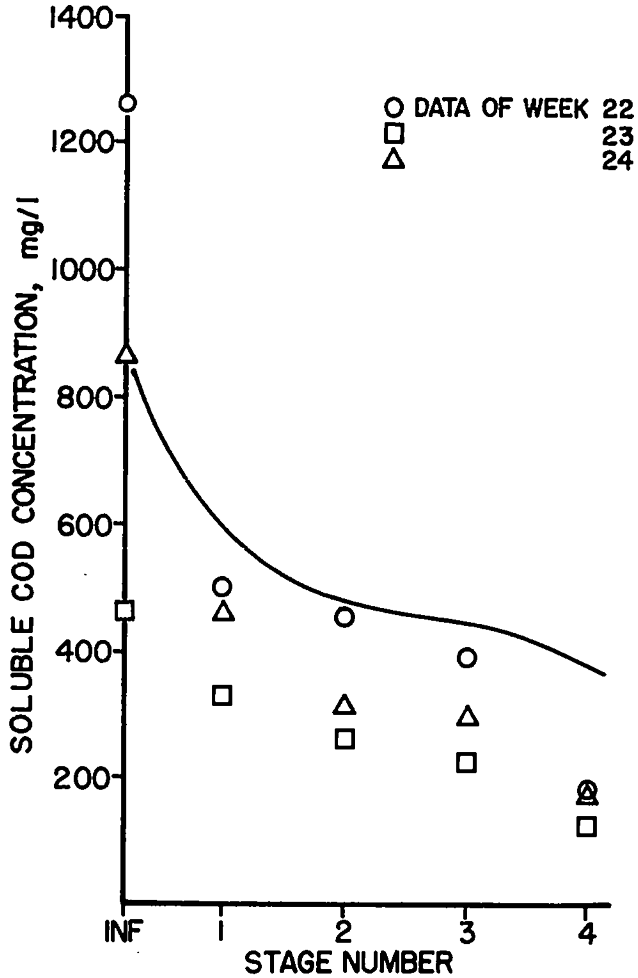

Figure 4. Simulation of the 3300 mℓ/min HLR steady-state data, using 0.1 mg/ℓ as the fourth stage value of K_s.

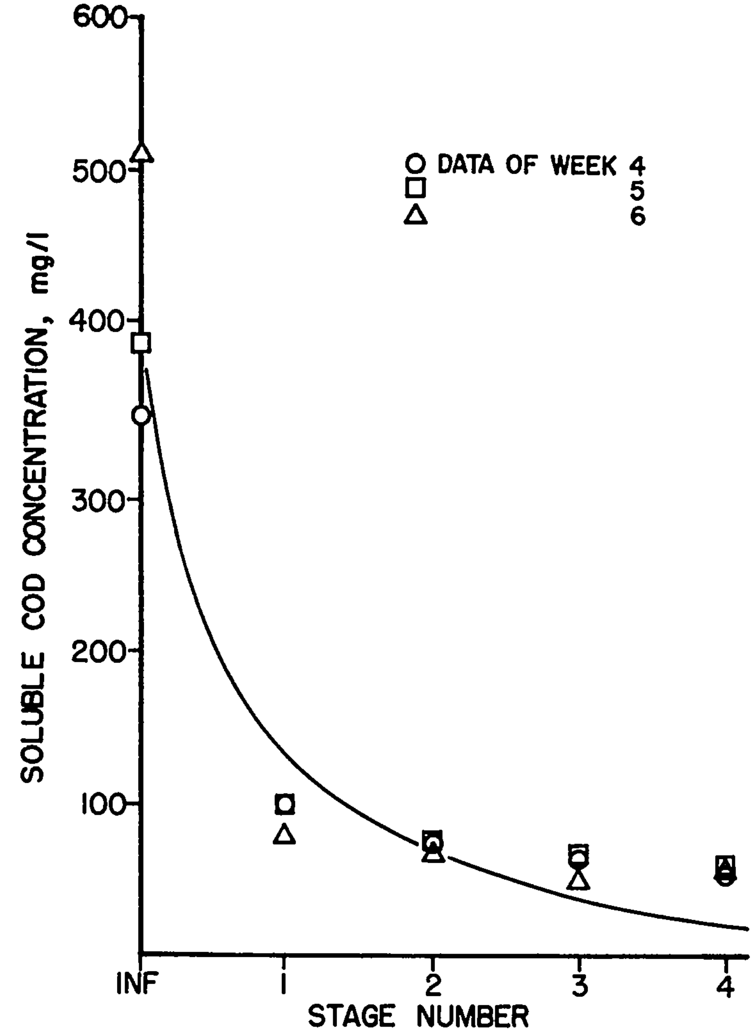

Figure 5. Simulation of the 850 mℓ/min HLR steady-state data, lumping stages 3 and 4.

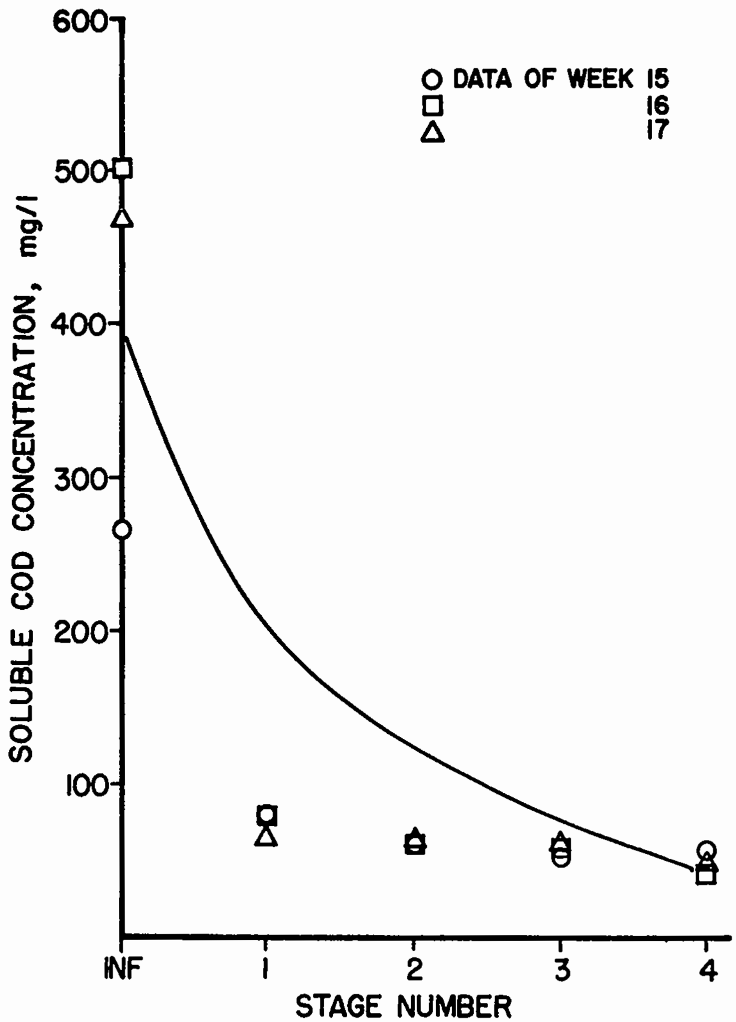

Figure 6. Simulation of the 1250 mℓ/min HLR steady-state data, lumping stages 3 and 4.

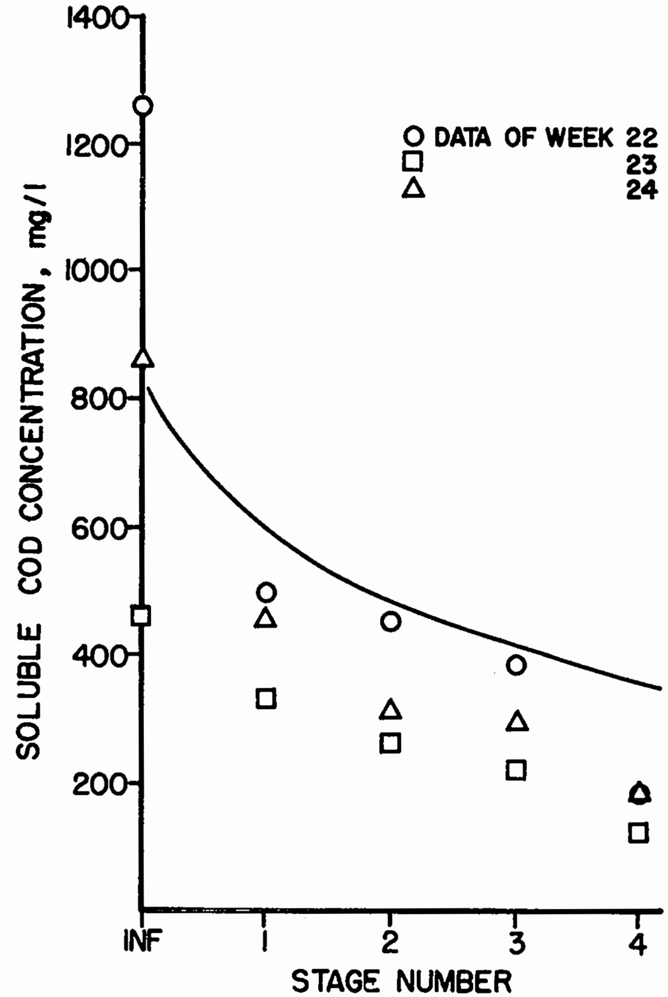

Figure 7. Simulation of the 3300 mℓ/min HLR steady-state data, lumping stages 3 and 4.

1. Soluble COD was chosen to represent the carbonaceous substrate. The kinetic parameters obtained are lower than those that would be obtained using soluble BOD_5 or soluble organic carbon as the carbonaceous substrate. Better correlations may have been achieved using these other parameters.

2. Inaccurate values for the growth yield, Y, may have been used. The steady-state values of growth yield used in the model varied from .16 to 1.66. Although high growth yields were expected due to maximum substrate utilization, the computed values greater than 1.0 probably should not have been used. Growth yields of greater than 1.0 probably occur instantaneously, or for short periods of time, but not at steady-state. The dynamics of sludge production in the RBC are quite complicated, and the attempts in this study may not have successfully established an empirical basis for Y.

3. The entire depth of attached biomass plus the suspended biomass was assumed to affect substrate oxidation. This assumed that diffusion of oxygen or substrate into the biofilm was not limited by temperature, film thickness, dissolved oxygen concentration, etc. In this study, however, low DO concentrations and high wastewater temperatures at the highest HLR could have limited substrate oxidation in the attached biomass. Other authors have used a partial depth of biofilm in modeling substrate removal [*Kornegay and Andrews* 1968 and *Williamson and McCarty* 1976].

The suspended biomass comprised only a small percent of the total mass of the system; therefore, its inclusion in the model probably had negligible effects.

ENGINEERING SIGNIFICANCE

The steady-state model (Equation 9) developed in this study can be utilized as a tool in designing RBC systems for similar applications. This is done by constructing a nomograph of flow rate versus required disc area for desired percent soluble COD removals. The isoconcentration lines (percent soluble COD removal lines) are found by solving Equation (9) for A_w, the disc area, as a function of the flow rate F. The resulting nomograph is shown in Figure 8. Equation (9) is solved in this manner using an average value for each of the kinetic constants.

In constructing this nomograph, the four stage RBC is treated as one stage, so that the disc area needed to obtain a specified percent COD removal at a specified flow rate is the total disc area for the entire multi-stage system. Nomographs could similarly be constructed for each stage.

Figure 8 was constructed based on the average influent soluble COD concentration of 575 mg/ℓ. The influent soluble COD concentration determines the position of the isoconcentration lines, since the required disc area is dependent on ΔCOD removed. Thus a different influent soluble COD concentration would shift the position of the

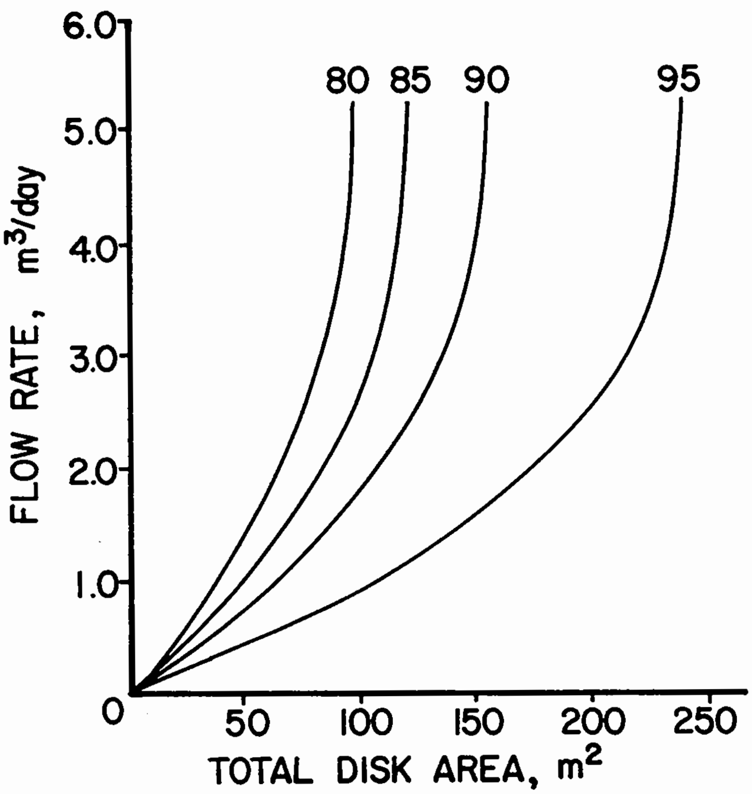

Figure 8. Design nomograph relating flow rate with total disc area for various percent soluble COD removals.

isoconcentration lines. Note that the isoconcentration lines are nonlinear because X_f, the attached biomass concentration, varies with the flow rate F. If X_f did not vary with F, the isoconcentration lines would be linear.

SUMMARY AND CONCLUSIONS

A four stage 1.2 m (47 inch) diameter Rotating Biological Contactor treatment system was operated continuously for six months in treating a cheese processing plant wastewater. The system was operated under flow rates of 850, 1250, and 3300 mℓ/min, and steady-state effluent carbonaceous substrate concentrations were achieved for each of the three flow rates. Treatment process performance was measured via weekly chemical analyses of 24 hour composite samples of the influent, stages, and effluent from the system. System physical parameters were monitored regularly at the field site.

Carbon was determined to be the growth limiting nutrient, and a steady-state Monod kinetics model was developed to simulate carbonaceous removal in the RBC. In the model, the temperature dependency of the maximum specific growth rate $\hat{\mu}$ was represented by an Arrhenius-type relationship. A nonlinear FORTRAN curve-fitting solution technique was used to solve for the stage kinetic parameters E_a, the activation energy, A, the frequency factor, and K_s, the half-saturation constant. The kinetic parameters found from the curve-fitting technique are listed in Table 4.

Leaving the kinetic parameters unbounded and free to "float" during the solution technique, E_a ranged from 17.78 to 17.85 kcal/mol in the four stages. Likewise, A varied from 1.5×10^{12} to $.26 \times 10^{12}$ day^{-1}, and K_s ranged from 186 to -36 mg/ℓ. The corresponding values of $\hat{\mu}$ ranged from .307 to .015 day^{-1}. The high correlations obtained for each stage indicate that the assumptions used in deriving and applying the model are valid.

From the results of this study, the following conclusions can be made:

1. In the lower concentration ranges encountered, soluble COD removal follows first-order kinetics, while at the higher concentration ranges, soluble COD removal is zero-order.

2. Carbon is the growth limiting nutrient in the wastewater, and the average C:N ratio for the steady-state data is 44.6:1.

3. Carbonaceous substrate removal can be best described by the Monod kinetics model of Equation (9).

4. The maximum specific growth rate, $\hat{\mu}$, of the microorganisms decreases with available substrate and temperature, among other factors, from .307 day^{-1} in the first stage to .015 day^{-1} in the fourth stage.

5. The maximum specific growth rate temperature dependency can be described by an Arrhenius-type equation.

6. The activation energy, E_a, is independent of stage, and has a mean value of 17807.5 cal/mol.

7. The half-saturation constant, K_s, changes with stage, and the values for the four stages are 32, 186, 59, and -36 mg/ℓ, respectively.

8. The summation model (Equation 9) becomes inaccurate at the fourth stage, as shown by the negative K_s value in this stage.

9. Using an empirical equation for Y, the growth yield, the steady-state values for Y varied from .16 to 1.66.

10. The percent soluble COD removal was 88.0, 88.0, and 71.1 percent for the 850, 1250, and 3300 mℓ/min flow rates, respectively.

ACKNOWLEDGMENTS

The equipment for the project was provided by the Environmental Systems Division of George A. Hormel and Company, Coon Rapids, Minnesota. Their assistance and encouragement are greatly appreciated. In addition, financial support was provided by the U. S. Environmental Protection Agency Graduate Training Program Grant Number T-900861, Manpower Planning and Training Branch, Washington, D.C.

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PART V: AIR DRIVE AND SUPPLEMENTAL AERATION

AERATED RBC'S - WHAT ARE THE BENEFITS

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INTRODUCTION

Since 1973 Autotrol Corporation has conducted extensive testing, evaluating the impact of the aeration process on the operation of RBC units. Earlier studies were conducted solely under the auspices of the Company at its Corporate laboratories and at test facilities located in the Milwaukee Metropolitan Sewerage Commission's Wastewater Treatment Facility South Shore plant. Other pilot studies have been performed in cooperation with private engineers such as Greeley and Hansen (Portsmouth, VA), Metcalf & Eddy Engineers (Southwest Water Pollution Control Plant, San Francisco, and Jenks and Harrison Inc. (Union Sanitary District, CA). The conclusions derived from these studies have been confirmed in more recent full-scale evaluations comparing the effluents of aerated biological contactors (ABC) and standard mechanically driven rotating biological contactors (RBC) at identical loadings. These tests have been conducted by Greeley & Hansen at the Alexandria, VA Wastewater Treatment Plant and Duncan, Lagnese & Associates at an industrial installation. The results of these studies will be more fully discussed in papers to be presented later in this conference. This presentation summarizes the benefits of subjacent aeration in the RBC process.

RBC - Unaerated Mechanical Drive units

To appreciate the benefits of aerating an RBC unit one should have some familiarity with typical process limitations and problems encountered by standard mechanically driven units. Overloading, anaerobic growth, beggiatoa and lack of operator control comprise the majority of problems and criticisms encountered at RBC facilities experiencing difficulties. Most of these criticisms are interrelated and can be discussed under a few general areas.

a) Overloading

For every RBC system, treating either municipal or industrial waste, there is a limit to the rate at which the waste is applied beyond which first stage RBC units will not be able to maintain aerobic conditions. Once pushed into this anaerobic region there are a number of problems which can develop.

Undesirable organisms will proliferate and retard oxidation of the BOD substrate. The two most common bacterial forms are anaerobic and sulfur oxidizing micro-organisms.

The anaerobic bacteria normally co-exist with aerobic micro-organisms forming the underlayer of the biofilm. Mathematical model studies have shown that the aerobic layer is extremely thin, its depth controlled in early stages by oxygen diffusion and in downstream stages by substrate diffusion. These same models indicate that aerobic layers average only .004 to .008 inches thickness with any greater depth attributed to anaerobic growth. When overloading occurs, the tremendous growths which tend to develop are primarily inactive anaerobic bacteria. This extra "dead load" potentially reduces the life expectancy of the equipment.

However, the anaerobic layer is not inactive. Any sulfate present in the waste is reduced to sulfide under anaerobic conditions. The sulfide becomes a food source for beggiatoa, or sulfur oxidizing bacteria. Although an aerobic bacterial form, this chemotroph is extremely successful in competing with carbon bacteria, but provides no removal of BOD. Further, the growth is gelatinous, effectively blocking oxygen transfer and further inhibiting carbon bacteria.

Even with a correctly loaded system, beggiatoa presents a potential problem if sulfide already exists in the system due to either septic conditions in the collection system or the primary clarifier, an industrial waste discharge having a high sulfide content or any other uncontrollable source of sulfide. In addition to causing reduced BOD removal through the system, beggiatoa growth has poor settling characteristics and will normally create a pinpoint white floc which easily crests the effluent weirs in the secondary clarifier.

b) Lack of Operating Control

This historically has been and continues to be one of the most serious arguments by engineers and operating personnel against the use of RBCs. It is especially strong in small applications where the operator frequently does not have the luxury of removable baffles for some measure of control. It is the unforeseen situation causing a short term problem, for which there is no easily applied correction, which makes this criticism valid. With a strictly mechanically driven RBC. The only potential controls available to the operator are rotational speed adjustment and removable baffles. However, the rotation of the RBC unit provides both aeration and shearing force for controlling biomass thickness.

Rotational speeds are fairly uniform in the industry between 1.4 and 1.7 rpms. These numbers cannot be lowered because of reduced oxygen transfer efficiency. This loss of oxygen transfer efficiency would further complicate an already overloaded first stage, while reduced shearing force would cause further proliferation of anaerobic biomass. Increasing rotational speed is not a viable alternative since an increase to ven 2.25 rpm would cause the RBC process to be non-competitive with other biological treatment systems. This seemingly small adjustmant would double the power consumption, doubling the operating costs. Baffle adjustment for staging alternative does not easily lend itself to short term problems. This option entails draining tankage to remove baffles and would most likely be accomplished after the problem had passed.

ABC - Aerated Biological Contactors

Understanding these major limitations of the traditional RBC unit should give greater appreciation for the following improvements which have been evaluated for ABC systems.

a) Thinner Biomass

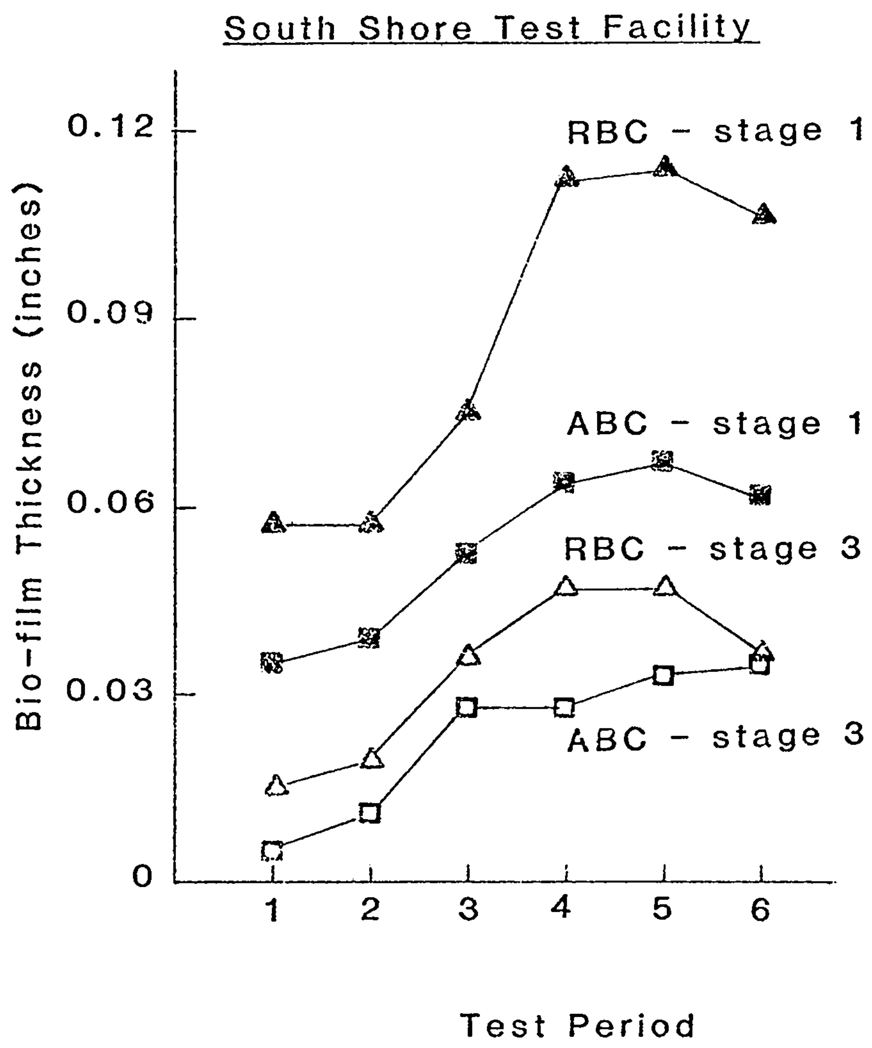
From the earliest research, the major concern of investigators attempting to optimize any biological fixed film treatment system has been control of biomass. Few fixed film processes, if any, offer an operator an "active" method for controlling film thickness.

As we have discussed, RBC systems effectively control the film thickness through a "passive" system of shearing forces created by the rotation of the unit.

Figure 1 presents data from six consecutive test periods of the South Shore test facility comparing biofilm thickness for stages 1 and 3 for RBC and ABC systems each loaded at identical rates. For only the first two test periods for Stage 3 of the ABC system and the first test period for Stage 3 of the RBC system, is the biofilm thin enough that a completely aerobic growth is indicated.

At every point, however, the ABC system is carrying less anaerobic growth than the RBC train. A dramatic disparity in growth thickness is especially obvious in the first stage. The two systems start to approach the same thickness in stage 3, with stage 4 showing little or no difference in biomass thickness.

Figure 1


South Shore Test Facility
0.12
0.09
0.06
0.03
0
Bio-film Thickness (inches)
RBC - stage 1
ABC - stage 1
RBC - stage 3
ABC - stage 3
1
2
3
4
5
6
Test Period

In these later stages the low organic loadings act as a control on the biofilm thickness without any need for further "outside" measures, or operator control.

The major implication of this benefit is that high density media, which provides more surface area per cubic foot, because of the closer spacing of the plastic media layers, can be used earlier in the treatment train without fear of bridging caused by excessive bacterial growth. The thinner biofilm can be attributed to the stripping action of the bubbles rising across the surface of the media. In addition to keeping the biomass to a thickness which can be kept primarily aerobic, the decreased weight carried by the shaft can only increase the life cycle of the equipment.

b) Higher D.O. Concentrations

Since the aeration arc of the RBC unit is augmented by an additional source of oxygen in an ABC system, comparisons of identically loaded side by side evaluations have always shown improved D.O. profiles for the ABC units. Figure 2 indicates such a comparison for the first stages at South Shore. Plotting the change in dissolved oxygen versus organic loading the ABC system, Figure 2 shows that the ABC units have excess oxygen transfer capability above loadings of 5 lbs of soluble BOD per day per 1000 ft^2 surface area. RBC systems' oxygen transfer capability, however, is exceeded above loadings of 3 lbs of soluble BOD per day per 1000 ft^2 of surface area.

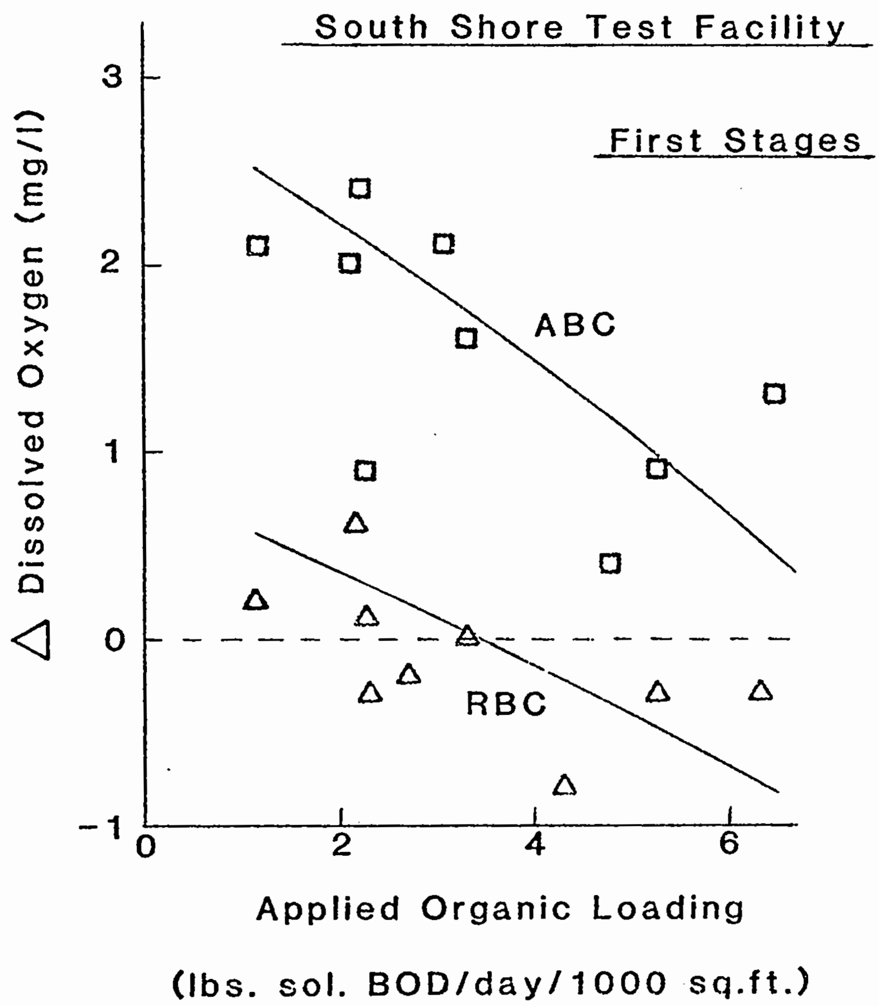
This advantage permits an ABC system to be loaded at higher levels and remain aerobic. This coupled with the higher ORP prevents septic conditions from prevailing and avoids the undesirable effects of an overloaded system. Further, ABC units can more easily control situations where sulfides already exist in the waste prior to entering biological treatment. This has been demonstrated in full-scale studies at Alexandria, VA and in pilot plant studies at San Fransicso. In both cases sulfide content of the influent stream is such that an RBC system during summer operation cannot meet effluent criteria without drastically lowering the organic loading to prevent beggiatoa poliferation.

Metcalf & Eddy recommended, in their Summary Report for the Southwest Water Pollution Control Plant Project, that "... RBC units be used following primary sedimentation and that they be mechanically driven with supplemental air provided or that they be air driven." This was based upon the Engineer's experience with a beggiatoa problem and his evaluation that the "... correction of the problem by supplemental air addition was demonstrated dramatically... where the addition of air to the wastewater flows under the discs eliminated the beggiatoa growth within a matter of hours. The air not only corrects the low dissolved oxygen condition but, through the shearing action of the air bubbles results in a thinner layer of biomass on the disc, allowing the entire biomass to remain aerobic."

c) Increased BOD Removal

Primarily due to the higher dissolved oxygen concentrations, denoting improved oxygen transfer, coupled with the control of undesireable bacterial forms, increased BOD removal rates have been noted for ABC systems. Figures 3 and 4 show side by side comparisons of an RBC and ABC systems at Alexandria,

Figure 2


South Shore Test Facility
First Stages
Δ Dissolved Oxygen (mg/l)
3
2
1
0
-1
0
2
4
6
ABC
RBC
Applied Organic Loading
(lbs. sol. BOD/day/1000 sq.ft.)

Virginia. The stage analysis for soluble BOD concentrations shows lower concentrations for each stage of the ABC units than the RBC stages, although both systems were loaded at similar rates.

This improved performance has been eveluated to a much greater extent at the South Shore test facilities and has resulted in Autotrol Corporation publishing a separate design curve for ABC systems. The major improvements have been noted in early stages, the more highly loaded units, demonstrating further that this advantage is primarily an oxygen transfer improvement phenomena.

This reduced surface area requirement, coupled with the more efficient use of higher density media, has resulted in the elimination of 10% to 20% of the number of shafts required, depending on the application.

d) Operating Flexibility

Because of the improved oxygenation and additional shearing forces for biomass control, the operator has available one more level of control with an ABC system - air flow rate. These can be adjusted as required during seasonal peaks or unanticipated discharges from heavy industrial users. Further, the system can be earily "turned down" during seasonal low periods.

While this would permit the operator to take advantage during low periods of additional power savings, the major thrust is the ABC's ability to safeguard against the unforeseen heavy load.

CONCLUSION

In summary by applying aeration equipment to an RBC unit, an engineer can offset the traditional criticisms of earlier applications. Further, by efficient application, an ABC system can offer a necessary safety factor to the total installation without increase in total life cycle cost.

Figure 3

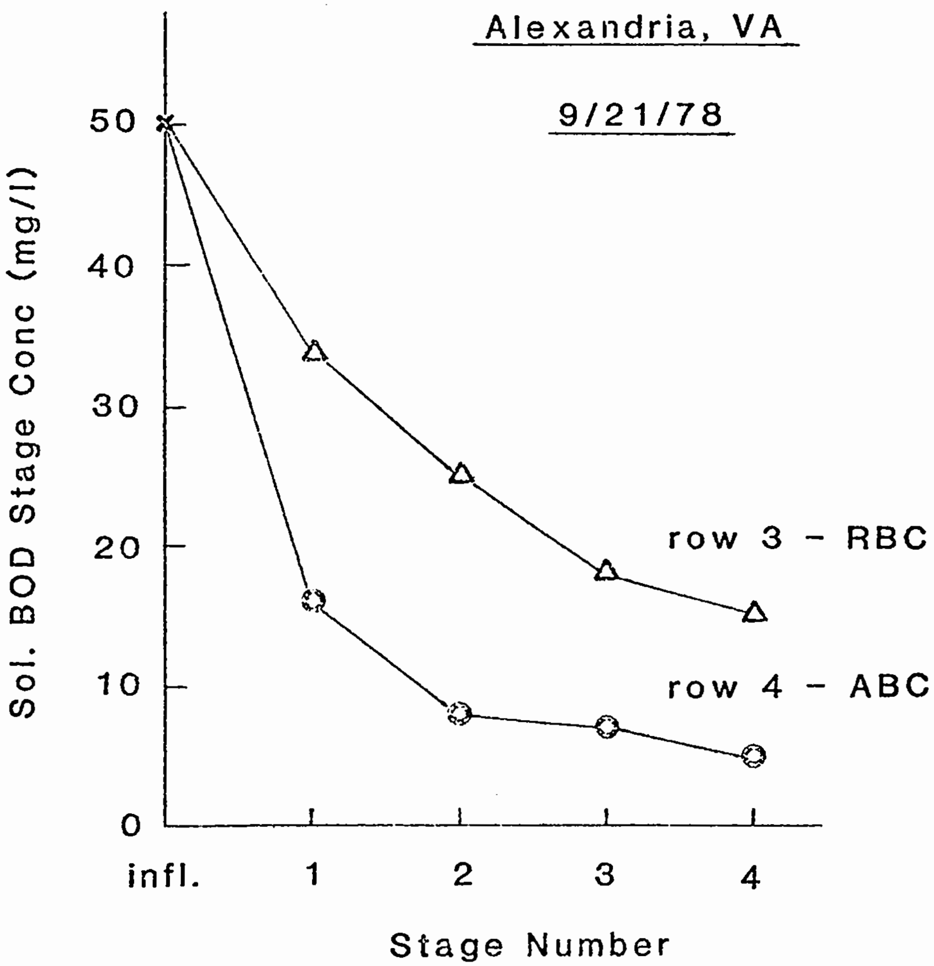
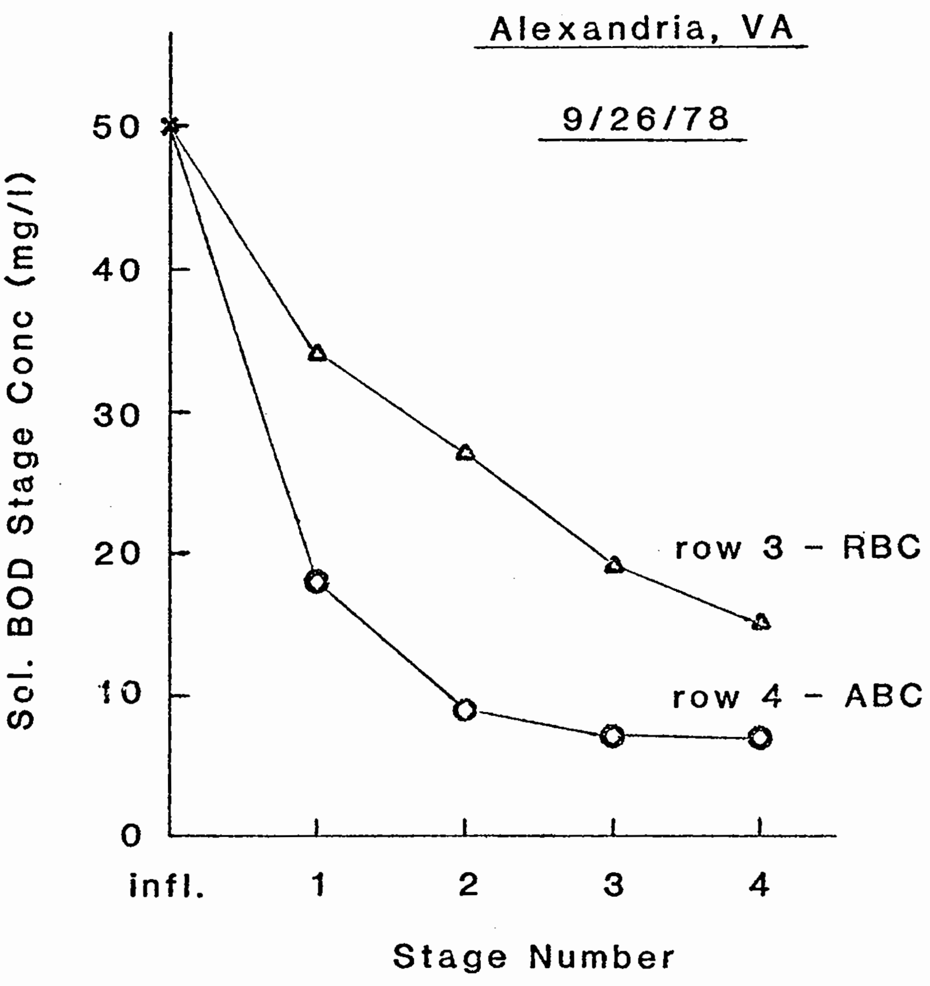

Alexandria, VA
9/21/78
Sol. BOD Stage Conc (mg/l)
50
40
30
20
10
0
row 3 - RBC
row 4 - ABC
infl.
1
2
3
4
Stage Number

Figure 4


Alexandria, VA
9/26/78
Sol. BOD Stage Conc (mg/l)
50
40
30
20
10
0
infl.
1
2
3
4
Stage Number
row 3 - RBC
row 4 - ABC

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PERFORMANCE EVALUATION OF AIR DRIVEN RBC PROCESS FOR MUNICIPAL WASTE TREATMENT

By

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INTRODUCTION

The existing Pinners Point Wastewater Treatment Plant is a primary plant treating an average flow of 12 mgd. The plant is to be upgraded to provide secondary treatment at a design average flow of 16.4 mgd. The effluent criteria for the secondary plant are 30 mg/l biochemical oxygen demand and 30 mg/l suspended solids.

Based on preliminary comparative process evaluation studies, the Rotating Biological Contactor (RBC) process has been chosen as the secondary treatment process at Pinners Point. A pilot plant study was conducted to establish bases of design for the secondary treatment facilities.

EXPERIEMNTAL SET-UP

Pilot Unit

An "aerosurf" RBC pilot unit was used in the study. The pilot unit was a four-stage air-driven RBC, 10 feet long and a 10.4 feet diameter. The total surface area of the unit was 23,000 ft^2. The disc was rotated by air provided through the diffusers located at the bottom of the tank.

The primary effluent was the feed to the pilot plant. The waste flow to the RBC pilot unit was controlled by using a throttle valve to obtain the desired rate. The wet well was aerated to reduce the sulfide level and provide a positive

ORP in the sewage to avoid the growth of undesirable organisms. The RBC effluent was settled for thirty minutes in a six-inch diameter experimental column to simulate the operation of the secondary settling tanks. A diagram of the pilot plant is shown on Figure 1.

Operating Conditions

The start-up period lasted for about three weeks in September, 1977. The test program lasted approximately six months and consisted of seven phases. Overall hydraulic loadings were adjusted to cover a range of 1.0 to 3.0 gpd/ft^2 during Phases I through V. The shaft speed was maintained, through the adjustment of air flow rate, at about 1.60 rpm through all phases except Phase V when the speed was 1.04 rpm.

Phases VI and VII studied the effect of instantaneous and diurnal variations in flow on RBC performance. Plant operating reports indicate an instantaneous flow fluctuation cycle to be 20 to 25 minutes long, with a flow variation of 5 to 30 mgd during the day and 3 to 15 mgd during the night. A cyclical flow pattern as indicated on Figure 2 was maintained to approximately simulate this condition in the pilot study. The flow fluctuation was accomplished by using three constant head tanks with outlet solenoid valves, each controlled by a timer.

A summary of the program schedule and the operating conditions are shown in Table 1. Settling and thickening tests were performed on the RBC effluent using an eight-foot settling column and one liter glass cylinder, respectively.

Data Collection

As indicated on Figure 1, samples were collected at five sampling points for the RBC influent, effluent, and wastewater in three intermediate stages. All samples from the RBC unit were composited. Grab samples were taken for settling tests. The weekly analytical schedule is shown in Table 2.

RESULTS AND DISCUSSION

Influent Characteristics

The characteristics of the influent to the RBC unit are shown in Table 3. The values shown are the means of the daily determinations. The percent coefficient of variation for each parameter is also shown to indicate the extent of influent quality fluctuation.

The overall average influent BOD_5 and $SBOD_5$ were 104 ± 34 mg/l and 42 ± 22 mg/l, respectively, and the SS and VSS concentrations were 79 ± 44 mg/l and 67 ± 40 mg/l, respectively. It can be seen that there were considerable fluctuations in BOD_5 and $SBOD_5$ in each phase as reflected by the high coefficient of variation values. Even greater amounts of variation were observed in SS and VSS concentrations in the influent.

Table 4 shows the ratio of BOD_5 to COD on both soluble and total bases. The ratio ranged from 36 percent to 59 percent, indicating a significant amount of not readily biodegradable fraction in the influent. The average COD to BOD_5 ratio did not change significantly between phases, averaging about 2.3.

The average DO concentration in the RBC influent after aeration in the wet well ranged from 2.3 mg/l in Phase I to 7.3 mg/l in Phase VI. This increase is believed to be due to the decrease in water temperature from September to March.

RBC Performance

Effect of Loading:

Loading to RBC has been expressed in three different ways:

1. Hydraulic, gallons/ft^2/day

2. Soluble Organic, pounds $SBOD_5$/1,000 ft^2/day

3. Total Organic, pounds BOD_5/1,000 ft^2/day

Each has been plotted against the performance parameters, $SBOD_5$ and BOD_5 pounds removed, effluent quality, and removal efficiency, and summarized in Table 5. All plots are based on linear regression calculations.

The total BOD_5 are 30-minute settled effluent values and the $SBOD_5$ are RBC effluent values. Total BOD_5 values are plotted only to obtain the order of magnitude of BOD_5 concentrations to be expected in the effluent. The thirty-minute settling test only grossly simulated the performance of a full-scale secondary clarifier. It is anticipated that the effluent quality would be better with a full-scale secondary clarifier.

- Hydraulic Loading

  The average hydraulic loading was varied from 1.0 to 3.0 gpd/ft^2 during the test program. The pounds $SBOD_5$ and BOD_5 removed, effluent $SBOD_5$ and BOD_5, and $SBOD_5$ and BOD_5 removal efficiencies are plotted as functions of hydraulic loading for all phases on Figures 3 through 8.

  It can be seen from Figures 7 and 8 that, in spite of high fluctuations, the treatment efficiency generally decrease with an increase in hydraulic loading. Figures 5 and 6 show the same result, but in terms of effluent quality. Overall the correlation is not satisfactory for performance prediction.

- Soluble Organic Loading

  The average soluble organic loading ranged from 0.5 to 1.2 pounds $SBOD_5$/1,000 ft^2/day. Since the influent $SBOD_5$ varied widely, individual data points were plotted in addition to the average phase data.

  An average phase value plot of pounds $SBOD_5$ removed versus pounds $SBOD_5$ applied is shown on Figure 9. Figure 10 is the same graph using daily data points. It can be seen that excellent correlation exists between the pounds of soluble organics applied and removed. The correlation coefficient is 0.97 which is extremely good for a biological system. The slope of the regression line which is representative of the percent removal efficiency indicates about 80 percent $SBOD_5$ removal.

  Figure 11 is a plot of pounds BOD_5 removed versus $SBOD_5$ loading. The correlation coefficient for this graph, 0.92, is also relatively high. Figures 9, 10 and 11 show that the pounds of organics removed increased linearly with increases in soluble organic loading, indicating nonoxygen limiting conditions.

  On Figures 12 and 13, the average percent $SBOD_5$ removal and BOD_5 removal are plotted against $SBOD_5$

loading for each phase. The regression lines indicate that the performance of the RBC decreases with an increase in soluble organic loading.

Figures 14 and 15 are plots of effluent $SBOD_5$ and BOD_5 against $SBOD_5$ loading. These plots indicate that effluent quality decreases as the soluble organic loading decreases.

- Total Organic Loading

The average total organic loading ranged from 1.1 to 3.5 pounds $BOD_5/1,000\ ft^2/day$. As with the soluble BOD_5, the influent total BOD_5 varied widely, and individual data points were plotted in addition to the average phase data.

Figures 16 and 17 are average phase value plots of $SBOD_5$ removed and BOD_5 removed versus BOD_5 loading. The same correlation is shown on Figure 18 using daily data points. The plots based on BOD_5 loading have high correlation coefficients: 0.89 for $SBOD_5$ removal and 0.95 for BOD_5 removal. The regression lines indicate nonoxygen limiting conditions.

The $SBOD_5$ and BOD_5 effluent concentrations and removal efficiencies are plotted as functions of BOD_5 loading for each phase on Figures 19 through 22. These graphs show that the treatment efficiency and effluent quality decrease as the total BOD_5 loading increases.

Effluent Characteristics

The characteristics of the RBC pilot plant effluent are shown in Table 6. Both BOD_5 and SS are the properties of the 30-minute settled effluent.

The effluent ammonia nitrogen (NH_3-N) concentrations were compared to the influent NH_3-N concentrations. The average NH_3-N concentration for all seven phases decreased from 30 $\pm$ 10 mg/l in the influent to 24 $\pm$ 9 mg/l in the effluent. This decrease is due to bacterial assimilation. The largest decrease in NH_3-N occurred during Phase II when the average NH_3-N concentration decreased from 36 to 19 mg/l across the RBC, indicating some nitrification.

In order to estimate the BOD_5 in the effluent, an attempt was made to correlate the particulate BOD_5 (total BOD_5 - soluble BOD_5) and SS removals during the settling process. Figure 23 is a plot of particulate BOD_5 removal versus SS removal, mg/l, during the 30-minute settling test. The particulate BOD_5 removal was determined by the difference between the total BOD_5 concentrations in the RBC effluent and 30-minute settled effluent, assuming there was no change in the $SBOD_5$ concentrations during the settling process. The slope of the regression line is 0.37, indicating 0.37 mg particulate BOD_5 per each mg SS.

Settling Characteristics

Settling tests were performed during each phase to determine the required surface overflow rate, gpd/ft^2, and surface area of settling tank. The tests indicate better settling characteristics during the lower soluble organic loading phases.

Figure 24 is a plot of settled effluent suspended solids as a function of soluble organic loading. The purpose of the plot is not to determine effluent SS at various loading but to observe the effect of organic loadings on general settling characteristics reflected by 30-minute settling test. The regression line indicates that the effluent quality decreases as the $SBOD_5$ loading increases.

Figure 25 shows a plot of percent SS removal versus surface overflow rate calculated from the settling tests conducted for all of the phases. A band is shown to illustrate the range of values observed.

Sludge Production

The RBC influent solids to effluent solids ratio was plotted to determine if there is any net sludge increase or decrease due to synthesis or oxidation. This is shown on Figure 26. No definite trend is apparent. Therefore, the mean value is chosen for the estimation of sludge production. The sludge production is approximately equal to 0.83 times the amount of suspended solids entering the RBC.

Effect of Rotational Speed

A shaft speed of about 1.60 rpm was maintained through all of the phases except Phase V when the speed was 1.04 rpm.

Comparing Phase V with Phase II and III which had similar soluble organic loadings (about 0.6 pounds $SBOD_5$/1,000 ft^2/day), there is no significant difference in effluent quality. The $SBOD_5$ removal was lower when the rotating speed was reduced, but this is probably due to a lower influent $SBOD_5$ concentration in Phase V.

Comparing Phase V to all of the phases, no significant difference is observed in the performance of the RBC. This is probably because oxygen limiting conditions did not occur during any of the phases.

Effect of Diurnal Flow Variation

During Phase VI and VII, the RBC influent flow was varied in a four-step cyclical flow pattern to simulate day and night flow variations. There is no significant difference between the results of these phases and the first five phases. This indicates that the diurnal flow variation of the magnitude and range experimented has no significant effect on the performance of the RBC unit.

SUMMARY

The pilot plant data generally seemed to correlate better with loading when loading is expressed as organic rather than hydraulic. The performance characteristics showed a direct dependence on the soluble BOD_5 loading.

The bases of design for the full scale RBC process is shown in Table 7. Based on these design conditions and the RBC pilot plant study, the performance of the RBC can be estimated as follows:

- Effluent $SBOD_5$

  From the regression equation developed based on data shown on Figure 10, it is estimated that the effluent $SBOD_5$ at the design loading rate of 1.4 pounds $SBOD_5$/1,000 ft^2/day would be 17 mg/l.

- Effluent SS

  Better than 80 percent of the RBC effluent suspended solids can be removed in the secondary clarifier at the design surface overflow rate of

740 gpd/ft^2 and a hydraulic retention time of 2.97 hours. Assuming an RBC effluent SS concentration of 88 mg/l, the effluent SS concentration would be 18 mg/l.

- Effluent BOD_5

  The ratio of particulate BOD_5 to SS in the effluent was found to be 0.37. An effluent with 18 mg/l SS will have a particulate BOD_5 of about 7 mg/l. Therefore, the total BOD_5, soluble and particulate, in the effluent is estimated to be about 24 mg/l.

- Sludge Production

  The secondary sludge production under the assumed design conditions is indicated to be approximately 0.83 times the influent suspended solids.

- Effect of Rotating Speed

  Comparing Phase V, when the rotating speed was changed from about 1.6 rpm to 1.04 rpm, with the other phases, no significant difference is observed in the performance of the RBC.

- Diurnal Variation

  The diurnal flow variation of the magnitude and range experimented during Phases VI and VII has no significant effect on the performance of the RBC unit.

TABLE 1

Study Phases and Hydraulic Loading Characteristics

Phase	Period	Number of Days at Steady State	RBC Influent Average Temperature °F	Shaft Speed rpm	Mean Hydraulic Loading gpd/ft^2
I	9/18/77- 9/30/77	13	81	1.61	3.00
II	10/ 9/77-10/30/77	22	73	1.71	1.00
III	11/ 6/77-12/ 2/77	27	69	1.62	1.79
IV	12/11/77- 1/ 4/78	25	61	1.66	3.03
V	1/ 8/78- 1/31/78	24	57	1.04	2.16
VI	2/22/78- 3/ 7/78	14	56	1.55	2.74
VII	3/ 8/78- 3/22/78	15	55	1.58	1.41

TABLE 2

Sampling and Testing

	NUMBER OF ANALYSES PER WEEK											
	pH	Temp.	D.O.	SBOD*	BOD	SS	VSS	COD	SCOD*	NH_3-N	S-	ALK.
Influent	5	5	5	5	5	5	5	3	3	1	2	1
Stage 1	-	-	5	5	5	5	5	-	-	-	-	-
Stage 2	-	-	5	5	5	5	5	-	-	-	-	-
Stage 3	-	-	5	5	5	5	5	-	-	-	-	-
Stage 4	5	5	5	5	5	5	5	3	3	1*	2	1*
30 Min. Settled Effluent	-	-	-	-	5	5	-	-	-	-	-	-

Settling Test - Once a week during steady state operation in each phase.

* Analyses on Filtered Samples

TABLE 3

RBC Influent Characteristics, mg/l

Phase	BOD5		SBOD5			COD		SCOD		SS		VSS		DO	
	Mean	% COV(1)	Mean	% COV	% SBOD5	Mean	% COV	Mean	% COV	Mean	% COV	Mean	% COV	Mean	% COV
I	144	25	49	62	34	328	25	126	21	92	29	75	34	2.3	16
II	146	20	71	35	49	262	32	124	29	108	29	97	34	3.1	16
III	93	30	41	51	44	202	23	126	45	70	21	54	28	3.8	14
IV	100	27	40	40	40	226	21	134	32	84	40	75	43	4.4	19
V	84	30	29	52	35	181	16	105	22	84	102	68	106	5.9	12
VI	90	22	34	36	38	211	22	133	31	64	33	54	43	7.3	6
VII	96	25	40	36	42	237	35	119	16	58	36	48	40	7.0	10
Overall Average	104	33	42	53	40	226	30	123	32	79	56	67	59	4.7	37

(1) % COV = Percent Coefficient of Variation = $\frac{\text{Standard Deviation}}{\text{Mean}} \times 100$

TABLE 4

Influent COD/BOD_5 Ratios

Phase	$\frac{COD}{BOD5}$	$\frac{BOD5}{COD}(100)$	$\frac{SCOD}{SBOD_5}$	$\frac{SBOD_5}{SCOD}(100)$
I	2.34	43	3.01	33
II	1.80	59	1.79	56
III	2.26	44	3.28	31
IV	2.40	42	4.00	25
V	2.32	43	4.11	24
VI	2.32	43	5.06	20
VII	2.80	36	2.74	36

TABLE 5

Performance of RBC Pilot Plant at Various Loadings[1]

Phase	Loadings			$SBOD_5$ Removed lbs/1000 ft^2/day	BOD_5 Removed lbs/1000 ft^2/day	Effluent $SBOD_5$ mg/l	Effluent BOD_5 mg/l	% $SBOD_5$ Removal	% BOD_5 Removal
	Lbs $SBOD_5$ / 1000 ft^2/day	Lbs BOD_5 / 1000 ft^2/day	gpd / ft^2						
I	1.22	3.53	3.00	0.98	2.73	9	35	81.6	75.7
II	0.57	1.14	1.00	0.54	1.08	8	17	88.7	88.4
III	0.65	1.42	1.79	0.51	1.11	7	19	82.9	79.6
IV	1.07	2.62	3.03	0.60	1.67	12	34	70.0	66.0
V	0.60	1.60	2.16	0.40	1.10	9	23	69.0	70.2
VI	0.78	2.05	2.74	0.59	1.72	9	18	75.1	80.5
VII	0.47	1.13	1.41	0.38	0.97	8	13	75.4	85.2

(1) BOD_5 concentrations are 30-minute settled effluent values; $SBOD_5$ concentration are RBC effluent values.

TABLE 6

Effluent Characteristics, mg/l

Phase	BOD_5 (1) Mean	BOD_5 (1) % COV	$SBOD_5$ (2) Mean	$SBOD_5$ (2) COV	% $SBOD_5$	COD (2) Mean	COD (2) % COV	SCOD (2) Mean	SCOD (2) % COV	SS (1) Mean	SS (1) % COV	VSS (2) Mean	VSS (2) % COV	DO (2) Mean	DO (2) % COV
I	35	26	9	77	26	220	25	89	17	32	42	73	25	1.7	35
II	17	40	8	92	47	141	24	65	30	24	74	88	35	4.2	7
III	19	47	7	43	37	155	21	83	40	20	75	49	31	4.3	6
IV	34	38	12	33	35	171	24	95	27	24	46	51	37	5.1	27
V	23	40	9	49	39	140	24	77	11	18	63	44	32	5.8	17
VI	18	27	9	47	50	159	18	89	34	11	61	36	36	7.3	5
VII	13	26	8	32	62	141	29	85	17	9	56	48	59	7.5	8
Overall Average	23	5	9	65	39	156	27	83	33	20	71	55	47	5.1	36

(1) 30-minute settled effluent values

(2) RBC effluent values

For effluent before settling, $BOD_5 = 50 \pm 19$ (Range: 20-135) mg/l
$SS = 71 \pm 31$ (Range: 8-224) mg/l

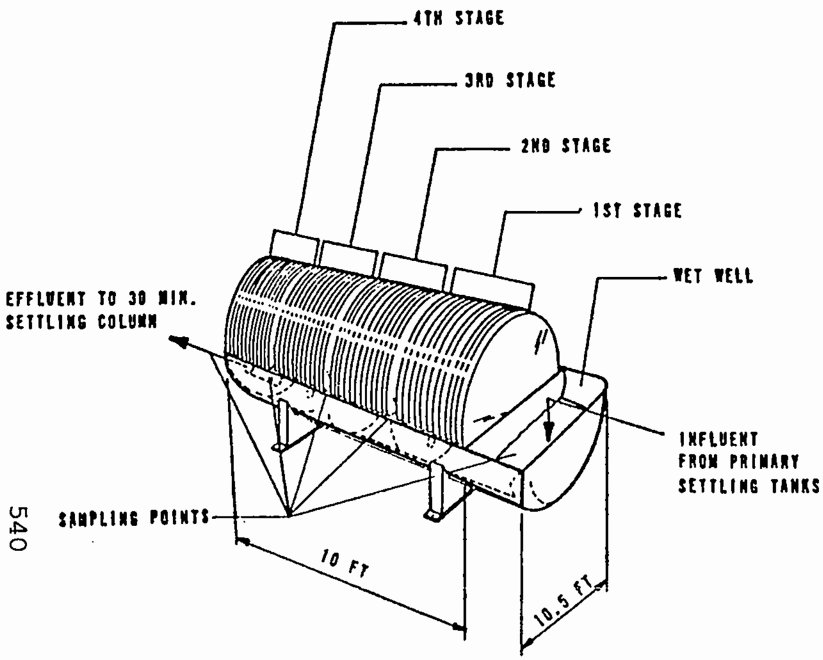
TABLE 7

RBC Bases of Design[1]
(16.4 mgd)

1.	RBC dimensions	
	Diameter, ft.	12
	Surface area per unit, ft^2	100,000
2.	Maximum shaft speed, rpm	1.6
3.	Number of shafts per tank	6
4.	Number of tanks	12
5.	Number of shafts	72
6.	Annual average loading	
	Total BOD_5, lbs/1,000 ft^2/day	2.7
	Soluble BOD_5, lbs/1,000 ft^2/day	1.4
	Hydraulic, gpd/ft^2	2.3
7.	Annual average influent concentration	
	Total BOD_5, mg/l	148
	Soluble BOD_5, mg/l	75
	Suspended Solids, mg/l	88

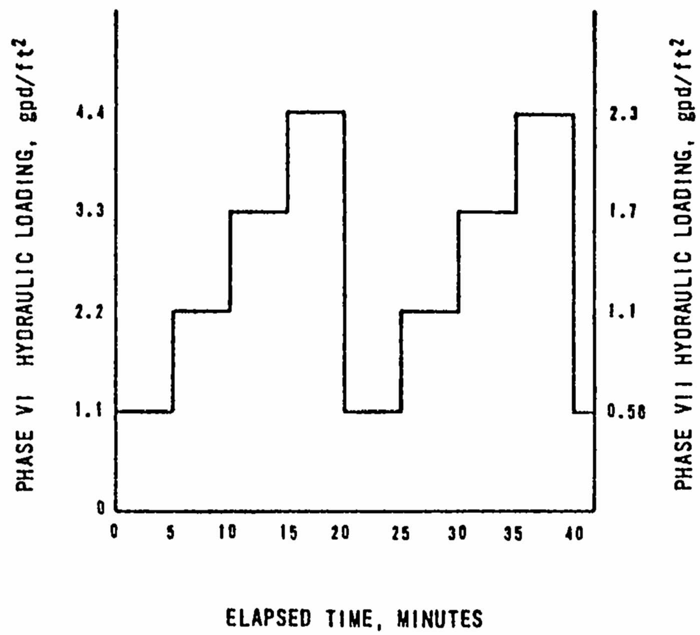
(1) Design year: 2010

FIGURE 1



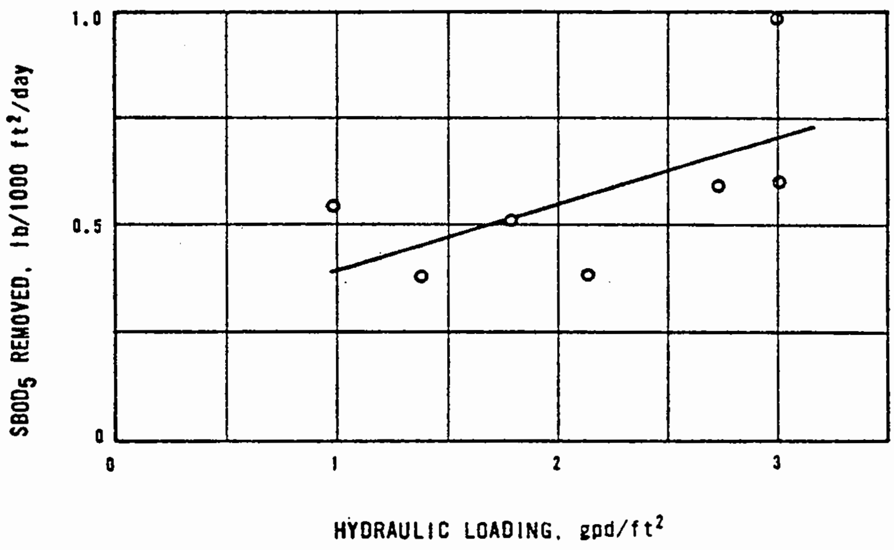
RBC PILOT PLANT

FIGURE 2



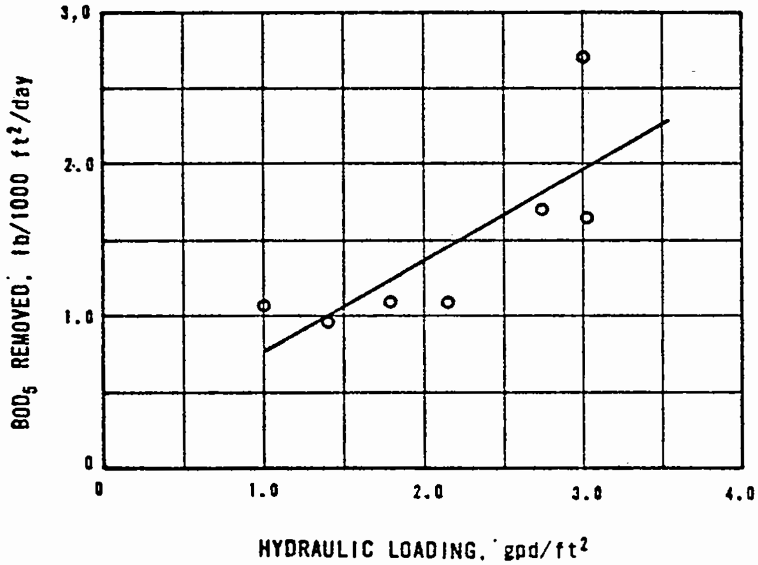
PHASE VI AND PHASE VII CYCLICAL FLOW PATTERN
SIMULATING DIURNAL VARIATION

FIGURE 3



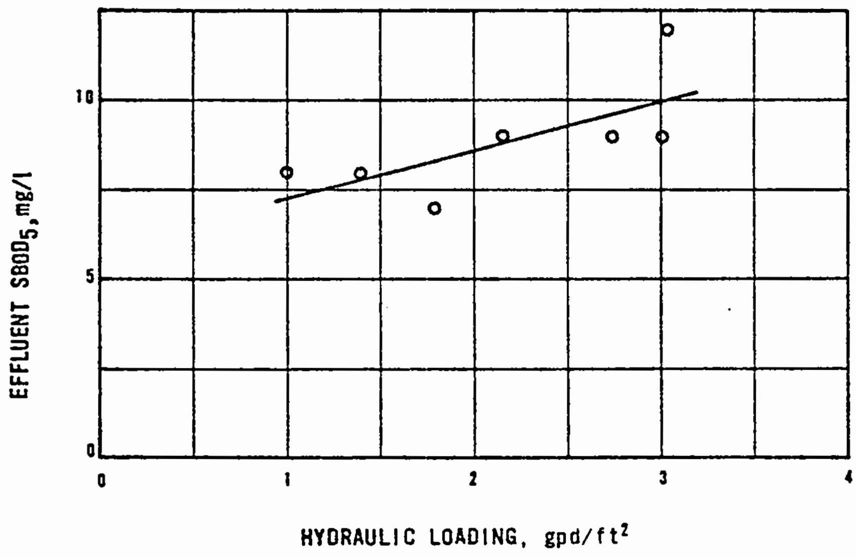
SBOD$_5$ REMOVED vs HYDRAULIC LOADING

FIGURE 4



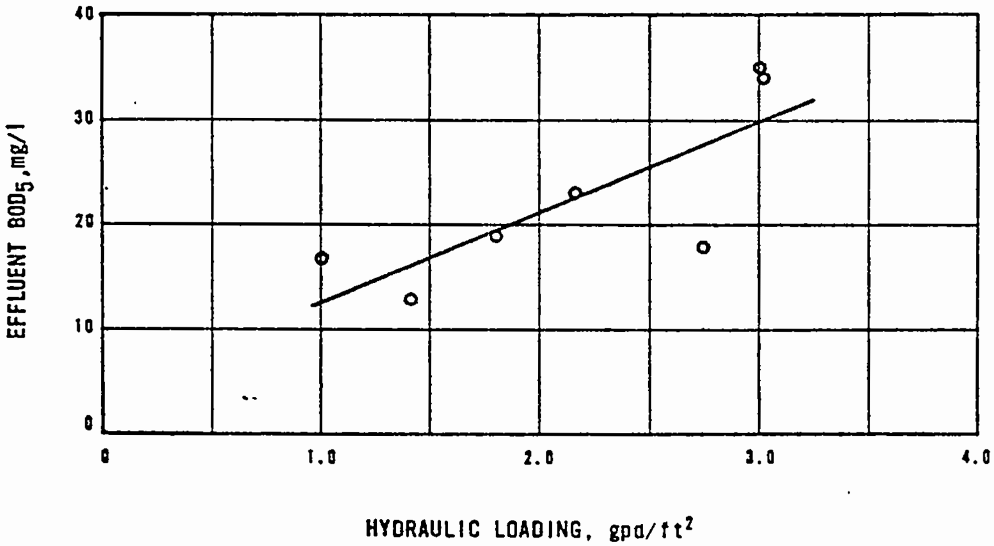
BOD$_5$ REMOVED vs HYDRAULIC LOADING

FIGURE 5



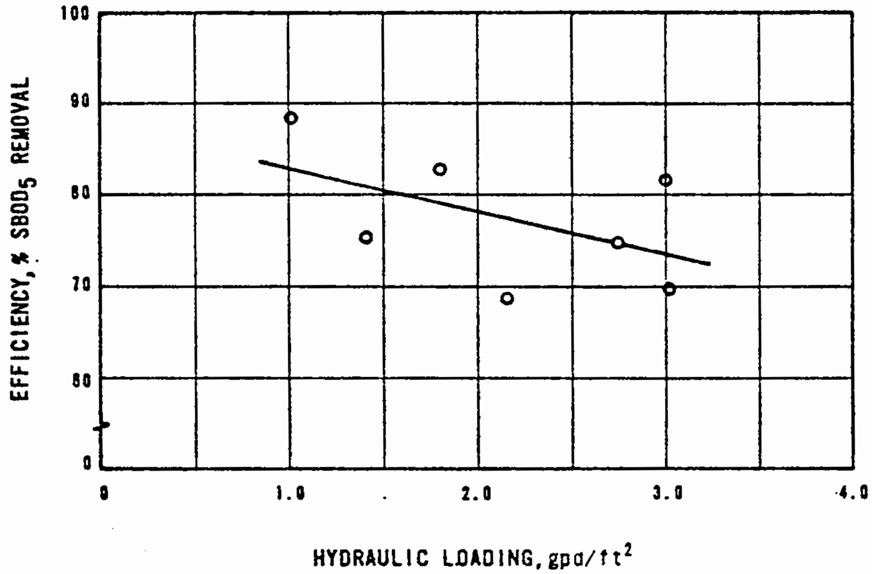
EFFLUENT $SBOD_5$ vs HYDRAULIC LOADING

FIGURE 6



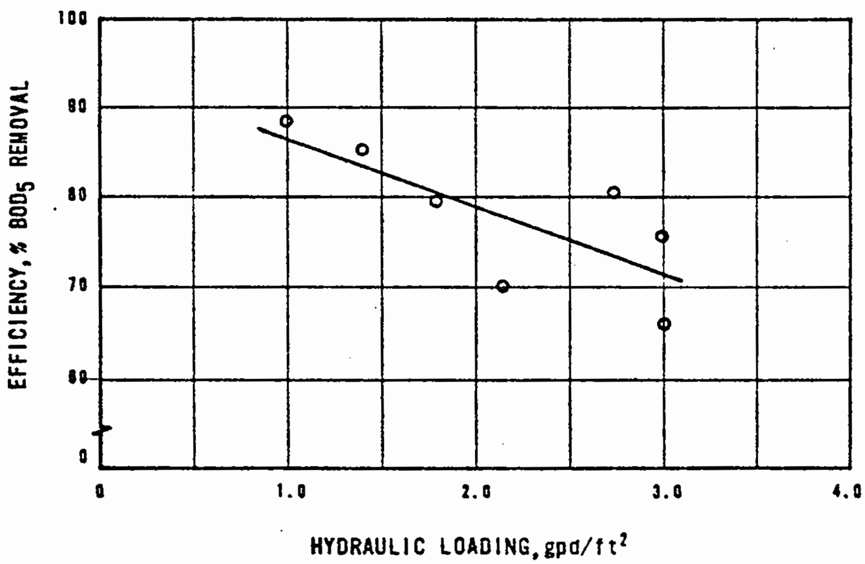
EFFLUENT BOD_5 vs HYDRAULIC LOADING

FIGURE 7



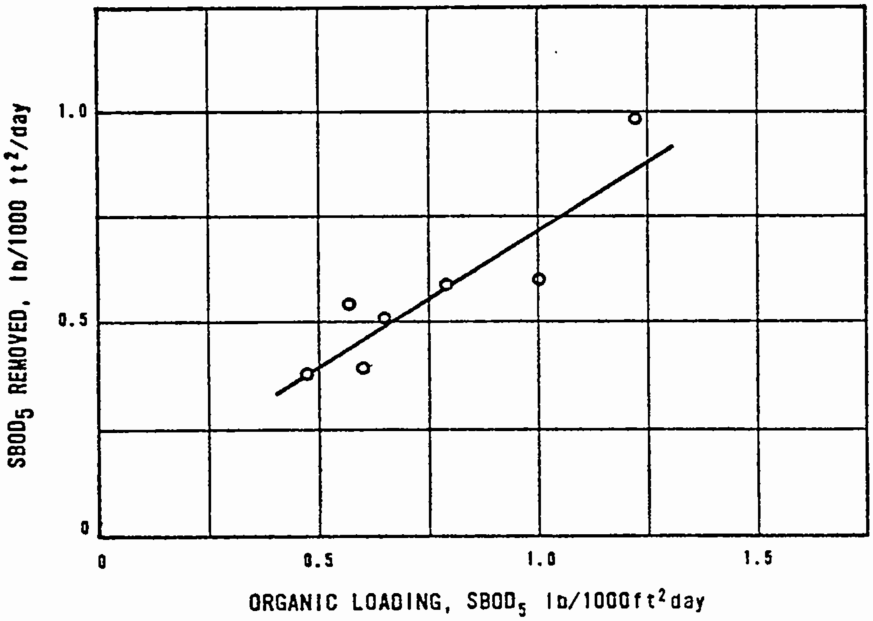
EFFICIENCY (% $SBOD_5$ REMOVAL) vs HYDRAULIC LOADING

FIGURE 8



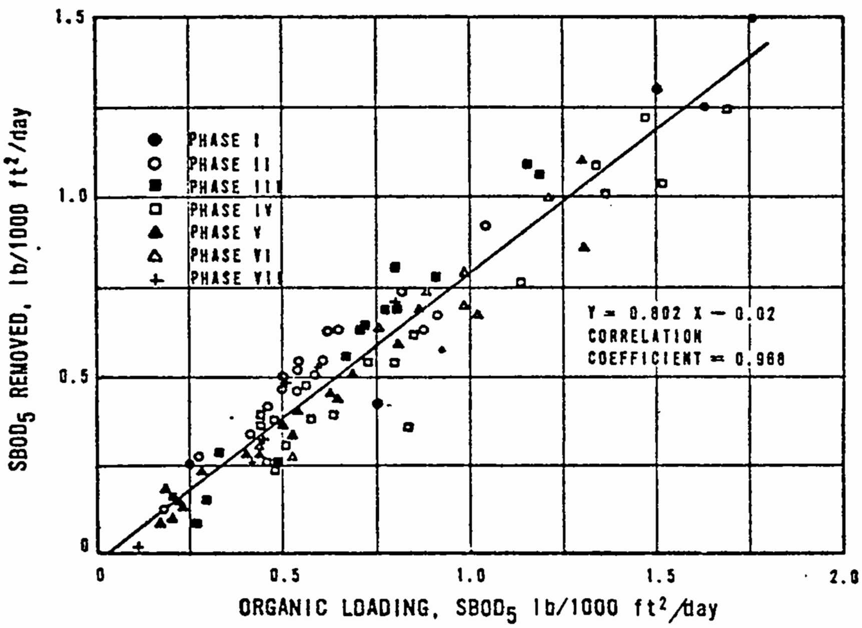
EFFICIENCY, (% BOD_5 REMOVAL) HYDRAULIC LOADING

FIGURE 9



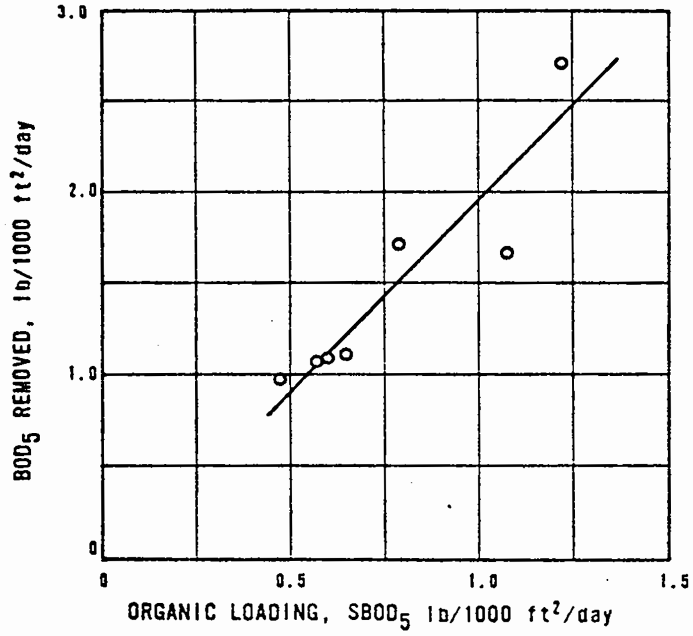
$SBOD_5$ REMOVED vs $SBOD_5$ LOADING (MEAN VALUES)

FIGURE 10



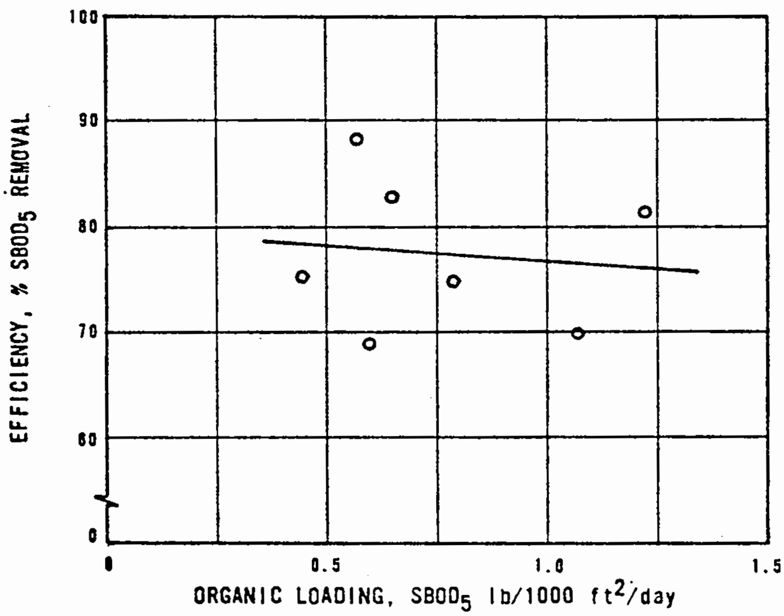
$SBOD_5$ REMOVED vs $SBOD_5$ LOADING (DAILY VALUES)

FIGURE 11



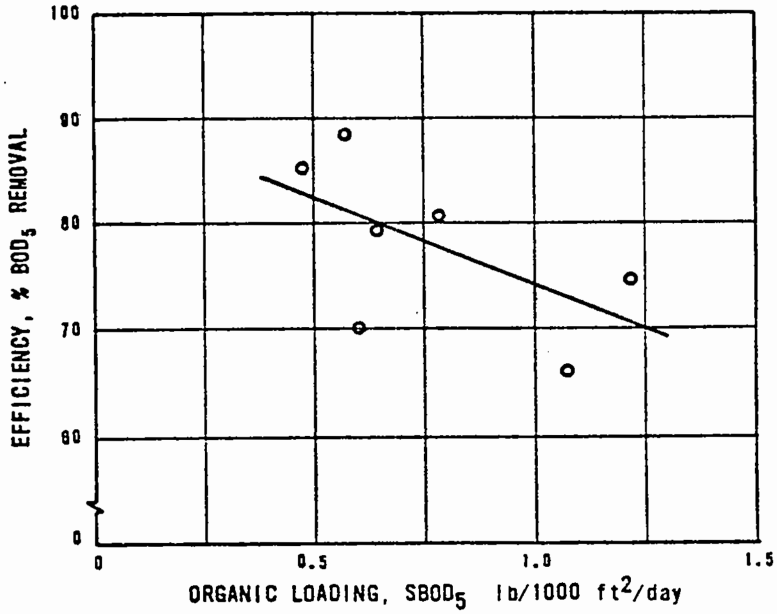
BOD_5 REMOVED vs $SBOD_5$ LOADING

FIGURE 12



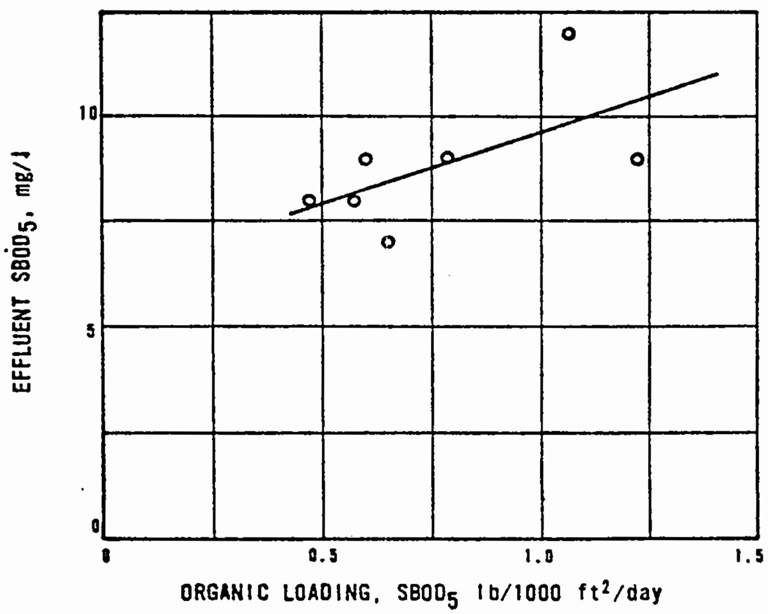
EFFICIENCY (% $SBOD_5$ REMOVAL) vs $SBOD_5$ LOADING

FIGURE 13



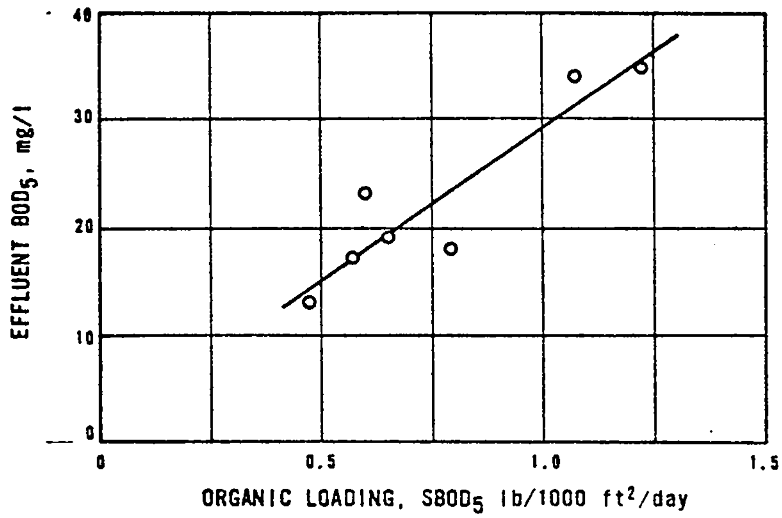
EFFICIENCY (% BOD_5 REMOVAL) vs $SBOD_5$ LOADING

FIGURE 14



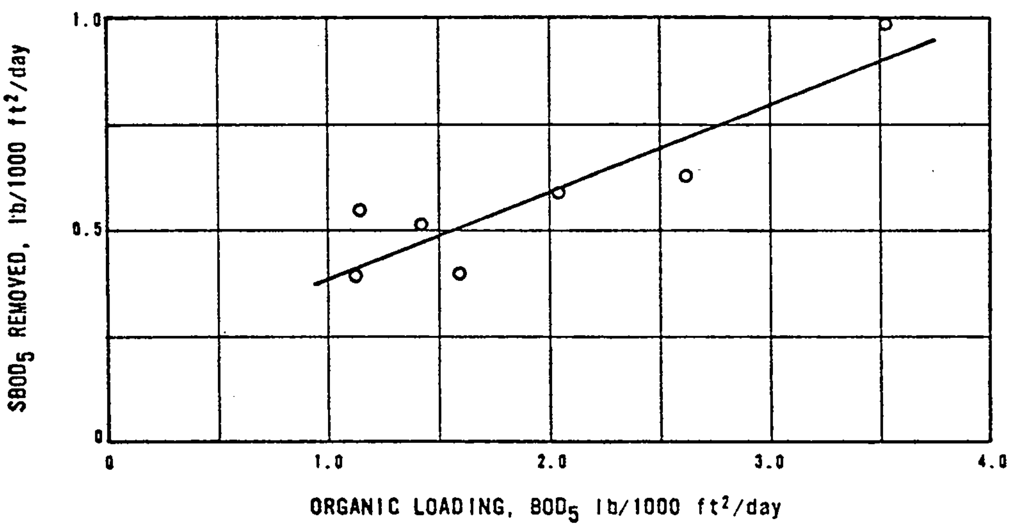
EFFLUENT $SBOD_5$ vs $SBOD_5$ LOADING

FIGURE 15



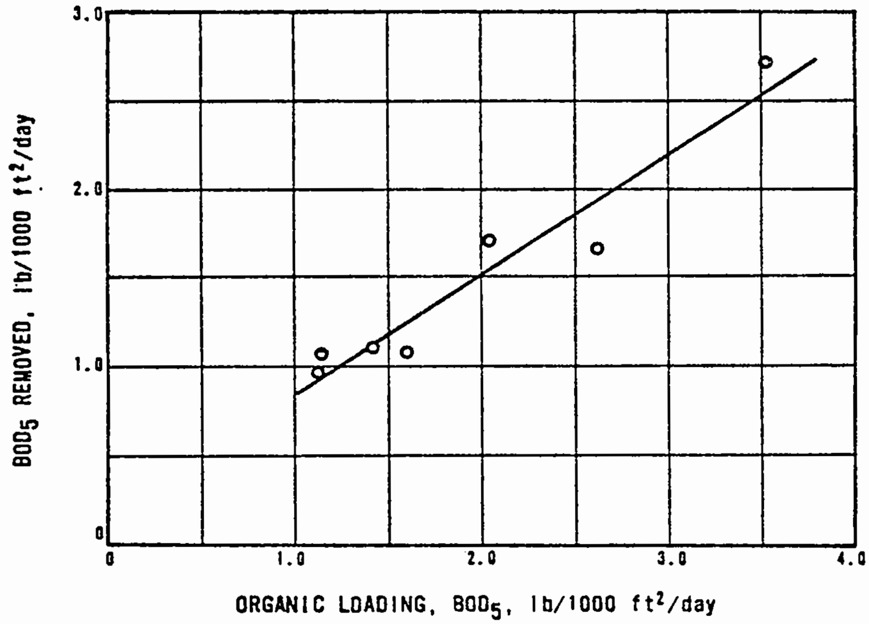
EFFLUENT BOD_5 vs $SBOD_5$ LOADING

FIGURE 16



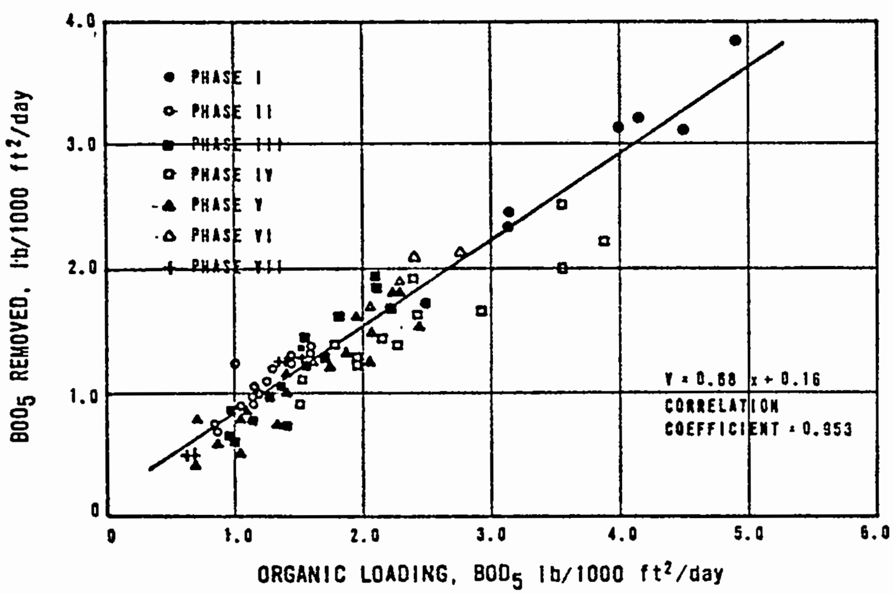
$SBOD_5$ REMOVED vs BOD_5 LOADING

FIGURE 17



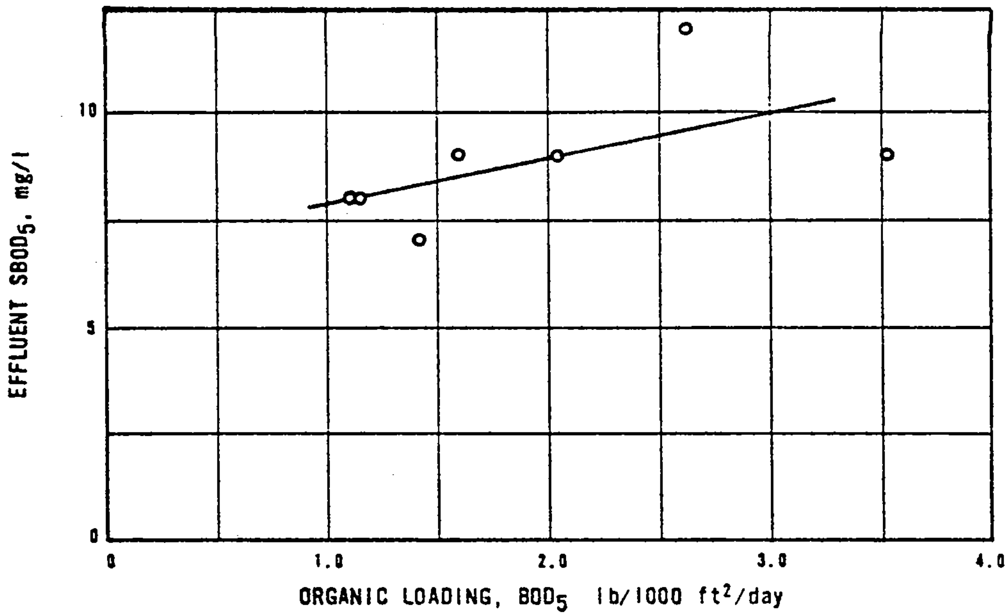
BOD_5 REMOVED vs BOD_5 LOADING (MEAN VALUES)

FIGURE 18



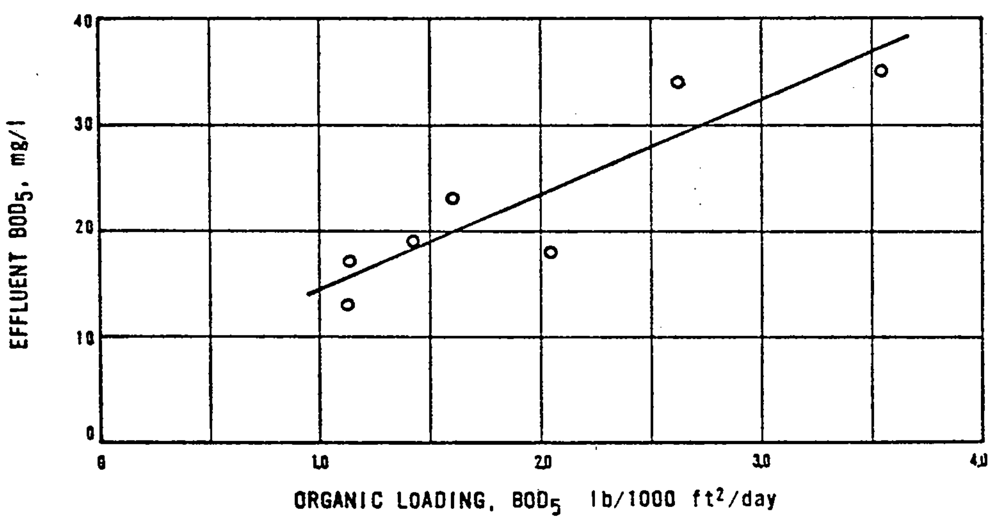
BOD_5 REMOVED vs BOD_5 LOADING (DAILY VALUES)

FIGURE 19



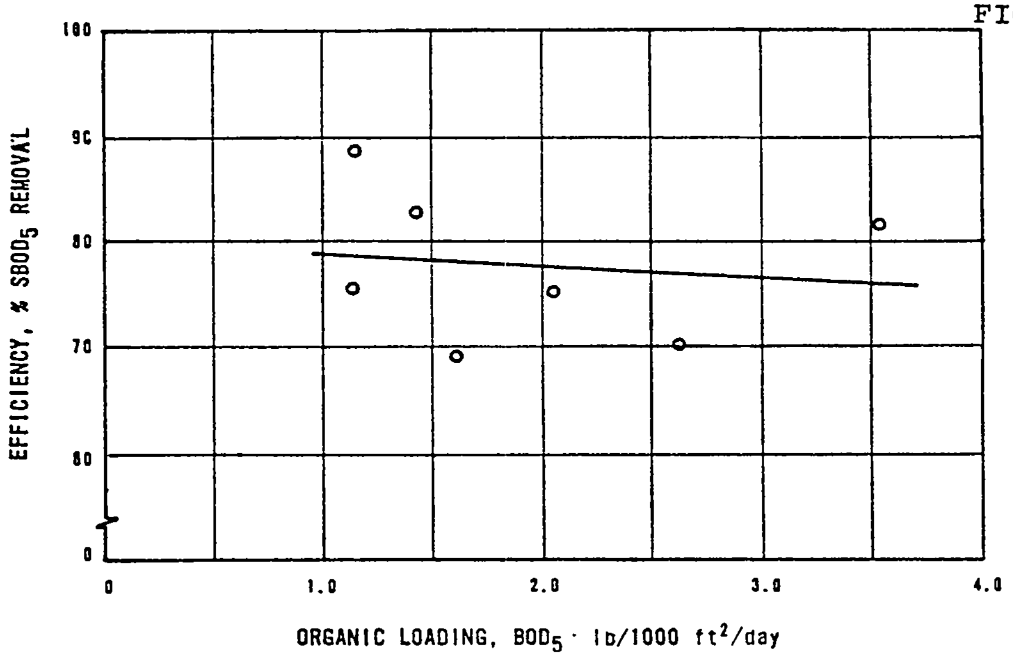
EFFLUENT $SBOD_5$ vs BOD_5 LOADING

FIGURE 20



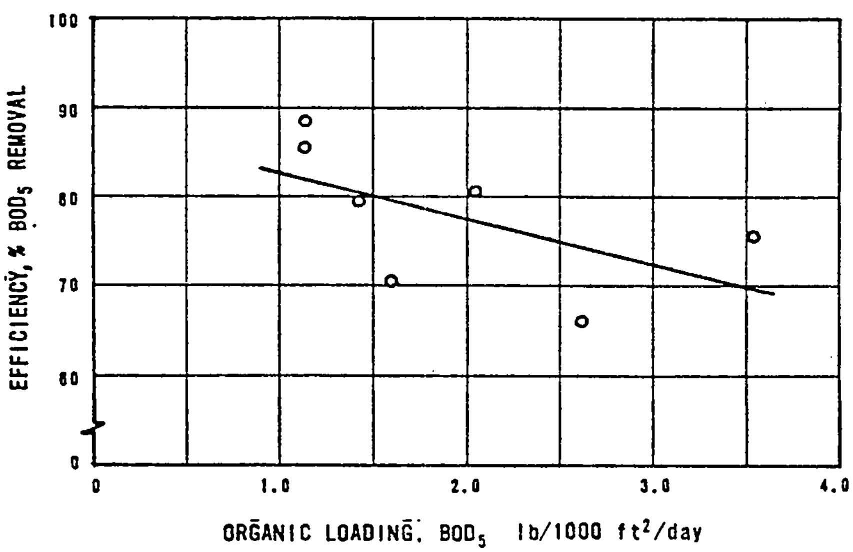
EFFLUENT BOD_5 vs BOD_5 LOADING

FIGURE 21



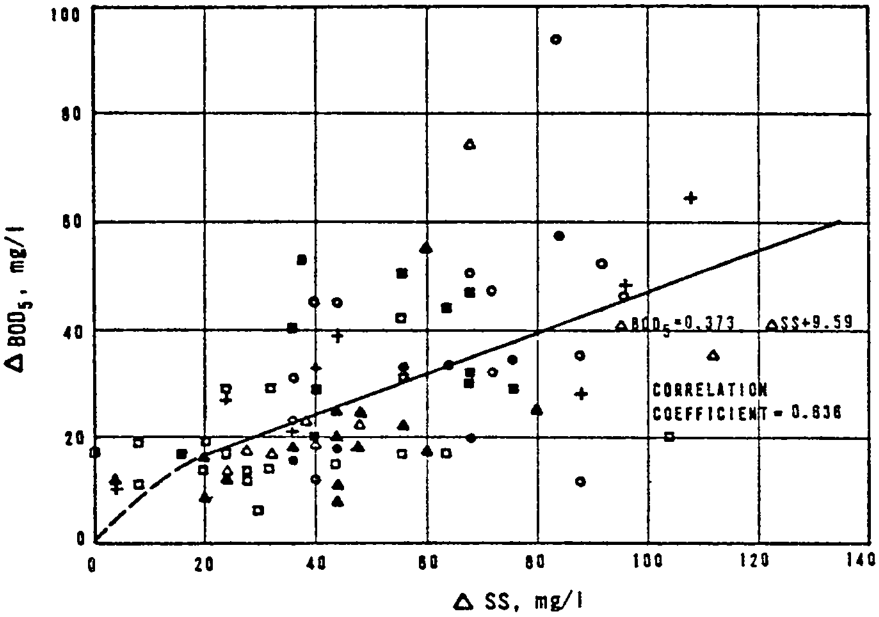
EFFICIENCY (% $SBOD_5$ REMOVAL) vs BOD_5 LOADING

FIGURE 22



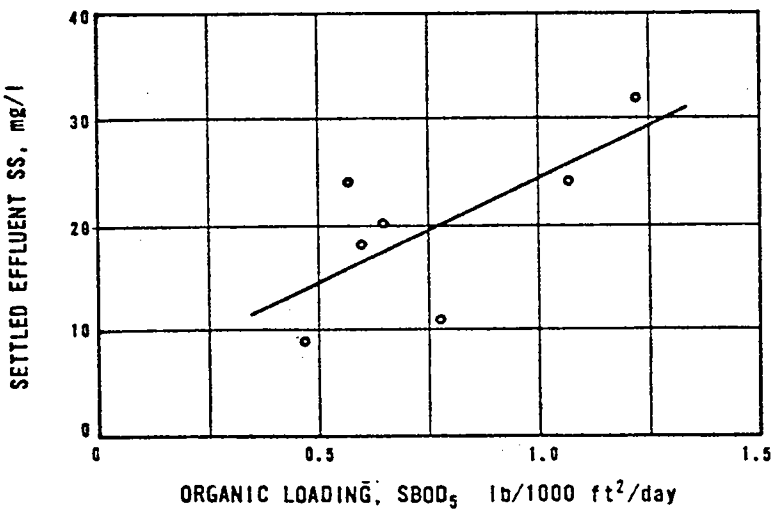
EFFICIENCY (% BOD_5 REMOVAL) vs BOD_5 LOADING

FIGURE 23



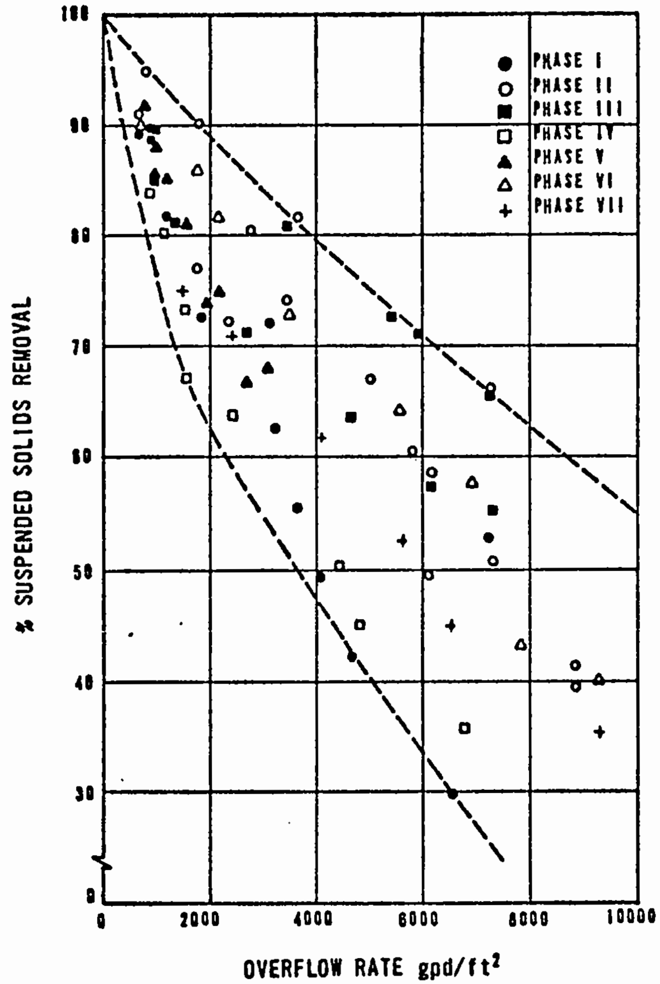
BOD$_5$ REMOVAL vs SS REMOVAL
DURING SETTLING PROCESS

FIGURE 24



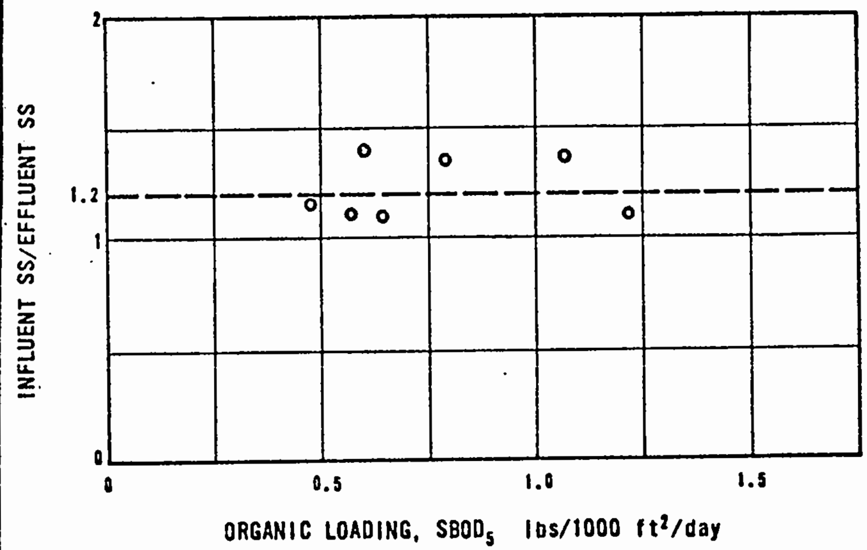
SETTLED EFFLUENT SS vs SBOD$_5$ LOADING

FIGURE 25



% SS REMOVAL vs SURFACE OVERFLOW RATE

FIGURE 26



SLUDGE PRODUCTION RATIO vs SBOD5 LOADING

SURFACT: CURRENT DEVELOPMENTS AND PROCESS APPLICATIONS

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INTRODUCTION

A new secondary wastewater treatment concept called the SURFACT Process has been developed for upgrading existing activated sludge plants to higher capacities, with no additional land usage and little or no increase in energy consumption. The Surfact process is a biological system resulting from the combination of a rotating fixed film contactor, called Bio-Surf, and an activated sludge system. The process can be applied to augment any overloaded activated sludge system or for new treatment facility construction. Revamping of existing systems is enhanced by minimal capital and operating expenditures while allowing for unusually rapid construction scheduling.

In the Surfact process, rotating biological contactor (RBC) media assemblies are installed in the aeration basin of an activated sludge system, as shown in Figure 1. The media for the rotating contactor is fabricated from polyethylene plastic. Air cups are then attached to the periphery of the media to capture the diffused air, which is the preferred means of driving the shaft. The air being diffused at the bottom of the aeration tank provides rotation of the shafts primarily through capture in the air cups, although some assistance is derived from the hydraulic roll of the tank fluids. A supplementary air header, using air from the main blower air supply, can be installed for additional control of rotational speed.

The combination of fixed film growth and suspended growth within a single tank, provides additional biological solids in the system, making higher treatment capabilities possible. The result can be a higher treatment level at the same flow rate or the same level of treatment at an increased flow rate, or a combination of increased flow rate at increased treatment levels.

By merging the two systems, the Surfact process increases the efficiency and stability of the existing activated sludge system. Inherent process benefits include:

- Higher Treatment Efficiency - Higher levels of treatment in excess of 90% result from increased "sludge age" by adding fixed biological culture to the existing suspended culture.

- Process Stability - Because of the high biological solids inventory on the media, the Surfact system is less susceptible to process upset from hydraulic or organic shock loads.

- Flexibility - The Surfact process lends itself to upgrading treatment facilities because of its modular construction and low hydraulic head loss.

- Maintenance and Power Consumption - Minimal maintenance is required on the RBC units. Utilization of the existing aeration system achieves higher treatment efficiencies with minimal additional power.

- Low Initial Cost - The Surfact concept permits upgrading facilities with minimal capital expenditures. Little or no plant modification is required to install the Aero-Surf units in existing plants.

- Ease of Nitrification - By providing the proper sludge age, the Surfact process allows for nitrification to proceed without costly separate aeration, settling and sludge recycle systems. Under proper loading conditions, the Surfact process provides both BOD_5 removal and nitrification in the tankage.

- Improved Sludge Characteristics - Resulting sludge has improved settling characteristics, permitting secondary clarifiers to be designed for relatively higher overflow rates. When upgrading plants for higher degrees of treatment, additional secondary clarifier surface area will not be required.

SURFACT PROCESS

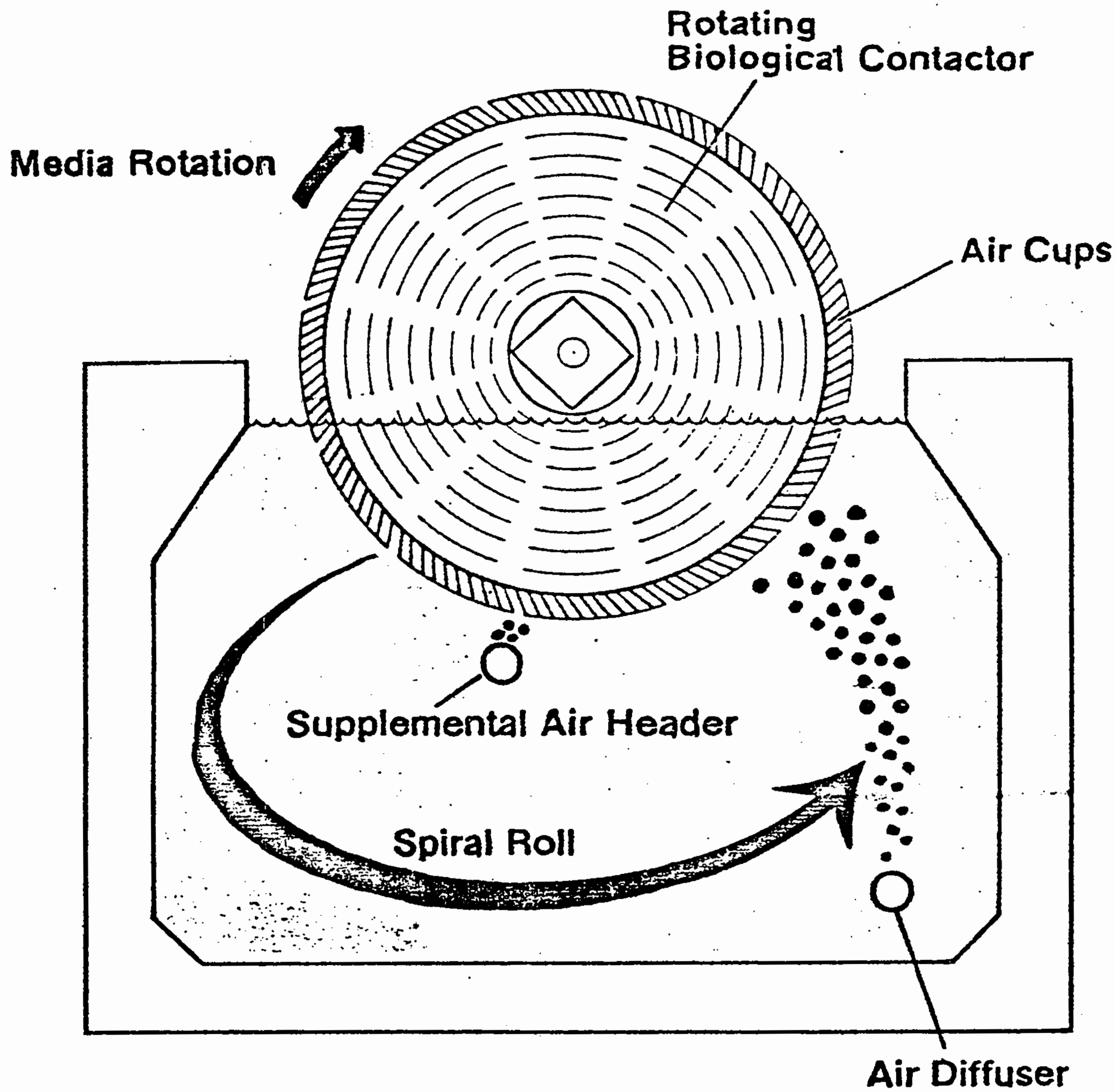


FIGURE 1

END VIEW SKETCH BIO-SURF MEDIA SUPERIMPOSED ON AN ACTIVATED SLUDGE AERATION TANK. HYDRAULIC ROLL AND AIR CAPTURE WITHIN THE TANK CAUSES ROTATION OF THE MEDIA.

AUTOTROL CORP.

PILOT PLANT OPERATIONS AND HISTORY

The Surfact process is currently undergoing pilot and full-scale testing. To date, both the full-scale demonstration plant testing and the pilot scale testing is proving the concept to be a very effective and economical means of upgrading existing activated sludge plants. The process has demonstrated capability to improve flow characteristics, improve loading rates and successfully provide single stage nitrification at loading rates in excess of those considered practical for single stage activated sludge nitrification systems.

Due partially to the success of Philadelphia Phase I studies, several other pilot facilities have been designed and placed in operation throughout the world. A detailed discussion of these investigations would be both cumbersome and incomplete at this point. However, data generated from some of these studies is included in tables and figures as noted.

A brief summary of the ongoing pilot studies appears in Table 1. Unfortunately, of the six pilot facilities, data is available fron only three. The data available from two of these units has only recently become available in a preliminary and incomplete fashion.

Philadelphia, Pennsylvania

Based on the predicted process advantages, the first pilot work was initiated by the City of Philadelphia, Pennsylvania in 1974. All testing was performed at the Northeast Water Pollution Control Plant in one of the existing aeration basins. The primary objective of inital studies was to determine whether the RBC units would rotate under the conditions of captured power from the aeration system. Other objectives included observation of the tendencies to support uniform biomasses and structural observations related to clogging of the media. The studies also included observation of the contactors rotational speeds during periods in which air was supplied by means of the special supplemental diffuser pipe directly under the Aero-Surf unit. The unit was tested at two tank locations and observations of the biological growth were made at each location. Preliminary results were favorable to the point that additional testing was conducted.

Since favorable indications were observed during the initial evaluation, additional testing of the Surfact system continued with the installation of one full-scale air driven shaft in the existing aeration tank. The air driven shaft incorporated a plurality of circumferential cups placed on the media periphery such that a major portion of shaft rotational power could be accomplished with existing aeration tank air supply and hydraulic motion. Due to the capture of aeration tank energy, minimal additional power is required to obtain shaft rotation.

After favorable operating results were obtained with the full-scale shaft, a prototype installation of 22 shafts incorporated in two passes of the existing tank was undertaken. This installation necessitated the isolation of an existing secondary clarifier to avoid the intermixing of the existing suspended activated sludge culture and the Surfact mixed liquor.

Table 1

SUMMARY OF ONGOING SURFACT PILOT STUDIES

Location	Type	Size	Unit Start-Up	Expected Completion	Data Included In Paper	Comments
Domestic						
Philadelphia, PA	Carbon Only	Full	9/77	Not Known	Yes	Heavy Industrial Influent
Milwaukee, WI (Autotrol)	Single Stage Nitrification	Bench	9/79	Winter '80	Yes	Influent Solids Control Difficult
Reno-Sparks, NV	Single Stage Nitrification	Bench	10/79	Summer '80	Yes	Phostrip process & Return Sludge Plugging
Foreign						
Japan	Carbon Only	Intermediate	Fall '79	Not Known	No	No Data Available
Japan	Single Stage Nitrification	Bench	12/79	Not Known	No.	No Data Available
Sweden	Carbon Only	Intermediate	12/79	Not Known	No	No Data Available

A settled sewage pumping station provided a known controlled flow to the isolated reactor. A separate return sludge pumping station was also provided. Because of the simplicity of Surfact retrofit, construction was completed in only sixty days.

Operation of the full-scale demonstration plant was initiated in mid-October of 1977. Stabilization of operation occurred by the end of December of 1977. The most significant start-up problems were related to return and waste sludge pumping facilities. Virtually no structural clogging problems were observed in relation to the internal structure of the disc units. By the end of eleven months of testing, several significant combinations of conditions had been observed and a generous amount of data had been recorded.

The conditions of testing and results generated appear as Tables 2 and 3, respectively. An in-depth discussion of the data accumulated during Phase I was presented by Michael D. Nelson, et al, at the 51st Annual WPCF Conference.

Following a brief down-time, a second testing phase was initiated. The aerated tank was dewatered and the existing coarse bubble diffuser system was removed from under the Aero-Surf units and replaced with a relatively high efficiency fine bubble, ceramic type diffuser system.

Data included in Tables 4 and 5 demonstrate the performance with fine bubble diffusers.

Milwaukee, Wisconsin

The pilot plant in Milwaukee, Wisconsin is operating at the Autotrol Corporate Testing Facility located at the South Shore Wastewater Treatment Facility. The unit is receiving primary clarifier effluent from the channel feeding the existing activated sludge reactors. The Surfact unit is operating in parallel with the activated sludge plant. Operating data summaries appear as Tables 6 and 7.

Two primary objectives were established for the Surfact Study in Milwaukee. These were to verify operating results from Philadelphia and to establish operational parameters for nitrification. In essence, verification of Philalelphia parameters would also establish a suitable environment for at least partial nitrification. This is due to the fact that under many of the test conditions and environmental conditions were adequate to allow the establishment of nitrifying cultures. It is believed that nitrification was not achieved at Philadelphia due to a significant degree due to toxic or inhibitory industrial influents.

The Period I study at South Shore was aimed at verifying the validity of a single stage nitrifying concept. The conclusions from this period are presented in Table 7. The overall conclusion is that single stage nitrification is viable and economically feasible. Definitive analysis of parameters and environmental conditions will be undertaken in Period III of the study.

TABLE 2

PHILADELPHIA, PA.

CHRONOLOGY OF OPERATION

PHASE I

PERIOD NO.	START DATE	STOP DATE	FLOW	SRT	DESCRIPTION AND COMMENTS
--	9/14/77	10/1/77	12 MGD	LOW	Preliminary start up and debugging.
--	10/1/77	12/4/77	12 MGD	3.3	Start up problem resolution period and debugging
I	12/23/77	1/9/78	12 MGD	3.3	Low tank suspended solids SRT and normal combined SRT. (1)
II	1/31/78	2/14/78	12 MGD	4.4	Normal tank suspended solids SRT and high combined SRT.
III	2/27/78	4/2/78	10 MGD	5.5	Normal suspended solids SRT.
IV	4/4 /78	5/12/78	15 MGD	4.4	Low tank suspended SRT and medium combined SRT.
V	5/26/78	7/16/78	10 MGD	7.0	High tank suspended SRT and high combined SRT.
VI	7/17/78	8/3/78	10 MGD	1.7	Operation without benefit of return sludge - non-equilibrium and non-acclimated culture.
VII	8/4/78	8/13/78	10 MGD	1.7	Operation without benefit of return sludge - pseudo-equilibrium and acclimated culture.

(1) Tank suspended solids refers to those solids contributed by the activated sludge process. Combined solids refers to those contributed by both the activated sludge process and air driven rotating biological contactor media surface slimes.

Source: Nelson, et al.

TABLE 3
PHILADELPHIA, PA
OPERATIONAL DATA
PHASE I

TEST PERIOD:		I	II	III	IV	V	VI	VII
START DATE		12/23/77	1/31	2/27	4/4	5/26	7/17	8/4
END DATE		1/9/78	2/14	4/2	5/12	7/16	8/3	8/13
REMOVAL:								
$SBOD_5$	(%)	83	93	93	86	95	87	88
$TBOD_5$	(%)	82	87	87	76	91	85	86
SS	(%)	72	84	84	67	82	79	83
INFLUENT:								
$SBOD_5$	(mg/1)	82	97	101	86	92	91	65
$TBOD_5$	(mg/1)	181	198	183	190	158	159	135
SS	(mg/1)	115	163	153	136	100	97	107
EFFLUENT:								
$SBOD_5$	(mg/1)	10	7	7	12	5	12	7
$TBOD_5$	(mg/1)	31	25	23	44	14	24	18
SS	(mg/1)	29	26	23	42	18	20	18
TOTAL REMOVAL:								
$TBOD_5$	(%)	85	88	89	78	92	89	91
SS	(%)	86	87	89	81	93	92	93
SECONDARY CLARIFIER								
SOR	(gpd/ft^2)	720	720	600	900	600	N/A	600
DETENTION TIME	(hrs)	2.8	2.8	3.3	2.2	3.3	N/A	3.3

TABLE 3 (Cont.)
PHILADELPHIA, PA
OPERATIONAL DATA
PHASE I

TEST PERIOD:	I	II	III	IV	V	VI	VII
START DATE	12/23/77	1/31	2/27	4/4	5/26	7/17	8/4
END DATE	1/9/78	2/14	4/2	5/12	7/16	8/3	8/13
REACTOR:							
LOADING							
(#$TBOD_5$A/1000ft^2)	8.40	8.99	6.94	10.83	6.01	6.69	5.12
(#$SBOD_5$A/1000ft^2)	4.03	4.68	4.08	5.49	3.63	3.83	2.54
(#$TBOD_5$A/1000ft^3)	67.80	73.94	57.09	89.09	49.43	55.05	42.14
(#$SBOD_5$A/1000ft^3)	32.50	38.48	33.54	45.16	29.86	31.48	20.89
F/M							
WITH RBC	0.50	0.34	0.30	0.55	0.27	0.79	0.55
WITHOUT RBC	0.99	0.56	0.75	1.74	0.43	10.50	8.28
SRT							
WITH RBC	3.28	4.42	5.54	4.44	6.97	1.07	1.72
WITHOUT RBC	1.77	2.82	2.70	1.92	4.35	0.08	0.13
DETENTION TIME (hrs)	4.0	4.0	4.8	3.2	3.3	3.3	3.3
AIR RATES: (NOTE BELOW)							
AIR/VOLUME (ft^3/gal)	0.9	1.2	1.2	0.8	1.1	0.6	0.4
AIR/LOAD							
(1000ft^3/#$TBOD_5$R)	0.65	0.70	0.87	0.51	0.90	0.49	0.40
(1000ft^3/#$SBOD_5$R)	1.35	1.35	1.48	1.01	1.49	0.45	0.81
OXYGEN/LOAD							
(#O_2A#$TBOD_5$R)	0.49	0.50	0.62	0.42	0.61	0.26	0.29
(#O_2A#$SBOD_5$R)	1.02	0.96	1.06	0.83	1.01	0.52	0.58
POWER CONSUMPTION:							
POWER/LOAD	0.61	0.59	0.76	0.50	0.71	0.46	0.58
(KWH/#$TBOD_5$R)	1.27	1.13	1.29	1.00	1.18	0.79	1.17
(KWH/#$SBOD_5$R)							

<u>NOTE</u>: All air rates quoted at a diffuser efficiency of 3%

TABLE 4
PHILADELPHIA, PA
CHRONOLOGY OF OPERATION
PHASE II

PERIOD NO.	START DATE	STOP DATE	FLOW	RPM	SRT	DESCRIPTION AND COMMENTS
I	11/14/78	12/7/78	15 MGD	(1)	4.5	
I A	11/14/78	11/29/78	15 MGD	1.5	4.2	
I B	11/30/78	12/7/78	15 MGD	1.25	4.5	
II	12/10/78	2/7/78	18 MGD	1.25	4.7	
III	3/1/79	4/25/79	6 MGD	(1)	5.7	
III A	3/1/79	4/12/79	6 MGD	1.25	6.2	
III B	3/13/79	4/25/79	6 MGD	1.25	5.9	
IV	5/26/79	6/7/79	7 MGD	1.5	4.2	
V	6/9/79	6/19/79	8 MGD	1.5	8.0	

(1) RPM during this period was changed to allow for treatment observations.

TABLE 5
PHILADELPHIA, PA
OPERATIONAL DATA
PHASE II

TEST PERIOD:		I	IA	IB	II	III	IIIA	IIIB	IV	V
START DATE		11/14/78	11/14/78	11/30/78	12/10/78	3/1/79	3/1/79	3/13/79	5/26/79	6/9/79
END DATE		12/7/78	11/29/78	12/7/78	3/7/79	4/25/79	4/12/79	4/25/79	6/7/79	6/19/79
REMOVAL										
$SBOD_5$	(%)	91	91	92	80	92	91	94	93	97
$TBOD_5$	(%)	87	86	88	64	88	87	90	87	91
SS	(%)	80	79	82	53	85	85	85	83	84
INFLUENT:										
$SBOD_5$	(mg/l)	95	93	101	89	92	94	89	69	79
$TBOD_5$	(mg/l)	183	181	187	168	176	181	165	129	144
SS	(mg/l)	148	148	149	158	220	236	168	107	116
EFFLUENT:										
$SBOD_5$	(mg/l)	8	8	8	17	7	7	5	5	2
$TBOD_5$	(mg/l)	24	25	21	57	20	22	16	17	13
SS	(mg/l)	28	29	26	68	25	26	22	18	17
TOTAL REMOVALS:										
$TBOD_5$	(%)	89	88	91	75	91	90	93	92	94
SS	(%)	90	90	92	73	90	90	91	93	93
SECONDARY CLARIFIER:										
SOR	(gpd/ft^2)	900	900	900	891	360	360	360	420	480
DETENTION TIME	(hrs.)	2.2	2.2	2.2	2.5	5.5	5.5	5.5	4.7	4.1

TABLE 5 (cont)
PHILADELPHIA, PA
OPERATIONAL DATA
PHASE II

TEST PERIOD:	I	IA	IB	II	III	IIIA	IIIB	IV	V
START DATE	11/14/78	11/14/78	11/30/78	12/10/78	3/1/79	3/1/79	3/13/79	5/26/79	6/9/79
END DATE	12/7/78	11/29/78	12/7/78	2/7/79	4/25/79	4/12/79	4/25/79	6/7/79	6/19/79
REACTOR:									
LOADING									
($\#TBOD_5A/1000ft^2$)	10.41	10.30	10.63	11.47	8.33	8.40	8.24	7.53	9.62
($\#SBOD_5A/1000ft^2$)	5.42	5.28	5.72	6.05	4.35	4.38	4.43	4.01	5.26
($\#TBOD_5A/1000ft^3$)	85.64	84.74	87.43	94.34	66.00	67.78	61.64	56.30	71.95
($\#SBOD_5A/1000ft^3$)	44.59	43.44	47.05	49.76	34.47	35.34	33.14	29.98	39.34
F/M									
WITH RBC	0.47		0.47	0.57	0.27	0.27	0.28	0.35	0.46
WITHOUT RBS	0.75	0.76	0.71	1.17	0.35	0.36	0.33	0.44	0.58
SRT									
WITH RBC	4.58		4.54	4.66	5.74	6.20	5.85	4.20	8.00
WITHOUT RBC	2.80	2.69	3.01	2.56	4.56	4.79	4.98	3.39	6.36
DETENTION TIME (hrs.)	3.2	3.2	3.2	2.7	4.0	4.0	4.0	3.4	3.0
AIR RATES: (NOTE BELOW)									
AIR/VOLUME(ft^3/gal)									
AIR/LOAD									
($1000ft^3/\#TBOD_5R$)									
($1000ft^3/\#SBOD_5R$)									
OXYGEN/LOAD									
($\#O_2A/\#TBOD_5R$)									
($\#O_2A/\#SBOD_5R$)									
POWER CONSUMPTION:									
POWER/LOAD									
($KWH/\#TBOD_5R$)									
($KWH/\#SBOD_5R$)									

An error in the reporting and computation procedure has prevented inclusion of this data at time of this printing.

NOTE: All air rates quoted at a diffuser efficiency of %.

TABLE 6
MILWAUKEE, WISCONSIN
CHRONOLOGY OF OPERATION

PERIOD NO.	START DATE	STOP DATE	FLOW	DESCRIPTION AND COMMENTS
-	5/16/79	7/16/79	Variable	Preliminary Start-up and debugging
I	7/26/79	8/20/79	950 gpd	Verification of single stage nitrification capability with Surfact. Process successfully completed.
II	9/4/79	Summer80	Variable	Verify Philadelphia Phase I results and determine approximate design parameters.
III	Fall 80	Summer81	Variable	Determine design parameters for single stage nitrification facilities.

TABLE 7
MILWAUKEE, WISCONSIN
CONDENSED OPERATIONAL DATA SUMMARY
FOR
TEST PERIOD I
SINGLE STAGE NITRIFICATION VARIFICATION

Average daily flow = 950 gpd
Operational Mode = diurnal flow reariation
Maximum flow = 1.200 gpd
Minimum flow = 290 gpd
Average RAS flow = 570 gpd
Average WAS flow = 50 gpd

Removal Efficiencies:
$TBOD_5$ ~~= 94%~~
$SBOD_5$ = 94%

NOTE: Due to operational difficulties with secondary clarifier samples are decanted 24 hour composits of Surfact Tankage Effluent. Biomass WAS and RAS samples were grab.

Av. Influent Characteristics:

$TBOD_5$	=	235 mg/1
$SBOD_5$	=	93 mg/1
TSS	=	236 mg/1
SCOD	=	135 mg/1
NH_3-N	=	26 mg/1
HCO_3	=	250 mg/1
PH	=	7.7
Temp.	=	19^{o} C
D.O.	=	0.9 mg/1

Av. Effluent Characterist.:

$TBOD_5$	=	15 mg/1
$SBOD_5$	=	6 mg/1
NH_3-N	=	5 mg/1

Plant Operating Conditions:
F/M overall = 0.29

Reactor:
MLSS = 3,051 mg/1
MLSS = 2,197 mg/1

SHAFT:
MLSS = 1,100 mg/1
MLSS = 891 mg/1

Test Period II is currently ongoing. Results appear to fall within the ranges predicted by Philadelphia. However, several conclusions and recommendations are becoming clear. Specifically, the process is seeing a break in $SBOD_5$ in the front end of the reactor basin, indicating that step feed or step aeration would allow for a significant increase in flow and load capability. Concurrent with these increases, effluent quality would be improved.

Reno-Sparks, Nevada

The pilot plant in Reno, Nevada is operating at the Reno-Sparks Wastewater Treatment Facility. The unit is operating in a similar fashion to the unit at South Shore. One major exception is noted in that phosphorus stripping process called PHOSTRIP is in use. The return method from this process appears to be affecting the activated sludge significantly but does not appear to have a major or adverse impact on the fixed film solids inventory. The effect on the reactor basin solids inventory is a significant reduction in concentration.

The activated sludge system appears to have entered a recovery mode shortly after initiation of the PHOSTRIP process. Further investigation into the performance reduction indicates possible plugging of solids flow lines within the return sludge system. Proper maintenance of these lines may significantly increase performance and effluent quality.

The data generated, although not presented here, indicates a significant stabilizing effect of the fixed film. In conjunction with the data from South Shore and Philadelphia, this data indicates an inherent capability of Surfact systems to provide stable treatment under conditions which would otherwise shut down conventional activated sludge systems.

CONTACTOR INSTALLATION

Several methods of contactor installation have been proposed. The two primary categories are shafts mounted in a "longitudinal" axis and those mounted in a "transverse" axis. Several advantages are common to both systems.

Longitudinal placement may be defined as locating an RBC unit parallel to the longest dimension of the tankage as shown in Figure 1. Thus, in a standard, spiral roll activated sludge tank, the shafts would be mounted end to end in such a fashion as to take advantage of hydraulic roll concurrently with air capture. This physical placement may be accomplished by means of either supporting the shafts off the bottom of the tank, or by hanging a cantilever across the existing wall structure.

Several advantages are seen for this type of installation. The shafts and mounting structure could be assembled exterior to the existing tankage. This allows for installation of a complete shaft and support structure in one crane lift. This reduces the amount of time required for installation and the length of time an individual basin would be out of service.

Transverse installation could be defined as mounting shafts 90° from the longest dimension of the tank. In essence, the simplest fashion of installing shafts is to secure the bearings to existing wall structures. This installation allows for the advantages of completely preparing the tank for installation without dewatering, shaft installation without a supplemental structure and a requirement for one single crane lift. Figure 2 is an example of this installation method.

Length of time out of service is of major importance to most wastewater treatment facilities requiring upgrading. In many cases the process of removing tanks from service seriously hinders effluent quality and must be avoided.

GENERAL DISCUSSION

Indepth study of the data generated indicates that Surfact systems function in a similar fashion to activated sludge processes. Similarities can be seen in SRT relationships, effectiveness of RBC biomass versus suspended biomass, oxygen uptake rates, F/M relationships, retention time and sludge characteristics.

Percent Removal Versus SRT

Figure III shows the relationship between SRT and percent removal of BOD_5, soluble BOD_5 and suspended solids. The relationship depicted is similar to that expected for activated sludge with percentage removal increasing with an increasing SRT. Significant increases in percentage removal occur below an SRT of 5 to 6 days. Above six days the expected leveling off occurs. Percentage removal of soluble BOD ranges from 83 - 97 percent with the majority of data in excess of 90 percent.

The consistent performance depicted is indicative of a strong biological process.

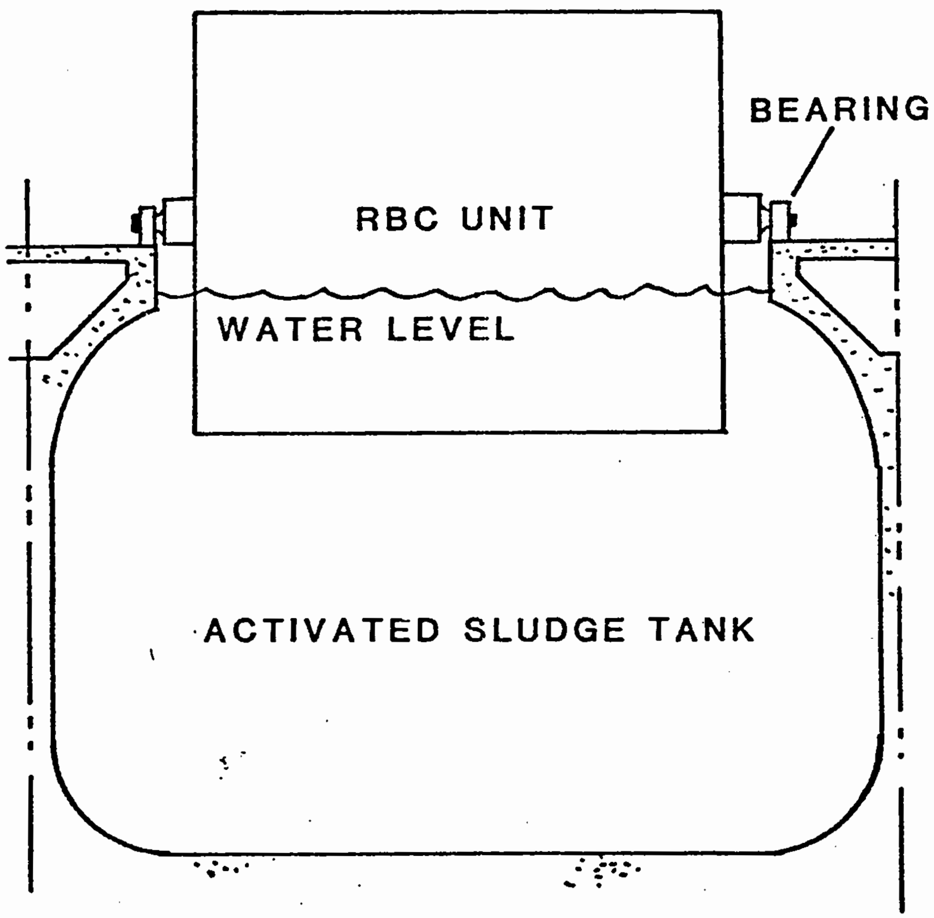
Effectiveness of Fixed Growth

One of the questions of primary importance is a determination of biological growth effectiveness. In specific, a determination of relative effectiveness of the suspended growth versus that maintained on the RBC surfaces is required. Table 7 contains data comparing operation during a period of no sludge return to periods high suspended mass concentrations. The indication is that fixed film masses are at least as effective as suspended masses.

Uptake Rate Studies - Biofilm Versus Suspended Culture

Several studies have been performed on the oxygen uptake rates of fixed film versus suspended film reactors. Table 8 is a compilation of the k rate determinations conducted at Philadelphia. As shown, the specific oxygen uptake rates are roughly equivalent. The work done by the disc and suspended cultures is thus shown to be roughly equivalent on a unit mass basis.

FIGURE 2



SURFACT PROCESS WITH RBC UNIT
INSTALLED IN TRANSVERSE POSITION

FIGURE 3

SURFACT PERCENT REMOVAL VS SRT

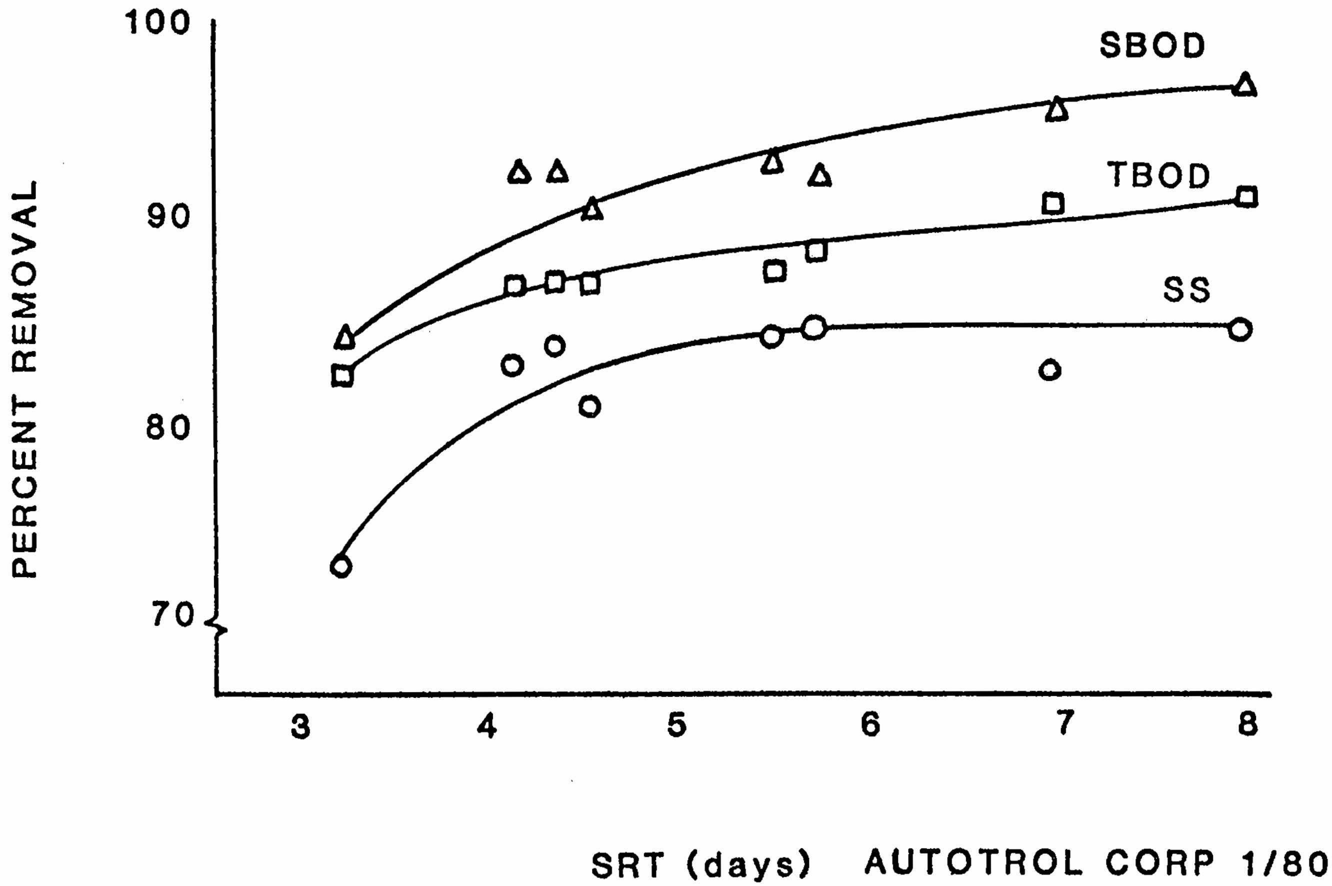


Table 8

Philadelphia, Pennsylvania

Biological Mass Effectiveness Comparison

PHASE	II	II	I
PERIOD	II	VI	VI
$SBOD_5$ Applied (#)	13,361	12,460	6,755
$SBOD_5$ Effluent (mg/1)	17	8	8
% $SBOD_5$ Removal	80	90	87
MLVSS - Total (#)	45,786	36,812	19,566
MLVSS - RBC (5)	44	35	93
# $SBOD_5$ REM/#MLVSS	0.24	0.31	0.30
F/M with RBC	0.57	0.66	0.69

Table 9

Philadelphia, Pennsylvania

Uptake Rate Studies

PERIOD	FLOW	SRT	LBS BOD/ 1000 cf	Oxygen Uptake Ratio mg O_2/mg VSS/hr Suspended	Oxygen Uptake Ratio mg O_2/mg VSS/hr Biofilm	Oxygen Uptake Ratio mg$_2$O /l/hr Suspended	Oxygen Uptake Ratio mg$_2$O /l/hr Biofilm	K rate K 10 Suspended	K rate K 10 Biofilm
III	10	5.5	57	.022	.022	35	33		
IV	15	4.4	89	.063	.048	76	78		
V	10	7.0	49	.039	.035	74	39	.160	.174

Data generated from several RBC installations confirms the uptake rates on fixed film reactors presented here. Sources of data are found in the history and bibliography prepared for this seminar by Ron Antonie and Dannette Lank.

Volatile Mass Comparison

In comparing the concentrations of volatile masses on the RBC shafts to the mixed liquor there is an average of 7.9 percent more volatile mass per pound on the disc surface area than appearing in the mixed liquor. Considering all cases, there is a range of 0 - 18 percent improvement in the volatile solids per pound of total solids maintained on the RBC as opposed to the suspended mass maintained within the aeration tank. Therefore, from a volatile solids concentration standpoint, the disc solids appear to be more viable. Based on Table 10, treatment capacity appears to be independent of the MLVSS location and primarily dependent on total biological solids inventory within the reactor system.

Thickening Test Results

A series of pilot-scale thickening tests were conducted on the waste activated sludge generated from the Philadelphia Surfact Facility. The investigations included both gravity and dissolved air floatation thickening. Several runs were performed on both secondary sludge and combined secondary and primary sludge.

The gravity thickening results fell within the normal expected ranges for activated sludge plants. Average results indicated concentration in execess of 2% with a 3 hour detention time.

Results from the floatation thickening study were impressive. Secondary sludge, introduced at 1.0% concentration, thickened to 4.5% without addition of supplemental chemicals. The resulting capture rate was 99% at an air to solids ratio of 0.008. These ranges are within extremely economical operating ranges while performing at well above expected concentration increase and capture rates.

CONCLUSION

Conclusions drawn from the data presented show the Surfact Process to be superior to a standard activated sludge process based on treatment reliability, adjustment to varying flow and load conditions and reduced dependence upon secondary clarifier operations. The process can readily be applied to overloaded activated sludge systems or to the construction of new treatment facilities. Upgrading is enhanced through the minimal construction time required, ease of operation and low power consumption.

The combination of fixed and suspended solid growth within a single reactor tank provides a significant increase in biological solids without increasing solids loads on the secondary clarifier. Thus, the capability for treatment is not dependent totally upon the capture, settling and return of solids to

Table 10

Philadelphia, Pennsylvania

Comparison of Treatment Based on MLVSS

PHASE	I	I	II
PERIOD	II	III	IV
F/M with RBC	0.34	0.30	0.35
$SBOD_5$ Removal (%)	93	93	93
MLVSS-Suspended (mg/l)	2163	1577	2184
MLVSS-Shaft (mg/l)	1373	1512	560
MLVSS-Shaft (% of Total)	39	49	29
$SBOD_5$ Applied (#)	9708	8423	8056
Effluent $SBOD_5$ (mg/l)	7	7	5

the activated sludge process. The result can be a higher treatment level at the same rate of flow or the same level of treatment at a significantly higher rate of flow or a combination of increased flow and treatment level.

The Surfact Process, by merging two systems, increases the efficiency and stability of the existing activated sludge system. Inherent process benefits include:

- Higher Treatment Efficiency
- Increased Process Stability
- Increased Flexibility
- Reduced Maintenance and Power Consumption
- Low Initial Cost
- Ease of Nitrification
- Improved Sludge Characteristics
- Increased Mass Without Increased Oxygen Provision

Ease of nitrification has been demonstrated at the South Shore Wastewater Treatment Facility. The use of single stage reactors enhances the upgrading of existing activated sludge systems. This concept is based upon carbon removal within the existing tank structure and nitrification primarily attained by the fixed film.

The data related to single stage nitrification and high load periods indicates that a Surfact system is significantly more stable than a standard activated sludge facility. Based on loading parameters, the Surfact system can be operated well beyond the bounds of standard activated sludge operation.

PLANT SCALE INVESTIGATION OF RBC PROCESS SUPPLEMENTAL AERATION

By

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1. BACKGROUND

The advanced wastewater treatment plant at Alexandria, Virginia includes preliminary treatment, primary settling, Rotating Biological Contactor (RBC) secondary treatment, carbon adsorption, phosphorus removal, filtration, ion exchange for nitrogen removal and chlorination. A process flow diagram of the plant is shown on Figure 1. This is the largest plant in operation in the U.S.A. employing RBC process at a design capacity of 54 mgd and was designed based on pilot study data. The secondary treatment consists of 56 motor driven RBC shafts in 14 tanks of 4 stages each. The design criteria is shown in Table 1.

The RBC process was placed on line in 1977. The operating results indicated less than expected performance. The RBC discs developed a very thick biological growth white in color. This was identified as beggiatoa. The beggiatoa predominance was evident in all stages of the RBC tanks. On the lead stages, where the discs were covered with beggiatoa, a black under layer was observed. It appeared that this under layer was not aerobic due to the thickness of the growth.

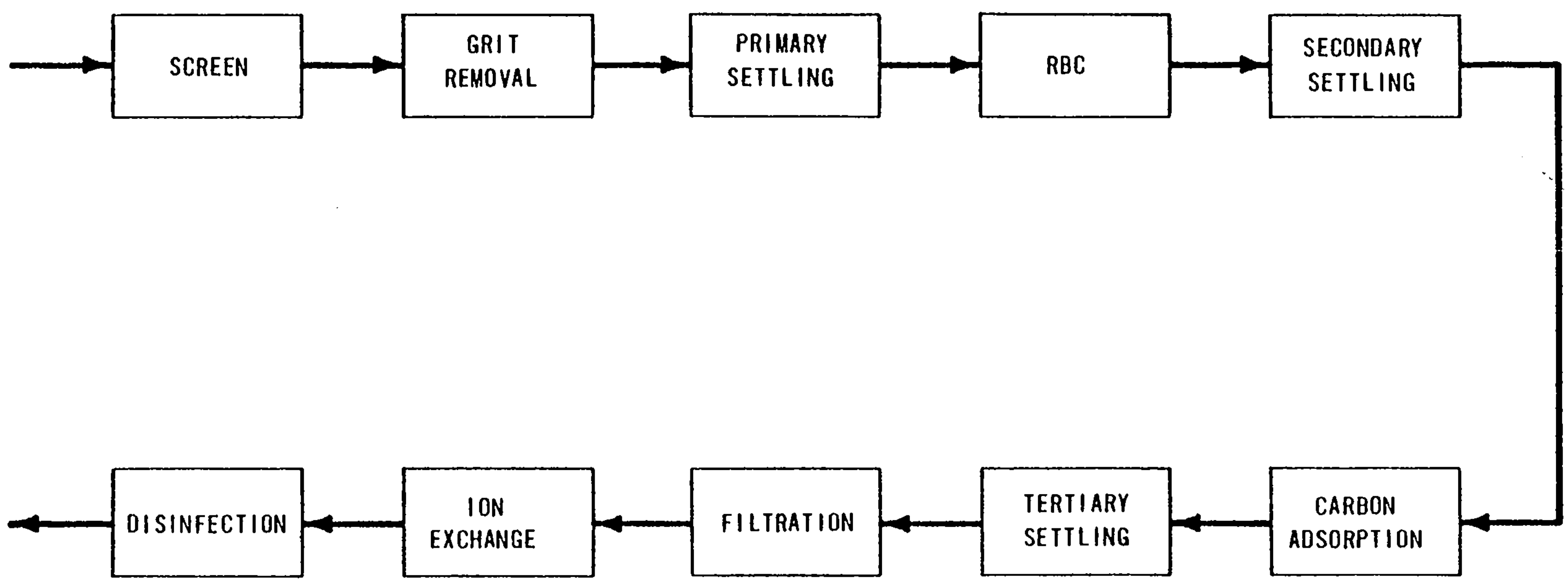

FIGURE I
ALEXANDRIA ADVANCED WASTEWATER TREATMENT PLANT
PROCESS FLOW DIAGRAM

TABLE 1

<u>ASA Plant Design Bases</u>
(mg/l)

Flow 54 mgd

Inf. BOD_5	220	Eff. BOD_5	3
Inf. $SBOD_5$	30	Eff. $SBOD_5$	1
Inf. N	42	Eff. N	1
Inf. P	15	Eff. P	0.2

The start-up data are compared with the pilot study data and design bases in Table 2. The average influent $SBOD_5$ during start-up period was significantly higher than during the pilot study period. It is noted that the $SBOD_5$ during start-up averaged 13 mg/l higher than pilot study data. Any increase in RBC effluent $SBOD_5$ would be removed by carbon adsorption process which would, therefore, require more than anticipated level of performance by the carbon columns and more frequent regeneration of the carbon bed.

An experimental program was developed to investigate the apparent differences in the RBC performance observed during the start-up and pilot study periods. This paper presents the information on tests conducted and the findings and conclusions of this dtudy.

2. EXPERIMENTAL PROGRAM

The experimental program was designed to eliminate unlikely hypotheses and converge on an approach to improve treatment efficiency. The first part of the experimental program comprised the following:

- o $SBOD_5$ tests on composite samples before and after treatment with coagulant chemicals. The effects of 100 mg/l $FeCl_3$ and 200 mg/l alum were studied.
- o Batch carbon adsorption tests on 24-hour composite samples of the RBC effluent. The initial and final SCOD concentrations were measured after two hours of contact with Calgon "Filtrasorb 400" activated carbon, 10 grams per liter.

The second part of the experimental program included on-line testing as follows:

The north RBC tanks were used as control units while the south RBC tanks were used as test units to study the operational modifications shown on Figure 2. The modifications were:

Tank No.	Description of Experiment
2	Addition of hydrogen perioxide to first stage (5 mg/l)
4	Supplemental aeration in each stage (180 cfm)
8	Operation in 1-1-1-1 mode
10	Operation in 3-1 mode
12	Operation in single-stage mode

The supplemental aeration equipment could not be installed in time to collect data in parallel with other on-line experiments conducted. The effect of supplemental aeration was studied at a different time on a long term basis subsequent to the completion of other on-line tests.

TABLE 2

RBC Pilot, Design and Start-Up Data

Parameter	Pilot	Design	Start-Up
Flow, mgd		54	30
Loading			
GPD/1,000 Ft^2	11.5	9.6	5.4
PPD $SBOD_5$/1,000 Ft^2	3.6	2.5	2.7
Influent $SBOD_5$, mg/l	38	30	60
Effluent $SBOD_5$, mg/l	12	12	25
Chemicals Addition	Alum	Alum	None

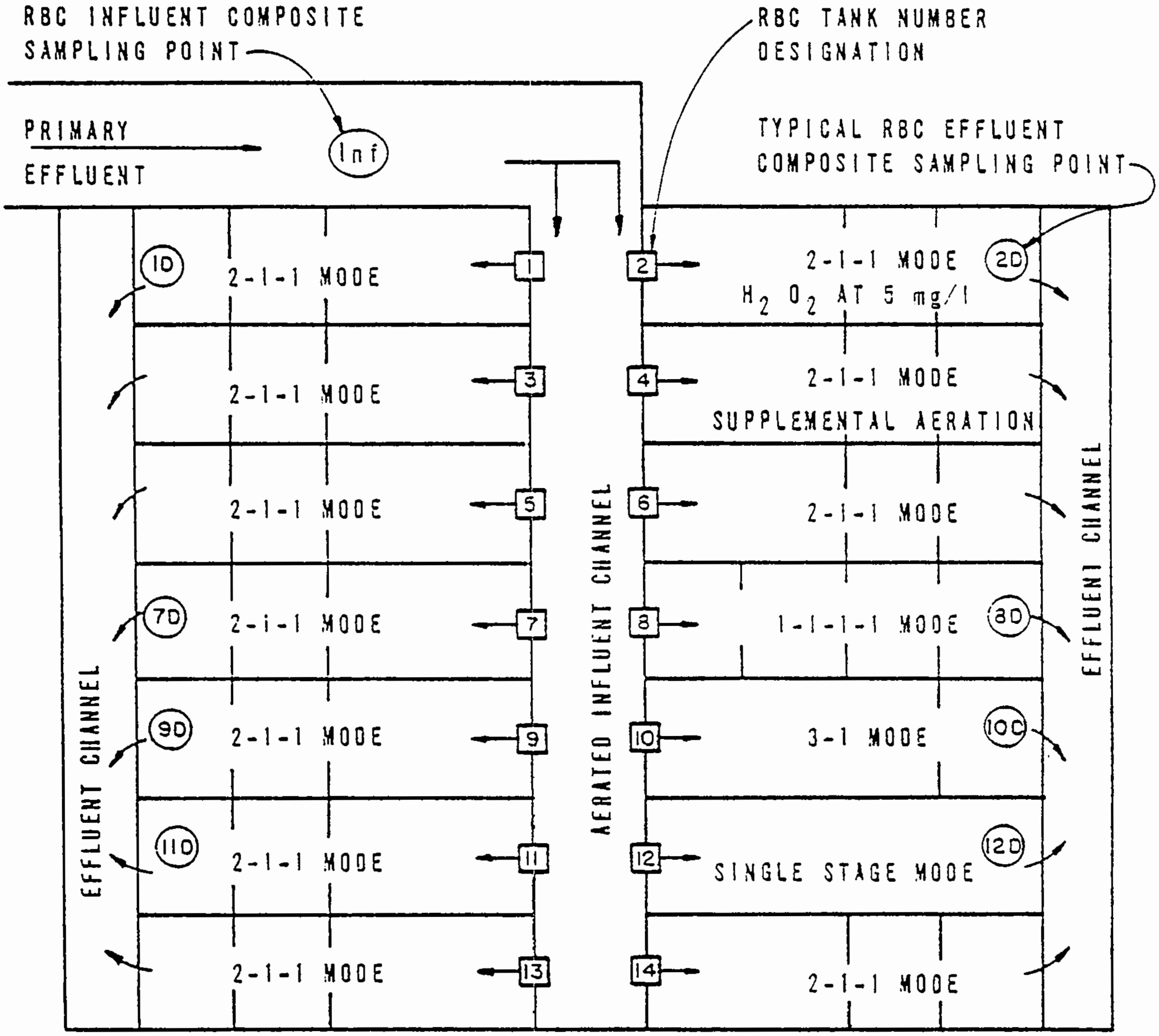

FIGURE 2
RBC ON-LINE EXPERIMENTAL SET-UP

During the test period, 24-hour composite samples were collected by automatic samplers at each sampling point shown on Figure 3. The samples were analyzed for $SBOD_5$, BOD_5, COD and suspended solids concentrations.

3. RESULTS AND DISCUSSIONS

a. Coagulation Tests

The $SBOD_5$ concentrations of coagulated and uncoagulated samples are shown in Table 3. There are significant differences between the values. The $SBOD_5$ concentrations in the samples analyzed were reduced by coagulation with both ferric and alum. Data are summarized as follows:

	Ferric	Alum
Number of sets	5	9
Reduction of $SBOD_5$, mg/l:		
Mean	16	11
Maximum	21	28
Minimum	12	4

Based on the mean of all samples, the reduction in $SBOD_5$ by coagulation with ferric was greater by 5 mg/l than by coagulation with alum. The expected $SBOD_5$ reduction by ferric coagulation is on the order of 15 mg/l.

b. Carbon Adsorption Tests

The SCOD concentrations were used to indicate the performance of carbon adsorption. The $SBOD_5$ concentrations are not considered reliable indicators in the low ranges expected. The SCOD values observed before and after contact with carbon in four tests are shown in Table 4. In all of the tests, the SCOD concentration was significantly reduced by carbon adsorption; the mean reduction was 77 percent. Based on a BOD/COD ratio of 0.33 (as was observed during the pilot work), an initial COD concentration of 40 mg/l and a 75 percent reduction, the estimated $SBOD_5$ in the effluent would be as follows:

Initial SCOD, mg/l	40
Percent reduction	75
Final SCOD, mg/l	10
BOD_5/COD Ratio	0.33
Estimated Final $SBOD_5$, mg/l	3

c. On-Line Testing

In this section, all on-line test data except the supplemental aeration test are discussed. The supplemental aeration test data are discussed in the next section.

An overall summary of the $SBOD_5$ results on composite effluent samples from the RBCs in each stage configuration, along with their control, is presented in Table 5. Hydraulic and organic loadings to all RBC tanks for this test period (Jan. 17 - Feb. 17, 1978) are summarized as follows:

TABLE 3

Coagulation Test Data Summary

Date	Sample Point	$SBOD_5$, mg/l Before Coagulation	After Coagulation	Reduction
$FeCl_3$, 100 mg/l				
2/4/78	RBC Inf.	58	40	18
	RBC Eff. - 1D	41	29	12
	RBC Eff. - 2D	30	16	14
2/5/78	RBC Inf.	44	23	21
	RBC Eff. - 2D	28	12	16
	Mean of all samples	40	24	16
Alum, 200 mg/l				
2/13/78	RBC Inf.	73	45	28
	RBC Eff. - 2D	37	24	13
2/14/78	RBC Inf.	39	35	4
	RBC Eff. - 1D	32	14	18
	RBC Eff. - 2D	26	18	8
2/15/78	RBC Inf.	63	53	10
	RBC Eff. - 1D	42	34	8
	RBC Eff. - 2D	28	24	4
2/16/78	RBC Eff. - 9D	23	12	9
	Mean of all samples	40	29	11

TABLE 4

Activated Carbon Test Data Summary

Date	Sample Point	SCOD[1], mg/l Initial	SCOD[1], mg/l Final	SCOD[1], mg/l Reduction	Percent Reduction
2/13/78	2D	44	3	41	93
2/14/78	2D	30	14	16	53
2/15/78	2D	43	14	29	67
2/16/78	9D	39	6	33	85
Mean		39	9	30	77
Median		43	9	34	79

(1) Coagulated with alum.

TABLE 5

Summary of On-Line Data[1]

Sample Point	Stage Configuration	No. of Samples	$SBOD_5$ mg/l						Remarks
			Max.	Min.	Range	Mid-range	Mean	Median	
Inf.	NA	20	93	41	52	67	61	55	
1D	2-1-1	20	71	24	49	48	47	44	Control: 2D
2D	2-1-1	20	92	12	80	52	41	36	Peroxide
7D	2-1-1	20	80	19	61	50	37	33	Control: 8D
8D	1-1-1-1	20	74	19	55	47	35	32	4-Stage Mode
9D	2-1-1	17	60	21	39	41	36	31	Control: 10D
10D	3-1	1	57	19	38	38	37	35	2-Stage Mode
11D	2-1-1	20	55	27	28	41	37	35	Control: 12D
12D	Single	20	55	19	36	37	36	34	Single-Stage Mode

(1) No coagulant chemicals were added to the plant or to the samples before analysis.

	Max.	Min.	Mean
Sewage quantity, mgd	54	36	44
Hydraulic loading, gpd/sf	9.6	6.4	7.9
Influent $SBOD_5$, mg/l	93	41	61
Organic loading, ppd $SBOD_5$/ 1,000 sf	6.5	2.6	4.0

The suspended solids concentrations were widely scattered. This is attributed to nonrepresentative solids samples obtained by the sampling equipment. For this reason, it is considered that the $SBOD_5$ concentrations are the primary performance indicators in this study.

The average effluent $SBOD_5$ concentration from the RBCs on the 1-1-1-1, 3-1 and single stage configurations and the respective control units which were operated on the 2-1-1 mode did not differ significantly, ranging from 35 to 37 mg/l. These results indicate that in terms of $SBOD_5$ removal, the performance of the RBCs is not affected by stage configuration.

Tank No. 2, where peroxide was added to the first stage, showed a mean effluent $SBOD_5$ of 41 mg/l compared to 47 mg/l for its control, Tank No. 1. The peroxide unit performed marginally better than its control unit with about 10 percent difference in $SBOD_5$ removal.

The poorer performance of Tanks Nos. 1 and 2 (average effluent $SBOD_5$ of 41 and 47 mg/l) compared to Tanks Nos. 7-12 (average effluent $SBOD_5$ of 35 to 37 mg/l) may be due to additional aeration of sewage along the RBC influent channel. It has been observed all along that the biology at the west end tanks included greater numbers of the higher order growths and less beggiatoa growth than the east end tanks. It apperas, therefore, that aeration may help in improving the biological growth and enhancing $SBOD_5$ removal.

Peroxide treatment did affect the biology observed on the first stage of RBC Tank No. 2 compared to its control tank. Before the test program began, the growth on the RBC media in the first pass of Tank No. 2 appeared similar to that observed at Tank No. 1, with patches of white biomass, beggiatoa, mixed with brown growth on top of a black underlayer. After two days of peroxide treatment, the white patches disappeared from the peripheral layers of the media where very thin areas and bare spots were observed. After ten days the white patches had entirely disappeared from the leading passes of Tank No. 2 and the bare spots were replaced with anew thin layer of brown growth.

The peroxide addition appears to have had a positive effect on biology as well as $SBOD_5$ removal in the tank where the beggiatoa growth was noted. While the improvement in treatment efficiency due to peroxide could not be quantified based on these results, it is anticipated that peroxide addition would control the beggiatoa growths especially during summer conditions.

d. Supplemental Aeration

The supplemental aeration test data analysis has been divided into two phases, related to the addition of $FeCl_3$ to the primary settling tanks, as follows:

- o No chemical addition
- o Daily chemical addition

(1) No Chemical Addition

During this phase, the effect of single stage aeration versus four stage aeration was studied. The average RBC effluent $SBOD_5$ values for the control and single stage supplemental aeration tanks were 19 and 16 mg/l, respectively. The data shows some marginal improvement in $SBOD_5$ removal with single-stage supplemental aeration. When the aeration was extended to all four stages of the RBC, the average effluent $SBOD_5$ decreased to 11 mg/l, while the control unit produced an effluent with $SBOD_5$ of 18 mg/l for the same period. The daily influent and effluent $SBOD_5$ values for this phase are shown on Figure 3. It is noted that fluctuations in the effluent $SBOD_5$ concentration in the supplemental aeration tank had been dampened considerably compared to the control unit which follows influent $SBOD_5$ closely.

The daily $SBOD_5$ removals are correlated to the applied load for the control and supplemental aeration units on Figures 4 and 5. It is evident from these plots that supplemental aeration unit performance was superior to the control unit. The improvement in performance is higher when aeration is provided to all four stages than single stage aeration. Therefore, only four stage aeration was studied in the subsequent phases.

(2) Daily Chemical Addition

During this phase, 20 mg/l of ferric chloride as Fe^{3+} and 0.5 mg/l polymer were added to the primary settling tanks. The influent $SBOD_5$ to RBC's averaged 32 mg/l, over 150 days of data collected. The average effluent $SBOD_5$ for the control and supplemental aeration units for the same period were 14 and 6 mg/l, respectively. Some statistical data on these values are shown in Table 6. The daily influent and effluent $SBOD_5$ values for the two units are plotted on Figure 6.

The $SBOD_5$ removal versus applied load relationship for the two units is shown on Figure 7. It is to be noted that the specific removal rate expressed in pounds of $SBOD_5$ removed per unit surface RBC area per unit time is clearly higher for the supplemental aeration unit than for the control unit. It is also observed that the difference in removal becomes more pronounced as the loading increases as indicated by the diverging regression lines shown on Figure 7. The correlation coefficients as noted on Figure 7 for the two lines are 0.83 and 0.93 testifying to the excellent correlation between the two parameters plotted.

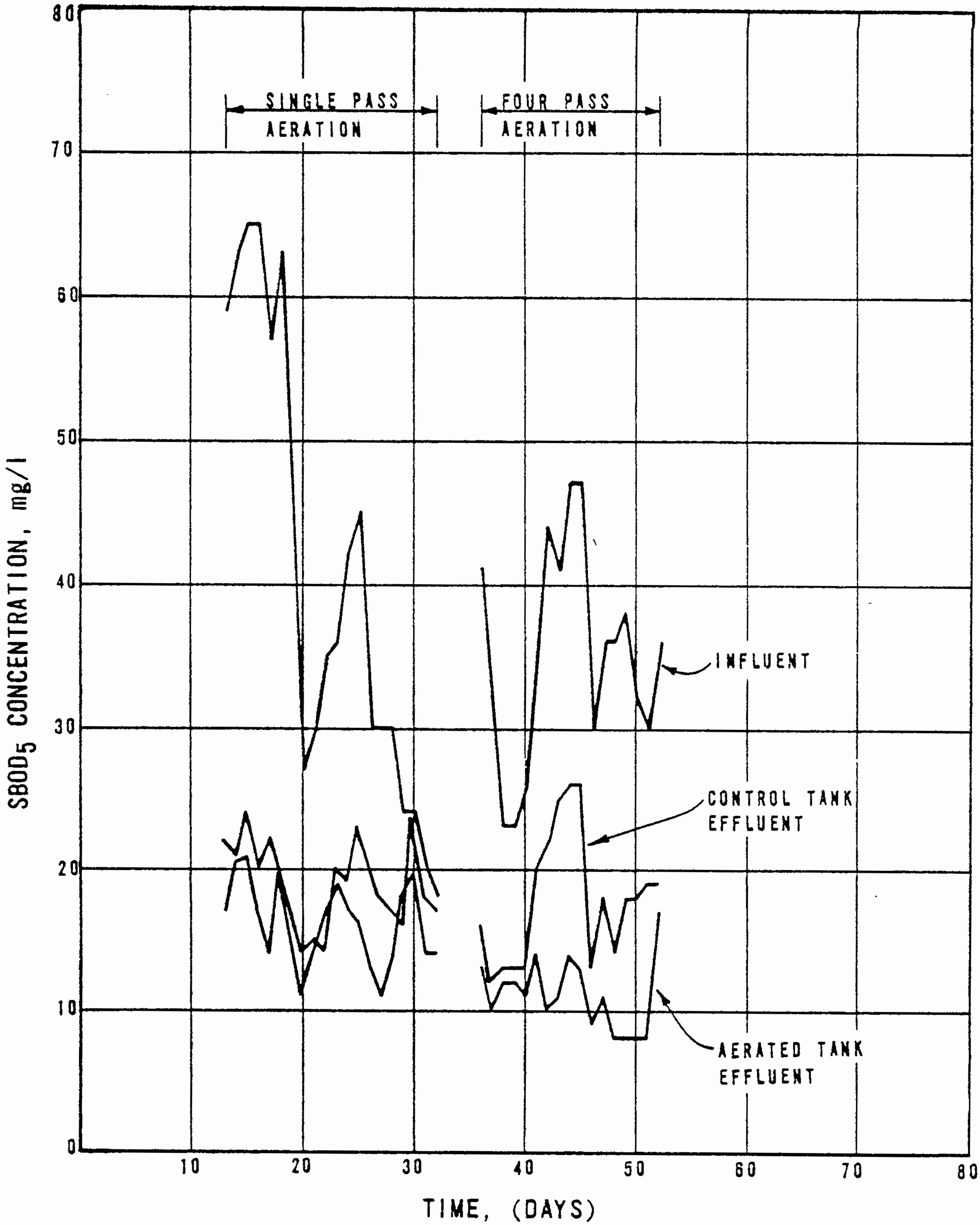

FIGURE 3
INFLUENT AND EFFLUENT $SBOD_5$ CONCENTRATION
WITH NO CHEMICAL FEED

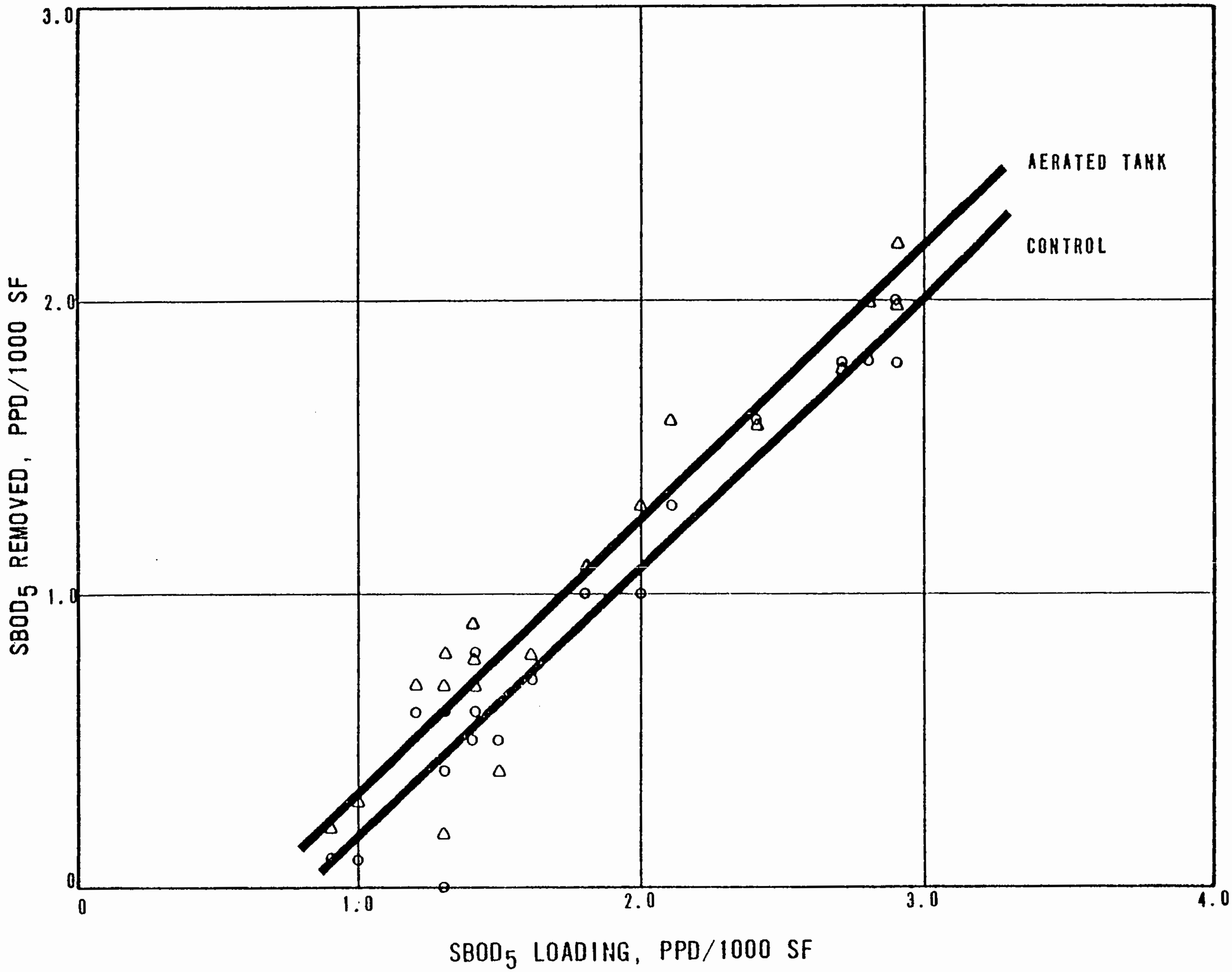

FIGURE 4
$SBOD_5$ REMOVED VERSUS $SBOD_5$ LOADING
FOR SINGLE STAGE AERATION

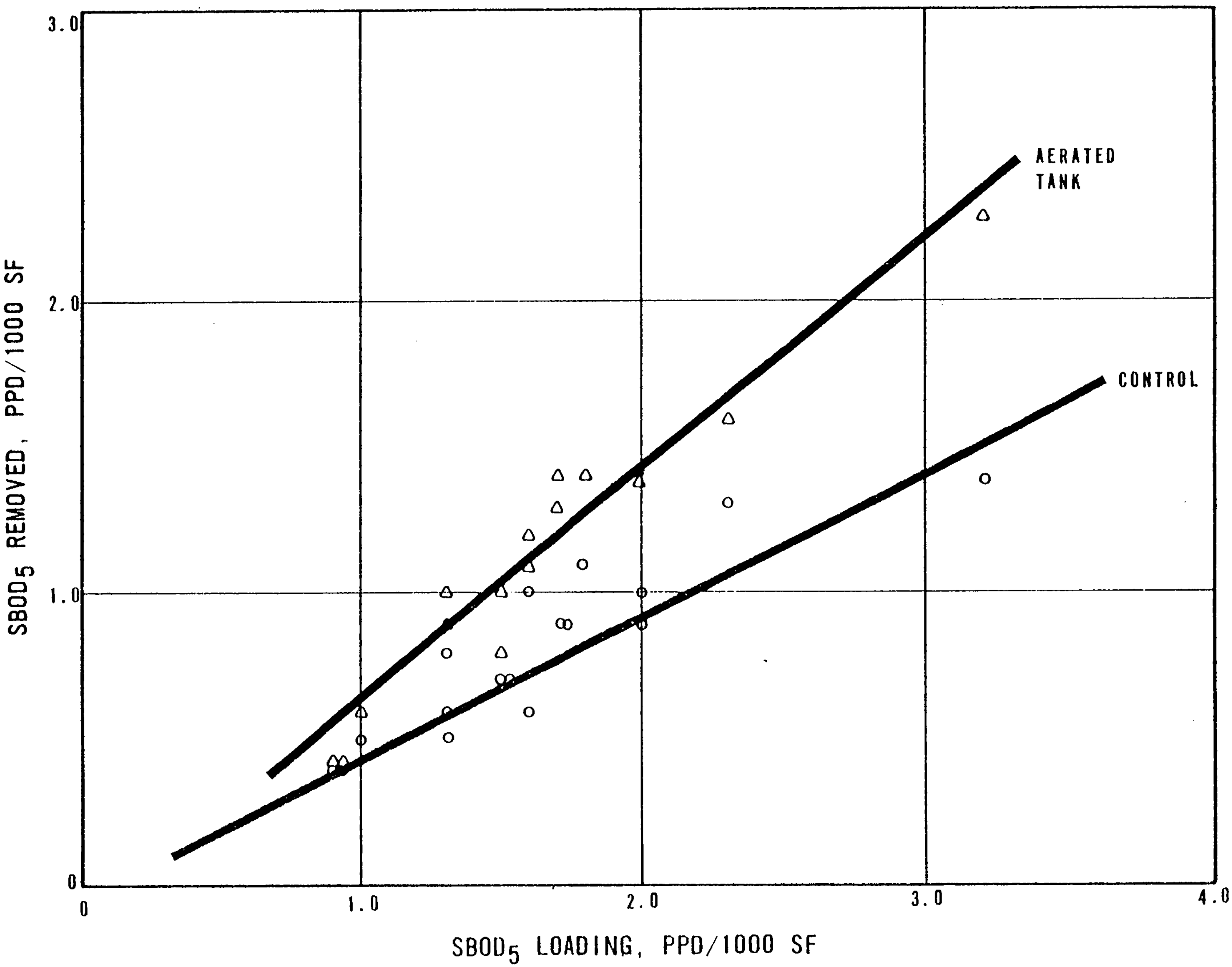

FIGURE 5

$SBOD_5$ REMOVED VERSUS $SBOD_5$ LOADING FOR FOUR STAGE AERATION

TABLE 6

On-Line Supplemental Aeration Test Data with Daily Chemicals Addition

Parameter	Minimum	Maximum	Mean	Standard Deviation	Mode	Median
Flow	18	51	27	5	28	27
Influent $SBOD_5$ Conc., mg/l	11	59	32	9	32	32
Control Unit Effluent $SBOD_5$ Concentration, mg/l	4	24	14	4	14	14
Supplemental Aeration Unit $SBOD_5$ Concentration	1	15	6	2	5	5
Control Unit $SBOD_5$ % Removal	10	88	54	14	50	56
Supplemental Aeration Unit % Removal	44	99	81	10	83	83

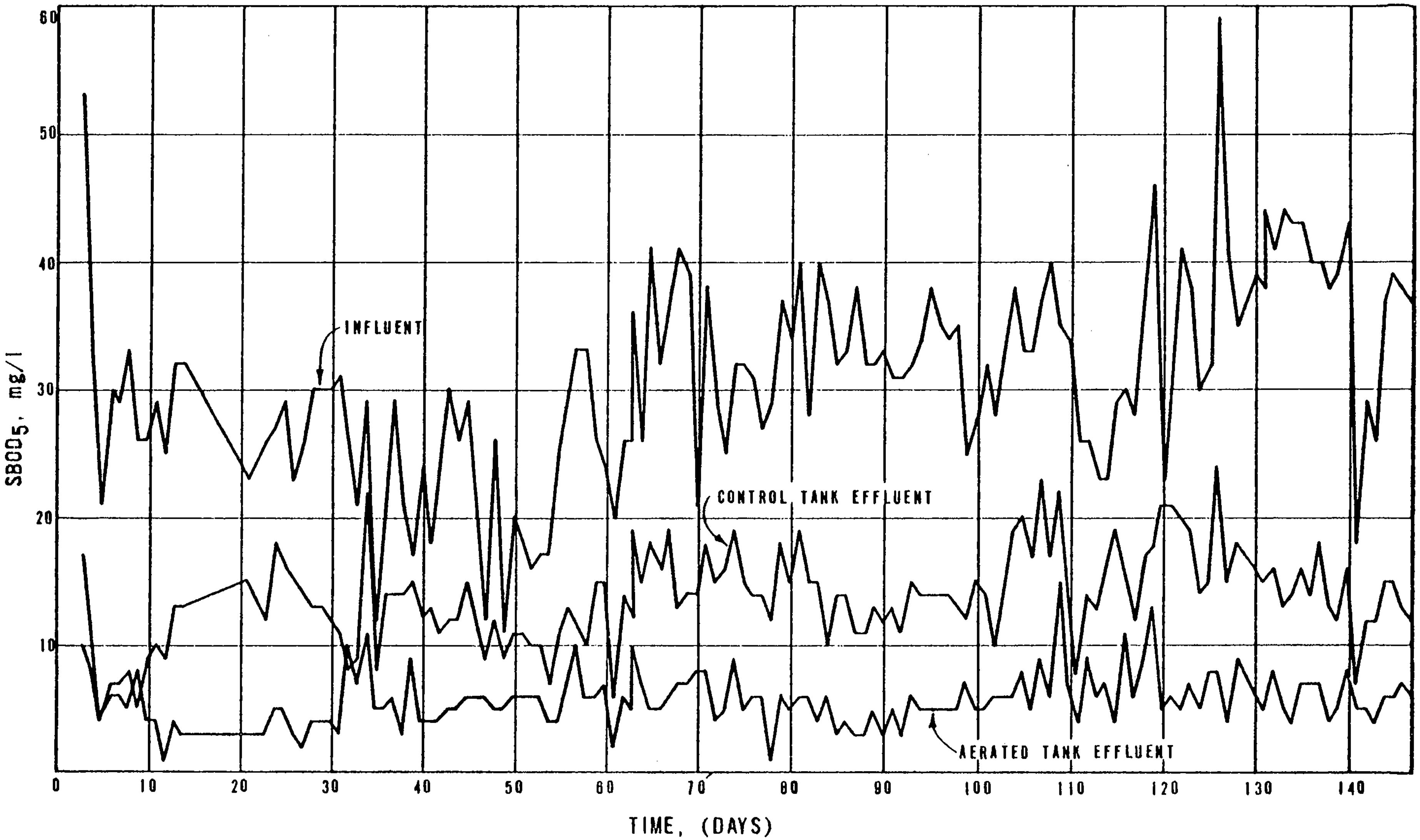

FIGURE 6
INFLUENT AND EFFLUENT
$SBOD_5$ CONCENTRATION
DAILY CHEMICAL FEED

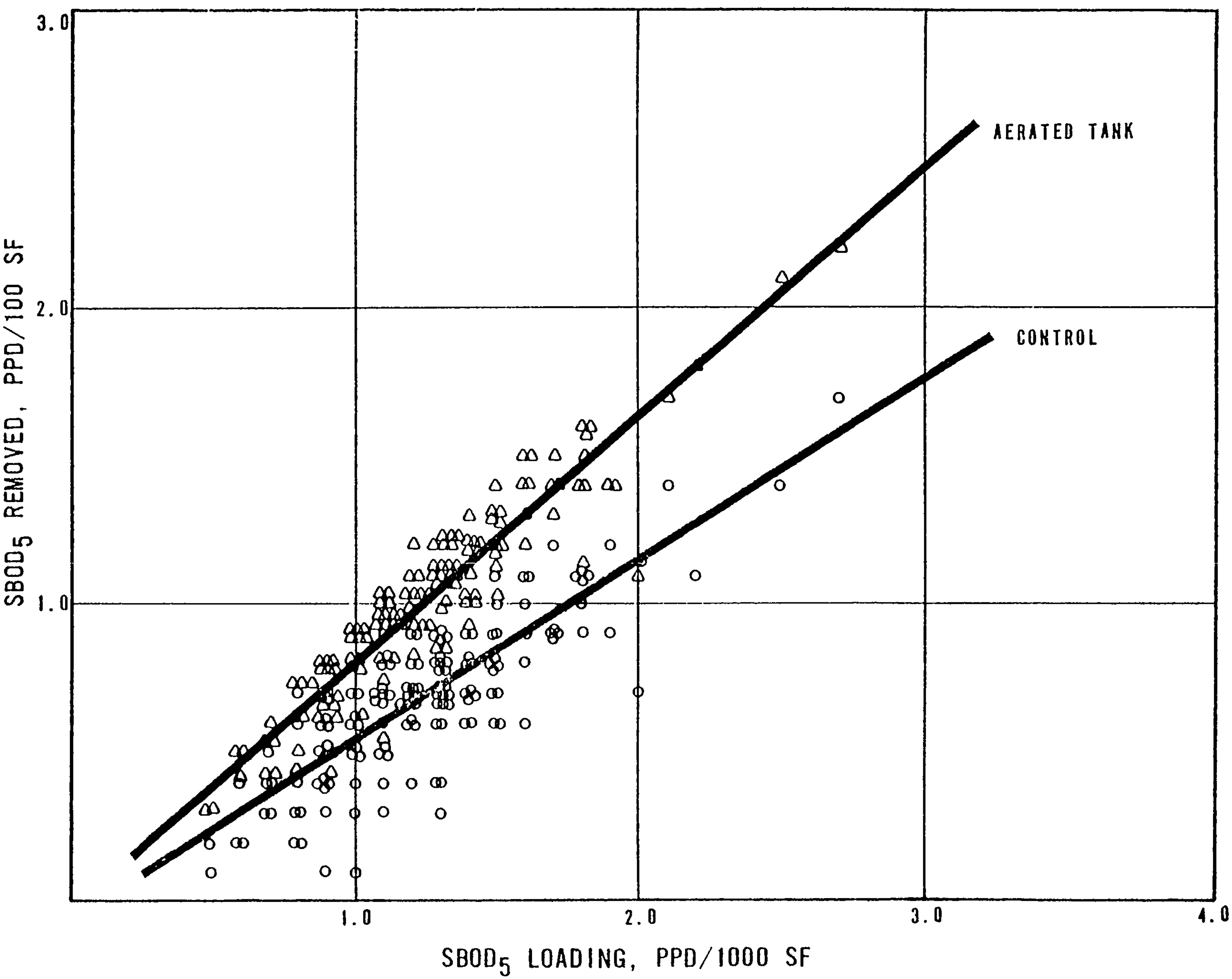

FIGURE 7

$SBOD_5$ REMOVED VERSUS $SBOD_5$ LOADING

FOR FOUR STAGE AERATION WITH DAILY CHEMICAL ADDITION

The effluent $SBOD_5$ concentrations for the supplemental aeration and control units are plotted against the loading on Figure 8. While the data are scattered, as anticipated, it is clear that the effluent quality from the supplemental aeration tank is superior to the control tank effluent in terms of SBOD concentration. It is also seen from Figure 8 that the effect of loading on effluent $SBOD_5$ is more severe for the control unit than on supplemental aeration unit.

(3) Summary

The supplemental aeration test data are summarized in Table 7. It is noted that the aerated unit produced effluent of better quality in every case. The average values of $SBOD_5$ removal for single stage and four stage aeration with and without chemical addition were 60, 69 and 81 percent, respectively, while average percent removal in the control unit was approximately 50 percent. The average $SBOD_5$ concentrations for these three modes of operations were 16, 11 and 6 mg/l.

4. COST COMPARISON FOR INSTALLATION OF THE RBC SUPPLEMENTAL AERATION

The treatment criteria require plant effluent BOD_5 and $SBOD_5$ concentrations of 3 and 1 mg/l, respectively. Removal of the $SBOD_5$ is achieved both in the RBCs and the activated carbon columns. Therefore, lower efficiency of the RBCs requires higher efficiency of activated carbon column and vice versa. The tests described above have shown that $SBOD_5$ removal efficiency of the RBCs can be improved by providing supplemental aeration to the existing motor driven RBC units. An analysis was performed to compare the costs of removing the $SBOD_5$ in RBC process versus its removal by carbon adsorption. The following assumptions were used in costs development:

RBC Performance

The mean effluent $SBOD_5$ concentrations of 6 mg/l from aerated tank and 13 mg/l from unaerated tank are assumed.

Carbon Regeneration

A uniform carbon exhaust rate of 8.2 lbs. per lb. $SBOD_5$ removed is assumed for regeneration requirements. A 6 percent loss of carbon during regeneration is assumed for make-up carbon requirements.

Operating Costs

The operating costs used for the cost estimate are as follows:

Labor, dollars per man-year	15,000
Make-up carbon, dollars per lb.	0.50
No. 2 fuel oil, dollars per gallon	0.525
Electricity, dollars per kilowatt-hr.	0.0317
Ferric chloride (28 percent), dollars per dry ton	128
Polymer, dollars per lb.	1.20

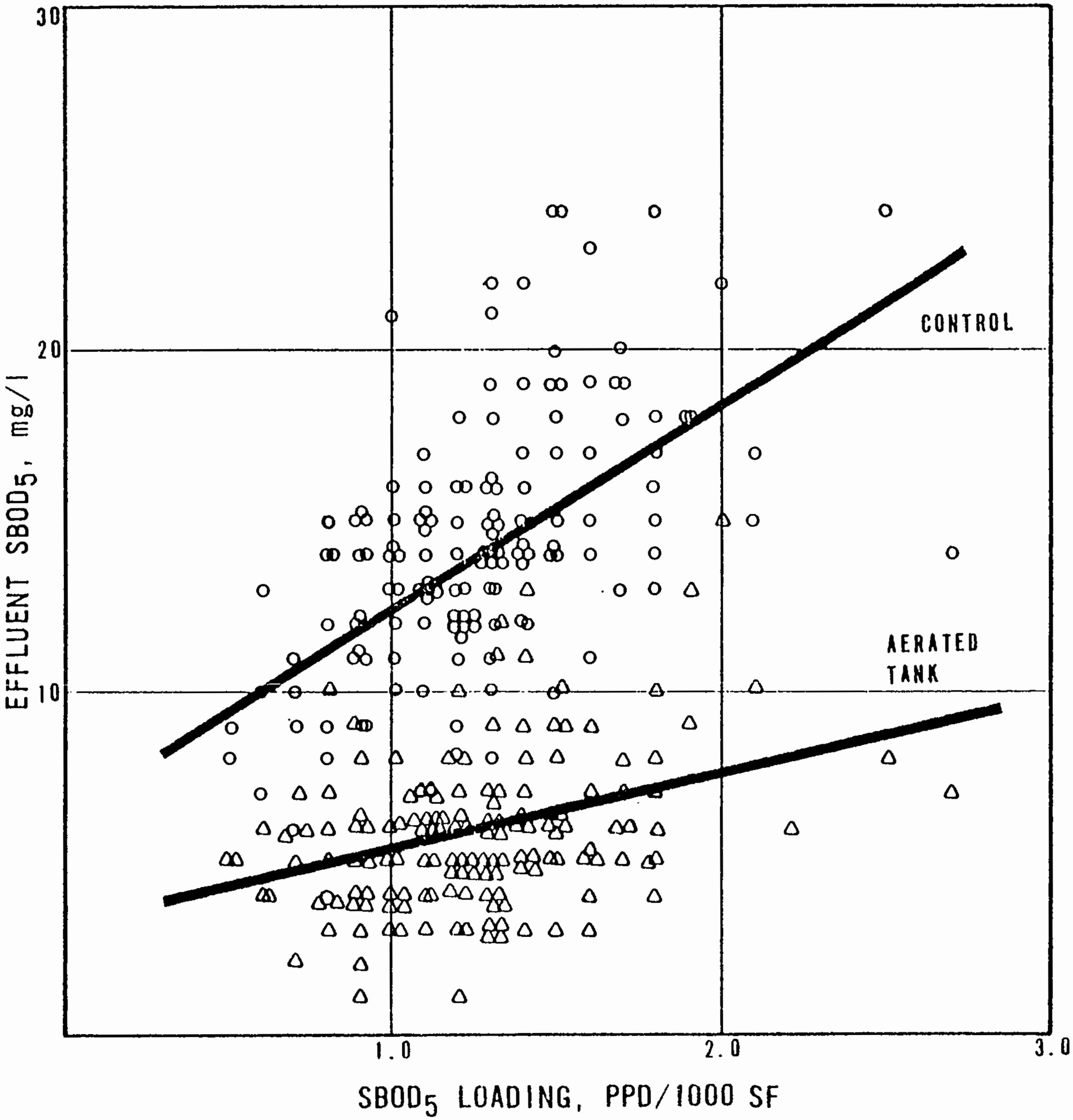

FIGURE 8

EFFLUENT $SBOD_5$ CONCENTRATION VERSUS $SBOD_5$ LOADING FOR FOUR STAGE AERATION WITH DAILY CHEMICAL ADDITION

TABLE 7

Summary of Supplemental Aeration
Test Data.

Mode of Operation	Average Plant Flow mgd	SBOD$_5$, mg/l			SBOD5 Removal, %	
		Influent	Effluent			
			Control	Supp. Air	Control	Supp. Air
Without Chemical Addition:						
Single Stage Aeration	30	40	19	16	53	60
Four Stage Aeration	30	35	18	11	49	69
Daily Chemical Addition:						
Four Stage Aeration	30	32	14	6	56	81

Maintenance Costs

The annual maintenance costs for equipment were estimated at 5 percent of the capital cost.

Supplemental Aeration Equipment Costs

The supplemental aeration equipment capital and installation costs were quoted by the manufacturer of the RBC units (Autotrol Corporation) at $3,125 per unit. The initial capital cost was amortized at a 7 percent interest rate over the assumed design life.

The alternatives investigated for cost analysis are as follows:

Alternative	
No. 1	No RBC supplemental aeration
No. 2	Supplemental aeration installed at four stages of each RBC tank (for total of 56 units installed). Five-year assumed design life.

Both alternatives include chemical addition to primary tanks with carbon adsorption following RBCs to produce same quality effluent.

A comparative cost summary of the two alternates is shown in Table 8.

The chemical cost is common to both alternates at $492,000 per year. The carbon regeneration and make-up cost is highest for Alternate 1 at $540,000 per year. This is due to higher exhaustion rate because of higher effluent $SBOD_5$ from RBCs without supplemental aeration. Carbon regeneration and make-up cost for Alternate 2 is $194,000 per year.

The initial cost of providing supplemental aeration to RBC tanks is estimated at $175,000 for 4 stage aeration. The annual costs including the amortized and O&M costs for supplemental aeration for Alternate 2 is $86,500.

The total annual costs for the alternates are $1,032,000 and $774,000, respectively. The corresponding unit costs per pound of $SBOD_5$ removed are 93¢ and 70¢, respectively.

The cost evaluation, under the assumptions used show the installation of RBC supplemental aeration equipment is cost effective.

TABLE 8

Cost Summary for $SBOD_5$ Removal

Basis: Flow 36.5 MGD

Item	Cost of Alternate[1]	
	No. 1	No. 2
1. Chemical feed to primary tanks, $1000 per year	492	492
2. Carbon regeneration and make-up, $1000 per year	540	194
3. RBC supplemental aeration equipment:		
Capital cost, $1000		175
Amortized cost, $1000 per year		43
Operating cost, $1000 per year		46
Total annual cost, $1000 per year		89
4. Total cost for $SBOD_5$ removal:		
$1000 per year	1,032	774
Dollars per lb. $SBOD_5$ removed by overall process	0.93	0.70
Dollars per 1,000 gallons sewage	0.077	0.058

(1) These costs do not include the cost of existing facilities such as RBC's, carbon columns, and regeneration equipment.

5. CONCLUSIONS

The following conclusions can be drawn from the tests conducted.

- A portion of the organic material observed in the filtrate of primary effluent and RBC effluent samples, and measured as $SBOD_5$ is due to a colloidal fraction. Based on the overall mean $SBOD_5$ reduction of primary effluent and RBC effluent samples treated with ferric chloride, this fraction appears to be about 15 mg/l.

- The three RBC operational modes tested did not indicate significant difference in performance.

- The peroxide treatment controlled beggiatoa predominance and established a healthy culture on the RBC discs. Increases in $SBOD_5$ removal on the order of 10 percent were experienced using perioxide.

- The supplemental aeration of the RBCs essentially eliminated beggiatoa population on the discs. The thickness of biomass growth on the aerated discs was much thinner compared to the unaerated units.

- The average $SBOD_5$ removal was increased by about 27 percent across the RBCs. The average effluent $SBOD_5$ concentration was reduced by 57 percent. The fluctuations in $SBOD_5$ in the effluent was attenuated considerably.

- The comparative cost analysis indicate that the removal of $SBOD_5$ would be cheaper by supplemental aeration of RBCs than by carbon adsorption for Alexandria Treatment Plant.

" EFFECT OF SUPPLEMENTAL AIR ON ROTATING BIOLOGICAL CONTACTOR PROCESS DOMESTIC WASTE"

By

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Introduction

The general acceptance of Rotating Biological Contactors for secondary and advanced wastewater treatment has led to the divulging of many different opinions concerning the use of various techniques to improve the wastewater treatment process with this equipment. One such modification that has received widespread interest among consulting engineers and designers has been the use of air to aid the biological oxidation that occurs on the media surface area. Much of the efficiency of the Rotating Biological Contactor can be credited to the relatively high concentration of biological growth that is exposed alternately to the food source in the wastewater and oxygen in the atmosphere. This vigorous contact in a short detention time is not matched in any other treatment process. We are going to look at the work that has been done thus far in the use of supplemental air provided for mechanically driven Rotating Biological Contactors.

DISCUSSION

Before we look at the direct application of supplemental air in varying quantities to wastewater of varying strengths, we are first going to review the work done by Warren Chesner, Roy F. Weston Company[1], concerning the use of air with the Rotating Biological Contactor process. In their paper presented at the New York Water Pollution Control Federation in January, 1977. it was stated that increased amounts of dissolved oxygen in the wastewater stream of the Rotating Biological Contactor process above 3 to 4 mg/l did not materially effect the performance of the Rotating Biological Contactor process. The data also showed that there was virtually no effect on the suspended solids nor was there any appreciable effect in the reduction in soluble or total BOD as a result of maintaining a high dissolved oxygen.

Chesner's conclusion supports earlier work by Welch[2] who found that a mixed liquor dissolved oxygen above 1.5 mg/l did not significantly affect the Rotating Biological Contactor process.

On the other hand, information developed from Alexandria, VA as shown on Table 1, indicates that copious amounts of supplemental air added to a mechanical drive Rotating Biological Contactor System, did affect the effluent quality.[3] At this installation, approximately 200 cubic feet per minute of air was added for each Rotating Biological Contactor unit with all of the air available to add oxygen rather than a substantial portion used for rotational force. The addition of air did reduce the soluble BOD, but had little or no effect on the total BOD. Refer again to Table 1. This installation is reported to be a side by side comparison of wastewater having similar characteristics of temperature, BOD, suspended solids and nutrients. In both of the cited examples, it should be noted that the treatment process for the Rotating Biological Contactor is being performed on what is commonly called domestic wastewater. By definition, domestic wastewater normally has influent strengths of 150 to 300 mg/l BOD and suspended solids, 15 to 35 mg/l ammonia nitrogen and a temperature range from 40°F to 75°F.

Now that we have examined both ends of a determined spectrum for this presentation, let us examine what is happening in between the extremes and see if a conclusion can be drawn for further study, investigation and use.

At the Purdue Industrial Wastewater Conference in 1979, Chou & Hynek[4] presented information on Autotrol's South Shore Pilot Testing Program in Milwaukee, WI. This program compared their mechanical drive and a prototype air drive treating domestic wastewater at relatively low soluble BOD loadings and when using their relatively closed media.[4] The media is termed as being relatively closed as the flow of wastewater and air throughout the interior surfaces of the media are dependent on a limited number of radial flow passages. The data presented shows that side by side runs for comparison did have equal influent concentrations, but different hydraulic and absolute organic loadings. Chou & Hynek concluded that the addition of air via the prototype air drive system did increase the soluble BOD removal.

STUDY AT COLD SPRING, NY

A study was started in June, 1979 at the Village of Cold Spring, NY Wastewater Treatment Facility, using a Clow Envirodisc Rotating Biological Contactor having a total nominal surface area of 11,000 SF. The tankage provided was designed to enable testing of each of the four stages independently of the other stages. The system was also designed to allow varying quantities of air to be discharged at the bottom of the tank in any stage and its effect measured independently of any of the other stages involved in the study. During the testing period from June through November, 1979, two parallel two stage flow paths were created, equal in all ways, except that air was added to one of the flow paths. The amount of air added was equivalent to 150% of the non-motive or biologically active air that is available in an air driven

Rotating Biological Contactor. (60 CFM/100,000 SF).[5]

Figure 1 shows the testing locations, sampling locations and a general outline of the flow diagram showing how independent and separate testing could be achieved on wastewaters that were identical in all significant characteristics. During the first phase of the study, relatively high loadings of BOD were applied to determine whether or not 3 lbs. soluble BOD/1000 SF/day is the relatively low maximum removal rate as stated by the Autotrol Corporation.[6]

Figure 2 shows loading rates of up to 14 lbs/1000 SF with resulting removal rates of up to 6 lbs/1000 SF. Upon completion of this first phase, successive phases at increasingly lower loadings were run.

Table 2 summarizes all of the data from June through November, 1979 and removals as a result of adding supplemental air to an open media at rates equal to 150% of the non-motive air in an air drive system. It is interesting to note and quite germane to the title of this presentation that by introducing supplemental air to wastewater of the strength that these wastewaters had, there was virtually no difference in removals of soluble BOD between the flow path that had supplemental air and the flow path that did not have any supplemental air, again as shown on Table 2. It is also noted, that the substrate loadings and removals in the initial phases were significantly above those which are normally encountered in waste treatment plants that have been designed for treatment of domestic wastewater, while latter phases of the testing program were run at loadings normally found in such facilities.

Figure 3 shows the percent of removals at various "normal" loadings with a one stage process, both for the Cold Spring Study using Clow's open media and a study using closed media. Again by looking at this data, it can readily be seen that no significant reduction in soluble BOD occurred as a result of adding supplemental air. Similarly, Figure 4 shows the percent of removals at various loadings with a two stage process. The points for the closed media are taken directly from Autotrol's South Shore, Milwaukee Test program published data, while the points for open media are taken directly from Clow's Cold Spring Study.

Conclusions of the South Shore program show that closed media with an air drive did reduce the soluble BOD loadings as compared with a closed media mechanical driven Rotating Biological Contactor. At Alexandria, VA as mentioned earlier, the addition of large amounts of supplemental air to the same type of closed media also reduced the soluble BOD, although not the total BOD. One could conclude then from those two studies that the addition of supplemental air to a relatively closed type media can and does improve the removal of soluble BOD.

On the other hand our studies at Cold Spring, using a relatively open media with supplemental air showed that there is no increase in the soluble BOD removal or advantage to providing supplemental aeration at the loading rates studied. If we then take the published data of pounds applied versus removal rates for the closed media and compare them at the same loading rates with Clow's open media as in Figure 3 and 4, we see that regardless of whether its mechanical drive or supplemental air to a mechanical drive or air drive, that the type of media that is open and that does not depend upon radial passages, appears to consistantly remove a greater percentage of the applied soluble BOD.

CONCLUSION

The data presented here clearly shows that with the use of a relatively open Rotating Biological Contactor media, the addition of supplemental air to enhance wastewater treatment with mechanically driven Rotating Biological Contactors does not offer any significant improvement over units that do not have supplemental air provided. Secondly, by comparing published data on soluble BOD removals using a closed media with our studies using an open media, the open type media Rotating Biological Contactor removed a greater percentage of soluble BOD regardless of whether additional or supplemental air was provided, and finally, the open type media as used in the Cold Spring Study does not have an upper removal rate limit of 3 lbs/1000 SF/day for soluble BOD as reported for the closed type media; removal rates of up to 6 lbs/1000 SF/day were observed.

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3. Strinivasaraghavan, R., Greeley and Hansen Engineers, Personal Communications, November, 1979.

4. Hynek, R.J. and Chou, C.C.S., Autotrol Corporation (1979) "Development and Performance of Air-Driven Rotating Biological Contactors." Presented at 1979 Purdue Annual Industrial Waste Conf.

5. Friedman, R., and Roeber, J., "Energy Reduction Consideration for the RBC Process." Presented at Energy Optimation of Waste and Wastewater Management for Municipal and Industrial Application Conf., December, 1979, New Orleans, LA.

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TABLE 1

SUMMARY OF DATA FROM ALEXANDRIA, VA

	MODE	No. OF STAGES	BOD_5 mg/l	$SBOD_5$ mg/l	SS mg/l
Avg of 5 month's testing	M		43	15	48
	MS		41	8	50
Without chemical addition	M	1		19	
	MS	1		16	
	M	4		28	
	MS	4		11	
With chemical addition	M	4		13	
	MS	4		6	

M = Autotrol Mechanical Drive R.B.C.
MS= Autotrol Mechanical Drive R.B.C. with 200 CFM supplemental Air added per shaft

TABLE 2

COLD SPRING, NY TEST PROGRAM (1)

AVERAGE OF ALL TESTING

June 28, 1979 - November 31, 1979

	Total BOD_5 mg/l	Soluble BOD_5 mg/l	Dissolved Oxygen mg/l	Total (2) Suspended Solids mg/l
Influent	157.1	86.3	4.67	113.2
Mechanical Drive 1st Stage	47.4	40.9	4.19	119.4
Mechanical Drive with supplemental air - 1st Stage	46.3	39.8	4.51	118.5
Mechanical Drive 2nd Stage	30.1	28.6	5.19	101.1
Mechanical Drive with supplemental air - 2nd Stage	30.2	27.6	5.21	106.1

(1) Clow Envirodisc Model A-10
(2) TSS - Prior to clarification

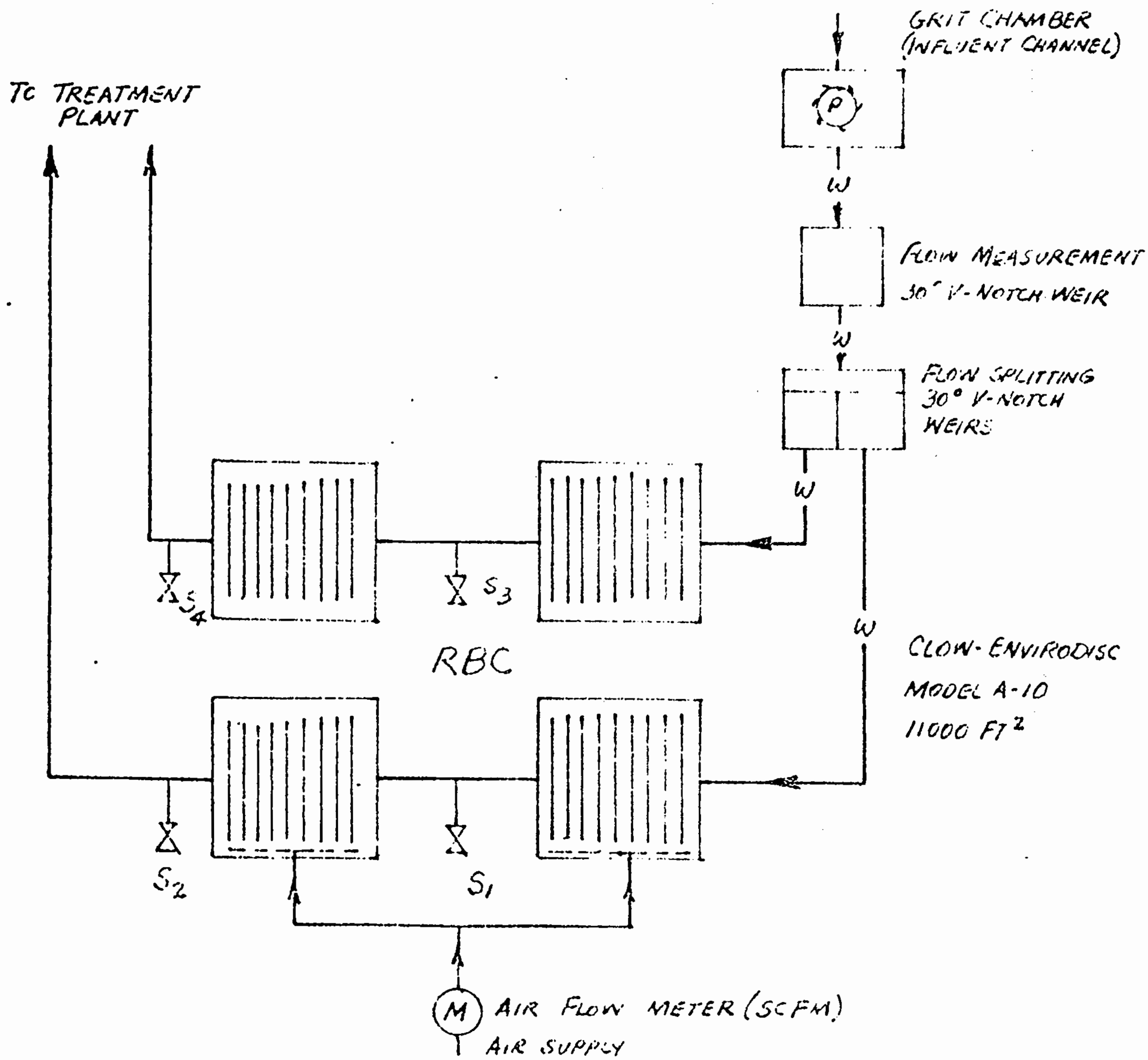

FIG 1 COLD SPRING, NY
PILOT PLANT

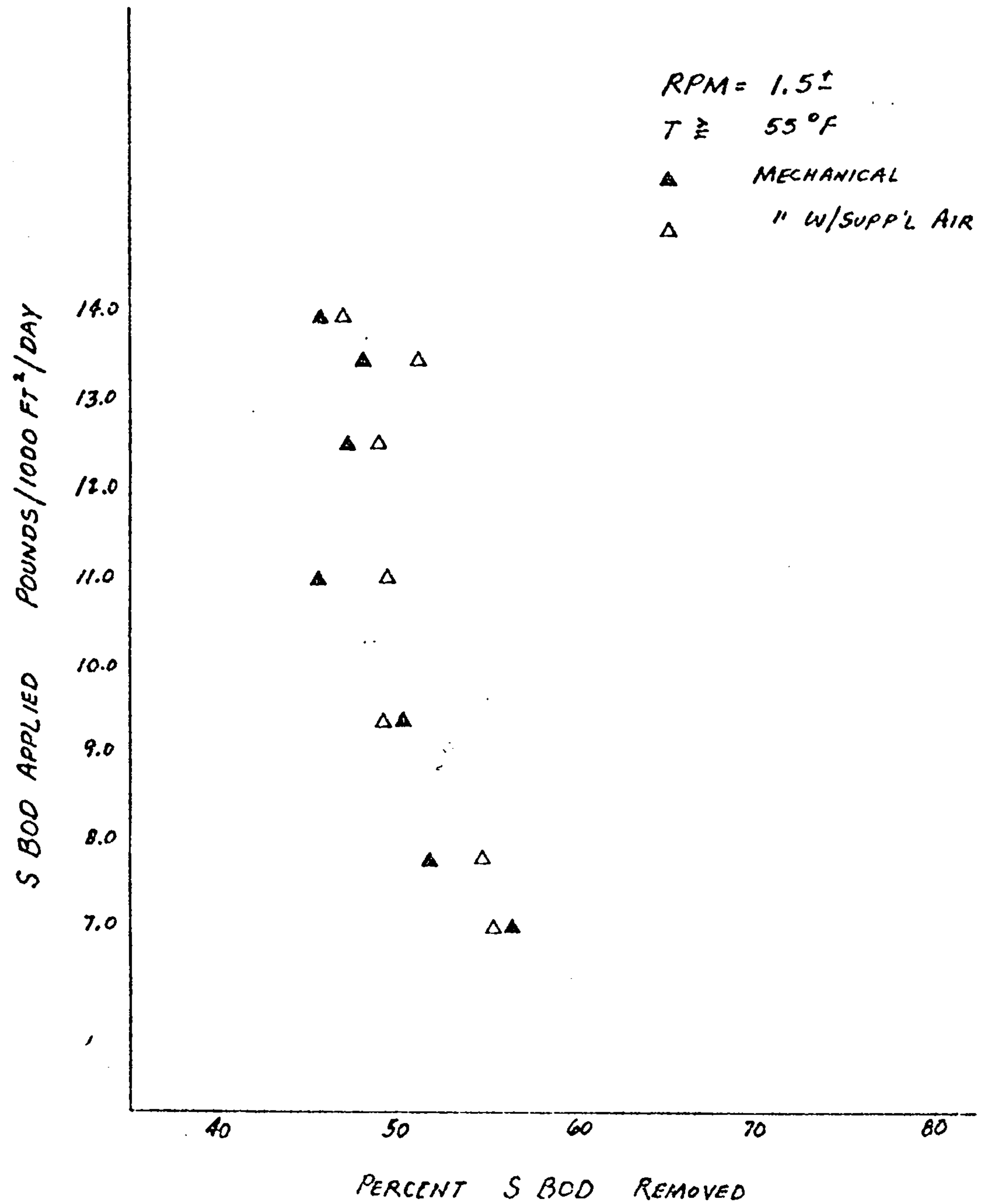

FIG. 2 REMOVALS BY CLOW RBC AT HIGH LOADING.

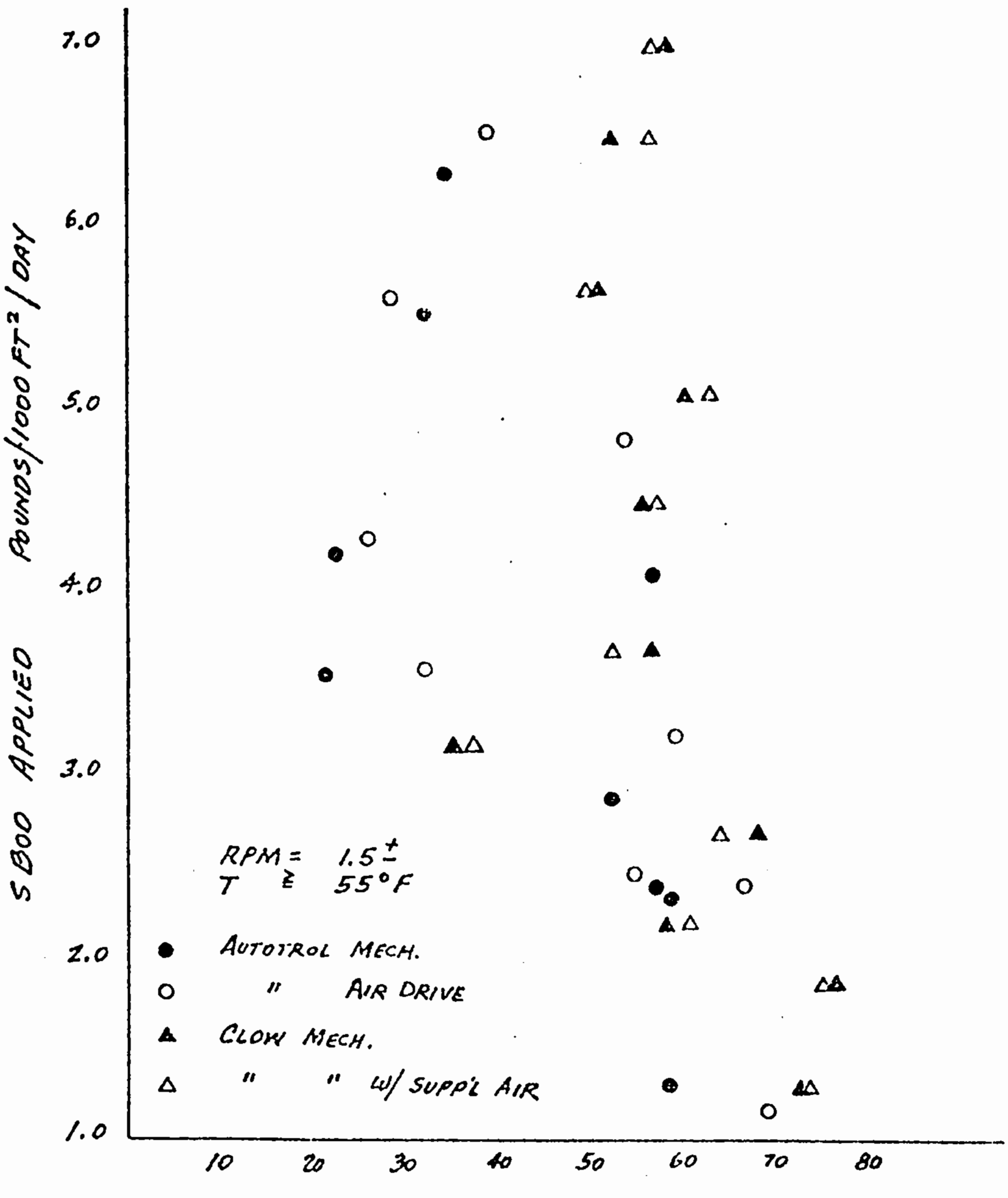

FIG. 3 FIRST STAGE REMOVALS

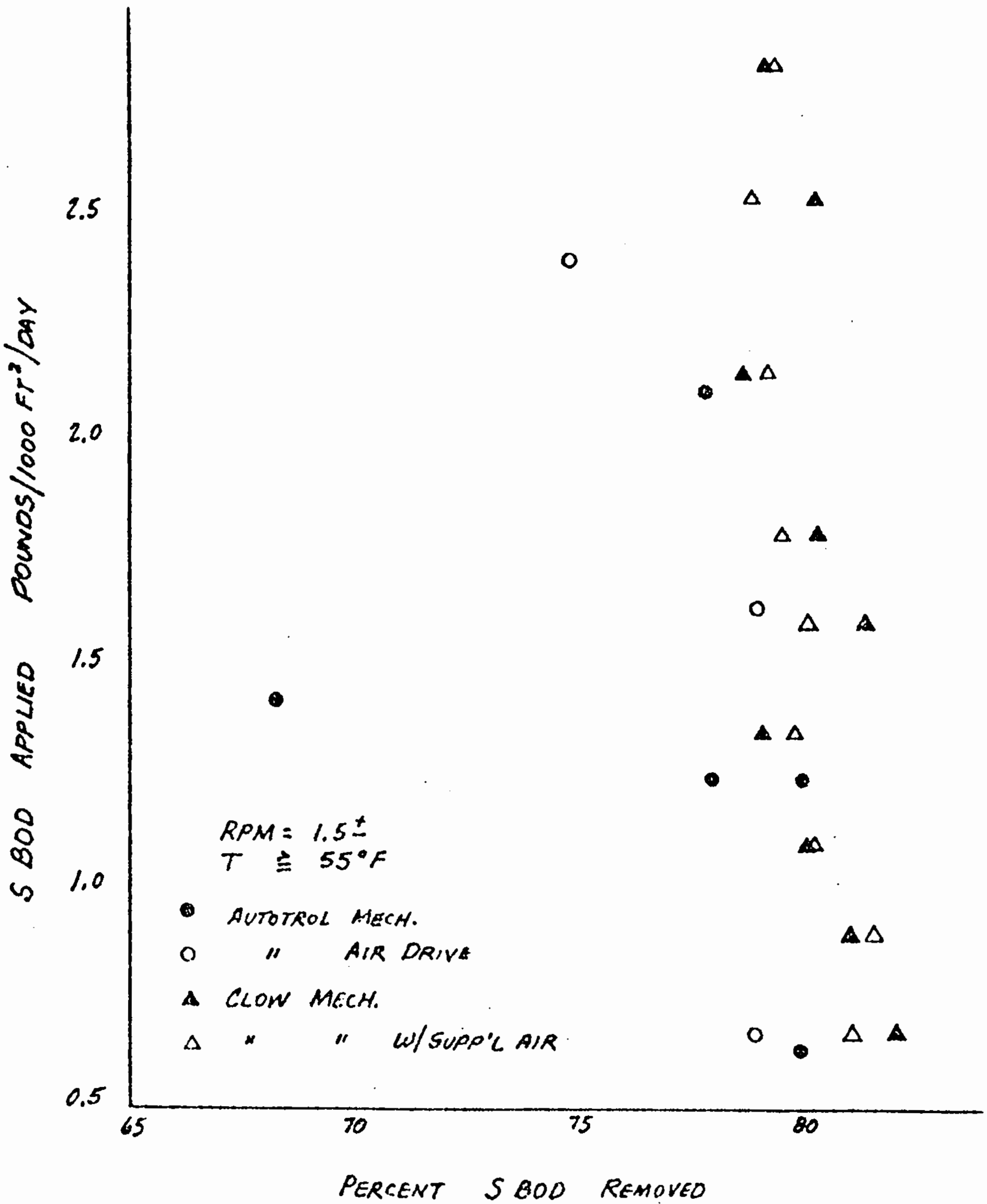

FIG. 4 FIRST & SECOND STAGE REMOVALS

USE OF SUPPLEMENTAL AIR TO CORRECT AN OXYGEN LIMITATION CONDITION OF AN OPERATING RBC SYSTEM

By

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INTRODUCTION

Howes Leather Company is a vegetable sole leather tanner with two operating plants, one in Curwensville, Pennsylvania and one in Durbin, West Virginia. The original treatment system for each plant relied upon a combination of primary lagoons for solids removal and storage, and secondary aerated lagoons for some biological reduction of organics. In 1975, a six month waste characterization and bench scale treatability evaluation of the Durbin plant wastes was undertaken in preparation for enhanced treatment capability at both plants.

This effort revealed the wastes to be of high lime content and high organic strength. Alhough amenable to biological treatment after significant pretreatment, the persistant tannin content of the wastes demonstrated adverse effect in the operation of the dispersed growth experiments, particularly with solids separation. Thereafter, in the late winter and summer of 1976, pilot plant studies of activated sludge and rotating biological contactors (RBC) were conducted at the Durbin tannery. This evaluation favored the use of the rotating biological contactors as the secondary biological process for the application required at the Curwensville, Pennsylvania tannery, which had the earliest compliance schedule of the two tanneries.

For the Curwensville plant, the State of Pennsylvania established effluent standards limiting total biochemical oxygen demand during warm weather conditions to 900 pounds per day, calculated as the total of 1.5 times the effluent five day biochemical oxygen demand (BOD_5) and 4.56 times the effluent ammonia nitrogen. Based on the estimated flow of 200,000 gallons per day and the predicted influent ammonia nitrogen level of 75 mg/l, not anticipated to be signicantly reduced through the treatment process, the effluent BOD_5 would have to be controlled to a level of about 131 mg/l. For the period of November through April, the State permits the BOD_5 to increase to 228/mg/l.

The two tanneries were considered to be similar in the tanning procedures utilized and in the resulting wastes. The only difference in the two plants was in the production level, with Curwensville processing about 1400 hides per day and Durbin, about 1900 hides per day.

PILOT PLANT EVALUATION

The pilot plant work at Durbin provided the design basis for the Curwensville treatment plant.

Pretreatment for the pilot plant was comprised of primary settling, equalization, pH adjustment and phosphorus enrichment. An 8200 square foot, four stage pilot plant RBC system, leased from the Autotrol Company, was used for the test work. It was operated at a rotary speed of 2.9 rpm and a peripheral speed of 60 fpm, the same peripheral speed as is standard for this manufacturer's full scale units. After preliminary experimentation with general flow rate ranges, a more detailed evaluation, upon which the design criteria were formulated, was carried out at flow rates in the general range of 0.25 gpd/sf, to 0.75 gpd/sf.

The characteristics of the influent to the pilot plant RBC are summarized in Table 1. The soluble portion (SBOD) of both the influent and settled effluent BOD_5 averaged 85 percent.

Graphical presentations of the pilot plant data are given in Figures 1 and 2. It is evident that as the SBOD loading increased, the specific removal rate increased, but the overall efficiency of removal decreased. At all test loadings, the SBOD removal efficiency remained above 88 percent.

Within the hydraulic and SBOD loadings tested, the major influence on performance appeared to be SBOD loading. For the limited amount of performance data obtained at comparable SBOD loading, a small improvement in SBOD removal was observed for decreased hydraulic loading, but there was too little data to determine with much certainty the quantitative significance of this indicated effect.

The level of performance for the first stage reactor at the lowest and highest hydraulic loading is shown graphically in Figure 3. Although showing a slightly less consistency of performance, the low specific hydraulic loading appears to show no dramatic sign of performance limitations due to either oxygen or biomass limitation. However, at the highest hydraulic and SBOD loading, the system appears to have reached a maximum SBOD removal rate of about 9 pounds per day per 1000 square feet. At this level of removal, it is apparent that the system was oxygen and/or biomass restricted. Having only limited and somewhat conflicting dissolved oxygen data for the first stage, it cannot now be determined with certainty the specific cause of this indicated removal limitation.

However, it is now apparent that the oxygenation capability of the pilot plant's first stage was disproportionally higher than for a full scale system at the same peripheral speed. This was unfortunately not realized at the time the pilot plant's overall performance was being evaluated for development of

TABLE 1

RBC INFLUENT CHARACTERISTICS
PILOT PLANT STUDY

	Average	Range
TBOD[1.]	939 mg/l	585 - 1290 mg/l
SBOD	798 mg/l	500 - 1080 mg/l
COD	1650 mg/l	1400 - 1870 mg/l
TSS	250 mg/l	140 - 790 mg/l
TKN	65 mg/l	37 - 88 mg/l
NH_3 - N	58 mg/l	35 - 81 mg/l
pH	7.4 (median)	6.4 - 8.2
Temperature	68°F	55 - 75°F

1. Total 5-day Biochemical Oxygen Demand

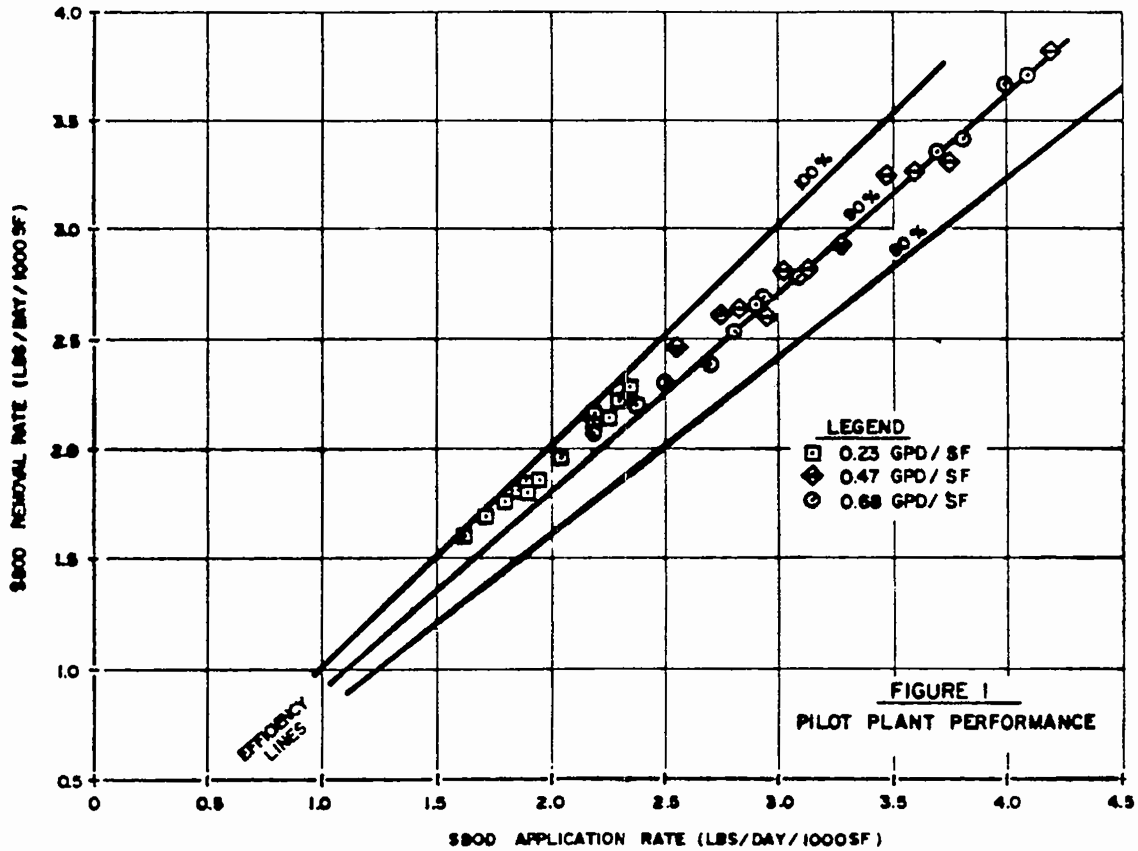


FIGURE 1

PILOT PLANT PERFORMANCE

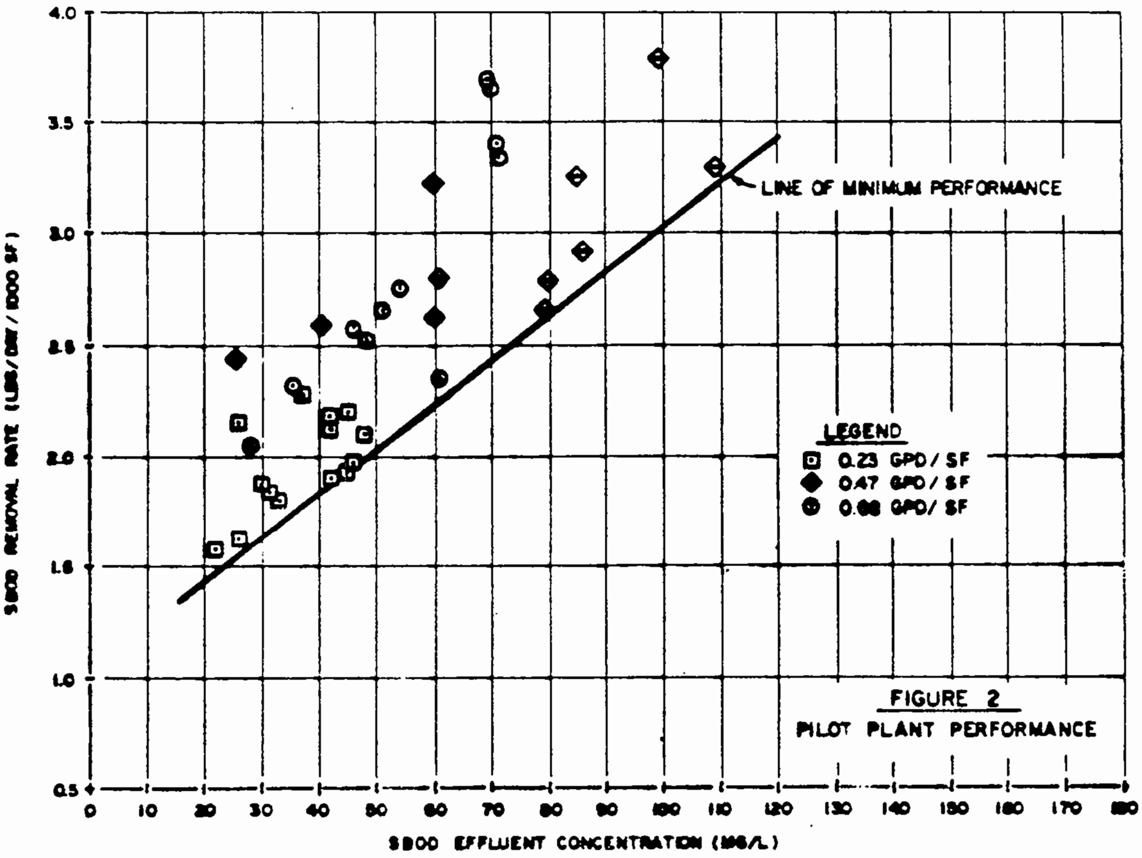
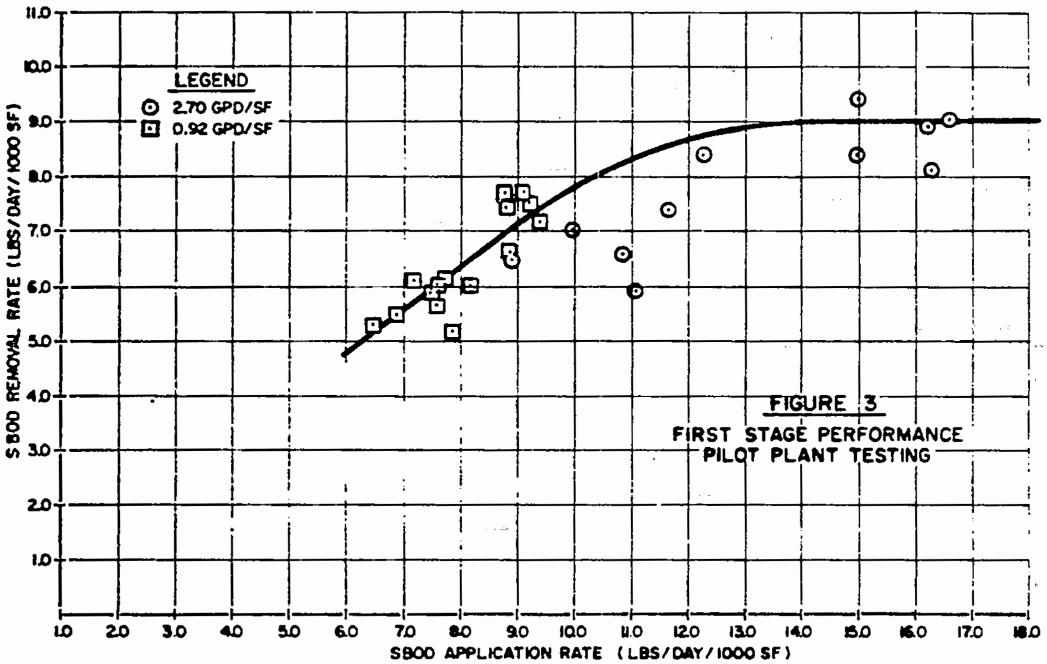


FIGURE 2
PILOT PLANT PERFORMANCE


LEGEND
2.70 GPD/SF
0.92 GPD/SF
SBOD REMOVAL RATE (LBS/DAY/1000 SF)
SBOD APPLICATION RATE (LBS/DAY/1000 SF)
FIGURE 3
FIRST STAGE PERFORMANCE
PILOT PLANT TESTING

design criteria for the full scale installation. The problems relating to scale-up limitations due to the higher oxygenation capability of the pilot plant's smaller disc sizing (regardless of comparable peripheral speed control) did not come to our attention until the full scale problem was upon us.

Although BOD was the most closely monitored parameter for performance measurement of the RBC, some performance data for chemical oxygen demand (COD), total suspended solids (TSS) and nitrogen was obtained. Effluent COD varied between 227 and 436 mg/l. Effluent TSS averaged 38 mg/l with a range of 20 to 96 mg/l. No nitrification was observed for any of the testing, although conversion of organic nitrogen to ammonia nitrogen through the RBC was consistently observed.

DESIGN BASES OF RBC FACILITY

The design bases for the Curwensville RBC facility is summarized in Table 2.

Allowing for no change in the predicted 75 mg/l influent ammonia concentration through the RBC, a plant discharge TBOD of 131 mg/l (219 pounds per day) would thus be required for the total oxygen demand to remain within the 900 pounds per day summertime limit of the DER and EPA permits. On a SBOD basis, the effluent concentration to be achieved by the RBC was set to be 112 mg/l. Based on the estimated influent SBOD concentration of 800 mg/l, the RBC would be required to remove 1149 pounds of SBOD per day.

Referring to Figure 2, a removal rate of 3.25 pounds of SBOD per day per 1000 square feet was indicated for this required effluent concentration. A total surface area of 353,000 square feet was indicated.

To provide a margin of safety and to work within the manufacturers' standard sizing of four stage systems, a unit comprised of four shafts, each 25 feet long, and each having a rated surface area of 100,000 square feet, was selected and installed. The total surface area of 400,000 square feet, related to the design hydraulic loading and required SBOD removal, provides a specific hydraulic loading of 0.5 gpd/sf and a specific SBOD removal rate of 2.9 pounds per day per 1000 square feet of total shaft area.

A flow diagram of the treatment system, as it was initially provided, is given in Figure 4.

INITIAL OPERATING EXPERIENCES

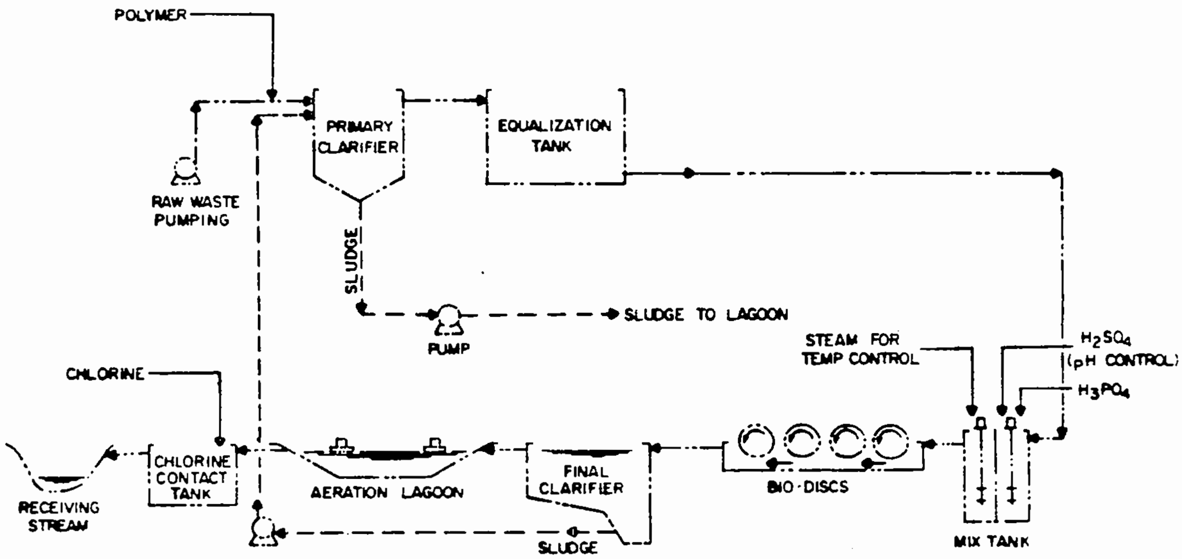
The construction of the plant was completed in early 1978, with all treatment units in operation by April, 1978. It became evident soon thereafter that the RBC facility was seriously stressed under the conditions which then prevailed. The symptoms were ominous - significant presence of hydrogen sulfide in and around the RBC facility, unfavorable growth conditions on the surface, and depleted dissolved oxygen levels in the first stage.

Concurrently, it was discovered that the waste load to the RBC was much greater than the design anticipated. A summary of this initial waste load characterization is given in Table 3.

TABLE 2

DESIGN BASES FOR
SIZING RBC FACILITY
CURWENSVILLE TANNERY

Flow (Equalized)	200,000 gals/day	
Soluble BOD		
Loading	800 mg/l	1336 lb/d
Required Effluent	112 mg/l	187 lb/d
Required Removal	688 mg/l	1149 lb/d
Minimum Sizing Criteria, Based On Pilot Plant Findings (Figure 2)		
4-Stage Removal Rate	3.25 lbs/day/1000sf	
Total Area Required	353,000 sf	
Selected RBC System		
4 - Stage, 25 ft. Shaft Per Stage		
Total Area Provided	400,000 sf	



NOTE ALL TREATMENT UNITS FROM PRIMARY CLARIFIER TO FINAL CLARIFIER ARE UNDER ROOF.

FIGURE 4
FLOW DIAGRAM
TREATMENT PLANT
CURWENSVILLE TANNERY

TABLE 3

RBC INFLUENT WASTE CHARACTERISTICS DURING PERIOD OF PLANT START-UP

Flow	210,000 gallons/day
TBOD	1490 mg/l
SBOD	980 mg/l
TSS	900 mg/l
Sulfide	27 mg/l
TKN	360 mg/l
NH_3-N	180 mg/l
pH	8

As indicated, the flow was higher, as was the SBOD load. Further, the suspended solids were greater, as is probably the reason for the higher proportion of non-soluble BOD than found in the pilot plant wastes. The presence of high sulfide concentration was also unexpected from our pilot plant work. Because of the total oxygen demand limitations of the permit, the influent presence of ammonia nitrogen higher than the effluent design expectations was another serious difficulty revealed by this evaluation of initial operating conditions.

Thereupon, a multi-faceted program was initiated to reduce waste load and increase treatment capability in order that permit requirements could be met. This was eventually accomplished with the full cooperation and support of the Meadville Office of the Pennsylvania DER. This discussion is mostly limited to the efforts made to enhance RBC performance.

The first relief attempt was to remove the baffle between the first and second shaft to reduce the first stage specific BOD loading rate. Also, flow reduction to the RBC unit was made. The major concern at this point was to relieve the hydrogen sulfide generation problem. The results of these efforts are tabulated in Table 4. Although the specific hydraulic and SBOD loads were reduced to or below design levels, the effluent SBOD concentration remained high as did the hydrogen sulfide levels in and around the RBC basins. A dissolved oxygen level could not be maintained in the mixed liquor of any of the stages.

The problem appeared to be more than oxygen limitations related to BOD conversion. Both the high sulfide and suspended solids content of waste to the RBC were significant contributors to the pervasivenss of the reduced conditions within the RBC. Influent sulfide levels varied between 30 - 60 mg/l, and lesser values persisted through the stages. Though each shaft assembly was equipped with external devices to increase turbulence and minimize solids deposition, substantial deposition did occur midway between the shafts providing a source for anaerobic conditions within the tank. High mixed liquor solids concentrations in the 1500 - 2000 mg/l range, with a 50 percent volatile content were observed, as were high oxygen uptake rates in the 2 - 3 mg/l per minute range.

The attached growth on the surfaces was gray to black and quite thick. Under the biomass, there was a thin crystalline inorganic layer on the media surface, assumed to be a calcium carbonate accumulation.

USE OF HYDROGEN PEROXIDE

As an expedient measure to temporarily relieve the sulfide problem, hydrogen peroxide was added to the RBC influent. At a reduced waste flow of about one-half of the design basis, hydrogen peroxide was used initially at the rate equivalent to approximately 35 mg/l of influent flow. It was increased later to higher levels sufficient to totally oxidize all sulfides. At even the lower dose of hydrogen peroxide which did not totally remove all sulfides, significant improved performance was observed. A summary of two test runs when hydrogen peroxide was used is given in Table 5.

TABLE 4

INITIAL PERFORMANCE OF RBC

Specific Hydraulic Loading (gpd/sf)	Specific SBOD Loading (ppd/1000sf)	Specific SBOD Removal (ppd/1000sf)	Percent SBOD Removal (%)	Effluent SBOD Conc. (mg/l)
0.45	3.0	1.67	56	353
0.45	2.1	1.00	48	300
0.45	1.9	0.90	47	267
0.25	1.6	1.18	74	221

TABLE 5

USE OF HYDROGEN PEROXIDE TO IMPROVE RBC PERFORMANCE

Specific Hydraulic Loading (gpd/sf)	Hydrogen Peroxide Conc. (mg/l)	Specific SBOD Loading (ppd/1000sf)	Specific SBOD Removal (ppd/1000sf)	Percent SBOD Removal (%)	Effluent SBOD Conc. (mg/l)
0.283	35	2.85	2.35	82	213
0.17	125	1.17	1.11	95	47

In the first test of higher flow and lower hydrogen peroxide dose, the sulfide in the RBC influent was reduced from an average level of 46 mg/l to 14 mg/l. For the most part, the RBC effluent showed only occasional signs of sulfide. The dissolved oxygen level remained consistently low at 0.1 mg/l in all stages. The heavy biomass accumulation in the RBC surfaces was, however, significantly reduced. At the higher hydrogen peroxide dose and lower flow, total sulfide control was attained. With this control and with the reduced SBOD loading of 1.17 pounds per day per 1000 square feet, good performance of the RBC was realized for the first time.

However, with the high cost ($100/day) of the hydrogen peroxide and the limited hydraulic loading, other methods of obtaining improved treatment performance were considered. These efforts were in two directions, one involving treatment facility alterations and the other involving waste production alternatives.

PRODUCTION CHANGES TO IMPROVE WASTE CHARACTERISTICS

Although detailed discussion of the production changes to improve waste characteristics is beyond the intent of this presentation, it should be noted that an interested and technically oriented plant manager worked diligently and effectively to reduce both the constituent and volumetric level of the wastes from the time of plant start-up to the present. This program is still continuing. It is pertinent to note that as changes were made in production practices to reduce flow and critical waste constituents, problems were often encountered in product quality. This was particularly problemsome and costly due to the lengthy time sequence involved in processing a raw hide. Therefore, any modification in processing which proved to impair product quality could be damaging to 15 - 20 days of hide processing, before the effect of re-adjustment could be realized.

Somewhat offsetting this risk for the tannery was the potential savings in chemicals and water such a conservation effort could provide. At this time, substantial reduction in lime, sulfide, ammonium chloride and water has been realized.

A summary of the reductions in raw waste quantity and quality which were effected from June, 1978 to September, 1979, at approximately the same hide processing level, is given in Table 6. The reduction is significant for all parameters, being highest for TBOD and TSS. A comparison of the settled TBOD and TKN for the two periods tends to somewhat decrease the significance of the constituent reduction, as relates to RBC feed. Nevertheless, the reduction is still significant and overall helpful to the intended improved performance of the RBC.

FACILITY ALTERATIONS FOR IMPROVED PLANT PERFORMANCE

From the beginning of the full scale operation, it was realized that the amount of suspended material entering the RBC and becoming deposited in the basins was contributing to the oxygen limitation problem of the RBC. The amount of suspended matter in the pre-settled and equalized wastes was also being increased by partial precipitation of soluble constituents, as a result of the acid adjustment of the pH to 7 - 9 just prior to the RBC. Laboratory experiments also revealed further pH adjustment to a level of 4 - 5 significantly reduced the BOD by coagulation of the protein. With the reduction of

TABLE 6

REDUCTION OF RAW WASTE CONSTITUENTS BY PRODUCTION CHANGES

	June 1978	September 1979
Flow	210,000 gal/day	130,000 gal/day
COD	15,000 lbs/day	10,000 lbs/day
TBOD	8,200 lbs/day (2,600 lbs/day)	3,300 lbs/day (1,900 lbs/day)
TSS	10,600 lbs/day	3,600 lbs/day
TKN	1,160 lbs/day (630 lbs/day)	620 lbs/day (485 lbs/day)
NH_3-N	460 lbs/day	315 lbs/day
Sulfide	170 lbs/day	110 lbs/day

Note: Data in parenthesis is for settled wastes.

this substrate portion, the resulting waste could be expected to be more amenable to biological degradation.

It was also evident that improved mixing and oxygenation in the equalization tank would be beneficial to the control of the sulfide generation problem in the wastes prior to the RBC and, more particularly, prior to the desired addition of the acidification and intermediate clarification processes. Because of the high pH of the primary settled wastes, the aeration of the equalization tank provided a potential further benefit of reducing the ammonia content of the wastes by air stripping, and concurrently reducing the alkalinity and subsequent acid requirements for protein coagulation.

Accordingly, an acid feed/mix system and an intermediate clarifier were installed between the equalization tank and the RBC unit, to permit the acid coagulation of the equalized wastes and the reduction of suspended matter in the feed to the RBC. Concurrent with placing these alterations into operation, the use of hydrogen peroxide was discontinued.

The use of supplemental air with the RBC also was considered a beneficial alteration, in that such use would decrease oxygen limitation conditions and create increased agitation to prevent the thick biomass accumulation and the sub-surface anaerobic conditions in which sulfur reduction occurred.

This facility alteration program was undertaken in stages and, in fact, was not totally implemented at the time this presentation was being prepared. The acidification system and intermediate clarifier were installed first, with operation commencing in November 1978. Supplemental air was added to the first shaft in January 1979, to the second shaft in April 1979 and to the third and fourth shafts in August 1979. The turbine aerators for the equalization tank are just now being installed. Initially, in late 1978, two surface aerators were taken from the existing aerated lagoons as an interim effort to aerate the equalization tank. Although reasonably successful in terms of maintaining an aerobic environment and stripping ammonia, this old equipment's operation was unreliable and had to be totally discontinued half way through the testing period.

All of the facility changes were innovatively detailed, fabricated and installed by plant engineering and maintenance staff.

The desired intent of developing comparative data for the variable conditions of production and treatment was made difficult by (1) the EPA pressure to expedite compliance of effluent standards, (2) the trial and error nature of the production changes being attempted for improved wastewater conditions and (3) the problems involved in retrofitting and altering existing facilities within a constrained space and time situation. Adding further to the burden of this evaluation was the required experimentation being carried on at this plant with a second stage biological system to achieve nitrification capabilities within a very expedited compliance schedule to meet EPA proposed BAT standards.

The test effort on the first stage RBC facility reached final frustration on October 31, 1979 when an RBC shaft broke. As of January, 1979, the RBC facility was still out of operation, with all four shafts being replaced by the manufacturer.

A flow diagram of the altered treatment system is given in Figure 5. A sketch of the supplemental air system installed in the RBC is given in Figure 6.

PERFORMANCE OF ALTERED SYSTEM, WITH SUPPLEMENTAL AERATION OF RBC

Prior to the completion of the installation of the diffused air system in the first two stages, but following the completion of the intermediate clarifier and aeration of the equalization tank contents, there was a short period of time when some limited observations were made of this change in RBC performance. Although the dissolved oxygen level in the first two shaft mixed liquor volumes remained absent, or at very low fractional level, the sulfide generation problem appeared to be greatly improved. At reduced hydraulic loadings of approximately .28 gpd per square foot of total shaft area, effluent SBOD levels of 100 - 120 mg/l were being achieved at SBOD removal rates of near 1.5 pounds per day per 1000 square feet of total shaft area.

The major impact on RBC performance, however, occurred with the use of supplemental air. A graphical representation of the testing performed with supplemental aeration of the RBC, is shown in Figure 7.

As shown, the best performance was achieved when the wastes were additionally pretreated by acid coagulation and clarification. For this series of tests, supplemental aeration was limited to the two shafts of the first stage, at a total air flow of approximately 600 cfm. This is equivalent to about 12 cfm per lineal foot of shaft.

The data for partially aerated RBC receiving acid treated wastes do not indicate for the SBOD range tested that the maximum SBOD removal rate had yet been attained. Without acid pretreatment, the maximum SBOD removal rates is indicated to be approximately 2.5 pounds per day per 1000 square feet of total shaft area.

Although there is some indication for the non-acid treated conditions that supplemental aeration of all shafts at an air flow of 1100 cfm results in better performance than when only the first stage is aerated, more data is required before a final judgement can be made on this point.

For the acid treated conditions, some experimentation was made with effluent recycling. Two data points shown in Figure 7 for where non-settled effluent from the RBC was recycled at a rate of 50% of the influent rate suggest somewhat reduced performance when compared to the non-recycle conditions of operation. It should be noted, however, that the influent flow rate (172,000 gpd) during the recycle testing was somewhat higher than the influent flow rate (138,000 gpd) during the non-recycle testing periods. On a specific hydraulic rate basis, the loading, including recycle flow, for the recycle operation was 0.68 gpd/sf as compared to 0.36 gpd/sf for the non-recycle operation.

A comparison of aerated and non-aerated operation of the RBC treating these tannery wastes is given in Figure 8. Interestingly, the best performance with supplemental aeration of the RBC with acid pretreatment was similar to the pilot plant performance of the nonaerated RBC unit, under similar hydraulic conditions.

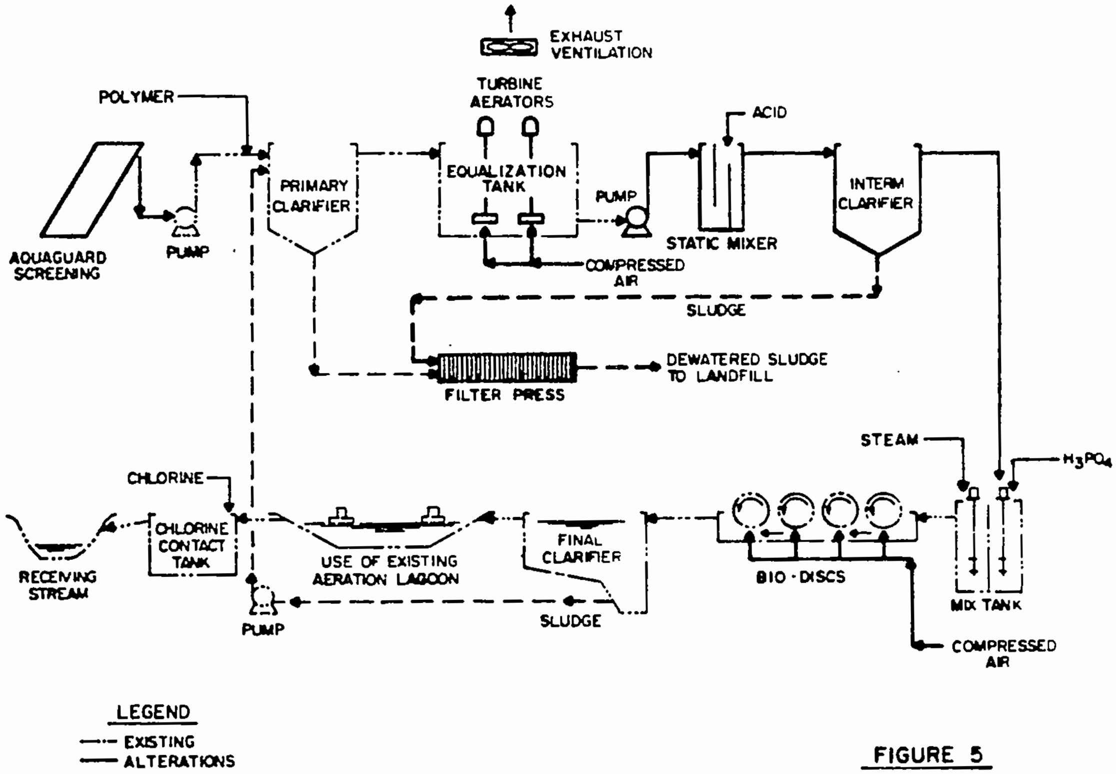


FIGURE 5

ALTERATIONS TO
TREATMENT SYSTEM
CURWENSVILLE TANNERY

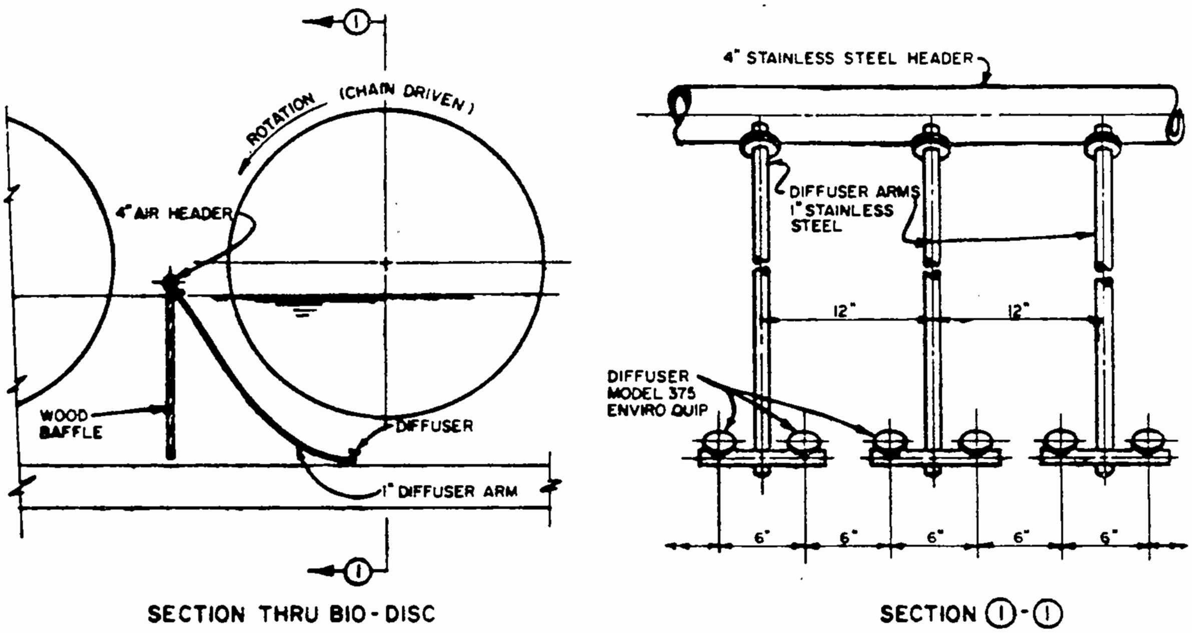


FIGURE 6

INSTALLATION DETAILS
RBC DIFFUSER ASSEMBLY

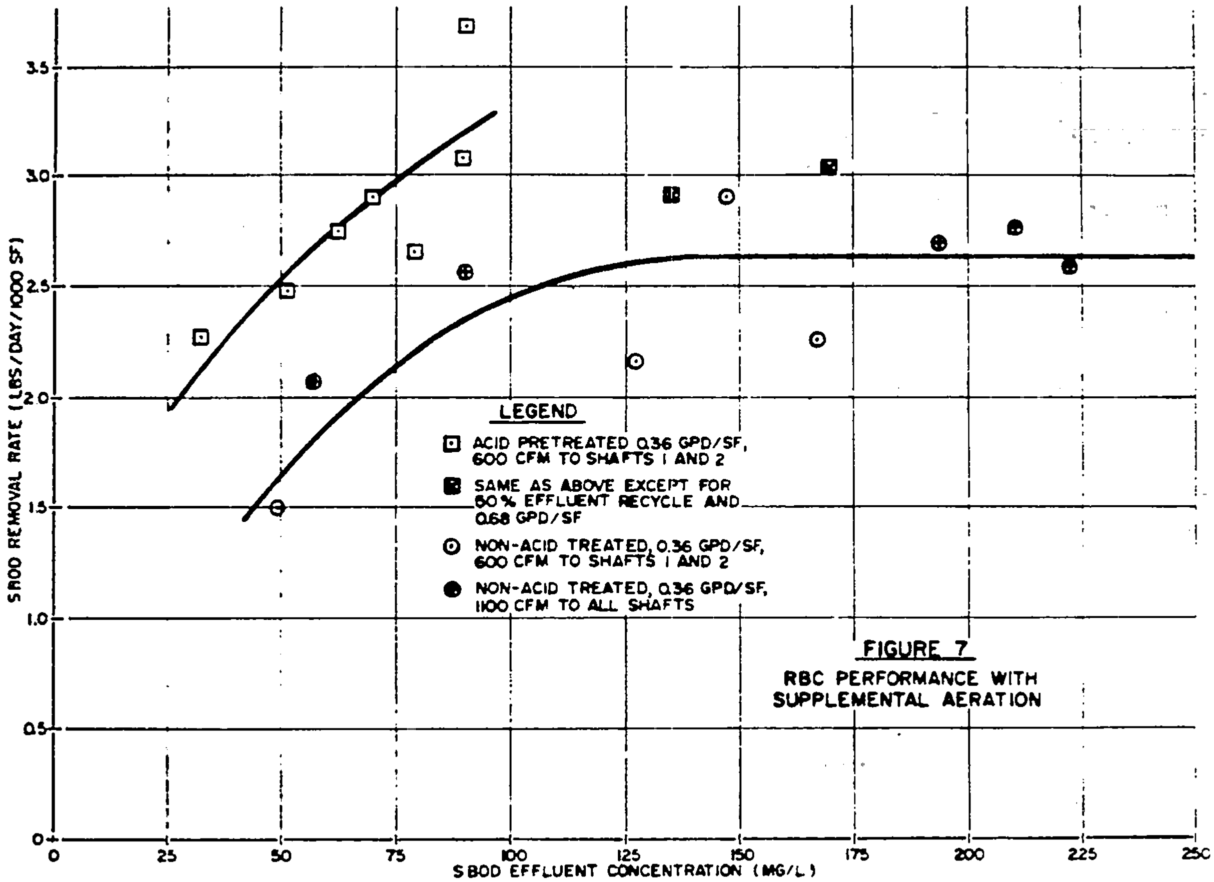


FIGURE 7
RBC PERFORMANCE WITH SUPPLEMENTAL AERATION

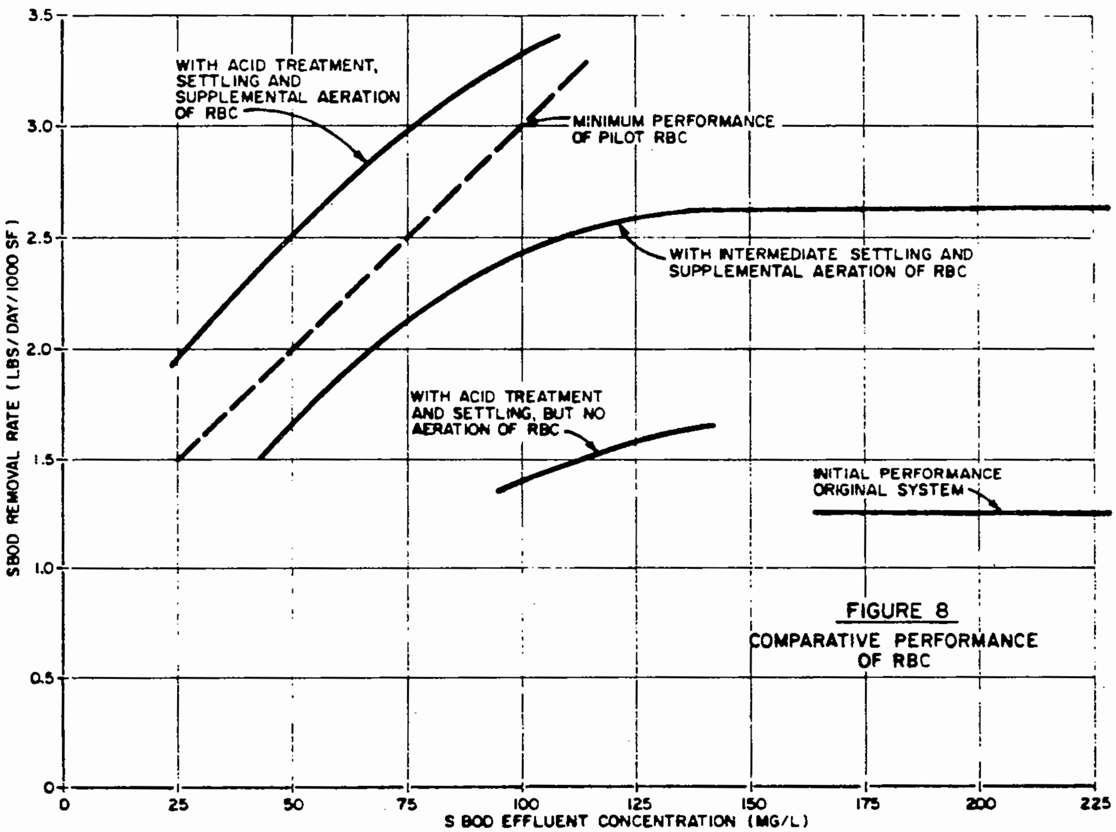


FIGURE 8

COMPARATIVE PERFORMANCE OF RBC

Examining performance of the RBC by stages, it appears that for the acid treated wastes the aerated first stage operation is not oxygen or biomass limited up to the maximum observed specific SBOD removal rate of approximately 6 pounds per day per 1000 square feet of first stage area. However, for the wastes not acid coagulated and settled, the aerated first stage operation is apparently biomass limited at a specific SBOD removal rate of about 4.25 pounds per day per 1000 square feet of first stage area. This is shown graphically in Figure 9. Based on the relatively significant level of dissolved oxygen in the first stage mixed liquor for this period of testing, it does not appear to be a problem of oxygen limitation. For the most part, the dissolved oxygen level in the aerated first stage averaged over 1 mg/l, varying between 0.6 and 2.1 mg/l. Surprisingly, the third shaft (second stage) and fourth shaft (third stage), during this testing of the RBC with only first stage aeration, maintained lower dissolved oxygen levels at the 0.5 mg/l level. It had been expected, based on the low specific SBOD removal rates in these last two stages as related to the normal oxygenation capability of the rotating discs, that the dissolved oxygen levels would increase. However, when aeration was added to the last two shafts, the dissolved oxygen level gradient changed to the more expected increasing type, with dissolved oxygen concentrations reaching levels of 3 to 4 under the third shaft and 4 to 5 under the fourth shaft.

It is evident that the major removal of SBOD occurs in the first stage, with only minor removal occurring in the last two stages. A summary of the stage removal rates for the acid treated and non-acid treated waste testing of the aerated RBC is given in Table 7. It is noted that both the first stage and the combined stages for wastes of reduced protein content have a greater specific removal rate of SBOD than do the wastes which did not have benefit of acid coagulation and settling for reduction of protein. Interestingly, the reverse is true for the last two stages, probably because some of the protein content in the lesser pretreated waste is degraded in the latter stages. In more closely evaluating shaft performance during the period of protein reduction and first stage aeration, the data indicates that as total SBOD loading and removal increases through the RBC unit, the percent of the total removed by the first stage reduces somewhat as the removal percentages by the last two stages increase. This is shown graphically in Figure 10.

For this same period of pretreatment to reduce the protein content, some analyses were made to determine the change in the BOD rate constant as the flow passed through the stages. The "k" value decreased from 0.26 for stage 1 influent to 0.15 and 0.08 for stages 2 and 3 influents, and 0.06 for stage 3 effluent.

Also shown in Figure 10 are two data points for the same pretreatment conditions and supplemental aeration except that 50% recycle of effluent has been provided. It is apparent that some of the SBOD removal has been transferred from the first stage to the second stage, by the effect of recycle. Presumably, as the recycle rate would be increased further, the specific removal rate would be more uniform throughout all the stages. This limited testing indicates, however, that the overall removal rate would be reduced somewhat by such effluent recycling possibly due to the increased hydraulic rate.

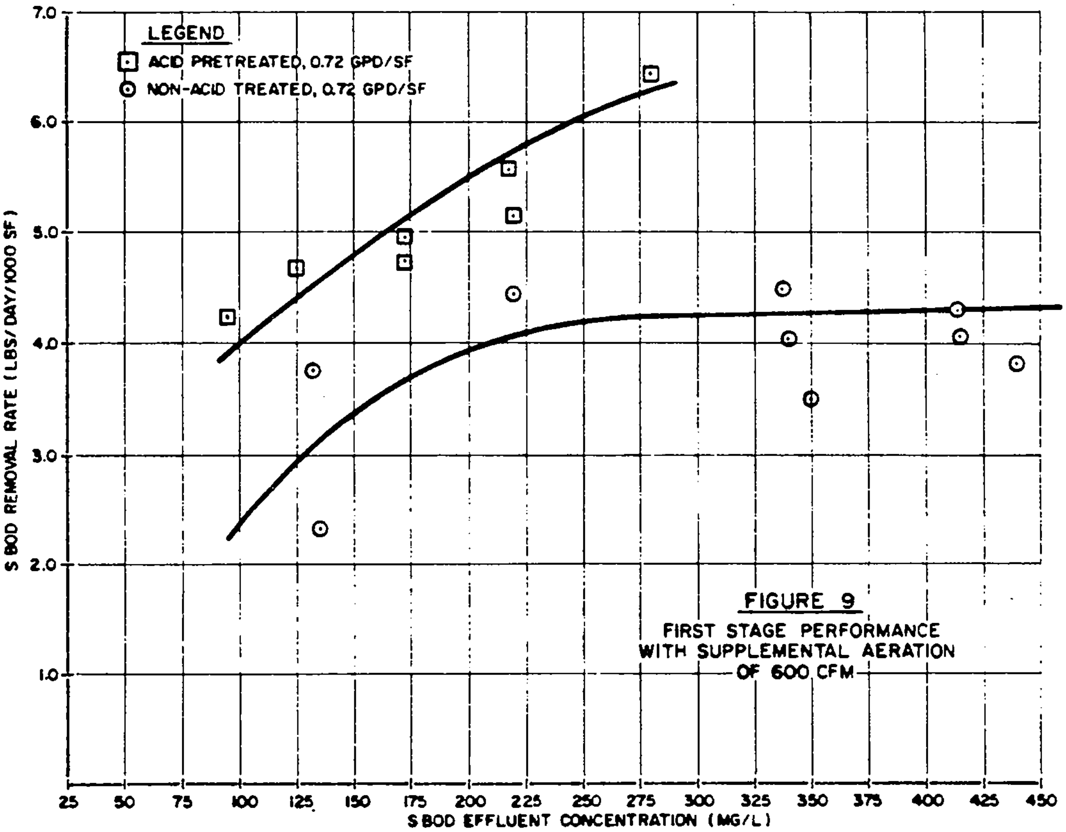


FIGURE 9
FIRST STAGE PERFORMANCE WITH SUPPLEMENTAL AERATION OF 600 CFM

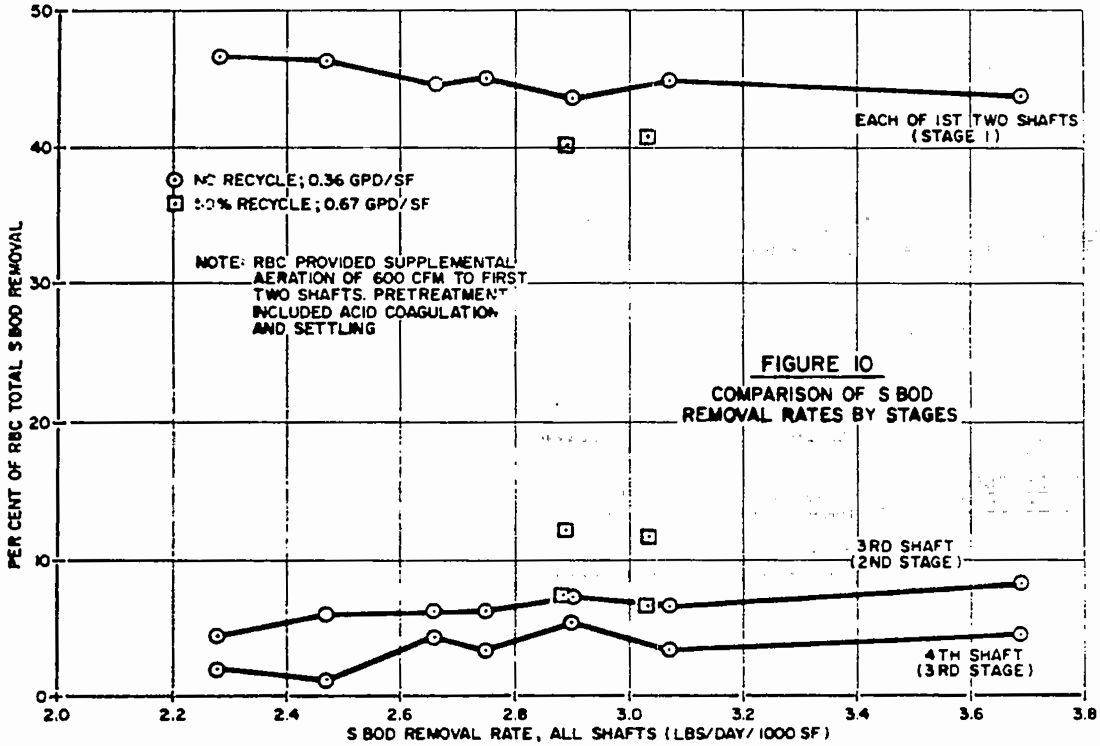


FIGURE 10

COMPARISON OF S BOD REMOVAL RATES BY STAGES

TABLE 7

AVERAGE SPECIFIC SBOD REMOVAL RATES FOR RBC STAGES

	Supplemental Aeration With Acid/Settling Pretreatment	Supplemental Aeration Without Acid/Settling Pretreatment
1st Stage	5.1 ppd/1000sf	3.7 ppd/1000sf
2nd Stage	0.7 ppd/1000sf	1.4 ppd/1000sf
3rd Stage	0.4 ppd/1000sf	0.7 ppd/1000sf
All Stages	2.83 ppd/1000sf	2.38 ppd/1000sf

Similar to the pilot plant, the full scale RBC with its best performance of SBOD removal still did not effect nitrification. The effluent ammonia nitrogen concentration remained within a range of 180 to 220 mg/l. Although COD removal was substantial through the entire treatment system, effluent values from the RBC remained in the 350 - 450 mg/l range.

Although the use of supplemental aeration with the RBC involves additional electrical energy to provide the compressed air, the SBOD removal rate per horsepower for the RBC in this application still appears to be reasonable. Based on the high performance curve of Figure 7 and a total of 20 draw horsepower for the blower and the four rotary drives, the SBOD removal rate per horsepower to meet summertime permit requirements of 112 mg/l is approximately 2.7 pounds of SBOD per hours per horsepower. Obviously, if diffused aeration is used for both supplemental oxygenation and shaft rotation, the energy efficiency should be enhanced.

SUMMARY AND CONCLUSION

The use of supplemental aeration with the RBC unit at the Curwensville Tannery significantly improved the performance of this process. Supplemental aeration, through limited diffusion to the basin contents, provides an additional source of oxygen for biomass respiration and for prevention/control of sulfate reduction. The diffused air flow below the rotating assemblies also assists to maintain a thinner biofilm on the plastic surfaces, which reduces the possibility of developing sub-surface anaerobic conditions which can promote sulfide generation with these wastes of high sulfate content. In combination with improved pretreatment and product processing modifications, the diffused air supplemented RBC facility was able to achieve design based pilot plant performance and to exceed present effluent permit requirements for BOD.

The serious problems caused by the generation of sulfide within the tank contents of the RBC and within the attached biomass should be viewed as a limitation to the use of non-aerated RBC for wastes of high sulfur or sulfate content.

Upon repair to the fullscale RBC facility and installation of the turbine aerators in the equalization tank, additional testing will be undertaken. This will be to determine the effect of equalization tank aeration in reducing the ammonia content of the waste by air stripping, and agglomerating and precipitating BOD constituents for removal in the intermediate clarifier prior to the RBC. This approach for reducing the BOD content to the RBC is preferred to the acid coagulation procedure, because of the difficulties involved in concentrating and dewatering the acid sludges.

It is also intended that a more quantitative evaluation be made of the effect of calcium accumulation on the rotary surfaces both in terms of structural stress and biological effect.

The problems in applying pilot plant experience to full scale design as reported herein is further evidence of the difficulty which can be encountered in scaling-up RBC pilot plant results. Although it is evident that the differences in waste characteristics were in part responsible for the performance problems of the fullscale system, significant contribution to the problem was also due to the much higher oxygen supported SBOD removal rate of

OPERATIONAL EXPERIENCE OF OXYGEN-ENRICHED ROTATING BIOLOGICAL CONTACTORS

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INTRODUCTION

The use of rotating biological contactors (RBC) for treating both municipal and industrial wasters has gained considerable popularity in recent years. This is partly because of its low energy need and ease of operation and maintenance. It is also partly because the attached microbial growth on the RBC unit can allow different groups of microorganisms to exist at different discs within a single treatment unit. This provides a valuable feature for achieving biological nitrification.

One of the major factors limiting the performance of an RBC unit is the availability of oxygen in the treatment system. In the literature it has been well documented that the rate of organic stabilization in such a system is generally limited by the oxygen penetration rather than by the substrate diffusion into the biological film.[1,2] It is only in a multi-stage system that the substrate diffusion may become a rate-limiting factor in the last stages of the system.[3,4] In fact, numerous field installations have experienced variable degrees of septic problems in their treatment units. Many researchers[3,5,6,7] have also found that the organic stabilization rate increases with the disc rotating speed. This is because a higher rotating speed would cause a greater oxygen transfer efficiency. Unfortunately, the rotating speed cannot be increased indefinitely without causing some major drawbacks. First, power requirement increases exponentially with the disc rotating speed.[3,8] Secondly, an excessively high rotating speed creates a high hydraulic shearing force which may interfere with satisfactory development of biomass on the disc surface. Antonie[8,9] has suggested that the optimal peripheral speed for treating domestic wastewater is about 18.2 m/min (60 ft/min). Therefore, in order to increase the oxygen penetration in the RBC system, some appropriate means other than unlimitedly increasing the rotating speed is necessary. There are at least two possible methods to accomplish this. One is to use an enclosed RBC system and replace air with pure oxygen as the feed gas. The other is to pressurize the enclosed RBC system using either air or pure oxygen. In either case, the partial pressure of oxygen in the gaseous phase is increased and the oxygen penetration into the biofilm can thus be increased. The objective of this study was to investigate the operational characteristics of several bench-scale, enclosed RBC units receiving oxygen as the feed gas.

EXPERIMENTAL STUDY

For the convenience of this study, a synthetic milk waste was used as the influent feed. Three bench-scale, enclosed RBC units were built and each of them was operated under a specifically designed condition. The operational characteristics of these units were evaluated in order to assess the advantages of using the oxygen-enriched RBC system.

Influent Feed. The synthetic milk waste was prepared at two different strengths, the higher one being twice as strong as the lower one. The single-strength waste was prepared by adding 0.900 g of a commercial dry milk powder (Kroger Brand, a product of Kroger Company, Cincinnati, OH) and 0.084 g of K_2HPO_4 to each l of tap water. The dibasic potassium phosphate was added to both supplement the phosphorus content and buffer the pH so that the feed solution would not drop to the isoelectric range of casein to cause protein precipitation. The chemical oxygen demand (COD) of this waste was about 1,000 mg/l. Its pH was 7.2. During the experimental study, a suitable quantity of this waste was prepared each day, which was then placed in three 55-gal drums stored inside a 4°C walk-in cooler. The waste was continuously stirred to keep undissolved milk solids in suspension.

Rotating Biological Contactors. The three bench-scale RBC units were built with plexiglas tubes, each 30.5 cm (12 in.) OD and 0.61 m (2 ft) long with a wall thickness of 3.18 mm (0.125 in.). Each tube was made to hold a series of discs submerged at one-half tube depth. The end plates as well as the rotating shaft-bearing assembly were so constructed that the entire unit could withstand a pressure of 68.95 kN/m^2 (10 psig) without any water or gas leaks. The detail construction drawing is shown in Figure 1.

The end plates, each 12.7 mm (0.5 in.) thick, were milled to have a groove 1.5 mm (0.06 in.) deep to fit the tubular wall with a clearance of 0.64 mm (0.025). The groove was filled with a rubber-based silicone seal to serve as an O-ring gasket. The two end plates were pressed against the tubular body and secured together with eight connector rods, each 4.76 mm (0.188 in.) and spaced at 45° intervals. With this construction, the unit was able to achieve a gas-tight condition, even with an applied pressure of up to 68.95 kN/m^2 (10 psig).

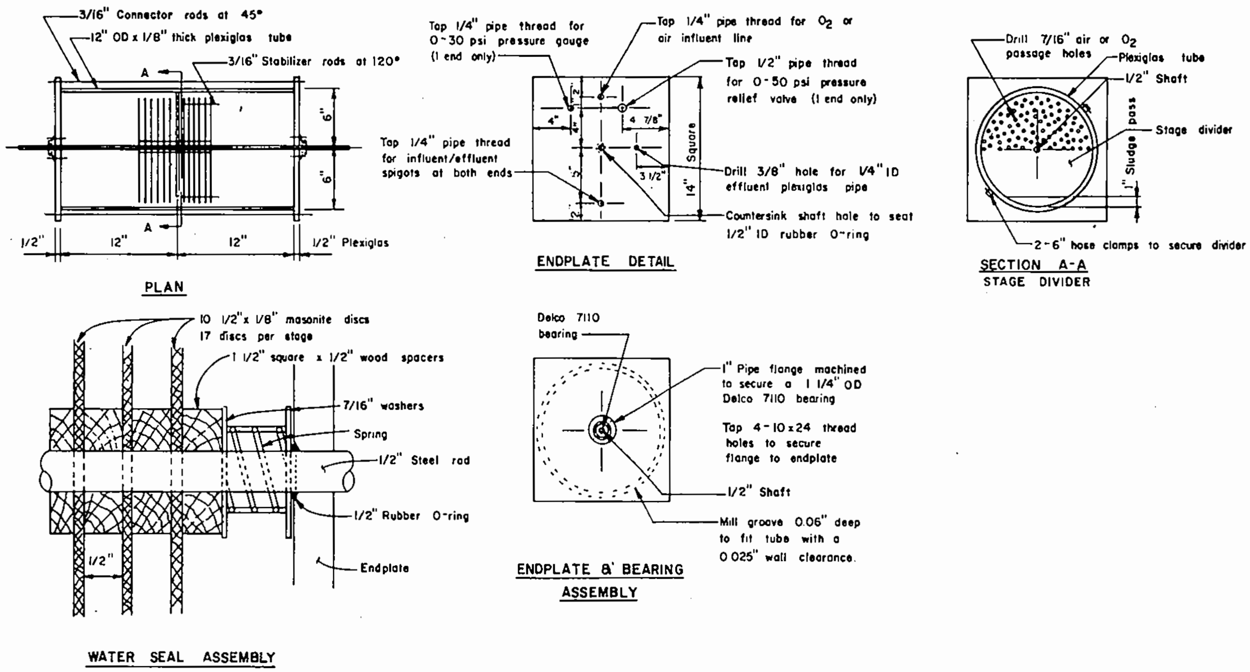


Figure 1. Construction Detail of the Experimental Rotating Biological Contactors

Inside the tublar body was a total of 34 circular discs, each 26.7 cm (10.5 in.) in diam x 3.2 mm (0.125 in.) in thickness. The discs were equally spaced with wood spacers to result in a face-to-face clearance of 12.7 mm (0.5 in.). Three threaded stabilizer rods, each 4.76 mm (0.188in.) in diam and spaced at 120° were passed through holes drilled 12.7 mm (0.5in.) from the outer edge of each disc, and then bolted at both sides of the disc to hold it in alignment.

The tubular body was divided at the midpoint into two equal sections with a divider so that each RBC unit would function as a two-stage system. The bottom 25.4 mm (1 in.) portion of the divider was cut off to facilitate passage of the liquid and sludge. Numerous 11.1 mm (0.44 in.) holes were drilled in the upper half of the divider to allow the passage of gas. The divider was cut to fit exactly into the tubular body and was then secured in place using hose clamp pressure around the circumference of the outer wall of the tubular body.

In order to make the entire RBC system air and water tight, a special shaft-bearing assembly had to be devised. The shaft holes of the end plates were countersunk to seat a 12.7 mm (0.5 in.) ID O-ring. A spring was then positioned to push the O-ring into the countersunk space and prevent leaks around the shaft. However, due to friction the O-rings were found to gradually wear out and had to be replaced at 2-3 wk intervals.

The end plates were properly tapped and threaded for installations of liquid influent and effluent lines, gas inlet and outlet, the latter being a pressure relief valve, and a pressure gauge or mercury manometer for accurately monitoring the internal operating pressure. Both influent and effluent lines were connected to positive displacement pumps so that the system could be maintained at any desirable pressure. Also, just before the stage divider, a spigot was tapped to facilitate the withdrawal of the first-stage effluent sample. The detail schematic of the three experimental RBC systems is shown in Figure 2.

During operation, the water level was maintained at half depth of the tubular body, and the rotating speed was set at 12 rpm, which provided a peripheral velocity of 10 m/min (33 fpm). The flow rate was rigidly controlled at 56 ml/min (2,105 gal/day). The following represents the pertinent information about the system operation:

Liquid flow rate	56 ml/min (21.5 gpd)
No. of stages in each RBC system	2
No. of discs in each stage	17
Disc diam	266.7 mm (10.5 in.)
Rotating speed	10 m/min (33 fpm)
Face-to-face spacing between discs	12.7 mm (0.5 in.)

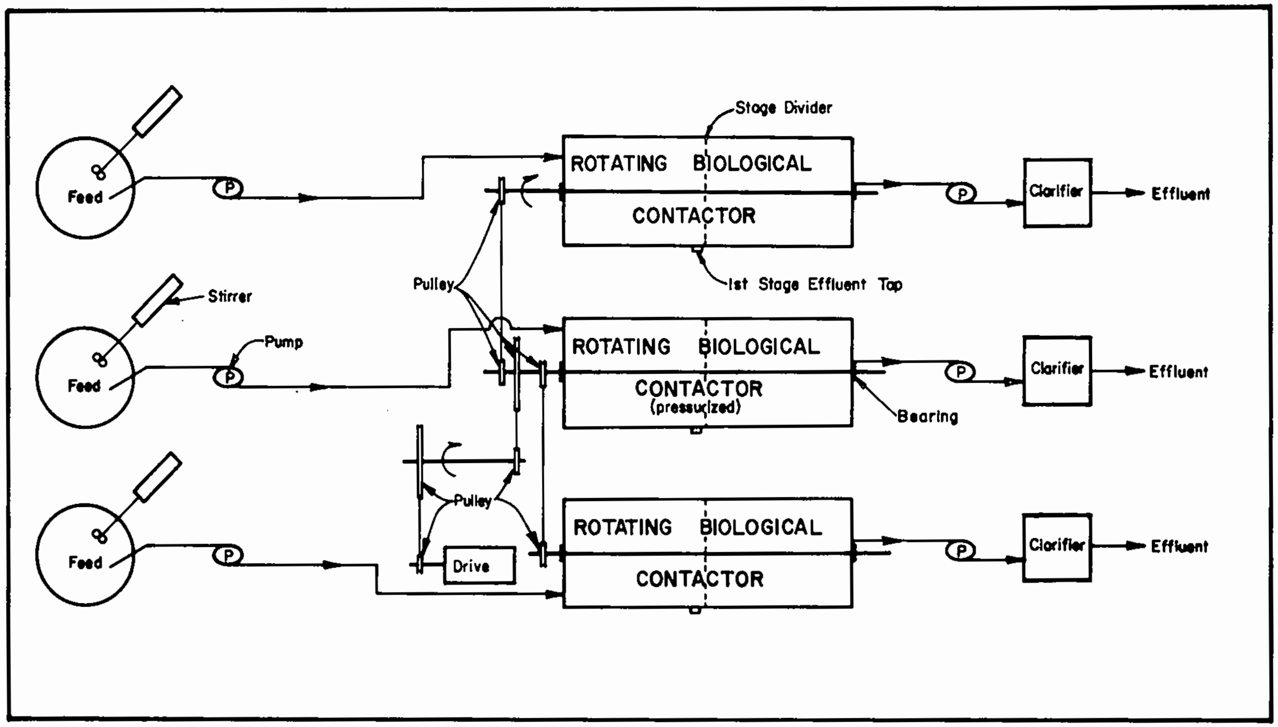


Figure 2. Detail Operational Schematic of the Three RBC Systems

Liquid volume in each stage	10.58 l (2.82 gal)
Gas volume in each stage	10.58 l (2.82 gal)
Total disc surface in each stage	1.89 m^2 (20.6 ft^2)
Ratio of vol/surface in each stage	5.60 l/m^2 (0.14 gal/ft^2)
Ratio of flow rate/surface in each stage	42.7 l/m^2-day (1.04 gpd/ft^2)
Detention time in each stage	3.15 hr

Experimental Approach. Three separate experimental runs were conducted in this study. The operating conditions designed for each RBC system in these three experimental runs are shown in Table I.

The first experimental run was conducted using air as the feed gas. The gas flow rate was controlled at 420 ml/min for all three RBC units. This was equivalent to 1.0 ft^3 air per gal of influent waste flow, similar to that commonly used by an activated sludge plant. The first RBC system was not operated under pressurization, and it was designated as Unit I. The second RBC system was operated under a gauge pressure of 27.6 kN/m^2 (4 psig). Since this unit received the same influent waste (i.e., single- strength or 1000 mg/l COD) as the first unit except that it was pressurized, this second unit was designated as Unit I-P. The notation "P" indicates operation under pressure. The third RBC system received a double-strength waste (i.e., 2000 mg/l COD) and it was also operated under a pressure of 27.6 kN/m^2 (4 psig). Thus, this unit was designated as Unit II-P; that is, the notation "II-P" indicates that the unit received the double-strength waste and it was operated under pressure.

The first experimental run was started in late April of 1978 when the three RBC systems were first placed in operation. During the first month the synthetic milk waste was mixed with gradually decreasing concentrations of sewage to provide microbial seedings and speed up the establishment of biomass on the disc surface. Thereafter, only the synthetic milk waste was used. By early August, all three RBC units had reached consistent operation as evidenced by a fairly uniform reduction of COD in the day-to-day operation. An intensive monitoring program starting August 7 and ending August 18, was undertaken to evaluate the operational characteristics of each RBC system. This included the measurement of such performance parameters as soluble chemical oxygen demand (COD), suspended solids (SS), ammonia nitrogen, nitrate nitrogen, and sludge settling rate. Also, the mixed liquor suspended solids (MLSS) and the dissolved oxygen (DO) concentrations in each stage of the three RBC system were also determined. All of these tests were performed according to the procedures set forth in Standard Methods.[10]

After the detail operational characteristics of the first experimental run were obtained, the three RBC units were switched to the new operating condition designed for the second experimental run. That is, the feed gas was switched to pure oxygen and its flow rate was reduced to 240 ml/min, which was equal to 0.57 ft^3 oxygen per gal of influent waste. A "transitional" period of slightly more than a week was observed before the three RBC units reestablish consistent operation. Thus, at the end of the third week (i.e., beginning September 11, 1978), a 10-day intensive monitoring program was again undertaken to assess the performance characteristics of the second experimental run.

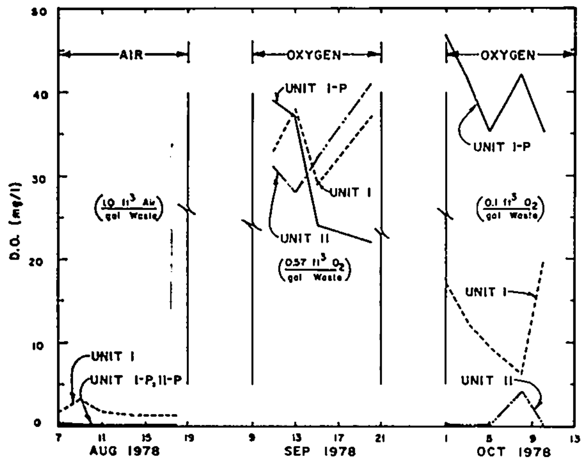
Similarly, at the conclusion of the second experimental run, the three RBC units were switched to the operation condition designed for the next phase, or the third experimental run. Actually, the only change in this phase was to reduce the oxygen flow rate from 240 ml/min to 42 ml/min. This low flow rate was attempted because in the second experimental run, the DO levels in all three RBC units were found to be excessively high. It took only a few days for the three RBC units to reestablish consistent performance. Thus, the intensive monitoring work was again started October 1, 1978. It lasted for 10 days before this phase of work was concluded.

RESULTS AND DISCUSSION

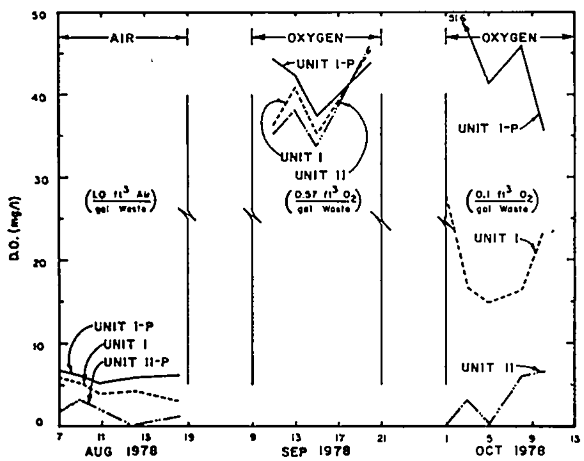
The main objective of this research was to evaluate the operational characteristics of parallel bench-scale RBC units under an oxygen-enriched environment with various organic loadings. The following will discuss the findings of various operational parameters examined in this study.

D.O. In this study, the main reason for using either pure oxygen or pressurization in an RBC unit was to increase the partial pressure of oxygen in the treatment system so that the oxygen availability would not become a limiting factor in the biooxidation process. In the three separate experimental runs, the first was conducted using compressed air, while the latter two were conducted using pure oxygen. Since the rate of organic loading on each specific RBC unit was the same for all three experimental runs, it was reasonable to expect that the use of pure oxygen would yield a higher D.O. level provided that the gas flow was adequate. This expectation was generally verified from the results obtained in this study (Figure 3).

In the first experimental run (conducted in August, 1978), the D.O. levels in the first stage of all three RBC systems were generally low. The two units under pressurization (Units I-P and II-P) were actually found to have D.O. from 0.1 to 0.5 mg/l, which were less than the value observed for the unpressurized unit (Unit I), approximately 2-3 mg/l. Since Unit II-P received twice as much organic loading as Unit I and the increment of oxygen partial pressure through pressurization amounted to only 27 percent (i.e., 4/14.7=27%), it was understandable that the D.O. in this unit would be lower than that in Unit I. However, it was difficult to explain why Unit I-P had a lower D.O. level than Unit I. The D. O. in the second stage did follow a reasonable trend, i.e., Unit I-P had the highest level, followed by Unit I, and then Unit II-P.



(a) First Stage - D.O.



(b) Second Stage - D.O.

Figure 3. Variations of D.O. in Each RBC System in the Three Separate Experimental Runs

In the second experimental run (September 1978), in which pure oxygen was injected at a rate of 240 m l/min or equivalent to 0.57 ft^3 oxygen/gal waste, all three RBC systems were found to have high D.O. (at least 20 mg/l) in both the first and second stages. The day-to-day fluctuations of D.O. in any specific units were partly due to variations of MLSS present in the systems (as will be shown later) and partly due to possible experimental errors encountered in the D. O. measurement using the probe technique. Because of the excessively high D. O. present in the withdrawn samples, they had to be diluted first with deoxygenated water to bring the D.O. down to a measurable range (maximum 15 to 20 mg/l) by the probe technique. Through this dilution, a portion of the over-saturated D. O. could be easily released.

In the third experimental run (October 1978), in which the pure oxygen flow rate was reduced to only 42 m l/min or equivalent to 0.1 ft^3/gal waste, the D.O. variations in all three RBC units generally followed an expectable pattern. That is, Unit I-P had the highest D.O., followed by Unit I and then Unit II. In fact, the D.O. in Unit II, particularly in the first stage, was either zero or close to zero in many observations, indicating that the supply of oxygen to this unit was not adequate to meet the biological demand.

Biomass Growth The periodic sloughings of biomass from the disc are normally due to the anaerobic activity developed at the deep layer of the biofilm[8,11] Thus, in any RBC system if the oxygen transfer is increased (through the use of either pure oxygen or pressurization), the development of biofilm would also be expected to increase. This expectation was substantiated in this study. It was consistently demonstrated in the three separate experimental runs that whenever the unit was supplied with pure oxygen, particularly in conjunction with pressurization, an extraordinarily heavy growth of biofilms was found. In fact, the biofilms in the first stage were so thick that the clearance between many discs were covered with biomass (Figure 4).

However, the observed heavy accumulations of biomass on the disc surface must not be misinterpreted as large productions of sludge mass from the RBC system. It has been reported[12] that the use of oxygen-enriched atmosphere tends to reduce sludge production in an RBC system. In this study no attempts were made to measure the biofilm thickness because of the difficulties involved in gaining an access to an enclosed RBC unit. However, routine determinations of the MLSS were made for all liquid samples withdrawn from both stages of the three RBC units, and the results are shown in Figure 5. It was found that the MLSS observed during the short monitoring period of each separate experimental run was quite variable and therefore, did not truly reflect the exact quantity of biomass present in the RBC system. This variation of MLSS was caused mainly by inconsistent sloughings of the biofilms during the three experimental runs. It appeared that the sloughings occurred in a somewhat cyclic manner. As such, there were times when the measured MLSS was low, but the thickness of biofilm on the disc surface was very high. Therefore, if it were intended to determine the relative sludge productions amoung different RBC units, it was necessary to conduct each experimental phase over a sufficiently long period of time so that the collected MLSS data can be analyzed statistically in order to obtain a realistic sludge production rate. Neverthless, the data

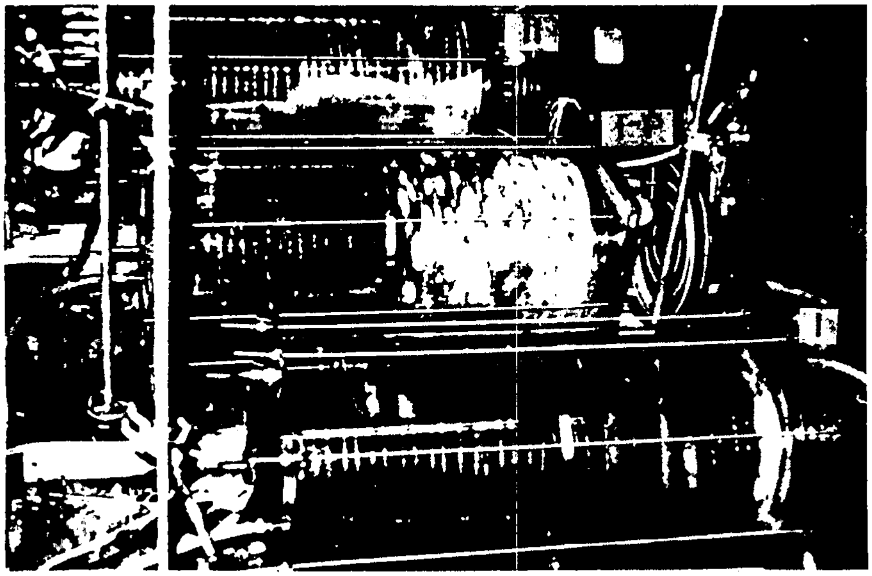
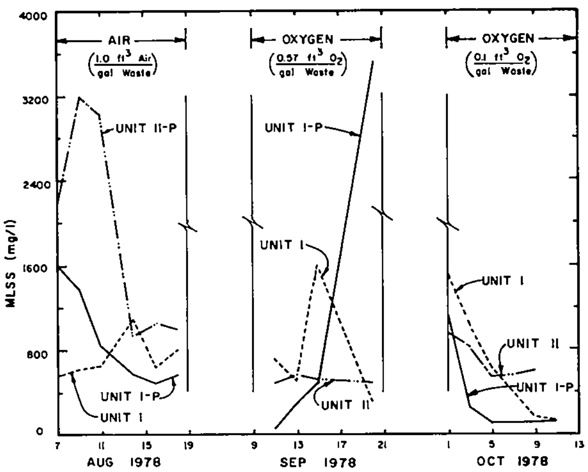
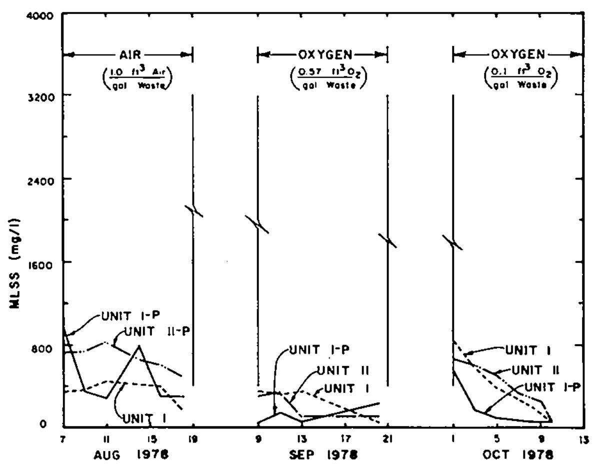


Figure 4. Development of Thick Biofilm in the Oxygen Pressurized RBC Unit



(a) First Stage - MLSS



(b) Second Stage - MLSS

Figure 5. Variations of MLSS in Each RBC System in the Three Separate Experimental Runs

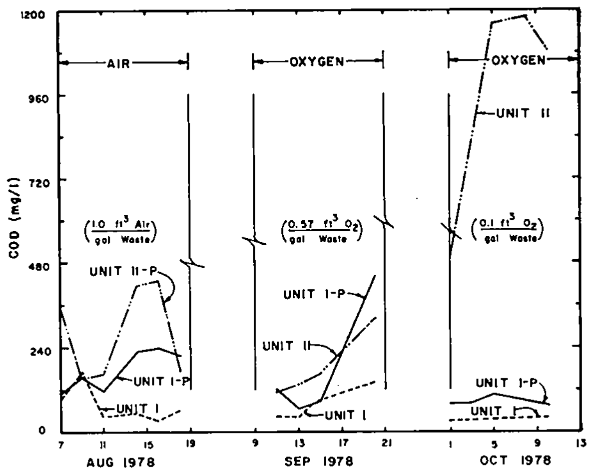
shown in Figure 5 did indicate that the quantity of MLSS in the first stage of each RBC unit was considerably higher than that in the corresponding second stage. This was, of course, because there were not much organics remaining in the second stage, and the growth of biofilms was limited by the availability of organic substrate rather than by the oxygen penetration.

pH Throughout the three separate experimental runs, the pH values in each of the RBC systems were found to stay within 7 + 1 units. In general, the pH level was slightly higher in the second stage than in the first stage. This could be due to deamination of milk protein in the bio-oxidation process. There was only one instance at which the pH did fall below 6. This occurred in Unit II of the third experimental run when the oxygen supply to this unit was not adequate to maintain a complete aerobic condition (Figure 3). As such, a buildup of organic acids could have occurred.

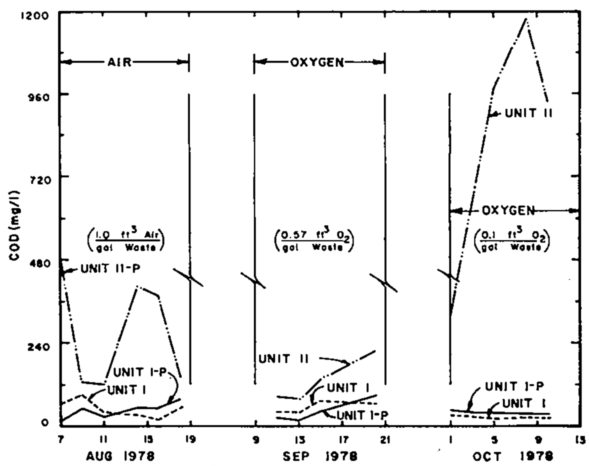
Soluble COD The reduction of COD in each RBC unit depended not only on the quantity of biomass present in the system, but also on the diffusability of organic substrate into the biomass. It was mentioned earlier that when an RBC unit was supplied with an adequate flow of pure oxygen, particularly under pressurization, it was able to develop an extraordinarily thick layer of biomass in its first stage. However, this heavy growth of biofilms had not always been accompanied by a higher level of COD removal compared to the other units, as shown in Figure 6 (a). In fact, the unpressurized Unit I showed a consistently better removal of COD than its pressurized counterpart (Unit I-P). It was believed that the over development of the biofilm and its subsequent bridging over the disc clearance actually reduced the available surface area for organic substrate to reach the bulk of the biomass. This kind of problem was not observed in the second stage of Units I and I-P, Figure 6 (b). Therefore, their relative COD reductions in the second stage were quite comparable and both were likely limited by the substrate diffusion. Because of this, in the future practical design of an RBC system using pure oxygen, it is important to provide adequate spacings (depending on the waste strength, but at least 1 in. for a waste having a COD of 1000 mg/L) between discs to aviod biomass-bridging over the discs.

In the RBC unit receiving the double-strength waste (i.e., Unit II), the COD reductions in both stages were considerably higher in the second experimental run (the unit received an adequate flow of pure oxygen) than those in the first and third experimental runs. This would apparently suggest that, at a high organic loading rate, the limiting factor for the COD removal was the oxygen flux. Since the second experimental run supplied more oxygen than the other two, it was able to achieve a better COD removal. Therefore, it is reasonable to conclude that use of pure oxygen in sufficient quantities was able to increase the COD removal in a heavily loaded RBC system. If the oxygen supply was not sufficient, such as that occurred in the third experimental run, the system could become anaerobic (Figure 3) and the COD removal would be drastically reduced.

Sludge Settleability and Effluent Suspend Solids. The overall treatment efficiency of any biological system depends greatly on the settleability of biological solids in the secondary clarifier. Since sloughings of biofilms in an RBC system generally occur in a cyclic manner, its secondary clarifier



(a) First Stage - Soluble COD



(b) Second Stage - Soluble COD

Figure 6. Comparative Reductions of Soluble COD in Different RBC Systems

would be subject to periodic shock loadings of biological solids. Thus, the sludge settleability would become critical in dictating the overall treatment efficiency. It has been reported by Bintanja, et al.[12] that the use of an oxygen-enriched atmosphere is able to improve the sludge settleability. This finding had also been substantiated in this study.

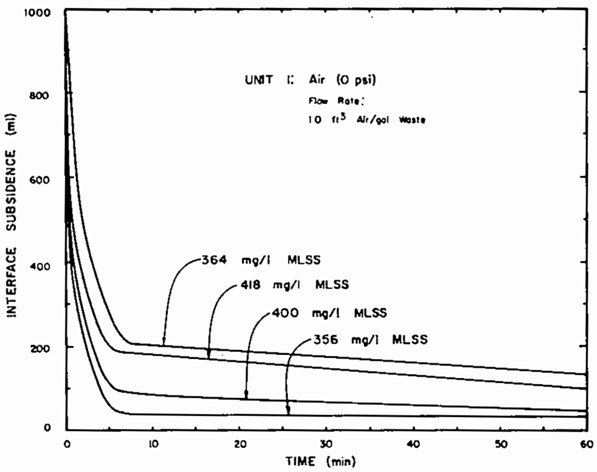
Figure 7 shows a striking improvement of the sludge settleability observed in Unit I when pure oxygen was used to replace air. Figure 8 further indicates that, even with pressurization alone, an RBC unit supplied with air could also achieve a significant improvement of the sludge settleability as compared to an unpressurized unit, i.e., Figure 7 (a) vs. Figure 8 (a). When pressurization was used in conjunction with pure oxygen, the sludge settleability was extremely good, and the settled sludge also compacted well, as shown in Figure 8 (b). In fact, the sludge produced from such a unit appeared to consist of very dense granules, as shown in Figure 9 . This figure compares the relative settling and compaction of the three MLSS samples taken from the three "oxygenated" units designated, respectively, as Units I (0 psi), I-P (4-psi) and II (0 psi).

In order to shed some light on the microbial nature of the sludges obtained from different RBC units, photomicrographs were taken throughout the course of the study. Typical results are shown in Figure 10 which indicates that Unit I (with unpressurized air) contained a large quantity of filamentous microorganisms, Photo (a). But when the air was replaced by oxygen, the extent of filamentous growth was significantly reduced, Photo (b). Pressurization of an RBC system also showed a reduction of the filamentous growth, Photo (c) and (d). The sludge in the oxygen-pressurized RBC unit appeared to be "chunky" and dense, as shown in Photo (d).

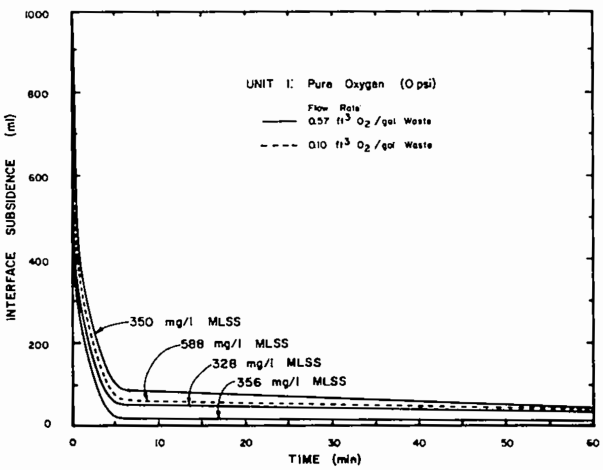
Because of the good sludge settleability, Units I and I-P almost consistently had lower effluent SS (measured after 1-hr settling) in the second and third experimental runs (i.e., using pure oxygen) than in the first experimental run (Figure 11). The effluent SS of Unit II were high in the third experimental run. This was because this unit had experienced anaerobic conditions from time to time, which had a deteriorating effect on the sludge settling quality.

Nitrification. Rotating biological contactors have been known to be effective in achieving nitrification in the multi-stage operation. This is possible because the slow-growing nitrifiers can grow separately from the saprophytes in the latter stages of the RBC system. Torpey, et al.[4] reported that use of an oxygen-enriched atmosphere was able to accelerate the BOD removal, thereby allowing nitrification to occur effectively in an earlier stage. In addition, the rate of nitrification was also increased by the oxygen enrichment.

In this study, it was found that when an RBC unit was supplied with air, pressurization was able to enhance nitrification. Figures 12 and 13 show that in the first experimental run (August 1978), both stages of Unit I-P had lower NH_3-N and higher NO_3-N than the unpressurized Unit I. Even in the double-loaded Unit II-P, considerable extents of nitrification were

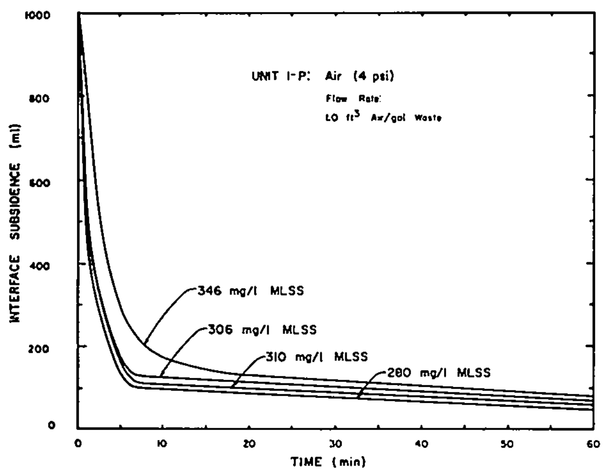


(a) Unit I - Air at 0 psi

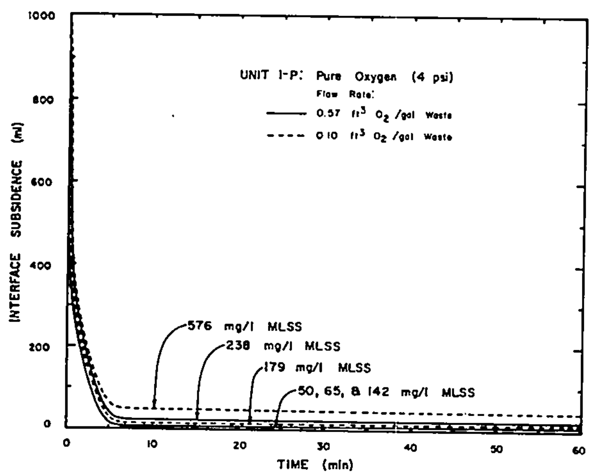


(b) Unit I - Pure Oxygen at 0 psi

Figure 7. Improvement of Sludge Settleability by Using Pure Oxygen

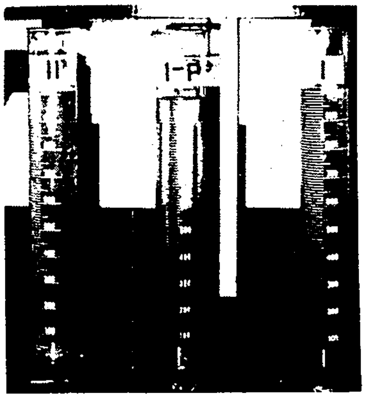


(a) Unit I-P - Air at 4 psi



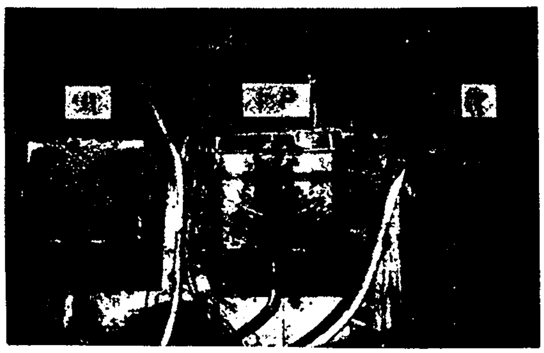
(b) Unit I-P - Pure Oxygen at 4 psi

Figure 8. Improvement of Sludge Settleability by Using Pressurization



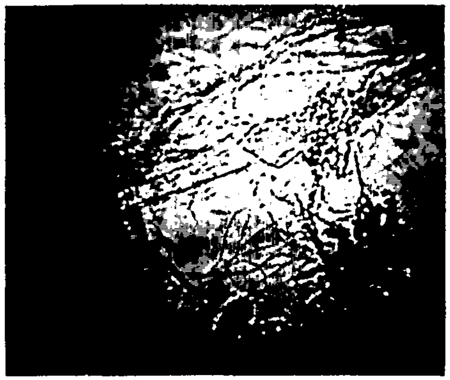
(a)
Beginning of Settling

(b)
One Hr After Settling



(c)
Secondary Clarifier

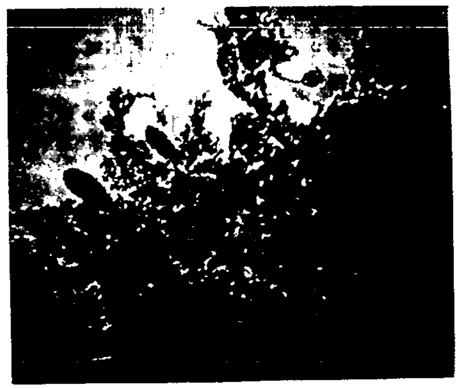
Figure 9. Relative Settling Qualities of Sludges from Different RBC Units with Pure Oxygen



(a) Air @ 0 psi

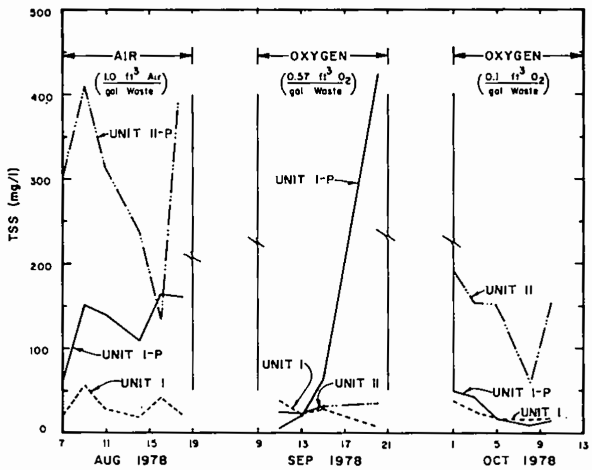
(b) Oxygen @ 0 psi

(c) Air @ 4 psi

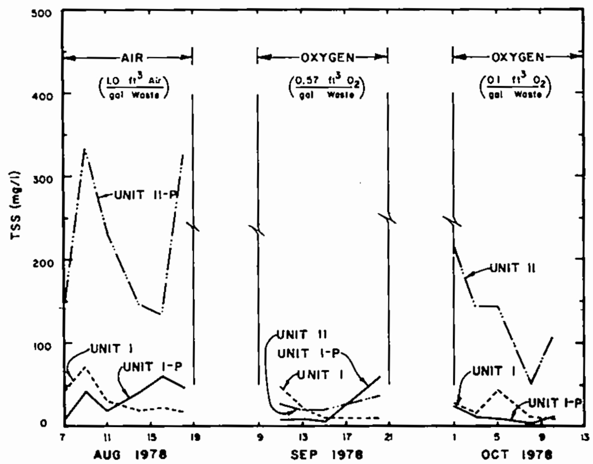


(d) Oxygen @ 4 psi

Figure 10. Photomicrographs (100X) of Sludge Cultures Obtained from Different RBC Units

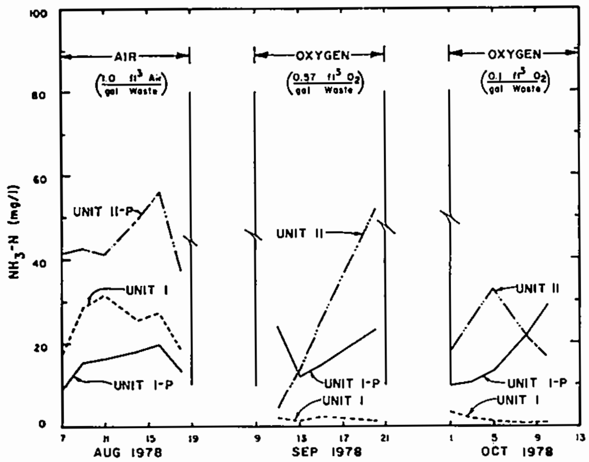


(a) First Stage - Effluent SS

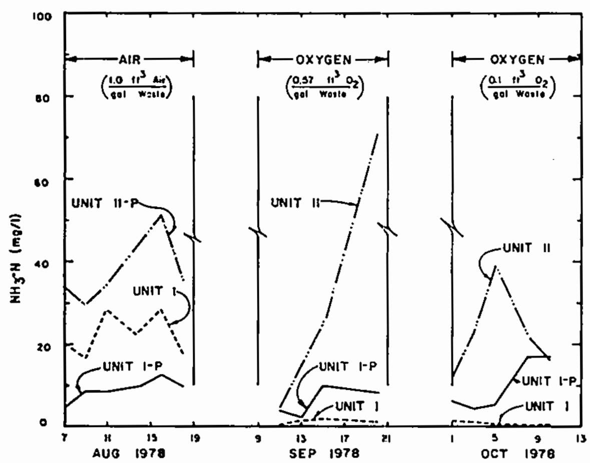


(b) Second Stage - Effluent SS

Figure 11. Variations of Effluent SS in the Three RBC Systems

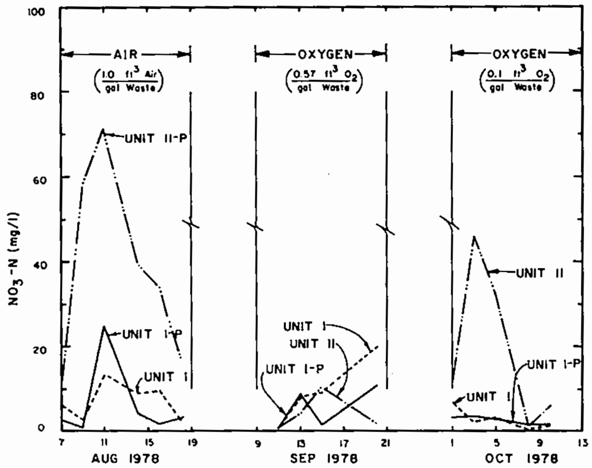


(a) First Stage - NH_3-N

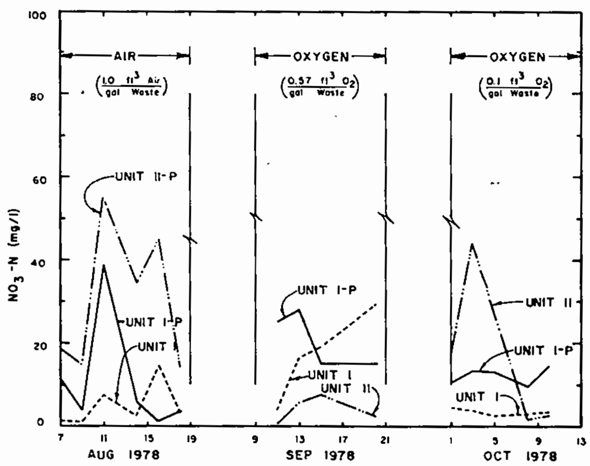


(b) Second Stage - NH_3-N

Figure 12. Variations of NH_3-N in the Three RBC Systems



(a) First Stage - NO_3-N



(b) Second Stage - NO_3-N

Figure 13. Variations of NO_3-N in the Three RBC Systems

also observed in both stages of the system. When air was replaced with pure oxygen in the second experimental run (September 1978), both stages of Units I and I-P appeared to have lower NH_3-N than those in the first experimental run. However, this was not always accompanied by greater amounts of NO_3-N in the treated effluent. There were two possible explanations for this. First, the heavy production of biofilms in this experimental run could have assimilated a greater amount of NH_3-N into the biomass. Secondly, the excessively high concentrations of oxygen (well above 20 mg/1) present in these RBC units could have imposed some inhibitory effects on the nitrifiers. The latter explanation may also apply to the data observed in Unit II, which showed a lower NO_3-N content in the second experimental run than in the first one.

The data of the third experimental run (October 1978) generally repeated the pattern of the second experimental run, except that more-extensive nitrification occurred in Unit II due to its "less than excessive" oxygen content present in the system. All of these seems to support the suggestion that an excessively high concentration of oxygen (20 mg/1 or above) in an RBC system can impose an inhibitory effect on nitrifiers. If this suggestion is indeed true, then in the future multi-stage design, use of pure oxygen should be limited to only the front stages of the system. The latter stages should use only regular air to allow effective nitrification to occur. This conclusion must be considered preliminary, and further research work has to be done to substantiate its fact.

CONCLUSION

Based on the findings of this study, the following conclusions can be drawn:

1. Use of pure oxygen, especially in conjunction with a pressurization of 27.6 kN/m^2 (4 psig), was able to allow a thick layer of biofilms to develop in an RBC system treating a synthetic milk waste having a COD of 1000 mg/1 . The biofilms were thick enough to bridge the 1.27 cm (0.5 in.) clearance between discs in the first stage of the RBC system.

2. When the bridging of disc clearance was too extensive, the available surface area for organic substrates to diffuse into the bulk of the biomass became significantly reduced. This would result in a reduction of the COD removal. Therefore, in the design of an RBC system using pure oxygen, an adequate spacing (at least 1 in.) between discs must be provided to avoid the biomass-bridging.

3. If there were no extensive biomass bridging and the substrate concentration was high, use of pure oxygen in sufficient quantities was able to improve the COD removal as compared to the use of air.

4. Use of pure oxygen in an RBC system showed a definite improvement of the sludge settleability. The sludge appeared to be very dense and was generally absent of filamentous growths. The use of pressurized air had a similar, but less noticeable effect.

5. Pressurization of an air-supplied RBC unit was able to enhance nitrification. However, when air was replaced with pure oxygen, the excessively high concentration of oxygen present in an RBC treatment system appeared to have an inhibitory effect on the nitrifiers.

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PART VI: INDUSTRIAL WASTEWATER TREATMENT

WASTEWATER TREATABILITY STUDIES FOR A GRASSROOTS CHEMICAL COMPLEX USING BENCH SCALE ROTATING BIOLOGICAL CONTACTORS

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Introduction

Pilot scale test units are normally recommended when evaluating the applicability of Rotating Biological Contactor (RBC) systems for the treatment of industrial wastes. However, insufficient samples, lack of funds and/or tight schedules often create circumstances which make it impractical to conduct treatability studies in the larger pilot test units. To accommodate these situations, the Environmental Systems Division of Catalytic, Inc. has fabricated bench-scale, modular design RBC systems and utilized them for industrial waste treatability studies where the benefits of the RBC concept appeared to have potential application.

This paper describes an application of this equipment on a project where pilot-scale operation was not possible; and where the treatment system design was part of a total design of a grassroots production facility.

Catalytic International in London, England was one of several contractors working on different process areas and off-sites of a large chemical complex for the client, Berol Kemi AB of Sweden.

During the engineering design phase, Catalytic's Environmental Systems Division provided consultation and laboratory services to Catalytic in London for treatment of the wastewaters that would be generated from the whole chemical complex. The plant was to be built in a virgin area on Swedens western coast. The combined discharge was estimated to be about 120,000 gpd including contaminated stormwater.

The site was in an area previously classified as non-industrial in Stenungshund, Sweden. Oxochemicals and phthalate plasticizers were the principal compounds scheduled for production. The wastewater treatment facilities had to be installed and operable at the time of plant start-up.

The test work included primary treatment of the phthalate ester waste stream and of the partial oxidation wastes, and biological treatment of the total combined wastewaters using a rotating biological contactor. The development program was initiated on 10 June 1977 and completed on 12 August 1977. These tests and the resulting conclusions are summarized herein.

APPROACH

A sample representative of the combined wastewaters projected for the new facility was not available at any single site. However, some of the chemical process routes included in the proposed design were in operation at various locations around the world as part of other manufacturing operations. Consequently, it was necessary to calculate the wastewater composition using material balances and process experience. And then, where it was possible to obtain access, a limited number of selected wastewater streams were gathered at operating process units similar to those under design. These samples were used for characterization by comparison with material balances and for evaluation of pretreatment options.

A "recipe" was derived and the combined "wastewater" was synthesized in Catalytic's Environmental Laboratory in the U.S. to conduct the concept development study for RBC treatment of the wastes. The wastewater fed to the units was prepared daily using purchased chemicals, where possible, or from chemical combinations synthesized in the laboratory using the same process technology as for the production plant design. After simulating all the preferred pretreatment steps, the mixture was evaluated using two parallel bench-scale (25 centimeter diameter disks) 5-stage RBC units. The units were operated at steady state conditions under various loadings; and performance data were collected for each stage.

Pretreatment studies on selected process streams, settling and thickening tests on the biosludges, and screening for sludge dewatering were part of the study. The need to have the treatment plant on line for the start-up of the manufacturing facility created a tight schedule. Due to lack of time the following activities were not evaluated:

- o Waste variability effect on treatment
- o Evaluation for chronic toxicity of the waste
- o Biological treatment plant limits
- o Dilution effects on the biological treatment plant
- o Sludge stabilization

Regulatory Limits

There was no permit available at the time of the study. The following limitations were obtained from the presiding regulatory board, and these numbers were used as design basis:

	Ton/year
Organics	25
BOD_7	18
COD	37
Total Solids	To be determined
Cyanide	To be determined
Specific Metals	To be determined

At design flow, this can be further summarized as follows:

	Avg Kg/day	Avg. mg/l @ * Avg. Daily Flow	
BOD_7	49	225 @ 450 m^3/day	320 @ 317 m^3/day
COD	101	110 @ 450 m^3/day	155 @ 317 m^3/day

The Swedish regulatory board requirements stipulated 7-day BOD analysis as opposed to 5-day BOD. All BOD's run during the study are BOD_7.

Waste Characterization

Material balance data was obtained through the client from all the contractors involved in the design of the whole complex. These were reviewed and additional data were then requested more specific to waste discharges, since some of the process materials balances were not entirely adequate for this purpose. Additional information, derived from their existing processes, was obtained from the client, and also from the licensee of another of the processes. Wastewater "formulas" were developed, including process washes and process pad wash downs and stormwater. Following sampling and analytical characterization, treatability studies of certain process streams were run in order to evaluate pretreatment. The formulas were then refined in order to characterize the waste after pretreatment steps. This was done in order to make a judgement on whether or not the final concentration of any components would be detrimental to the biological process at concentrations present in the combined wastewater, or would exceed effluent limitations following biological treatment.

*450 m^3/day is the projected wet weather flow (includes storm water)
317 m^3/day is the projected dry weather flow

Table A summarizes some of these formulations. Due to their possible proprietary nature, the * components are not identified specifically. As can be seen from the table (Column A), there was a considerable quantity of "unidentified organics" that were derived from the material balances. The rest of the columns have these quantities distributed proportionally among the other organic components. Many of these unknowns may be byproducts of the process reactions. Some of the components had to be chemically synthesized in our laboratory using the proposed process technology. This necessity had one possible advantage. By synthesizing the chemicals, some of these unknowns are possibly included in the mix.

INVESTIGATIVE PROGRAM

Some of the process streams contained pollutants that could be harmful to biological treatment; or that might be of regulatory concern, but not significantly removed by biotreatment. Considerable characterization and treatability screening was conducted on the samples that were obtained from similar existing processes.

Acute microbiological toxicity studies were conducted separately on most of these constituents, in the form that they were added to the synthetic wastewater used in the "recipe". All the levels of acute toxicity were significantly greater than the levels that were expected in the wastewater derived from the material balance. Chronic toxicity was not observed, although the study period was too short for a meaningful evaluation of this aspect. Pretreatment studies were run, however, in order to gain some assurance of the feasibility of removal of certain of these pollutants from the process discharge, should it become necessary. These pretreatment studies are discussed briefly followed by discussion of the RBC study.

Pretreatment Screening

During an on-site visit to two existing chemical plants, samples were taken and jar tests were conducted on a "Sour Water" wastewater that was a product of one of the process steps.

The chemicals evaluated for removal of the suspended solids, metals, and sulfide in these two streams were lime (CaOH); aluminum sulfate ($Al_2(SO_4)_3 \cdot 18H_2O$); potassium hydroxide (KOH) in combination with ferric chloride ($FeCl_3 \cdot 6H_2O$); potassium hydroxide (KOH) in combination with ferrous sulfate ($FeSO_4 \cdot 7H_2O$); and lime (CaOH) in combination with ferric chloride ($FeCl_3 \cdot 6H_2O$).

The results of the jar tests conducted on-site were not conclusive. These tests were repeated in the laboratory, and based on these repeated tests and the improved treated suspended solids level, ferric chloride ($FeCl_3 \cdot 6H_2O$) was found to be the best chemical treatment scheme for these waters. However, the dosage of ferric chloride was not optimized for metals or sulfide removal, and would not be until the plant was in operation. A high-molecular weight polymer was also needed to improve the settleability. The following table summarizes the results of settling without chemicals and the ferric chloride treatment.

TABLE A

Summary of Waste Characteristics Tables

	A	B	C Dry Weather	C Wet Weather	D
Flow, cubic meters per day	317	317	317	450	-
*Unidentified organics, ppm	750	-	-	-	-
*Oxochemical, ppm	60	95	95	67	95
*Oxoalcohol, ppm	6	10	10	7	10
*Oxochemical, ppm	246	395	35	25	35
*Plasticizer intermediate, ppm	650	1035	35	25	35
*Plasticizer, ppm	-	-	70	50	70
Isopropanol, ppm	246	395	395	278	395
Butraldehyde, ppm	-	-	50	35	50
Naphtha, ppm	4	5	10	7	10
Sodium Formate, ppm	54	86	90	63	90
Hydrazine, ppm	11	11	11	8	11
Trisodium Phosphate, ppm	11	11	11	8	11
Sodium Sulphate, ppm	3763	3763	3763	2655	3763
*Heavy Metals, Catalysts & Iron	1-2	1-2	0.2-0.4	0.2-0.3	0.4
Sulfide, ppm	35	35	10	7	10
Chloride, ppm	44	44	150	106	150
Cyanide, ppm	1	1	0.2	0.1	0.2
Ammonia, ppm	35	35	35	25	35
Settleable Solids, ppm	123	123	1	Trace	1
Dissolved Solids, ppm	4000	4000	4000	2818	4000
BOD, ppm	-	-	1100	775	700
COD, ppm	-	-	1900	1340	1200

A - Waste characteristics from material balance

B - Synthetic Waste Composition (no pretreatment)

C - Synthetic Waste Composition (with pretreatment)

D - Synthetic Waste Composition Fed to RBC

Sample 1

	Raw	Gravity Settled 2 hrs.	$FeCl_3$ Treated
BOD	8.0 mg/l	5.0 mg/l	5 mg/l
COD	-	39.0 mg/l	23 mg/l
SS	1412.0 mg/l	43.0 mg/l	14 mg/l
TOC	128 mg/l	73.0 mg/l	31 mg/l
NH_3	62.0 mg/l	62.0 mg/l	62 mg/l
CN	0.74 mg/l	1.7 mg/l	0.13 mg/l
*Heavy Metal	0.6 mg/l	0.2 mg/l	0.2 mg/l
*Heavy Metal	0.39 mg/l	.05 mg/l	0.1 mg/l
Fe	5.8 mg/l	0.9 mg/l	2.9 mg/l
Phenol	.01 mg/l	.01 mg/l	.01 mg/l
Sulfide	0.23 mg/l	0.23 mg/l	0.28 mg/l

Sample 2

	Raw	Gravity Settled 2 hrs.	$FeCl_3$ Treated
BOD	2.0 mg/l	20.0 mg/l	7.0 mg/l
COD	237.0 mg/l	256.0 mg/l	169.0 mg/l
SS	60.0 mg/l	34.0 mg/l	13.0 mg/l
TOC	132.0 mg/l	109.0 mg/l	23.0 mg/l
NH_3	500.0 mg/l	500.0 mg/l	500.0 mg/l
CN	15.2 mg/l	10.5 mg/l	13.2 mg/l
*Metal	73.0 mg/l	1.9 mg/l	0.2 mg/l
*Metal	0.98 mg/l	0.34 mg/l	0.55 mg/l
Fe	0.25 mg/l	0.25 mg/l	0.25 mg/l
Phenol	7.25 mg/l	9.6 mg/l	7.0 mg/l

In addition to the jar tests conducted above, a sample was spiked with NH_3, sulfide, CN, heavy metals, and Fe. A portion of the spiked sample was allowed to settle for one hour untreated and the supernatant was submitted for analysis. A portion of the spiked sample was treated by the addition of 400 mg/l ferrous sulfate with the pH adjusted from 8.2 to 10.0 with 7 mg/l of a 10 N solution of sodium hydroxide. Dow A-23 was also added at 5 mg/l. The sample was allowed to settle for thirty minutes and the supernatant submitted for analysis. Another portion of the spiked sample was treated with the same chemicals, however, ten minutes was allowed to elapse between each chemical addition to simulate separate mixing tanks. The remaining portion of the spiked sample was treated by the addition of 400 mg/l ferric chloride with the pH adjusted from 8.1 to 10.0 with 7 mls/l of a 10 N solution of sodium hydroxide. The results of these tests are summarized in the following table.

	Raw	Spiked	Settled 1 Hour	$FeSO_4$ Single Addition	$FeSO_4$ Multiple Addition	$FeCl_3$
NH_3	47 mg/l	2,000 mg/l	1,600 mg/l	1,450 mg/l	1,450 mg/l	1,300 mg/l
Sulfide	3.4 mg/l	0.9 mg/l	0.9 mg/l	1.1 mg/l	1.1 mg/l	1.1 mg/l
CN	0.23 mg/l	30.23 mg/l	-	7.5 mg/l	7.2 mg/l	7.55 mg/l
Metal	0.35 mg/l	60.05 mg/l	12.1 mg/l	2.77 mg/l	1.56 mg/l	1.63 mg/l
Metal	1.77 mg/l	80.8 mg/l	41.7 mg/l	30.4 mg/l	26.0 mg/l	30.1 mg/l
Fe	3.9 mg/l	46.9 mg/l	4.38 mg/l	5.09 mg/l	4.12 mg/l	5.68 mg/l

Data from the preceding jar test indicate that either chemical treatment procedure would be effective in removing the metals. The quantities of chemicals used in the jar tests were not optimized; this would require more testing after plant start-up.

During an on-site visit to one of the plasticizer plants, wastewater samples were taken from the water phase of the plasticizer catch tanks and submitted for analysis. The results of these analyses are contained in the following table.

	Tank #1	Tank #2	Tank #3
BOD	16,000 mg/l	-	-
Soluble BOD (filtered)	13,500 mg/l	8,000 mg/l	8,000 mg/l
COD	25,800 mg/l	15,000 mg/l	13,400 mg/l
Soluble COD (filtered)	21,600 mg/l	13,600 mg/l	12,500 mg/l
Suspended Solids	578 mg/l	338 mg/l	414 mg/l
Total Solids	2,600 mg/l	1,800 mg/l	2,300 mg/l
pH	6.7	5.5	6.7

Gravity separation tests were conducted on the plasticizer catch tank wastewater. The results of a typical test where the water phase from the catch tank was settled for 2 additional hours are shown in the following table.

	20°F	80°F	155°F
COD	17,008 mg/l	17,401 mg/l	15,951 mg/l
BOD	11,500 mg/l	9,600 mg/l	9,900 mg/l
TOC	4,600 mg/l	4,260 mg/l	4,200 mg/l

Distillation/stripping screening tests were conducted on the catch tank wastewater, and also the use of vacuum was evaluated. This testing confirmed that adequate allowance for ambient gravity separation should be included in the design, and that further removal or recovery by enhanced physical separation techniques would not be fruitful.

Biological Treatability

Evaluation of the rotating biological contactor was the main thrust of the treatability studies. Two identical bench-scale RBC's were run in parallel to gain the most data within a minimum time period.

Each unit consisted of five stages. The first stage contained five circular styrofoam discs, 0.25 meters in diameter, with a total surface area of 0.51 square meters; the second stage contained four discs with a total surface area of 0.408 square meters; and the remaining three stages each had three discs for a surface area of 0.306 square meters in each stage. The unit was constructed using a 12-inch diameter PVC pipe, cut longitudinally in half, as the basin. PVC sheet was cut and formed to section off each individual stage. The discs were fastened to a wooden shaft. Each disc was spaced ½-inch apart on the shaft. The actual volume of water contained in the combined disc sections of the unit was 11.5 liters. Contact times through the 5-stages ranged from 5 to 11 hours for the different loadings. The flow pattern through the biocontactor was plug flow; determined by dye testing. The flow entered the head end of the shaft, and criscrossed through the various stages through alternately placed overflow weirs. The clarifier section of the unit was the last stage of the unit without discs. Sludge was syphoned from the bottom of the clarifier section each day. The drive unit was a variable speed motor and controller. Nutrients were added in the form of ammonium phosphate to the synthetic wastewater in a slight excess of what would be needed to support a biological culture. The tests were carried out over a nine-week period. The first week and a half was needed to start an active biological culture growing on the discs. During this period, the synthetic wastewater fed to the discs was a composition containing small amounts of the chemicals determined in the material balance. After the acclimation and start-up period, the biosystems were run at steady state using the "recipe", with samples being taken at each stage to gain the information required to predict the surface area necessary to meet the effluent limitations. Three runs were made at design concentration at three different hydraulic loadings. The performance data for these are summarized in Table B. The data were analyzed statistically to ensure steady state and normal distribution. An additional run was made at twice the design influent concentration at the same hydraulic loading as the intermediate loaded system of the first 3 runs. The data from that system are not summarized, but were used to test some of the design correlations that will be discussed and graphically presented.

The use by others of peripheral disc speed as a design parameter has led to higher RPM's and excessive aeration in smaller units which can lead to scale-up problems. That is, full-scale performance something less than bench or pilot-scale predictions. To avoid that problem, all our study work is performed by operating the units at the minimum rotation speed that will maintain a dissolved oxygen concentration of 1.0 mg/l in the liquid of the first stage. During the different runs the rotation of the shaft varied according to loading from 6 rpm to 10 rpm.

Settling and thickening tests using a one-liter graduated cylinder were conducted. The sludge that sloughed from the rotating biological contactor during the test period did so in very large gelatinous pieces; uncharacteristic of other industrial RBC sludges we have encountered. Much of it had to be physically removed from each stage because it would not pass through the small channels of the bench-scale biocontactor. In addition, a whole stage would slough all at once. No two stages sloughed simultaneously, however, and the units operated without any excessive loss of efficiency through a sloughed stage.

The settling and compaction time for the sludge was approximately one hour, and the sludge compacted to a concentration of approximately one percent. The effluent TSS concentration in the biocontactor averaged about 20 mg/l. A typical settling time is shown in Figure 9.

TABLE B

RBC #1 RUN #1
32.98 liters/total m^2/day

	$CODgm/m^2/day$	CODmg/1	% Reduction	$BODgm/m^2/day$	BODmg/1	% Reduction
Influent	-	1160	-	-	708	-
Stage 1	75.1	556	52.1	-	-	-
Stage 1,2	41.7	352	69.7	-	-	-
Stage 1,2,3	31.3	173	85.1	-	-	-
Stage 1,2,3,4	25.0	73	93.7	-	-	-
Stage 1,2,3,4,5	20.9	45	96.1	12.7	17	97.6

RBC #2 RUN #1
12.24 liters/total m^2/day

	$CODgm/m^2/day$	CODmg/1	% Reduction	$BODgm/m^2/day$	BODmg/1	% Reduction
Influent	-	1154	-	-	706	-
Stage 1	49.0	371	67.9	-	-	-
Stage 1,2	27.2	223	80.7	-	-	-
Stage 1,2,3	20.4	114	90.1	-	-	-
Stage 1,2,3,4	16.3	62	94.6	-	-	-
Stage 1,2,3,4,5	13.6	50	95.7	8.7	6	99.2

RBC #1 RUN #2
48.24 liters/total m^2/day

	$CODgm/m^2/day$	CODmg/1	% Reduction	$BODgm/m^2/day$	BODmg/1	% Reduction
Influent	-	1210	-	-	687	-
Stage 1	114.3	585	51.6	-	-	-
Stage 1,2	63.5	463	61.7	-	-	-
Stage 1,2,3	47.6	274	77.4	27.0	125	81.8
Stage 1,2,3,4	38.1	165	86.4	21.6	71	89.7
Stage 1,2,3,4,5	31.7	77	93.6	18.0	22	96.8

Sludge stabilization was not one of the items included for laboratory evaluation, however, brief screening tests were conducted on unstabilized sludge for dewatering applications. Centrifugation screening tests were conducted to determine the feasibility of the application of centrifugation for dewatering of sludges generated by the installation of a rotating biological disc. A gravity thickened sludge with a suspended solids concentration of 10,200 mg/l was screened. The sludge could only be concentrated by centrifugation to 30,600 mg/l which is a sludge containing 96.3 percent moisture. Gravity thickened sludge would not filter in a vacuum filtration screening test. Separate chemical additions of lime, alum, $FeCl_3$, nonionic, cationic, and anionic polymers did not improve either process. At this point the study ended, and a recommendation was made that due to the relatively low volume of sludge projected, the design of any dewatering equipment be done after the plant had been started-up, and the actual sludge generated could be evaluated after stabilization, using on-site pilot equipment.

RBC DESIGN

Several of the design bases that were available at the time of the study were evaluated, as well as a number of additional empirical approaches of our own, based on fitting a line(s) to the data (rather than vice-versa). The design chosen was coincidentally equivalent to one of the laboratory runs. However, the data from that run falls on a straight line that includes the data from the other loadings that were run. These several approaches and the recommended design parameters will be explained in this section and presented graphically.

For several reasons that will also be further delineated, COD was used as the design parameter.

Design Parameters

Of the parameters mentioned in the regulatory limits section, dissolved organics and the resultant oxygen demand parameters were those considered for the RBC design.

Of the others, total solids (essentially total dissolved inorganic species) would not be treated in a biological process and should future limits require treatment, other unit processes and/or source controls would have to be evaluated separately. Since the ultimate discharge of this effluent is to the sea, it is not expected that the projected dissolved solids concentrations would be in excess of any limits. (Also effluent suspended solids are anticipated to be less than 30 mg/l).

Projected cyanide and heavy metals concentrations are low enough not to effect the biological system, and would not be expected to be in excess of future limitations. However, should the material balance projections be inaccurate for these paramters, or if regulatory limits dictate, space has been left near the major process sources of these pollutants for future pretreatment to reduce these parameters before going to the RBC system.

As to the design parameters, the "organic" limitation is somewhat elusive, however, the following approach was applied:

In Column C of Table A showing the projected wastewater constitutents, a calculated BOD and COD are presented. The COD values were calculated using theoretical oxygen demand. For the major components (those accounting for 90% of the organic loading) the ratio of weight of calculated oxygen demand to weight of organic was never lower than 2.4. That is, one gram of "organic" exhibits a theoretical COD of 2.4 gms. From Column D of Table A the actual COD was slightly lower. Even reducing the ratio by that much still results in COD to organic ratio of 1.52. The ratio of "organic" to COD as derived from the regulatory limits is 1.48. This is all based on the influent to biotreatment, but it appears safe to assume that if the required COD reduction is obtained, the "organic" will be sufficiently removed. Also, although not discussed in detail here, should this parameter ultimately prove to be regulated using "TOC" or "oil and grease" measurements, the data also project adequate removal.

An examination of Table C and Figure 1, derived from that table shows that BOD removed at all loadings never falls below 80%. At all reasonable COD removals, BOD will be more than adequate to meet effluent targets as discussed in the next section. Because of this, and as COD is a more efficient operating analysis to obtain, the interstage data and the resultant correlations were derived using COD as the design parameter. About 40% of the COD analyses runs had accompanying BOD analyses. The samples were not filtered, but were very low in suspended solids. Some data were taken on filtered vs. unfiltered COD and the difference was not significant, and is not of concern for the RBC design.

TABLE C

% COD Removed	COD Applied $gms/m^2/day$	No. of Stages	Eff. mg/l COD	Eff. mg/l BOD	BOD:COD Ratio	% BOD Removed
96.1	20.9	5	45	17	.38	97.6
95.7	13.6	5	50	6	.12	99.2
*77.4	47.6	3	274	125	.46	81.8
*86.4	38.1	4	165	71	.43	89.7
93.6	31.7	5	77	22	.29	96.8
95.3	39.6	5	102	22	.22	98.3
92.1	47.5	4	171	71	.42	94.3
80.0	59.3	3	431	221	.52	82.4

Design Targets

A summary of the data used to determine the design effluent targets is shown in Table D.

TABLE D

Design Parameters

<u>Wastewater Characteristics</u>

Flow - Dry weather	317 m^3/day
- Wet weather (Design)	450 m^3/day
BOD_7	222 kg/day
COD	380 kg/day

<u>Influent Concentrations</u>

	Dry Weather	Wet Weather
BOD_7	845 mg/l	495 mg/l
COD	1200 mg/l	700 mg/l
BOD:COD ratio	0.58	

<u>Effluent Requirements</u>

BOD_7	18 metric tons/year (49 Kg/day-avg.)
COD	37 metric tons/year (101 Kg/day-avg.)

<u>Extrapolated Effluent Requirement Concentrations</u>

	Dry Weather	Wet Weather
BOD_7	155 mg/l	110 mg/l
COD	320 mg/l	225 mg/l
BOD:COD ratio	0.485	

Based on the projected numbers, an average of 73% COD removal and 78% BOD removal would appear to be adequate for meeting the regulations, and would keep capital expenditures as low as possible. However, there are several reasons, some rather obvious, why such a design basis would be improprietas.

The primary reason is that such a design is at the <u>limit</u> of the requirements. Also, the resultant loading would surely place the system in a log growth mode and any shift in efficiency relative to loading and variability would necessarily be drastic. Although the projected waste is quite biodegradable, such a loading would be in excess of prudent and accepted design loadings with no allowance for other adverse factors such as:

1. The raw waste load is a projection and some allowance might be made for the probability that the actual concentration and loading will not be exactly as predicted.

2. Although every effort was made to minimize conditions that might adversely effect scale-up, some allowance might be made for this aspect.

3. It might reasonably be assumed that the final regulatory permit would require daily averages and <u>maximums</u>.

As mentioned earlier, wastewater variability, chronic toxicity, and low loading conditions were not part of the laboratory study either, and any "allowances" for any of these or the factors listed above would be purely subjective. However, a COD removal of 85% and the resultant BOD_7 removal of 90% or greater was chosen as the lower limit of a reasonable operating range, and recommendations were based on the minimum requirements to operate at that average condition. Although the loading of BOD per area of RBC at that condition may still appear high to some, the extent of the "allowances" such as design will afford will be further detailed as the design is discussed.

Design Evaluation

As a point of reference in the following discussion and figures, "loadings" are expressed generally as gms COD/sq meter of surface area (eq. Kg/1000 m^2), and the hydraulics are expressed as cu meters/day. Among other evaluations the following linear plots were prepared using the study data:

	Applied COD	vs.	COD
Figure 2	wt/cumulative area		% removed
Figure 3	wt/cumulative area		Concentration Remaining
Figure 4	wt/cumulative area		Wt removed/area (of 1st stage)
Figure 5	wt/cumulative area		*Wt removed/cumulative area
Figure 6	wt/cumulative area		**Wt removed/cumulative area
Figure 7	wt/area for each stage		Wt removed/area for each stage

Further explanation of the above summary table is contained in the figures themselves, as well as the following text.

Figures 2 and 3 are variations of plots that were part of two design approaches that were generally considered viable at the time of design. The percent removal basis shows no real pattern. The concentrations plot is not usable either. If the effluent limitations were concentration based, it may have been of some use; however, the first stage data did not plot in the realm of the other stage data and a single line did not emerge, rather three separate lines. If substrate concentrations were dependent of loading, one would expect a single line to emerge. Also, since the effluent limitations were based on weight (mass), a "mass removal" approach might be better suited if the data fit.

*Points for each run drawn.

**Points for each stage of the different runs drawn.

In Figures 4,5,6, & 7, the mass applied is plotted against mass removed; however, they are normalized in several different manners. In Figure 4 the removal at each stage is shown as a function of the loading applied to the 1st stage (gms/cm^2/stage 1) as that load is spread out over a larger area (loading is expressed gms/cumulative cm^2). Again a family of curves developed with the first stage data are askew from the others. In looking at the percent removal at each stage it was noted that although percent removal varied, the points from the same stages (except stage 1) of the different systems when connected formed a line, and that all these when extrapolated, intersected at a single point (Figure 4). This exercise still did not provide a usable design approach. Figures 5 & 6 show the same plots except, that the loading and removal are normalized based on the cumulative area. Figure 6 was used for design. It shows that mass removal is a function of mass loading, as well as number of stages, and that the number of stages continues to diminish in importance down to a level, in this case of 20 COD gms/cm^2 applied, where the number of stages would no longer appear to affect mass removal at a steady state. It also shows that the enhanced effect of staging diminishes with increasing stages; with 4 stages being the limit beyond which little is gained by increased staging. This last piece of information, of course, is no discovery, but it does show an agreement of this approach with established fact.

This plot was used for design projections. Beyond a single stage system, removal can be projected for a design loading based on number of stages. This plot was tested by using the data from the very highly loaded fourth system to see if concentration difference may have an effect. (This fourth run was at double the concentration of the previous runs). In Figure 6A, the first and second stage data for that system does not fall on the curves, but actual performance is better than projected. Third stage data fall in line, but is again slightly better than projected. It cannot be determined with certainty whether these differences are due to higher concentration or to the overall higher loading, but in either case the design approach seems to be a safe one. In lieu of such study data, an attempt to look separately at removals at lower concentrations was made using Figure 7. Here, each stage is plotted separately; gms applied vs. gms removed, or in 7A, gms applied vs. gms remaining. There is some scatter particularly on the higher loaded runs. A least squares regression was made using the two lower loaded systems (see Figure 7B). This line had a correlation coefficient of 0.961 and intercepted the Y-axis at 4.4 (less than 4.4 gms/m^2 applied per stage removes zero gms COD). Using the slope and plotting 3 stages in the line shows increased system removal as the initial loading is decreased. Less mass is removed, but a greater percent of the applied load is. Low loading could not be evaluated with actual data.

Design

Catalytic's Environmental Systems Division advised Catalytic International, who was preparing the final design, of our determination of the minimum area and number of stages that should be included based on the scope of our study and the discussed evaluation.

Although Figure 6 indicates that a two-stage design could theoretically meet the effluent requirements, there were two primary reasons why at least 3 stages were recommended.

These studies were at steady state and as loading might shift up and down, considerable efficiency could be gained at higher loadings for the same area of contactor surface using a 3-stage vs. a 2-stage design. Three stages or greater would assure system operation even if one stage were malfunctioning. This was of specific concern because of the way the sludge sloughed in our study work. It appeared possible that a whole stage might be "denuded" at once and although we saw no great reduction in overall efficiency from this occurrence on the lab scale, it appeared prudent to allow for 2 fully covered stages to be in operation at any one time.

Looking at Figure 6 again, 85% COD removal for a three stage system plots out to be 30 gms/cm^2 applied. If the system were at a constant steady state, the average regulation limit would be just met at 40 gms/cm^2 applied or 32 gms/cm^2 removed. Restated, using minimum area recommended (for 85% COD removal) of 12,667 cm^2, 8 gms/cm^2 remaining x 12,667 cm^2 is 101 Kgs/day in the effluent. The <u>recommended</u> minimum design would leave 4.5 gms cm^2 remaining at steady state conditions.

It was difficult to obtain quantitative sludge production data in the confines of this study. Some of the difficulties with sloughing sludge were discussed earlier. However, a range of numbers were obtained from the study work. We used the conservative end of that range (the highest sludge yields). The number used was still within the realm of experience (0.33 gms of solids per gm BOD_7 removed).

The RBC study was conducted at temperatures ranging from 18°C to 28°C and no data for evaluation of cold temperatures operation was collected. Since the biological unit is preceded by considerable equalization and the units have a rather long contact time, it was recommended that the systems be covered and heated.

SYSTEM DESIGN

The final design as completed by Catalytic International in London is summarized in Figure 8, and includes surge capacity, oil removal, equalization, 2-stage neutralization, RBC, and final clarification.

Other recommendations supplied from the laboratory data were clarifier sizing, degree of pre-treatment required, and solids (bio-sludge) production.

The final design is a 3-stage system; two air-driven shafts with the second shaft baffled to provide the second and third stages. The system totals 20,000 sq. meters of effective area. The shafts are covered and the influent water is heated as it comes from the equalization basin by direct steam injection. The effluent then flows to a single circular clarifier which overflows to a basin with about 8-hours hydraulic retention time where it will be continually monitored for flow and quality. If necessary, it can be diverted from this basin back to the surge basin to go through the system again.

The final design also provides 15 day heated aerobic digestion of sludge followed by centrifugation for dewatering.

SUMMARY

The use of bench-scale RBC systems broadens the scope of applicability for this unit process. It becomes another option in areas where completely mixed systems could only be evaluated before, due to the logistic limitations of pilot scale studies. Bench-scale units can establish feasibility and provide design data, allowing technical and economic comparisons to other bio-processes and a rational design when it is the unit process of choice.

The study herein discussed illustrates that there is a reasonable approach to defining and testing wastewaters and process plant effluents when they do not actually exist on an entity. This approach allows a treatment plant design that can better meet the needs of the grassroots process plant once it is constructed and operating.

FIGURE 1

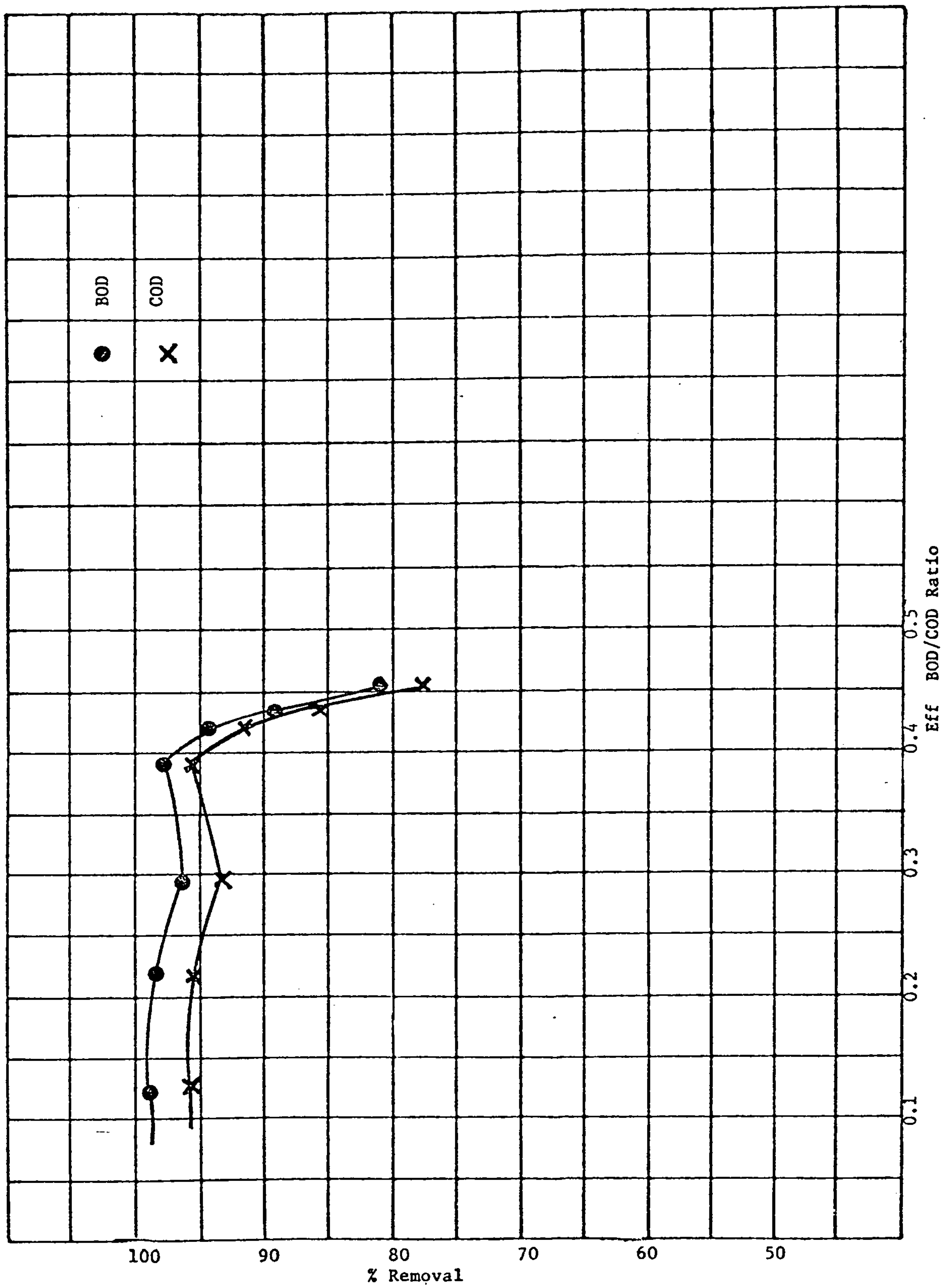
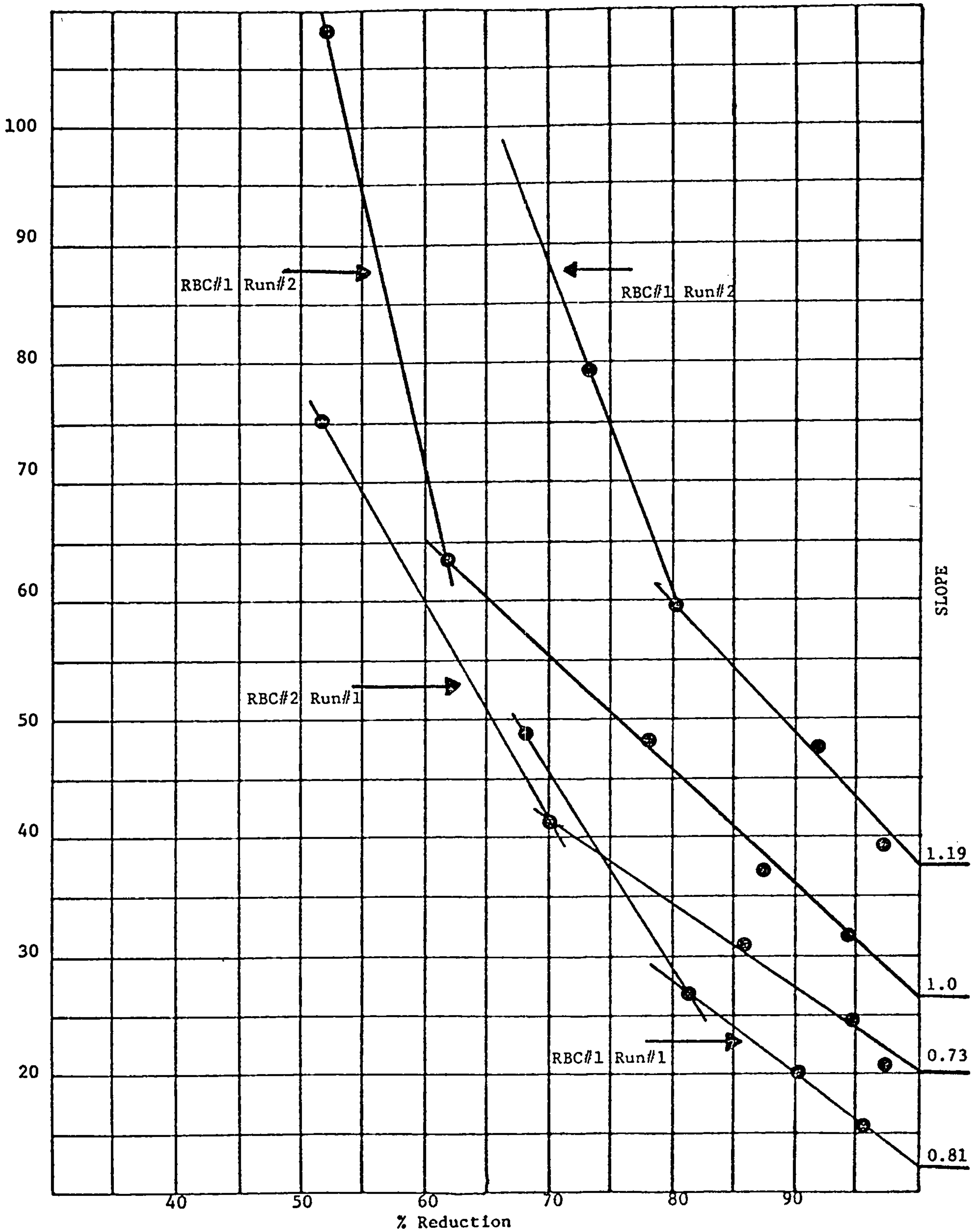

BOD
COD
0.1
0.2
0.3
0.4
0.5
Eff BOD/COD Ratio
100
90
80
70
60
50
% Removal

FIGURE 2



100
90
80
70
60
50
40
30
20
RBC#1 Run#2
RBC#1 Run#2
RBC#2 Run#1
RBC#1 Run#1
SLOPE
1.19
1.0
0.73
0.81
40
50
60
70
80
90
% Reduction

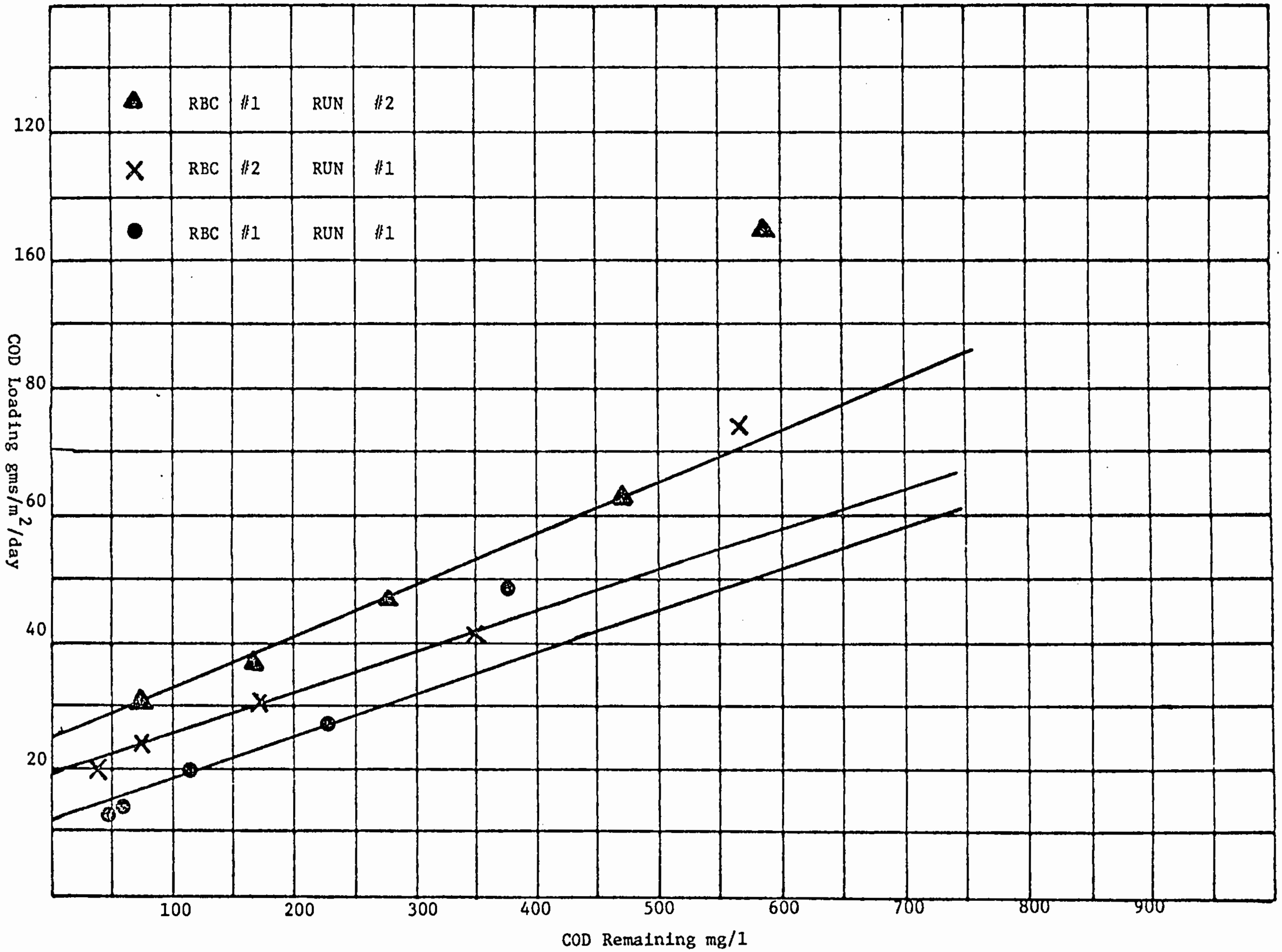
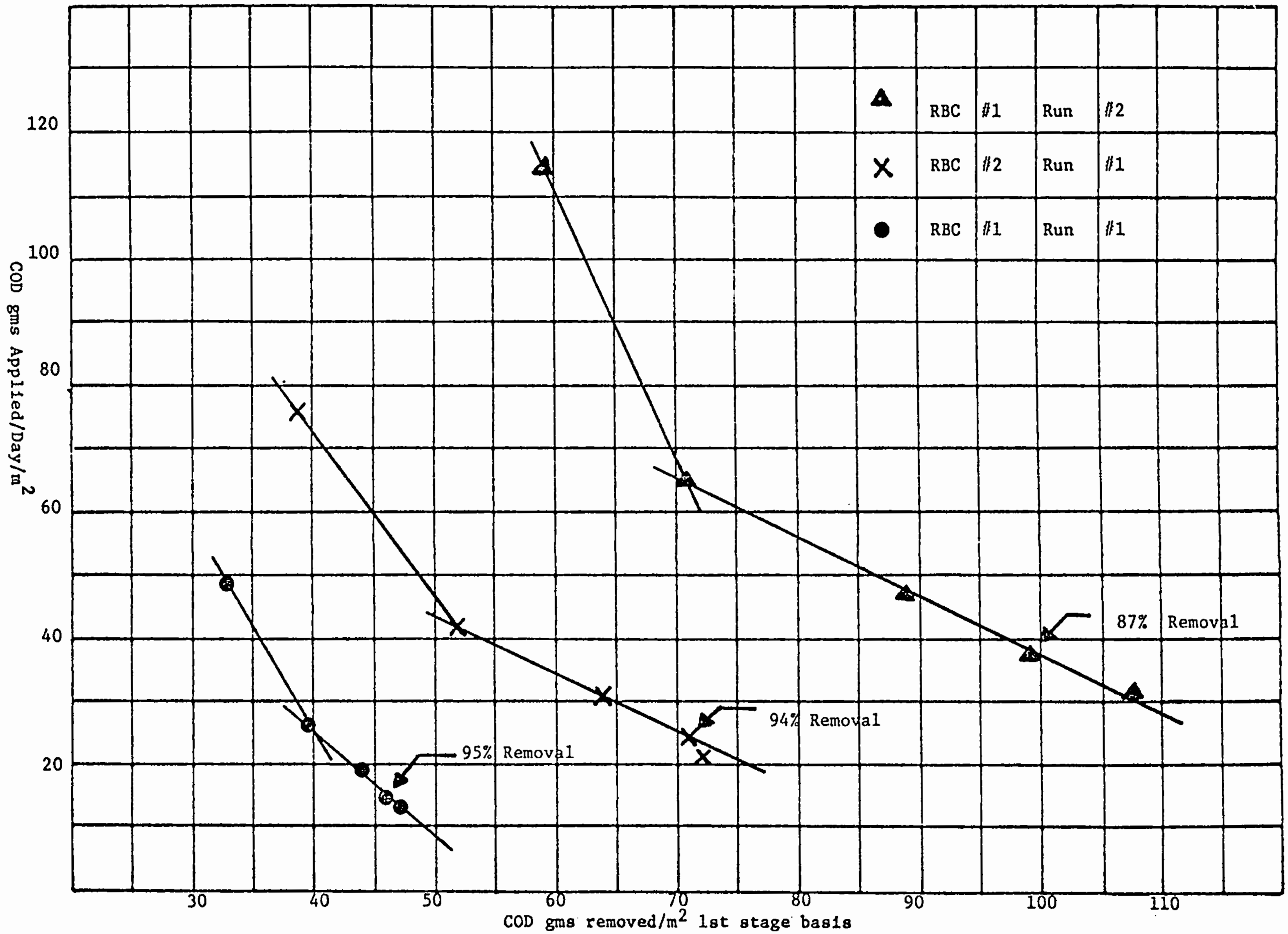

RBC #1 RUN #2
RBC #2 RUN #1
RBC #1 RUN #1
120
160
80
60
40
20
COD Loading gms/m2/day
100
200
300
400
500
600
700
800
900
COD Remaining mg/l

FIGURE 3



RBC #1 Run #2
RBC #2 Run #1
RBC #1 Run #1
87% Removal
94% Removal
95% Removal
120
100
80
60
40
20
30
40
50
60
70
80
90
100
110
COD gms Applied/Day/m2
COD gms removed/m2 1st stage basis

FIGURE 4

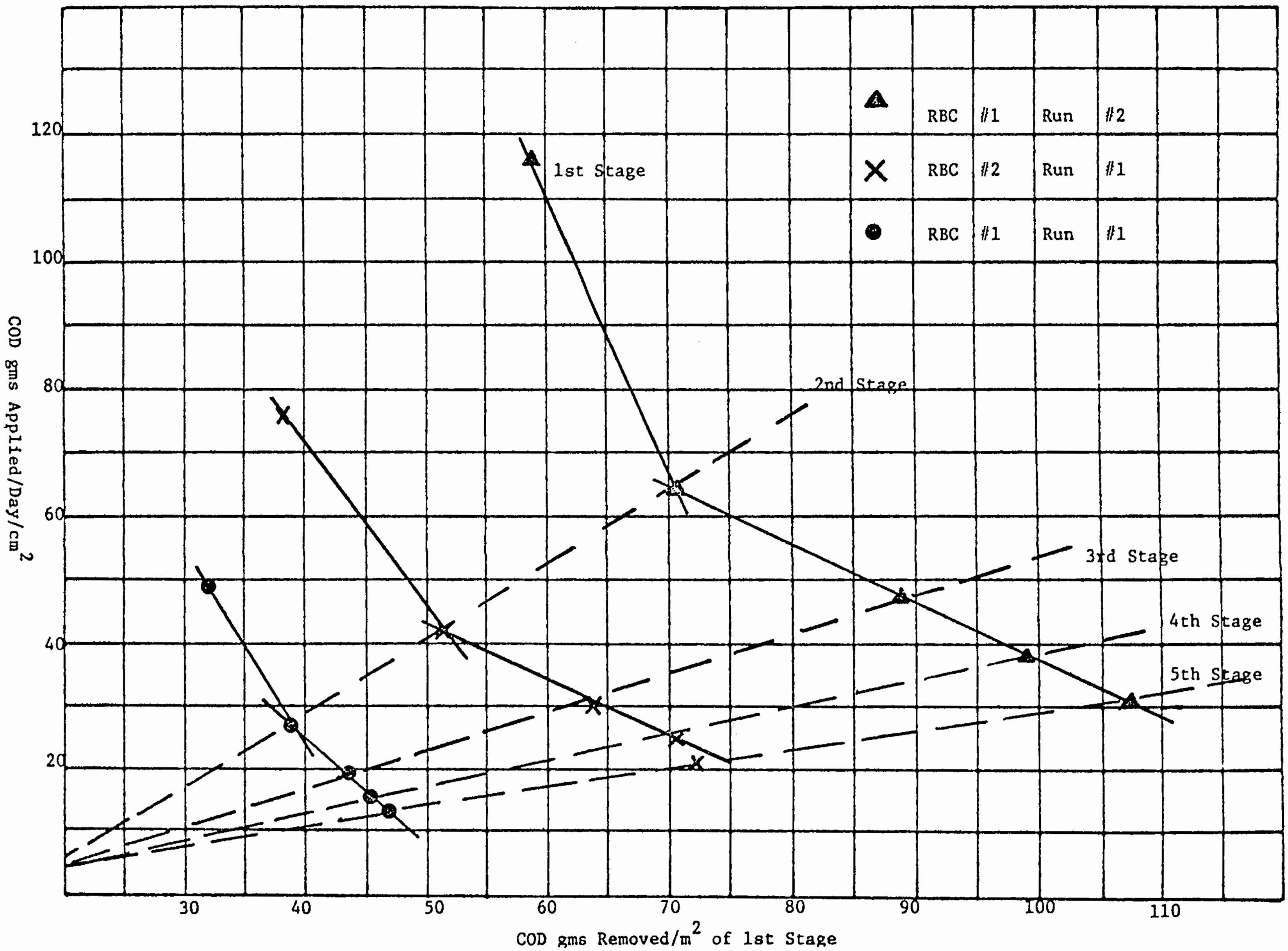
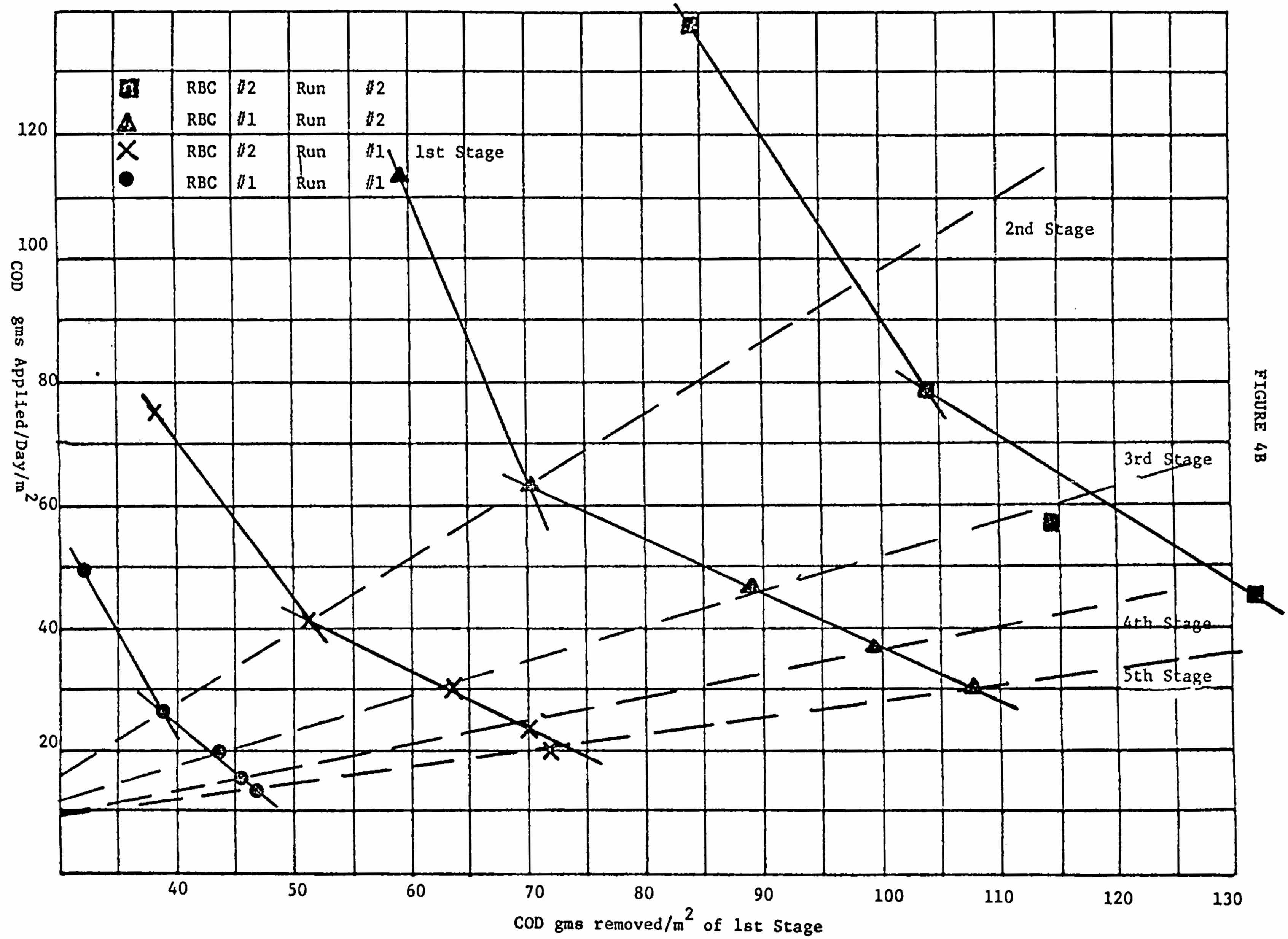

1st Stage
2nd Stage
3rd Stage
4th Stage
5th Stage
RBC #1 Run #2
RBC #2 Run #1
RBC #1 Run #1
COD gms Applied/Day/cm²
COD gms Removed/m² of 1st Stage
120
100
80
60
40
20
30
40
50
60
70
80
90
100
110

FIGURE 4A



RBC #2 Run #2
RBC #1 Run #2
RBC #2 Run #1 1st Stage
RBC #1 Run #1
2nd Stage
3rd Stage
4th Stage
5th Stage
120
100
80
60
40
20
COD gms Applied/Day/m²
40
50
60
70
80
90
100
110
120
130
COD gms removed/m² of 1st Stage

FIGURE 4B

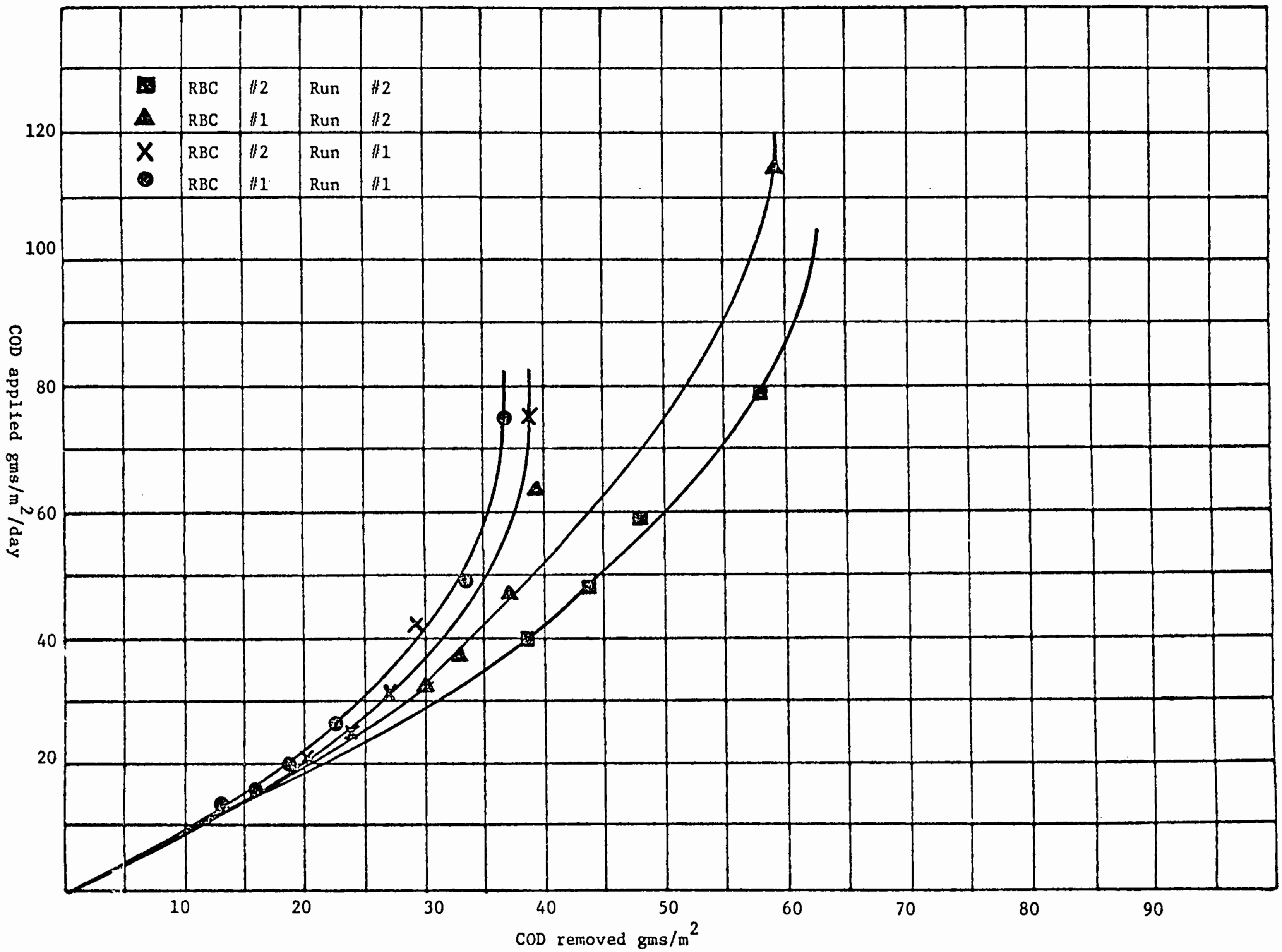

RBC #2 Run #2
RBC #1 Run #2
RBC #2 Run #1
RBC #1 Run #1
120
100
80
60
40
20
10
20
30
40
50
60
70
80
90
COD applied gms/m^2/day
COD removed gms/m^2

FIGURE 5

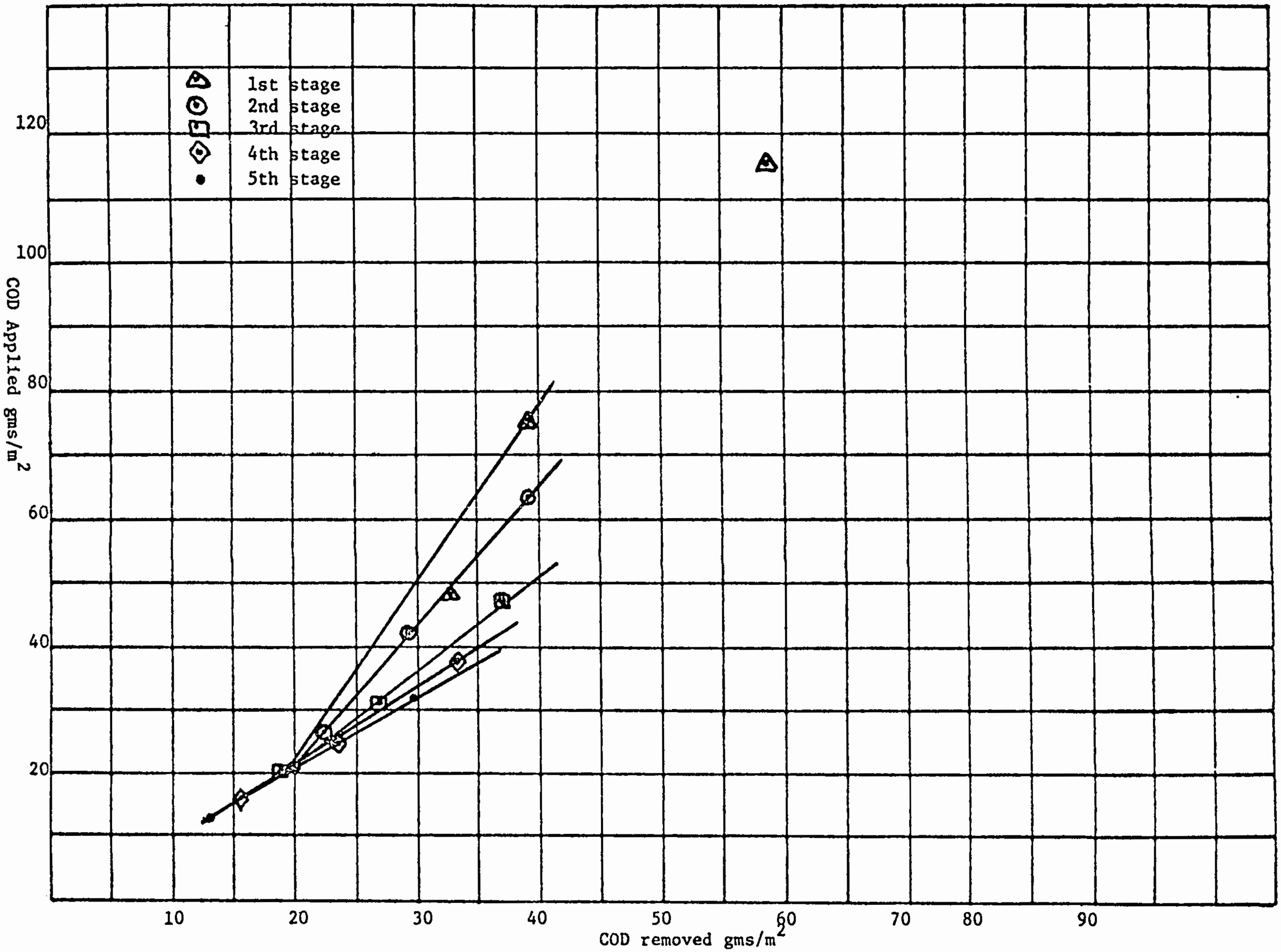
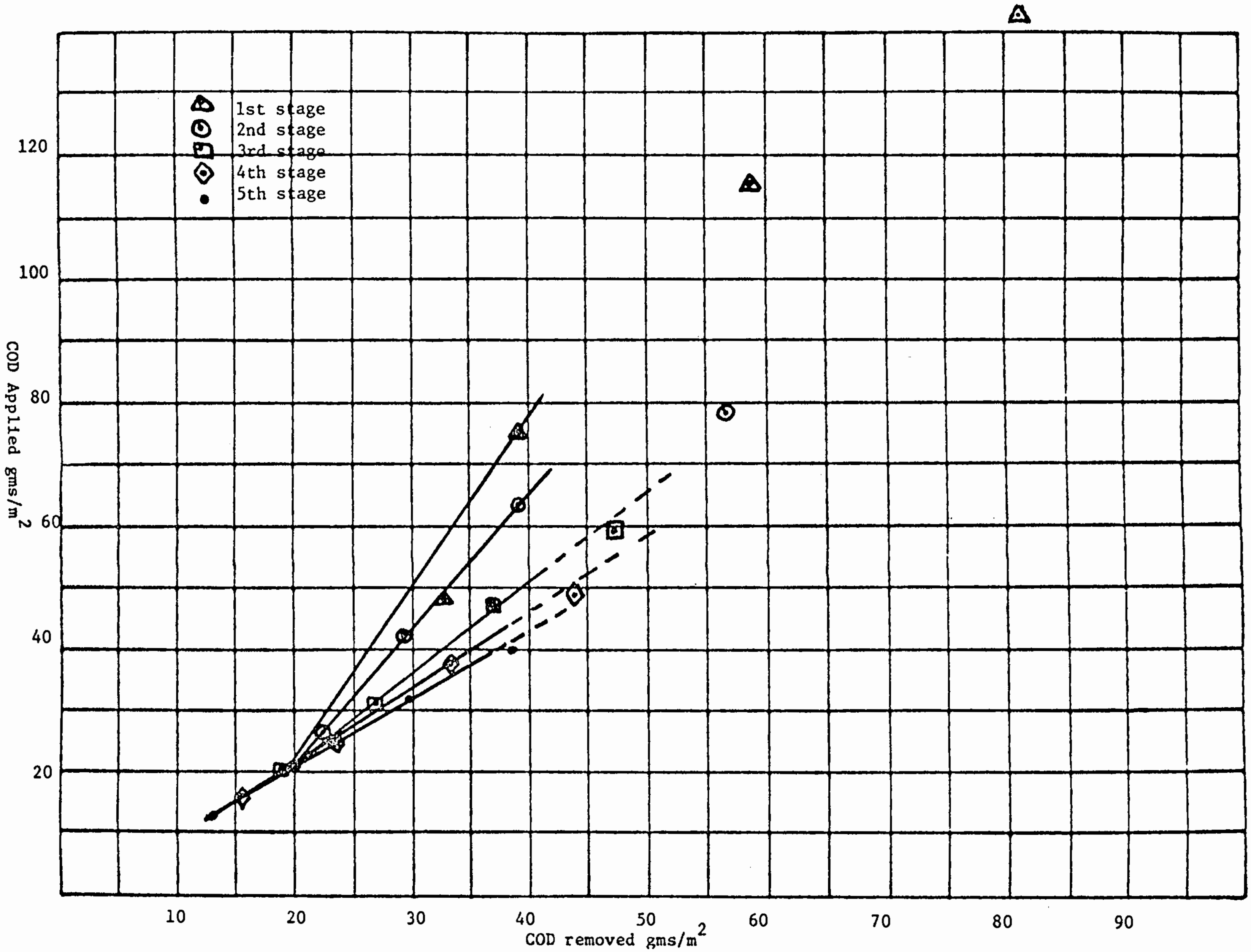
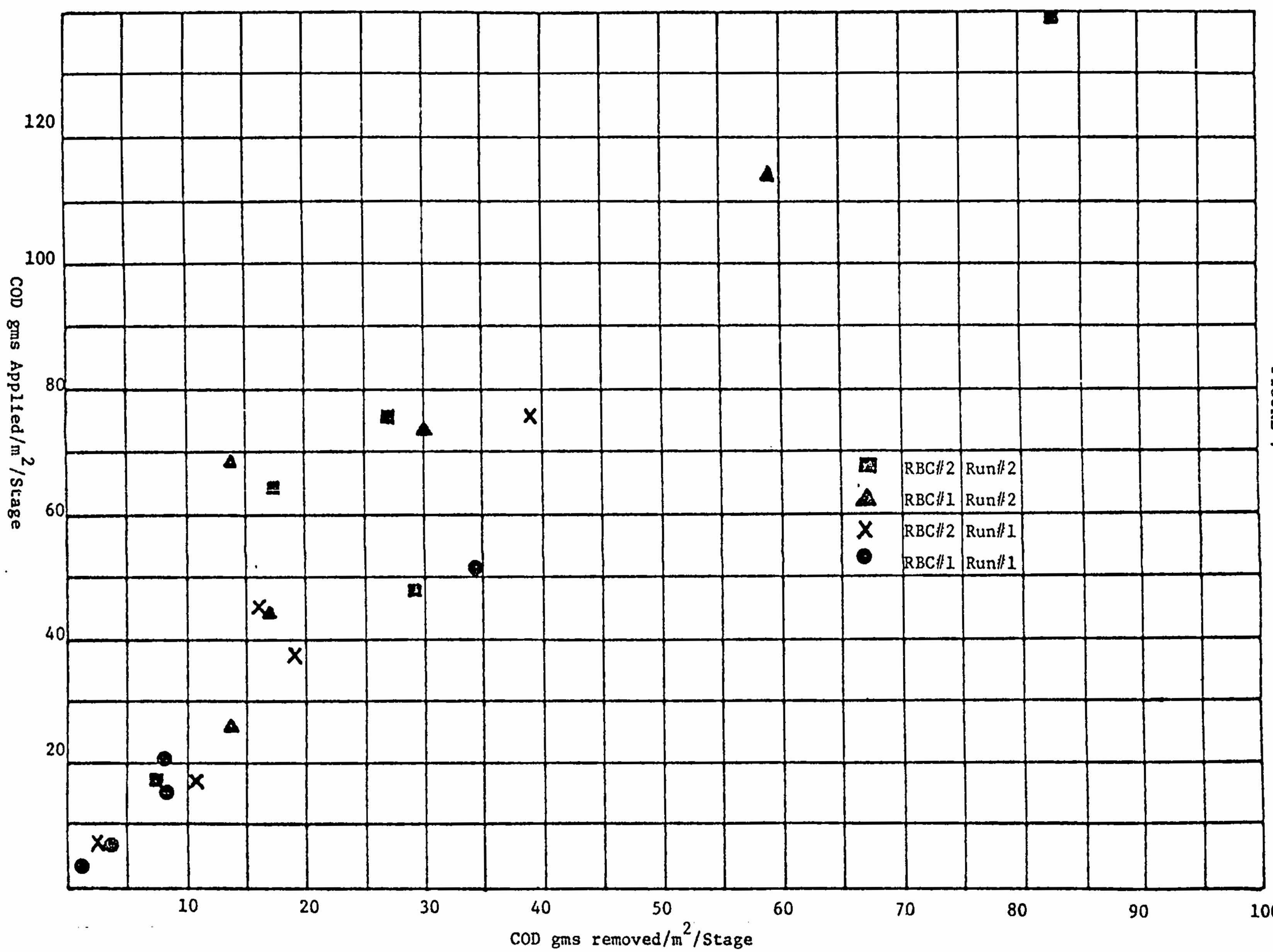

1st stage
2nd stage
3rd stage
4th stage
5th stage
COD Applied gms/m^2
COD removed gms/m^2
120
100
80
60
40
20
10
20
30
40
50
60
70
80
90

FIGURE 6



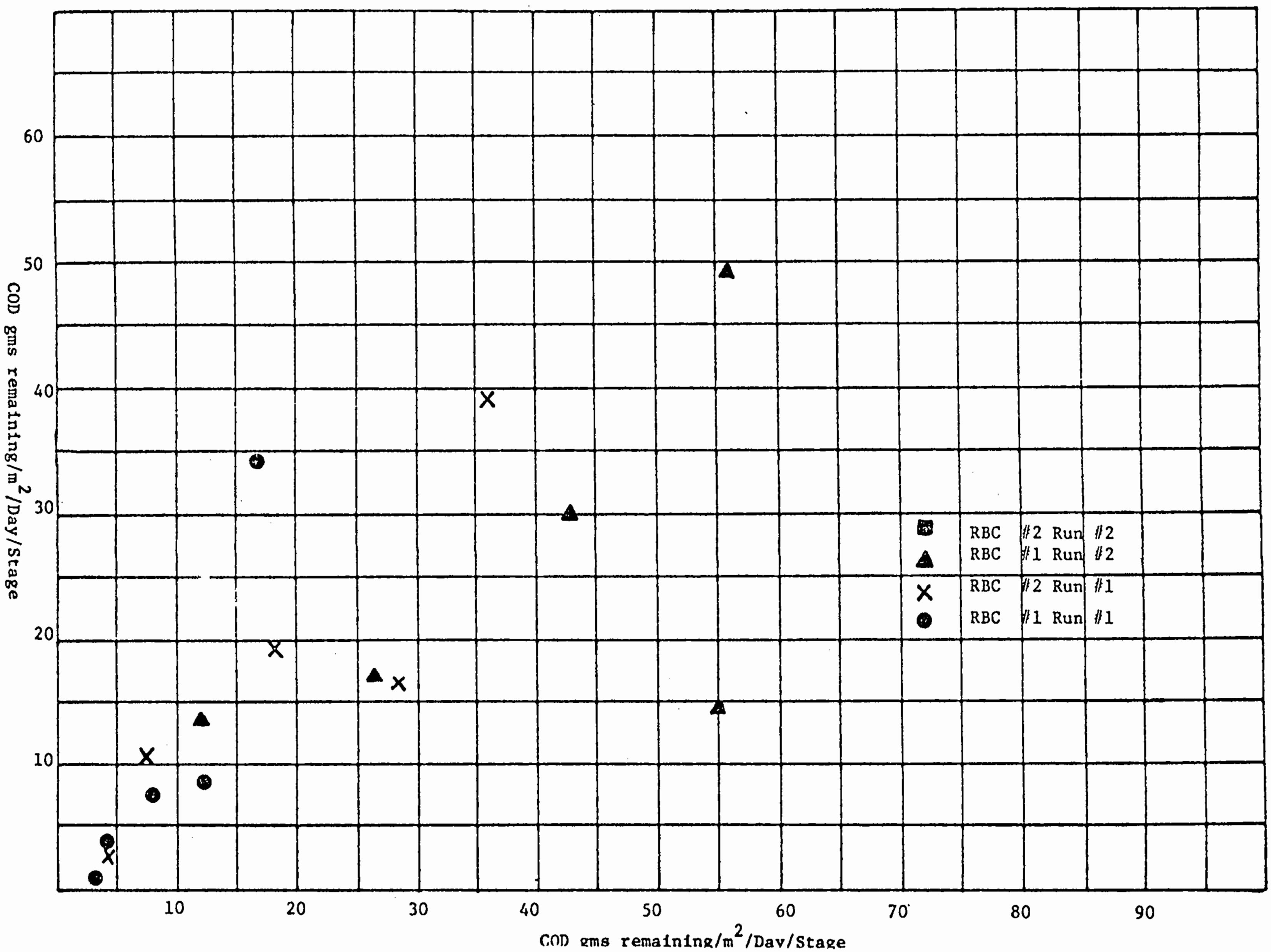
1st stage
2nd stage
3rd stage
4th stage
5th stage
120
100
80
60
40
20
COD Applied gms/m2
10
20
30
40
50
60
70
80
90
COD removed gms/m2

FIGURE 6A



COD gms Applied/m2/Stage
120
100
80
60
40
20
10
20
30
40
50
60
70
80
90
100
COD gms removed/m2/Stage
RBC#2 Run#2
RBC#1 Run#2
RBC#2 Run#1
RBC#1 Run#1

FIGURE 7



60
50
40
30
20
10
COD gms remaining/m2/Day/Stage
10
20
30
40
50
60
70
80
90
COD gms remaining/m2/Day/Stage
RBC #2 Run #2
RBC #1 Run #2
RBC #2 Run #1
RBC #1 Run #1

FIGURE 7A

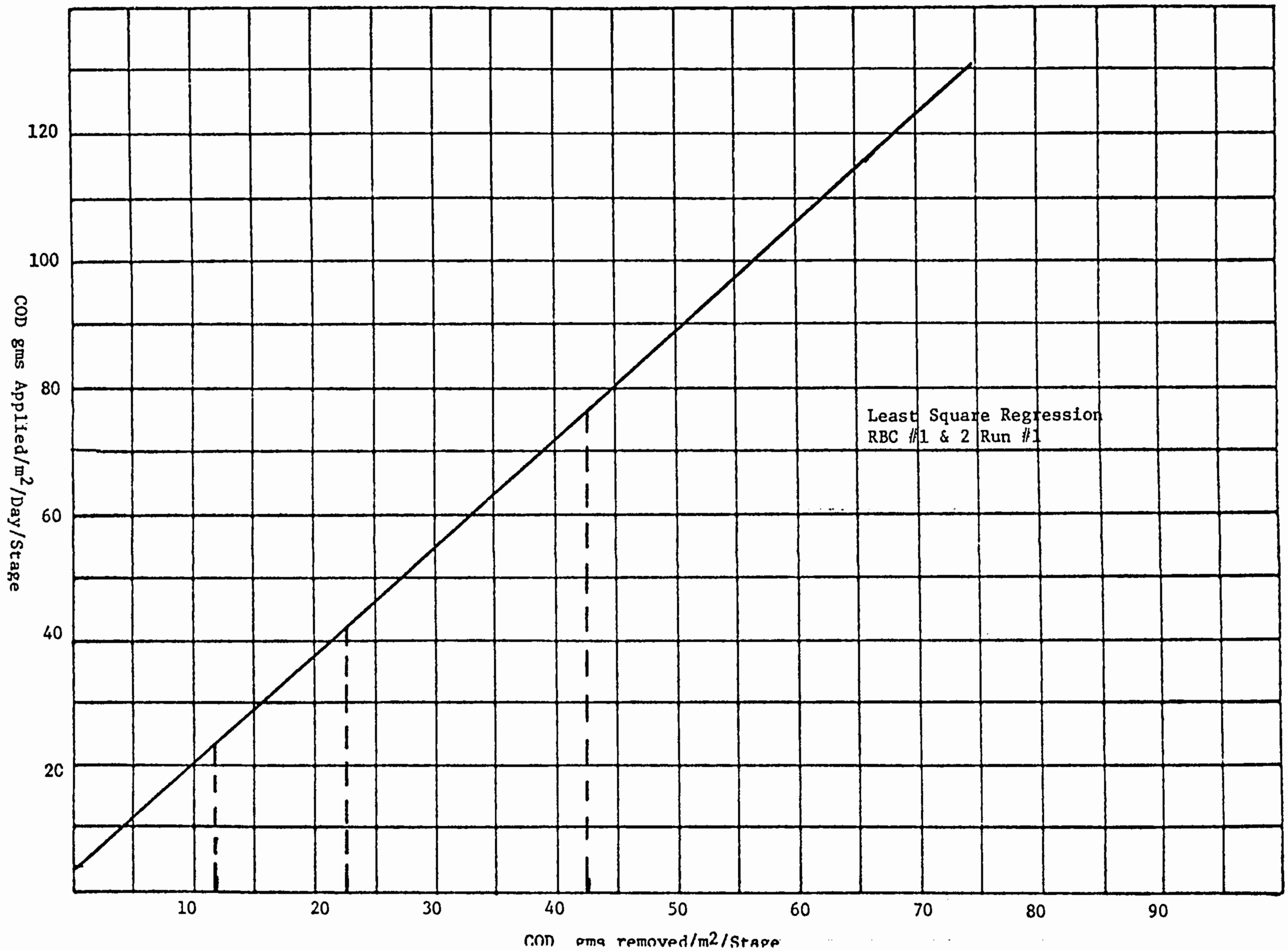

Least Square Regression
RBC #1 & 2 Run #1
COD gms Applied/m2/Day/Stage
COD gms removed/m2/Stage
120
100
80
60
40
20
10
20
30
40
50
60
70
80
90

FIGURE 7B

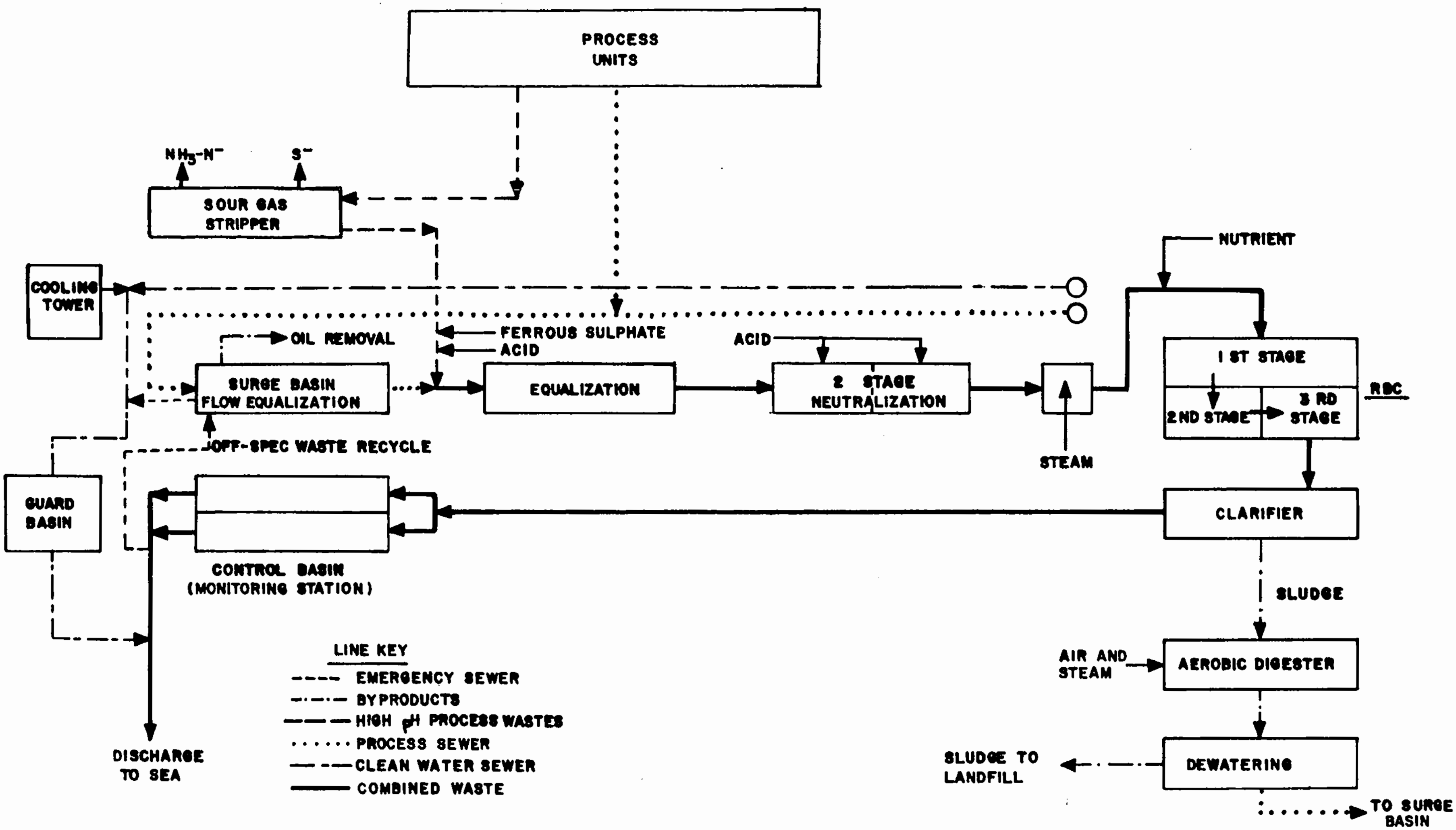
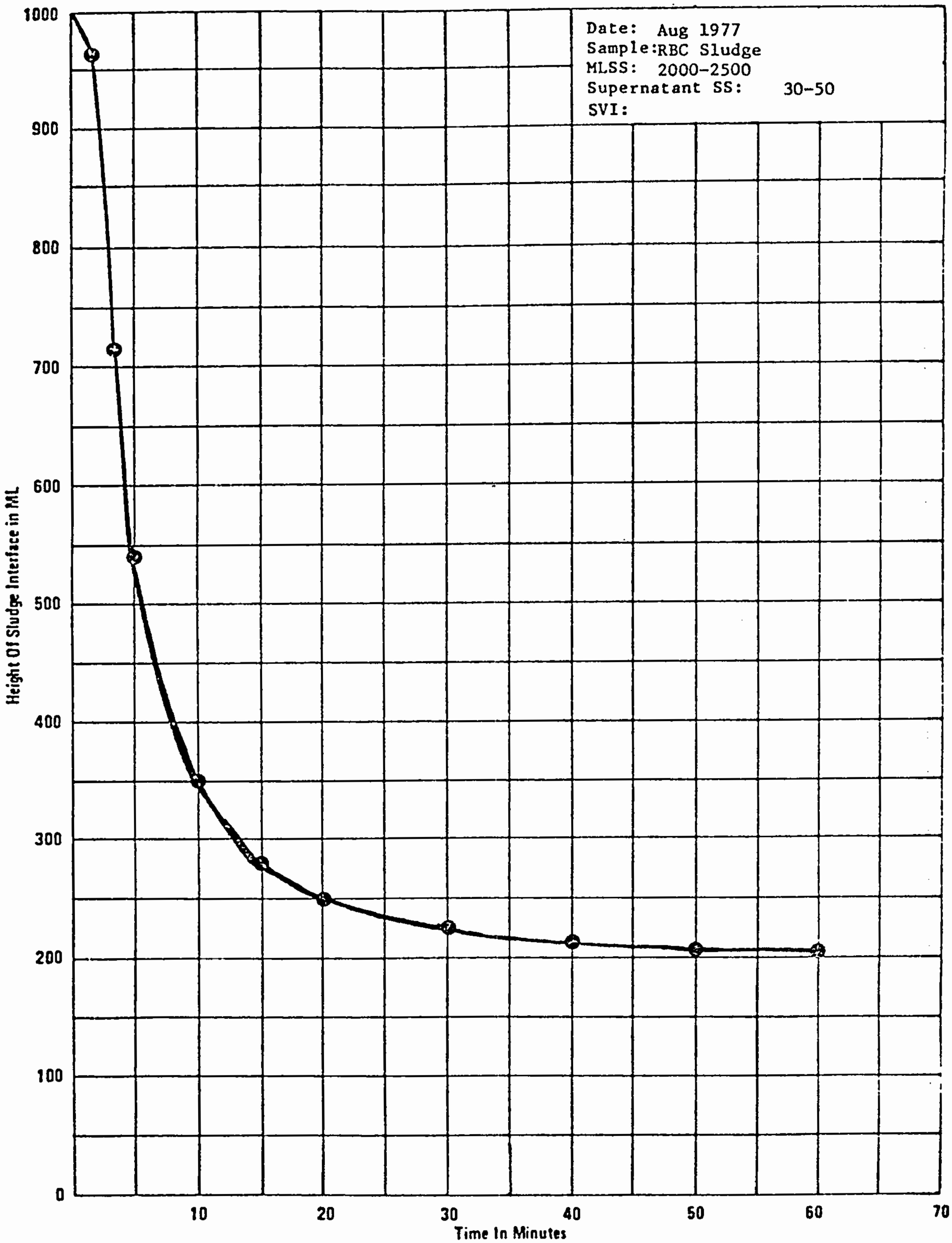

PROCESS UNITS
NH_3-N⁻
S⁼
SOUR GAS STRIPPER
COOLING TOWER
OIL REMOVAL
FERROUS SULPHATE
ACID
ACID
NUTRIENT
SURGE BASIN FLOW EQUALIZATION
EQUALIZATION
2 STAGE NEUTRALIZATION
1 ST STAGE
2 ND STAGE
3 RD STAGE
RBC
STEAM
OFF-SPEC WASTE RECYCLE
GUARD BASIN
CLARIFIER
CONTROL BASIN (MONITORING STATION)
SLUDGE
AIR AND STEAM
AEROBIC DIGESTER
DEWATERING
SLUDGE TO LANDFILL
TO SURGE BASIN
DISCHARGE TO SEA
LINE KEY
EMERGENCY SEWER
BYPRODUCTS
HIGH pH PROCESS WASTES
PROCESS SEWER
CLEAN WATER SEWER
COMBINED WASTE

FIGURE 8

FIGURE 9
SETTLING STUDY



THE TREATMENT OF SALINE WASTEWATERS USING A ROTATING BIOLOGICAL CONTACTOR

By

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INTRODUCTION

Recently aquaculture has become increasingly useful for the production of food for world-wide consumption. Aquaculture is the raising and harvesting of aquatic organisms in a controlled environment. As the use of aquaculture increases, the need for the development of treatment methods for the wastewaters generated becomes more apparent. The method of treatment to be evaluated by this study is rotating biological contactors (RBC).

The type of aquaculture of interest in this study is the raising and harvesting of *Macrobrachium Rosenbergii* (prawns). Prawn larvae are cultured in a 30 percent sea water solution. As the prawns mature they are able to survive in decreasing saltwater concentrations until, at the adult stage, the prawn are able to survive in freshwater. In a recirculating aquaculture facility, certain aspects of water quality must be kept within prescribed limits in order to provide optimum conditions for prawn growth and development. The most significant parameters include dissolved oxygen (DO), temperature, pH, unionized ammonia, and organic substances. The rotating biological contactor will be evaluated for its ability to maintain these parameters at levels allowing maximum prawn develpoment.

The rotating biological contactor is an aerobic treatment process that consists of a series of circular discs connected to a common horizontal shaft. The discs rotate partially submerged within the wastewater. A biological film (biofilm) is allowed to form on the discs as they rotate through the wastewater contained in the contactor basin. While submerged in the wastewater, the biofilm adsorbs organic matter and as the discs rotate, the biofilm containing the adsorbed organic matter is exposed to the atmosphere. The microorganisms, comprising the biofilm, use the oxygen from the atmosphere and the organic matter from the wastewater in a biodegradation-biotransformation process. Waste organics are entirely or partially reduced to their basic components (including ammonia, carbon dioxide, and water), and the microorganisms suspended in the water are subsequently removed from the system by sedimentation.

Rotating biological contactors are becoming increasingly popular in the field of wastewater treatment for a number of reasons which include: (1) lightweight and compactness, (2) low power consumption, (3) combination of functions as a trickling filter and an activated sludge process, (4) high efficiency in oxygen transfer, and (5) ability to achieve nitrification easily.[1] Success has been found in using the contactors to treat freshwater, domestic wastewaters. This study, however, has been designed to evaluate the treatment efficiency of rotating biological contactors in a closed marine system.

Previous studies have been concerned with the treatment of saline wastewaters, and one study performed at the University of Rhode Island dealt with the use of rotating biological contactors to treat saline domestic wastewaters. This study will be somewhat different than others previously performed in that nitrification of saline wastewaters will be evaluated. Ammonia is toxic to aquatic organisms at extremely low concentrations. These toxic concentrations are dependent upon pH, temperature, and ionic strength and will be determined for each of the sampling programs in order to evaluate the amount of nitrification necessary at the different salinity levels.

In summary, it has been hypothesized that rotating biological contactors can be used effectively to treat wastewaters such as those generated in an aquaculture facility. This pilot study will concentrate on the technical feasibility to achieve nitrification in a closed saline system using rotating biological contactors.

LITERATURE REVIEW

Biological Treatment Cost Comparison

As previously stated, one of the reasons for the increased popularity of rotating biological contactors is the low power consumption when compared to other biological treatment methods. Poon et al., based on work from a pilot study, estimates rotating biological contactor power requirements to be 45 percent lower than an activated sludge unit of equivalent capacity (0.8 MGD).[2] In a study of winery wastes, La Bella et al., found the capital costs of a RBC and activated sludge to be equal. Labella, however, found the yearly operational costs to be approximately $6,000 less for the rotating biological contactor than the activated sludge for flows between 0.34 and 0.44 MGD.[3]

System Loadings

Rotating biological contactors have been sized using a number of methods including hydraulic loading, detention time, and total organic loading. The method utilized in this study was total organic loading.

The organic loading is the total organic mass applied to the system over a period of time and is determined by multiplying the hydraulic loading rate and the influent organic concentration, yielding a unit of mass per time per area. Cook and Kincannon, in evaluating trickling filters as a fixed-film biological treatment process, found the total mass of BOD and/or COD applied to the system to be of importance when designing a process.[4] The total mass applied takes into account both the flow rate and the organic concentration of wastewater. The process performance evaluated as COD removal efficiency was dependent upon the total COD applied to the system as gram/hr/m^2, rather than its concentration or flow rate.[5] The COD removal efficiency remained constant at constant total loadings regardless of whether the load was caused by high organic concentrations and low flow rates or low organic concentrations and high flow rates.

Poon and Mikucki, while testing rotating biological contactors for the treatment of saline wastewaters, agreed with Cook and Kincannon in concluding that although the hydraulic loading is important in the rotating biological contactor process design, organic loading should be considered equally as important. High hydraulic loadings applied to low BOD influents will yield the same removal efficiency as low hydraulic loadings applied to high BOD influents.[6] The work performed at Colorado State University was based on the total organic load in order to monitor the effect of both the hydraulic load and organic concentrations on the rotating biological contactor.

NITRIFICATION

As previously mentioned, the main objective of this stidy was to obtain an optimum loading rate for nitrification of a saline wastewater. Nitrification is the biological transformation of ammonia to nitrate and nitrite. Ammonia, even at low concentrations, can have acute toxicity effects on aquatic organisms and deplete the dissolved oxygen concentration while being converted to nitrates in receiving waters. The toxicity of aqueous solutions of ammonia can be attributed to the NH_3 species. The toxicity of ammonia is dependent upon the pH and concentration of total ammonia ($NH_4^+ + NH_3$).[7] There are other factors which also affect the concentration of the NH_3 in solution; the most important of which are temperature and ionic strength. The Red Book states that the concentration of the NH_3 increases with increasing temperature and decreases with increasing ionic strength.[8]

In a review of the EPA Red Book, the American Fisheries Society states that the NH_4^+/NH_3 ratio is a function of the activity of the changed speices and the total ionic strength of the solution. Thruston et al., state that there is a slight decrease in the unionized fraction of the total ammonia as the ionic strength increases in a dilute saline solution (less than 40 percent sea water).[9]

The American Fisheries Society also states that a decrease in the dissolved oxygen concentration will increase the toxicity of ammonia. It is hypothesized that a reduction in the dissolved oxygen concentration would be accompanied by an increased ventilation rate by organisms, increasing the exposure to unionized ammonia.[10]

Nitrogen in the form of ammonia is converted to nitrate in two steps by autotrophic nitrifying bacteria: Nitrosomonas and Nitrobacter. The reactions as presented by Metcalf and Eddy, Inc. are as follows:

$$NH_4^+ + 1.5\ O_2 \xrightarrow{\text{Nitrasomonas}} NO_2^- + 2H^+ + H_2O$$

$$NO_2^- + 0.5\ O_2 \xrightarrow{\text{Nitrobacter}} NO_3^-$$

These reactions can be combined to read:

$$NH_4^+ + 2O_2 \longrightarrow NO_3 + 2H^+ + H_2O$$

The nitrifying organisms needed to convert ammonia to nitrate are present in almost all aerobic biological treatment processes. In many instances, however, the numbers are limited. Nitrification is brought about or encouraged by suitable adjustment of the operating parameters, namely the reduction of total applied loads.[11] Antonie et al., state that nitrification in a rotating biological contactor begins when the BOD approaches 30 mg/ℓ. At this concentration, nitrifying organisms are able to compete with the more rapidly growing carbon oxidizing organisms and establish themselves in the process. With the establishment of nitrifying organisms, nitrification is allowed to proceed rapidly until at a BOD concentration of approximately 10 mg/ℓ the nitrification is complete.[12]

Weng and Molof found using a pilot scale RBC and an ariticial substrate, that the chemical oxygen found (COD) must be below 50 mg/ℓ for nitrification to occur.[13] Using the artificial substrate, this 50 mg/ℓ COD corresponded to a BOD concentration of approximately 14 mg/ℓ. This is less than half the concentration stated by Antonie for nitrification (approximately 30 mg/ℓ). Weng and Molof's results show that increasing the disc surface area increased the rate of nitrification. In the tests conducted, Weng and Molof found that nitrification took place only in the stages where the mixed liquor dissolved oxygen was greater than 2 mg/ℓ. From a review of the literature, it may be stated that nitrification can be accomplished using a rotating biological contactor operated under the proper conditions.

EFFECTS OF CHLORIDE CONCENTRATION

Researchers have found that wastewaters with high chloride concentrations may be treated satisfactorily.[14,15,16] The chloride concentration does, however, play a major role in the treatment of wastes.

There is a marked difference between biodegradation-biotransformation rates of the suspended fraction in saltwater and freshwater environments.[17] It is believed that, for a microbial population such as those present in freshwater, domestic wastes, certain exoenzyme action may be inhibited in an environment with high chloride concentrations. This exoenzyme action is generally considered necessary for the metabolism of insoluble substrates, and its inhibition may result from a change in the surface configuration of the suspended material or in the configuration of the enzymes themselves.[18]

With regard to nitrification, Ludzack and Noran state that the nitrification during high-chloride operation was approximately 10 percent of that expected for the same operation at lower chloride concentrations.[19] This lower rate of nitrification may bring about the need for longer hydraulic detention times in order to achieve the desired levels of ammonia removal.

Treatment of Saline Wastewaters with RBC

Previous work has been performed on the use of rotating biological contactors to treat saline wastewaters.[20] This work, performed at the University of Rhode Island, focused on the highly saline, domestic wastewaters generated on the island of Kwajalein in the South Pacific. Poon and Mikucki performed the plot studies to obtain a secondary treatment method to meet the provisions of the 1972 amendment of the Water Pollution Control Act (PL 92-500), along with regulations implemented by the US Environmental Protection Agency.[21]

The major concern of the pilot plant study was to demonstrate that rotating biological contactors could meet the standards set by the national pollutant discharge elimination system (NPDES) while treating a highly saline wastewater. In order to do this, Poon and Mikucki conducted a series of experiments with and without sea water. Initial tests were conducted without the addition of sea water and high chloride concentrations in order to demonstrate that the pilot plant could successfully treat a wastewater of Kwajlein's strength to meet NPDES standards. Poon and Mikucki then evaluated the performance of the rotating biological contactors using various hydraulic and organic loadings at different chloride concentrations.

The two major findings of Poon and Mikucki were that: (1) the change in chloride concentrations had no effect on the RBC treatment efficiency after the microorganisms were allowed to acclimate and (2) the performance of the RBC is dependent upon the combination of the hydraulic and organic loadings. This combination of loading results in a loading term of mass per area per time.[22]

From the literature review, it can be seen that there is some controversy within the field as to the effects of various parameters on rotating biological contactors. These differences do not mean that RBC's cannot be properly managed; rather, that continued research is necessary in certain aspects of design and operation.

This literature review has shown that although the objectives of this study have been met in other studies, they have not been incorporated into one. Studies have shown that the rotating biological contactor can achieve nitrification and can successfully treat saline wastes. This study has been designed to determine the ability of a rotating biological contactor to achieve nitrification in saline wastewaters, such as those generated by a closed aquaculture system.

METHODOLOGY

The section on methodology is concerned with the approaches used in the pilot study in order to obtain values. The methodology of the study is important since it will have direct effects on interpretation of the data collected.

The section has been divided into four segments. The segments describe the pilot rotating biological contactor, the entire pilot system, the tests performed on the system, and the quality control program followed during the secondary program. The segments have been arranged in order to describe the entire system from its initial design to tests performed to analyze the systems treatment efficiency.

Description of the Rotating Biological Contactor

The rotating biological contactor used for the study was a relatively small four-stage unit having a capacity of 0.92 ft^3 (0.026 m^3) with discs in place. Each stage contained five discs having a total surface area of 26.40 ft^2 (2.45 m^2), and a specific surface area of 28.7 ft^2/ft^3 (93.47 m^2/m^3).

The rotating biological contactor was constructed at the Engineering Research Center of Colorado State University. The tank was fabricated from a section of 12 inch (30.5 m) PVC pipe. The tank ends, staging baffles, and discs were all fabricated from pelxiglass. The discs have been placed perpindicular to the flow with 36 percent of the discs submerged. The discs were rotated clockwise using a 1/4 horsepower Dayton gear reduced motor. The discs rotational speed was 14 revolutions per minute yielding a peripheral velocity of 40.8 ft/min (12.4 m/min).

Description of the RBC System

The RBC was fed from a 1.01 ft^3 (0.03 m^3) holding basin which was kept at a constant head using an overflow wier and two 4.0 ft^3 (0.11 m^3) storage tanks. This system was incorporated into the study so that the RBC could be fed using gravity rather than a pump which could possibly fail after extended use.

The two storage tanks were used to hold the saline synthetic wastes to be treated by the system. The synthetic sea water was produced using Instant Ocean Sea Salts obtained from Aquarium Systems of Eastlake, Ohio. The system was fed a balanced minimal media with a carbon source, sucrose, serving as the growth limiting nutrient. The composition of the synthetic feedstock is shown in Table 1. Compressed air was introduced into the two storage tanks to insure adequate mixing.

Table 1. Composition of Synthetic Feedstock

Constituent	Concentration
SUCROSE	100 mg/ℓ
$(NH_4)_2\ SO_4$	25 mg/ℓ
$MgSO_4 \cdot 7H_2O$	10 mg/ℓ
K_2HPO_4	6 mg/ℓ
$Mn\ SO_4 \cdot H_2O$	1 mg/ℓ
$CaCl_2$	0.76 mg/ℓ
$FeCl_2 \cdot 6H_2O$	0.05 mg/ℓ

After leaving the rotating biological contactor, the effluent flowed to the drain of the Research Center. No recirculation was performed during testing. The main advantages of recirculation are the delay of plant expansion along with the ability to supress shock loads, neither of which fall into the scope of this study. With regard to nitrification, Lue-Hing et al., found that recirculation of clarified secondary effluent did not significantly increase nitrification.[23] For these reasons, recirculation was not included in the study.

Tests Performed on the RBC System

A series of tests were performed on the rotating biological contactor system in both fresh and saline waters. The testing was more extensive on the saline wastewaters. The major tests conducted were as follows:

Ammonia
Chemical Oxygen Demand
Specific Conductivity
Dissolved Oxygen
Nitrate
Nitrite
pH
Salinity
Temperature

All tests, with the exception of ammonia, were performed in accordance with Standard Methods for the Examination of Water and Wastewater, 14th Edition.[24]

Ammonia concentrations were determined using two methods: (1) Orion specific ion electrode model 95-10 for ammonia was used on the freshwater samples, and (2) the Nesslerization Method 418B of Standard Methods was used on the saline samples. The ammonia probe was not used in the saline samples because membrane fouling caused unstable results. These tests were conducted in order to monitor the performance of the rotating biological contactor under various ammonia loadings in both freshwater and 10 percent sea water environments.

LABORATORY QUALITY CONTROL

A quality assurance program should be an important component of any water quality study involving chemical or biological analysis. In order to analyze the quality of data being collected, replicate and spiked samples were run where applicable on a minimum of 15 percent of all samples.

Although this program of laboratory quality control does not entail any quality control charts, the program was able to act as a satisfactory warning system when any problems developed in sampling or analysis. The significant problems encountered during the sampling and analysis segments of the study will be discussed in the results.

The sampling program was similar for both the freshwater and 10 percent sea water wastes. The two major exceptions being the method of determining ammonia and the number of samples taken. More tests were performed on the 10 percent sea water wastes than were performed on freshwater. The freshwater data was obtained to verify that the contactor was able to achieve adequate nitrification treating freshwater. This could be done without extensive sampling. The saline wastewaters, being the primary interest of the study, have been monitored more closely in order to observe the various stages of treatment. A discussion of the information obtained in both fresh and saline wastewaters will follow in the section on results.

RESULTS

As previously discussed, the sampling program was divided into two main segments: freshwater and saltwater. The freshwater segment was undertaken in order to show that the rotating biological contactor could successfully nitrify a freshwater waste, and to achieve loading rate-removal efficiency relationships to compare to those of saline wastewaters. The 10 percent sea water saline segment was undertaken to determine the loading rate-removal efficiency relationships for a wastewater similar to those generated in a *Macrobrachium Rosenbergii* (prawn) rearing aquaculture facility.

Freshwater Results

The rotating biological contactor was first acclimated under freshwater conditions using the synthetic feed material. This testing was conducted to monitor the contactors' ability to nitrify wastewater under "nonsaline" conditions. The results obtained during the freshwater testing may be seen in Table 2.

Table 2. Freshwater Results

Influent Ammonia (mg/ℓ)	Effluent Ammonia (mg/ℓ)	Hydraulic Loading (liter/100 m^2/day)	Ammonia Loading (g/100 m^2/day)	Maximum Ammonia Removal (%)
5.77	2.29	6,100	30	60
7.28	2.36	6,100	40	67
7.68	3.68	6,100	50	52
9.70	3.10	6,100	60	68
12.13	4.39	6,100	70	64
12.46	4,53	6,100	80	64
13,42	7.16	6,075	90	47
14.15	7.71	6,075	100	46
16.29	8.35	6,075	110	48

The removal efficiency of the contactor was found to be dependent upon the applied ammonia load. A plot has been prepared of ammonia load versus maximum ammonia removal. This plot may be seen as Figure 1. Under the freshwater conditions, the optimal ammonia loading rate (from Figure 1) was found to be approximately 60 grams/100 m^2/day. The 60 gram/100 m^2/day loading rate resulted in a maximum ammonia removal efficiency of 68 percent, yielding an effluent total ammonia ($NH_4^+ + NH_3$) concentration of 3.1 mg/ℓ - NH_3 (2.6 mg/ℓ - N). The pH of the system's effluent was 7.1 at a temperature of 18°C. This pH and temperature, in combination with the total ammonia concentration of 3.1 mg/ℓ yields an unionized ammonia concentration of 0.013 mg/ℓ - NH_3. This unionized value has been calculated using a percent unionized ammonia of 0.430 based on work performed by Thruston et al., at Montana State University.[26] As previously stated, this data demonstrates the contactors ability to successfully nitrify a "nonsaline" wastewater.

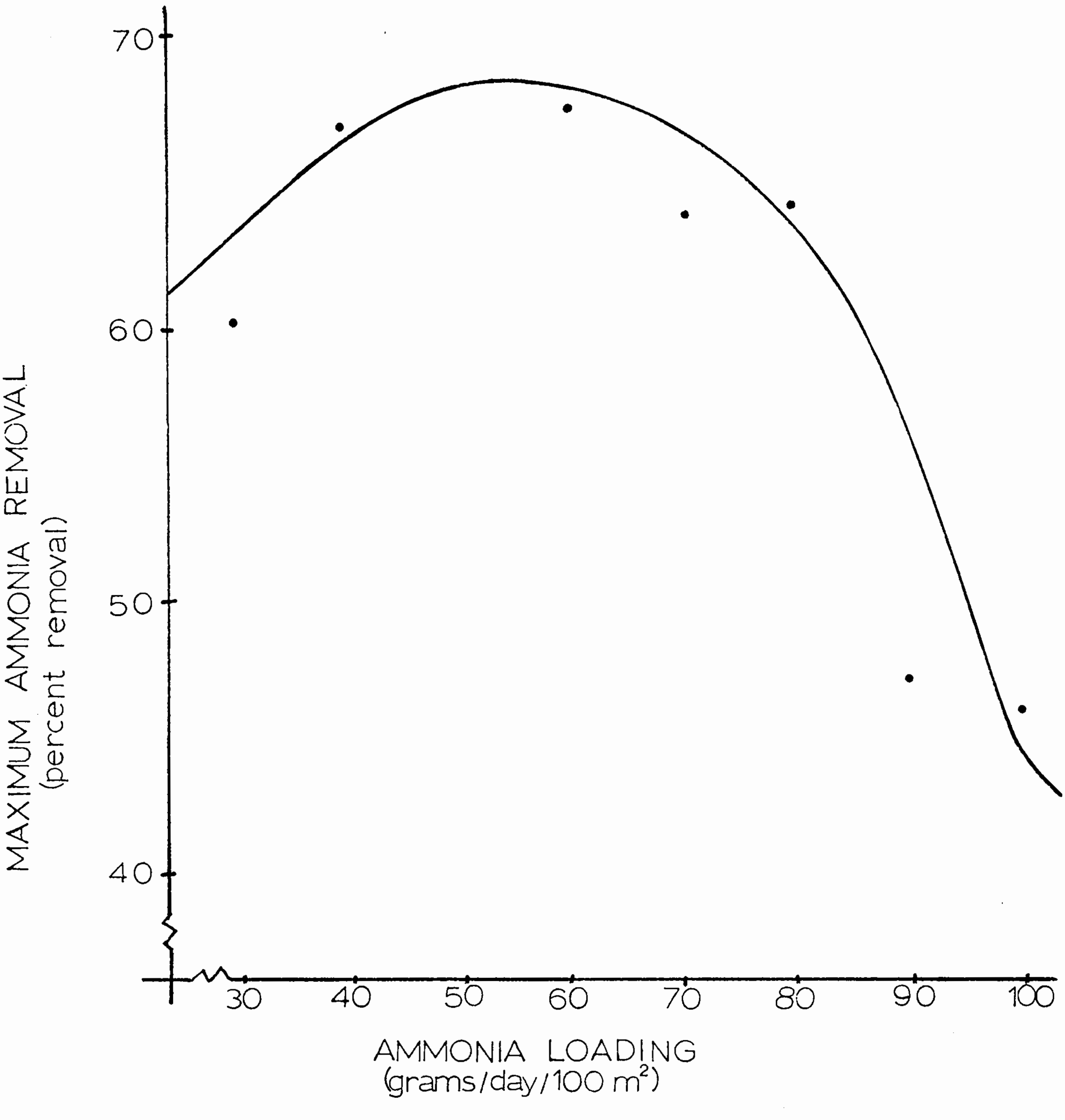

FIGURE 1. AMMONIA LOAD VS MAXIMUM REMOVAL

In order to more precisely determine the optimum ammonia loading, the model used by Weng and Molof in their nitrification studies will be used.[27] The model presented by Weng and Molof is as follows:

$$\text{Log } F = \text{Log } k + \text{Log } L_o + b \text{ Log } Q + c \text{ Log } S + d \text{ Log } t + e \text{ Log } A + f \text{ Log } d + g \text{ Log } T$$

Where:

F = fraction of influent loading remaining in the effluent
k = the intercept value
L_o = influent loading
Q = flow rate
S = rotational disc speed
t = detention time of the liquid in the BFFRD system
A = effective disc surface area
D = submerged disc depth
T = liquid temperature
a, b, c, d, e, f, g, = the partial regression coefficients

Since all of the variables, with the exception of influent loading, remained constant (assuming the variation of flow rate to be negligible), the following constant may be determined:

$$\text{Log } k_2 = b \text{ Log } Q + c \text{ Log } S + d \text{ Log } t + e \text{ Log } A + F \text{ Log } D + g \text{ Log } T$$

Combining the two equations yields:

$$\text{Log } F = \text{Log } k + \text{Log } k_2 + a \text{ Log } L$$

If:

$$\text{Log } k + \text{Log } k_2 = \text{Log } k',$$

then

$$\text{Log } F = \text{Log } k' + a \text{ Log } L_o$$

In order to use the equation with the data collected, Table 2 has been prepared.

Table 2. Freshwater Ammonia Results

Ammonia Loading L_o (grams/100 m^2/day)	Log L_o	% Removal	F	Log F
30	1.48	60	.40	-0.40
40	1.60	67	.33	-0.48
50	1.70	52	.48	-0.32
60	1.78	68	.32	-0.49
70	1.84	64	.36	-0.44
80	1.90	64	.36	-0.44
90	1.95	47	.53	-0.28
100	2.00	46	.54	-0.27
110	2.04	48	.52	-0.28

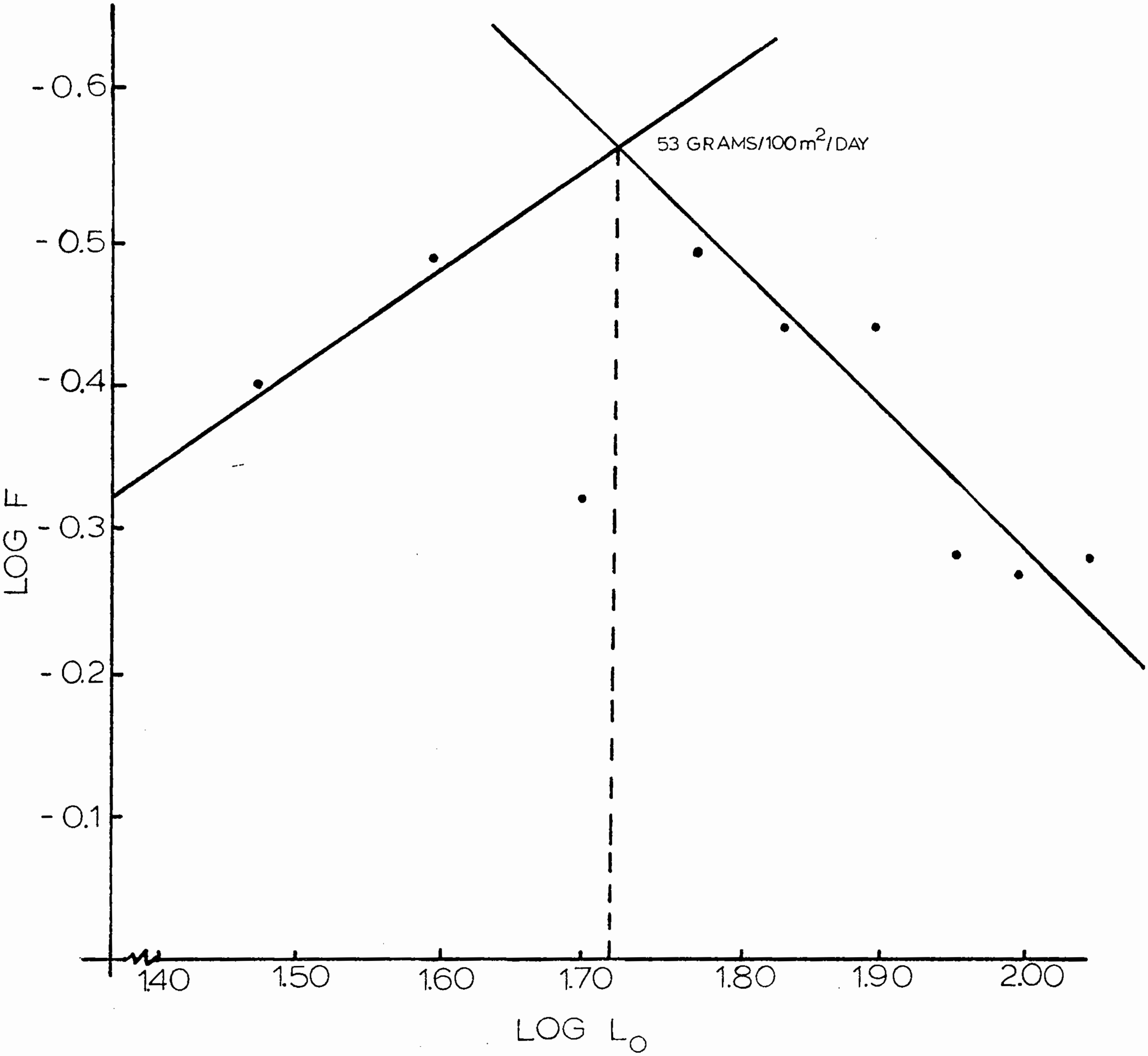

FIGURE 2. LOG F VS LOG L_o IN FRESHWATER

A plot of Log F vs Log L_o may be seen as Figure 2. The plot shows two distinct straight lines, the intersection of which represents the optimum loading rate. In this case, the optimal ammonia loading rate for freshwater was found to be 54 grams/100 m^2/day which yielded an ammonia removal efficiency of 73 percent. This determination of the optimal ammonia loading represents a more precise method than plotting ammonia loading versus ammonia removal to determine the optimal loading.

Ten Percent Sea Water Results

After completion of freshwater data collection, the RBC system was allowed to acclimate to a 10 percent sea water salt concentration. As anticipated, the system's nitrification capacity was altered by the addition of saltwater.

At an ammonia loading rate of 60 grams/100 m^2/day (optimum for the freshwater system from visual inspection), the maximum ammonia removal was 56 percent in the 10 percent sea water waste. This removal efficiency is 13 percent less than that of freshwater at the same loading rate.

Using the model described by Weng and Molof, the optimal ammonia loading for the 10 percent sea water was found to be 60 grams/100 m^2/day. This loading rate resulted in a maximum ammonia removal of 56 percent. The data used to obtain these values may be seen in Tables 3 and 4 and has been presented graphically in Figure 3.

Table 3. Saltwater Results

Influent Ammonia (mg/ℓ)	Effluent Ammonia (mg/ℓ)	Hydraulic Loading (liter/100 m^2/day)	Ammonia Loading (g/100 m^2/day)	Maximum Ammonia Removal (%)
4.5	3.1	5,470	25	34
7.5	4.6	5,470	40	41
9.5	4.3	5,470	50	50
10.4	4.6	5,470	60	56
14.8	8.4	5,470	80	43
16.2	9.2	5,470	90	40

Comparing these results with those from freshwater testing, it can be seen that although the optimal ammonia loading remained essentially the same the contactor's ability to nitrify decreased in the 10 percent sea water. This decrease in nitrification in the 10 percent sea water coincides with the conclusions of Ludzack and Noran that the nitrification will decrease with increasing chloride concentrations.[28]

The ammonia loading rate was the only parameter that varied throughout the monitoring program. The artificial substrate described in methodology

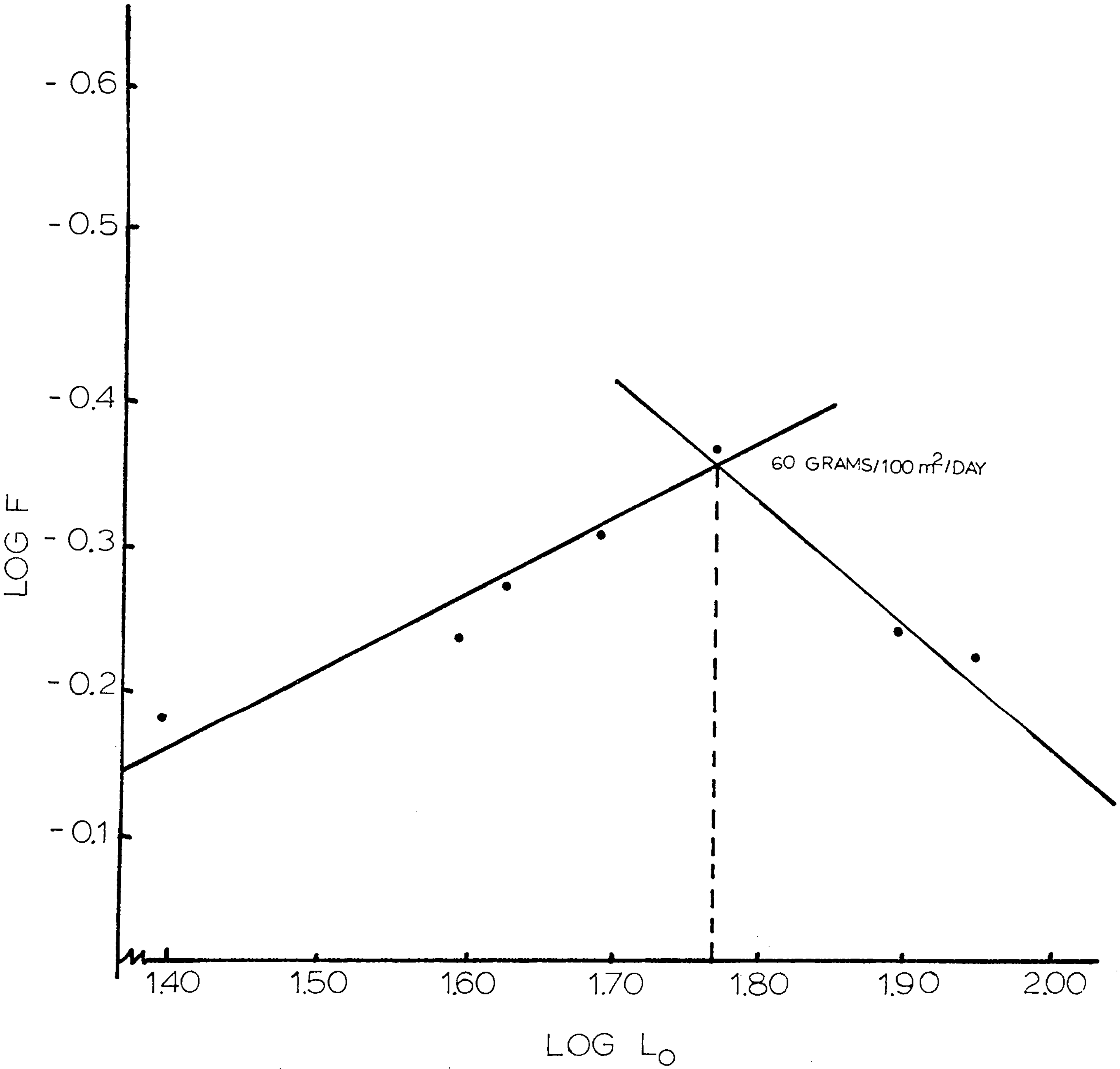

FIGURE 3. LOG F VS LOG L_O IN SALTWATER

section yielded an influent chemical oxygen demand of approximately 115 mg/ℓ. The 10 percent sea water system had the same organic removal efficiency as the freshwater system. Both systems achieved a 70 percent reduction in chemical oxygen demand.

Table 4. Saltwater Ammonia Results

Ammonia Loading L_o (grams/100 m^2/day)	Log L_o	% Removal	F	Log F
25	1.40	34	0.66	-0.18
40	1.60	41	0.59	-0.23
50	1.70	50	0.50	-0.30
60	1.78	56	0.44	-0.36
80	1.90	43	0.57	-0.24
90	1.95	40	0.60	-0.22

The chemical oxygen demand was monitored at each stage of the contactor during saltwater testing. The results of this monitoring may be seen in Figure 4. These results show a rapid decline in the COD in the first stage, approximately 50 percent, followed by lower removal efficiencies in the remaining three stages. Based on these results, it can be said that the design of a rotating biological contactor for saline wastewaters is limited by the ammonia load to the system.

The 10 percent sea water concentration yielded a salinity of 3.5°/∞ (grams/kilogram). This corresponds to a chloride concentration just under 2,000 mg/ℓ (1,920 mg/ℓ). This chloride concentration is considerably lower than others used in previous saltwater testing. The 10 percent sea water salt concentration was used in this study because it is within the range of salt concentrations required by *Macrobrachium Rosenbergii* as they develop from the juvenile to adult stage.

The specific conductance of the 10 percent sea water increased slightly as it passed through the contactor. The influent specific conductance was 5,500 µmho/cm @ 20°C. It increased to 5,800 µmho/cm @°20 C in the first stage and remained constant through the remaining three stages. This increase of 300 µmhos/cm represents an increase of dissolved ionic matter of approximately 4 percent based on the assumption that dissolved ionic matter in mg/ℓ is equal to the specific conductance multiplied by an empirical factor of 0.8.[29]

The dissolved oxygen and temperature were also monitored within the contactor during saltwater testing. The influent dissolved oxygen was 0 mg/ℓ and increased to a concentration of 7.6 mg/ℓ in the effluent of the fourth stage. This increase represents a reaeration capacity of 1.6 mg/ℓ/hour. The temperature of the wastewater decreased as it passed through the contactor. The influent temperature ranged between 28° and 30°C (82° and 86°F) and dropped to between 18° and 15°C (64° and 59°F) in the influent of the fourth stage. The dissolved oxygen-temperature profile may be seen as Figure 5.

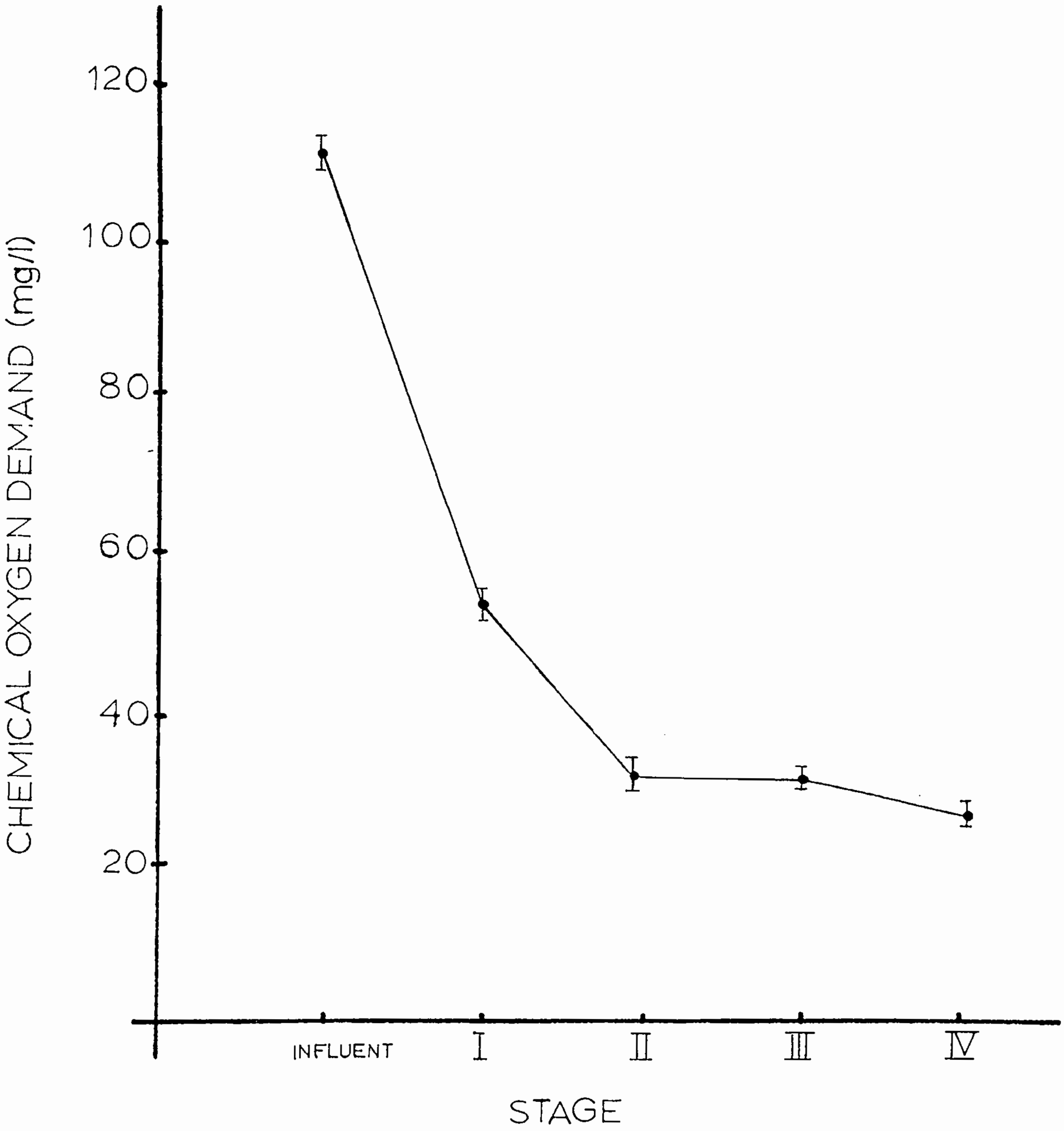

FIGURE 4. CHEMICAL OXYGEN DEMAND PROFILE

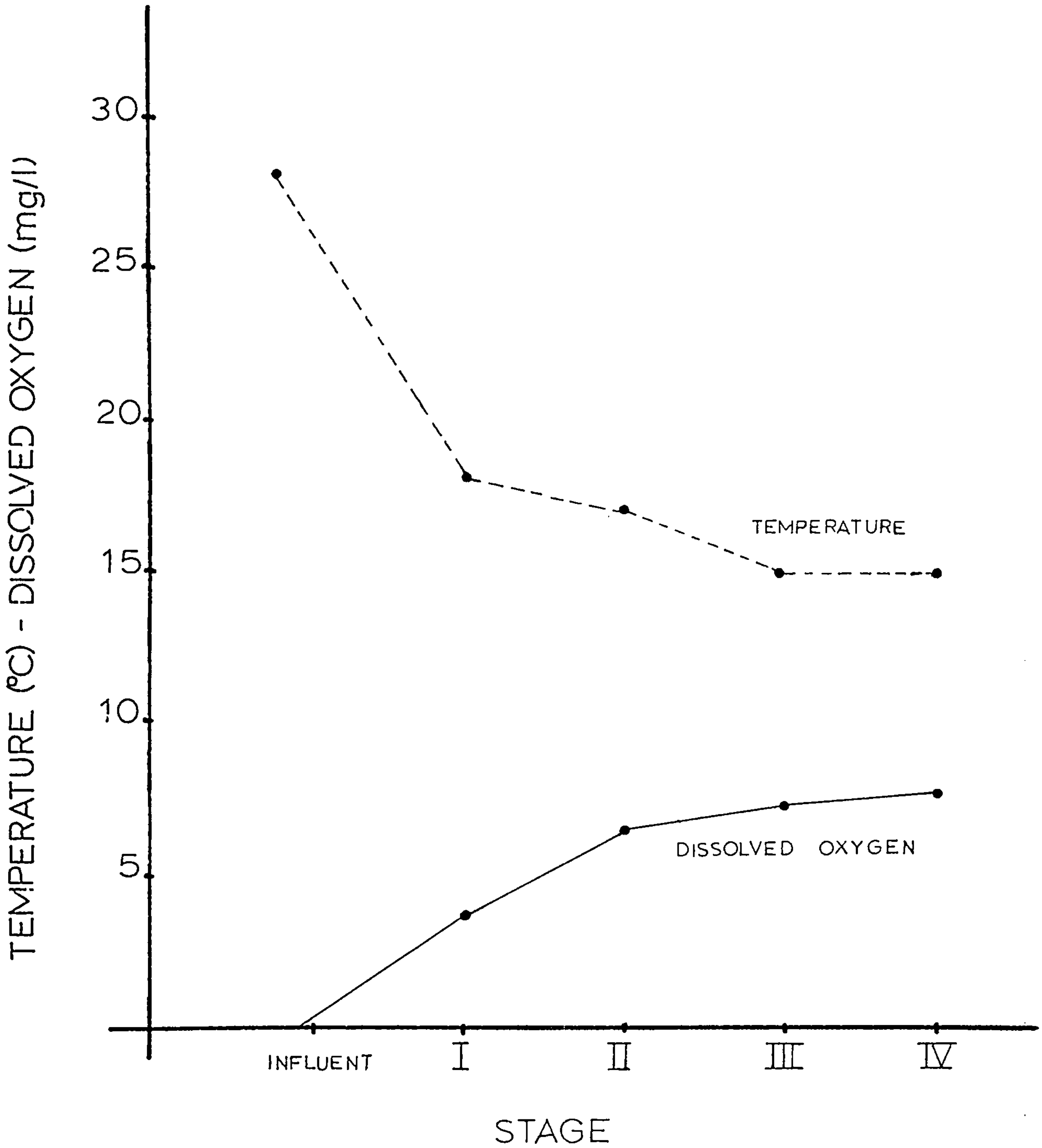

FIGURE 5. DISSOLVED OXYGEN--TEMPERATURE PROFILES

In summary, the results show that although the increased salinity did not seem to affect the contactors' ability to remove organic material measured as the chemical oxygen demand. The increased salinity did, however, alter the system's ability to nitrify ammonia. This decreased nitrification will become the limiting parameter in the design of a rotating biological contactor for use in an aquaculture facility.

CONCLUSIONS

The following conclusions can be made based on observations and analyses of the data collected in this plot study:

(1) The nitrification of ammonia can be accomplished in a 10 percent sea water waste.

(2) The ability of the rotating biological contactor to nitrify decreased with increased salinity.

(3) The organic removal efficiency of the contactor measured as COD was not affected by increased salinity (to 10 percent sea water).

(4) In designing a rotating biological contactor to nitrify a saline wastewater, the ammonia loading will be the limiting design parameter.

(5) The reaeration rate of the contactor was satisfactory to maintain a level of dissolved oxygen required within an aquaculture facility.

ACKNOWLEDGMENTS

Special thanks are extended to Anibal Alarcon for his insight during the period of testing. This investigation was supported by the Department of Civil Engineering, Environmental Engineering Program, Colorado State University, Fort Collins, Colorado and Klemetson Engineering, Fort Collins, Colorado.

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RBC FOR MUNITIONS WASTEWATER TREATMENT

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Introduction

The United States Army operates one of the largest industrial complexes in the United States. Just as other large chemical industries, the Army is subject to all Federal laws, including EPA regulations in the form of National Pollution Discharge Elimination System (NPDES) allowances. A part of the Army's chemical production is a group of products known as explosives. Since most of the manufacturing operations involved with this group of products are unique to the military industrial complex, it has been necessary for the Army to pursue a vigorous pollution abatement program, often being on the forefront of technology development.

This study was instituted to determine if a rotating biological contactor (RBC) could be used to treat wastes from a proposed new RDX-HMX manufacturing facility. Several problems were posed for this study. Initially, what would be the composition of the waste stream from the proposed facility? Secondly, what types of treatment would be effective? Finally, would there be any residual problems not handled by proper sequence of treatment?

The question of chemical composition of the waste stream was handled by Armament Research and Development Command (ARRADCOM), Dover, NJ. From pilot plant studies at Holston Army Ammunition Plant (HAAP) and the plans and material balances for the proposed X-Facility, two ranges of waste streams were predicted. The two ranges, designated A stream and B stream, differ only from dilution by condensate from a cooling tower. The predicted chemical compositions of the two waste streams are shown in Table 1. Table 2 gives the actual values chosen for use in the bench scale study. One concern was that the toxic concentrations of the constituents of the waste stream might not allow biological degradation to take place. Results of preliminary testing showed BOD removal taking place indicating biological degradation. A result which had impact on further testing was the wide disparity between anticipated and actual values of BOD and COD. Projected values were roughly 1/6 the initial readings of these two parameters. The stream was unique in composition and clearly the need existed to determine the most effective treatment system.

In determining the type of treatment, current technology was examined. Work had been conducted previously at Holston Army Ammunition Plant on wastes similar to those proposed. It was determined at that time that a biological system could be the most effective treatment for their operation, but the system chosen was not a RBC. Radford Army Ammunition Plant had also investigated a biological system and had used a rotating biological contactor for its wastes, but these wastes were not similar to those expected at the proposed facility. The designer of the new manufacturing facility reviewed both sets of data and proposed a treatment system which contained, as a key element, a RBC. The proposed treatment scheme is shown in Figure 1. To establish the validity of the treatment scheme, it was decided to perform bench scale studies at Mobility Equipment Research and Development Command (MERADCOM) and pilot scale studies at Atlantic Research Corporation. The authors of this paper were involved with the bench scale testing, though close proximity of the two organizations allowed good communication. It was determined that a complete treatment train at the bench scale level would be set up. Emphasis would be given to the biodisc component since this would contribute the major portion of the removal. Due to a number of system operational problems, it was quickly discovered that operation of more than the aerobic biodisc would prove fruitless, so work was limited to a detailed study of the aerobic bench scale biodisc. The immediate experimental goal was the determination of the optimum flow rate of formulated wastewater to the biodisc to achieve maximum BOD removal in conditions A and B.

A critical concern in the system was the explosive components because of possible residual toxicological and mutagenicity problems. For the past several years, the Surgeon General has had an extensive toxicology study underway for TNT, RDX, and HMX among other explosives. These materials have been found to be toxic in varying levels, depending on the conditions to which the explosives have been subjected. TNT, for instance, is photolyzed by sunlight and its composition is altered if the solution is basic. The RDX and HMX do not photolyze but do undergo some hydrolysis reactions. In both cases, the toxic properties are altered but not eliminated. The RDX and HMX went through the system almost unchanged, while the TNT degraded before treatment by the disc. From the data, it is clear that treatment, other than biodisc, of the explosives in the contaminated wastewater will be necessary. Carbon studies have been done and if a properly designed carbon system is employed, tests have established that the explosive components can be removed to acceptable levels.

EXPERIMENTAL DESIGN AND OPERATION

As previously mentioned, the original flow scheme is illustrated in Figure 1 and constituents of the hypothetical waste stream are set out in Table 2. Initial mixing of the waste stream took place in a covered 1100 litre tank on a weekly basis. PH adjustment was made using ammonium hydroxide on a batch basis to raise the pH to approximately 7.4. The wastewater was pumped into a smaller feed tank, 120 litres, for controlled flow into the biodisc system. Continuous additional pH adjustment was made in this tank by an automatic control system. Uniform mixing in both tanks was accomplished by use of submerged pumps.

From this feed tank, the flow went into the aerobic biodisc under control of a Masterflex pump. A picture of the aerobic biodisc unit appears as Figure 2. Effluent from the aerobic biodisc flowed into the anerobic biodisc unit which was covered and airtight. Beyond the anaerobic unit, the flow was pumped into an aeration chamber where air was bubbled through the effluent. This process enhanced settling of sludge which consisted primarily of biomass. A clarifier followed in the flow pattern, and effluent was pumped next into a multi-media column and on through a carbon column. The liquid was then discharged into the drain.

A number of alterations in the physical set up took place over the course of the experiment. The first was removal of the multimedia column and carbon column due to growth of microorganisms which appeared similar to those present on the discs. Backwashing the columns was not effective in removing these microorganisms, particularly in the multimedia column, as they clumped together, forming flake-like particles which clogged the column and caused substantial back pressure.

The second alteration in the system was the removal of the anaerobic unit. At a point early in the testing, it was necessary to dewater the entire system and it was noted that virtually no growth had occurred on the discs rotating in the anerobic unit. This fact had been anticipated due to the lack of gas evolution from the unit. Possible explanations include the relative delicacy of anerobic organisms and the fairly wide fluctuations in pH and flow rate to which the system was subjected at start up.

The third alteration was in feed tank size. After two months of operation, it became apparent that use of the 1100 litre tank for feed mixing was not an experimentally sound procedure. Extensive biological growth had taken place there, and the symptom of that growth was consumption of COD by these organisms. That is, COD was significantly higher for the tank mixture immediately after mixing than it was later in the week. For this reason, mixing was done in the smaller 120 litre feed tank on a more frequent basis, in the hope that the shorter retention time would inhibit growth. As of this writing, four additional months into the testing program, it appears that growth in the feed tank is again a problem, calling for further modification of the feed flow system.

At this point, then, the system has been pared down to functioning units as illustrated in Figure 3.

It is noteworthy that very early in the testing program, growth on the aerobic disc unit was limited, as was COD removal. It was necessary to balance the ingredients required for good biological growth by the addition of two chemicals. Addition first of nitrogen, as ammonium hydroxide, and then of phosphorous, as sodium phosphate, as supplemental nutrients proved to be the necessary changes, and substantial growth appeared within a week; COD removal efficiency increased dramatically as well. Amounts of nitrogen and phosphorus to be added were calculated using a molar ratio of carbon:nitrogen:phosphorus::106:16:1 which is the approximate ratio at which the microorganisms are thought to synthesize those elements.

At start up, projected pH levels were in the range 5.5-7.3; by contrast, it was found that the actual pH of the mixture in the feed tank was approximately 3. It was necessary to add significant amounts of ammonium hydroxide during the use of each batch. This was above and beyond the initial addition and must be considered in the actual plant operational costs.

In bench scale testing, it was important that the three explosives be completely dissolved to be sure that they were actually included in the feed stream to the biodisc. In order to reduce times and facilitate complete mixing, it was necessary to enhance the solubility of the RDX and HMX by the addition of more cyclohexanone than was called for in the hypothetical waste stream. Sufficient acetone was present in the formula to insure the solubility of the TNT. For reasons to be discussed later, RDX, HMX and TNT were deleted from the feed stream for the second half of the experimentation.

Once the waste stream composition was fixed, analytical methods were used to follow its progress through the treatment process. Analyses of the wastewater were performed using techniques described in Standard Methods, 14th edition. The specific parameters which were determined and the procedures used are shown in Table 3. Analysis of the microbiological growth was done at Natick Research and Development Command, Natick, Massachusetts, using culture techniques and visible identification. The Ames test was run by Atlantic Research Corporation with the five tester strains of Salmonella typhimurium.

RESULTS AND DISCUSSION

Figure 4 shows a curve which represents the COD and BOD removal efficiencies from start up through attainment of optimum removal. The initial goal of 95% BOD_5 removal efficiency has been shown to be attainable consistently and repeatably. In spite of the high levels of formaldehyde, it is safe to conclude that the microorganisms were able to survive and to degrade the organics in the waste stream effectively.

For A stream with BOD_5 near 1600 mg/L, optimum removal efficiency was found to occur at a loading rate of 3.3 lb BOD_5 per day per square foot of disc surface area. Experimenters on both bench and pilot-scale levels were reassured that very similar removal efficiencies were attainable with both systems when comparisons were made using loading rates per square foot of disc surface area. This allowed confidence in scale-up factors from bench to pilot-scale and from pilot to plant-scale for design purposes.

Data generated at the bench-scale level was the result of feeding the biodisc unit on a 4-stage basis, that is, flow from compartment one went to compartment two, and so forth. When COD analysis was done on a stage by stage basis, it became clear that a substantial majority of the removal (around 70% of the total) took place in stage one. Atlantic Research modified the pilot plant scheme to have influent flow directly into stages one and two. In so doing, they simulated three stage operation and were able to increase COD removal.

It is interesting that pH measurements in the chambers of the biodisc unit gave indication of COD removal efficiencies. If the pH dropped precipitously in the stage 2 chamber, COD removal could be expected to be less than adequate. Investigation of this situation could yield an on-line monitoring technique.

It became apparent as the testing progressed that the RDX, HMX and TNT were not being processed effectively by the system. RDX and HMX were found in identical concentrations in the influent and effluent and TNT was transformed as previously mentioned. In both cases, it is clear that treatment other than by biodisc will be necessary and it is likely that TNT will not be part of the ultimate influent to the biodisc. For these reasons, the decision was made to leave the explosives out of the stock mix for the second half of the experimentation. It should be noted that though the TNT was transformed, it is not reasonable to conclude that the contamination problem had been dealt with. It is known that TNT is readily transformed into compounds as toxic as TNT. It has been found subsequently that the mix without the explosives is not as toxic. Clearly, disposal of the explosives is a problem yet to be dealt with. Treatment by carbon will probably be used, but the designers must allow for the fact that biological growth can hinder the operation.

In the course of the investigation, it appeared that the microorganisms were not as differentiated as is often found in sewage treatment plants using RBC's. Evidence was sought that the strain found was not so pure in culture as to be susceptible to total kill should some toxic agent enter the system. The test results showed that this was not a pure culture, and in fact, consisted of two strains of fungi and seven different colonial morphologies. Two fungi identified were Fusarium sp. and Geotrichum sp. Three pseudomonads were isolated, one from the pseudomonas genus and two pseudomonad organisms. Two common bacillus organisms of a ubiquitous nature rounded out the lot of microbes found. No further analysis was done to classify the organisms once it was apparent that they were typical of normal sewage system organisms, and obtaining seed material would not be difficult should a massive kill take place. Roughly 10-14 days were required to go from clean disc start-up to optimum removal, though clean disc start-up is not a likely occurrence due to the hardiness of the microbial population in the face of adversity.

CONCLUSIONS

As shown previously, the biodisc is effective in reducing the BOD and COD significantly. This is true even if there are explosives present which are known to be toxic. Ames tests run on biodisc effluent show that the explosives produce a mutagenic effect. When explosives are not included in the waste stream, the Ames test results are negative. These results were anticipated from previous work done by the Surgeon General's office. The most toxic of the explosives is TNT and its conversion products. This leaves two choices, either pretreat or posttreat to remove the explosives. Both solutions may be feasible but must be evaluated from an economic and a safety standpoint. Further investigation is needed concerning these toxic by-products and their removal.

One production line in the X-Facility had been originally designed with supplemental pollution abatement equipment to deal with high levels of nitrogen in the effluent. Since the bio-system needs nitrogen, that separate pollution abatement process can perhaps be eliminated; this would provide a substantial cost savings for the plant. Before the process can be changed it will be necessary to determine if the nitrogen is in a usable form for the biosystem and whether that stream contains any potential toxicants.

The small bench scale model has proven to be linear in scale-up when considering pounds of BOD per square foot of disc surface area. This means that small bench scale systems can be used to perform tests which should be valid for full scale systems. Many variations and conditions can be investigated such as changes in food sources and flow rates, potential plant chemical surges, and other anticipated problems. The effect on the organisms as well as on the efficiency of removal can be determined. The system is cost effective. A small scale unit takes less chemicals, requires less power to run, and can be conditioned more rapidly to changing parameters. Therefore, the bench scale unit is a mini-ecosystem which provides fast, cost effective, and reliable data for the investigation of large scale operations.

TABLE 1

PREDICTED QUALITY OF INFLUENT TO INDUSTRIAL WASTEWATER TREATMENT PLANT - X FACILITY

	CONDITION A		CONDITION B	
Total Flow, Gallons/Day	1,539,800		993,600	
Contaminants:	LB/DAY	MG/L	LB/DAY	MG/L
NO_3-NO_2	230	18	225	27
Ammonia	19-46	2-4	19-46	2-6
RDX	60-147	5-11	60-147	7-18
HMX	20	2	20	2
TNT	64-155	5-12	64-155	8-19
Acetic Acid	420-1052	33-82	420-1052	51-127
Hexamine	464-576	36-45	464-576	56-70
Cyclohexanone	518-648	40-51	518-648	63-78
Propyl Alcohol	653-816	51-64	653-816	80-99
Methyl Acetate	250-312	20-24	250-312	30-38
Propyl Acetate	77-96	6-7	77-96	9-12
Formic Acid	2246-2808	175-219	2246-2808	272-339
Nitromethane	250-312	20-24	250-312	30-38
Formaldehyde	6912-8640	539-674	6912-8640	836-1045
Phosphate	66	5	56	7
Sulphate	1102	86	829	100
Acetic Anhydride	400	37	400	48
Amine	60	5	60	7
Organic Nitrogen	69	5	69	8
Toluene	38-48	3-4	38-48	5-6
Stearic Acid	12-24	1-2	12-24	1-3
Acetone	566-696	43-54	556-696	67-84

Condition A: Total wastewater includes heat exchanger condensate

Condition B: Total wastewater without heat exchanger condensate

TABLE 2

EXPERIMENTAL PARAMETERS
FOR CONDITIONS A AND B

CHEMICAL	CONDITION A (MG/L) INCLUDES HEAT EXCHANGER CONDENSATE	CONDITION B (MG/L) INCLUDES HEAT EXCHANGER CONDENSATE
Formaldehyde	674	1045
Formic Acid	219	339
Sulfate	86	100
Acetic Acid	85	184
1-Propanol	64	99
Acetone	54	84
Cyclohexanone	51	78
Hexamine	44	70
Methyl Acetate	24	38
Nitromethane	24	38
n-Propyl Acetate	7	12
Phosphate	5	7
Toluene	4	6
Amines	5	7
Stearic Acid	2	3
TNT	12	19
RDX and HMX	13	20
COD	1650	2300
BOD	1390	1660
pH	3	3

TABLE 3

LIST OF PARAMETERS TO BE MEASURED
AND ANALYTIC METHODS OR PROCEDURES

PARAMETERS	METHOD
Biochemical Oxygen Demand	per *Standard Methods*[1]
Chemical Oxygen Demand	Dichromate Reflux method (*Standard Methods*[1], p. 550)
Total Organic Carbon	Dohrmann TOC Analyzer (*Standard Methods*[1], p. 532)
pH	Beckman pH meter
Temperature	Thermometer
Ammonia, Nitrites, Nitrates	Hach tests
TNT, RDX, HMX	Waters Liquid Chromatograph

[1]American Public Health Association, *Standard Methods for the Examination of Water and Wastewater*, 14th edition, APHA, Washington, D.C. (1975).

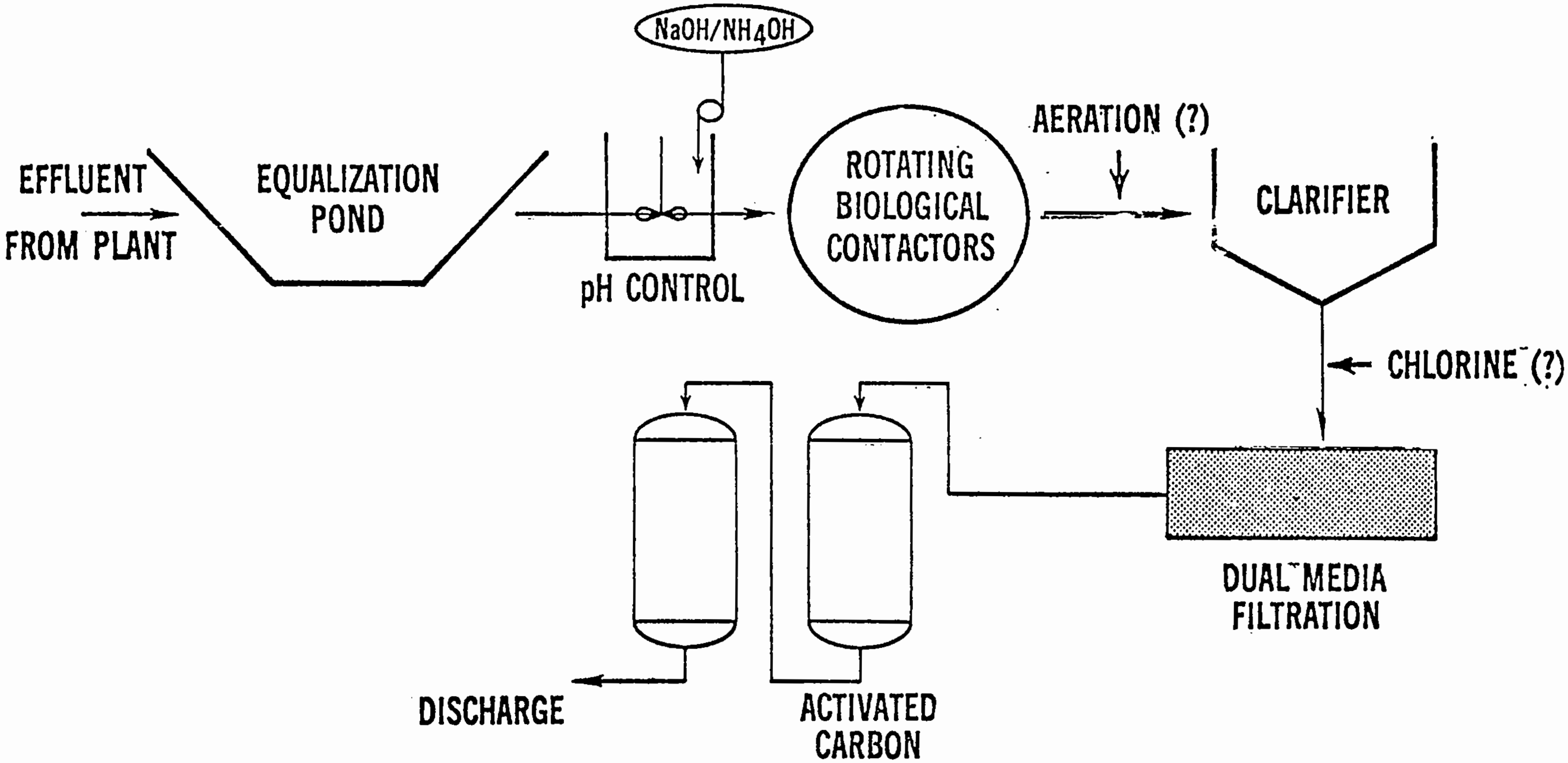
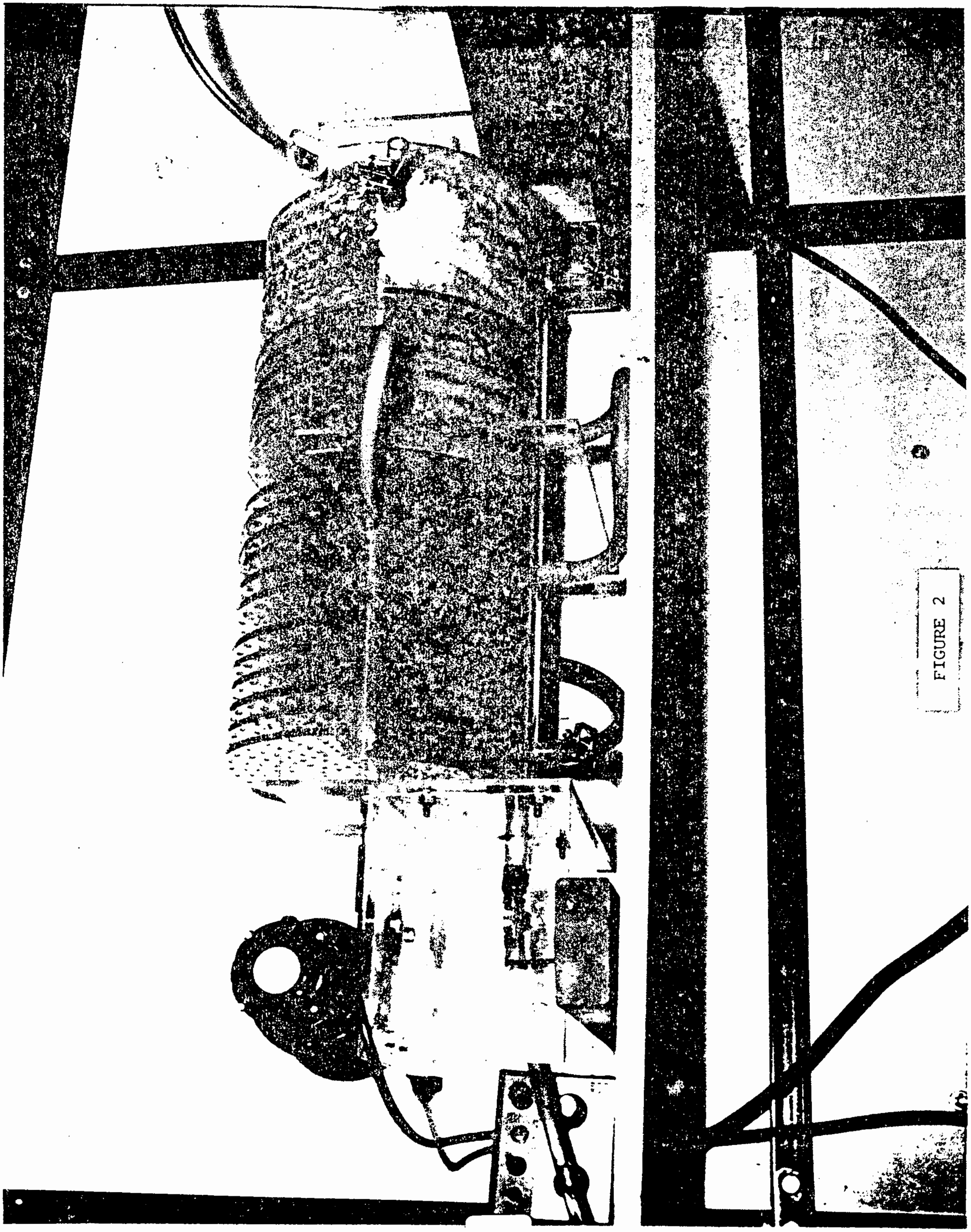
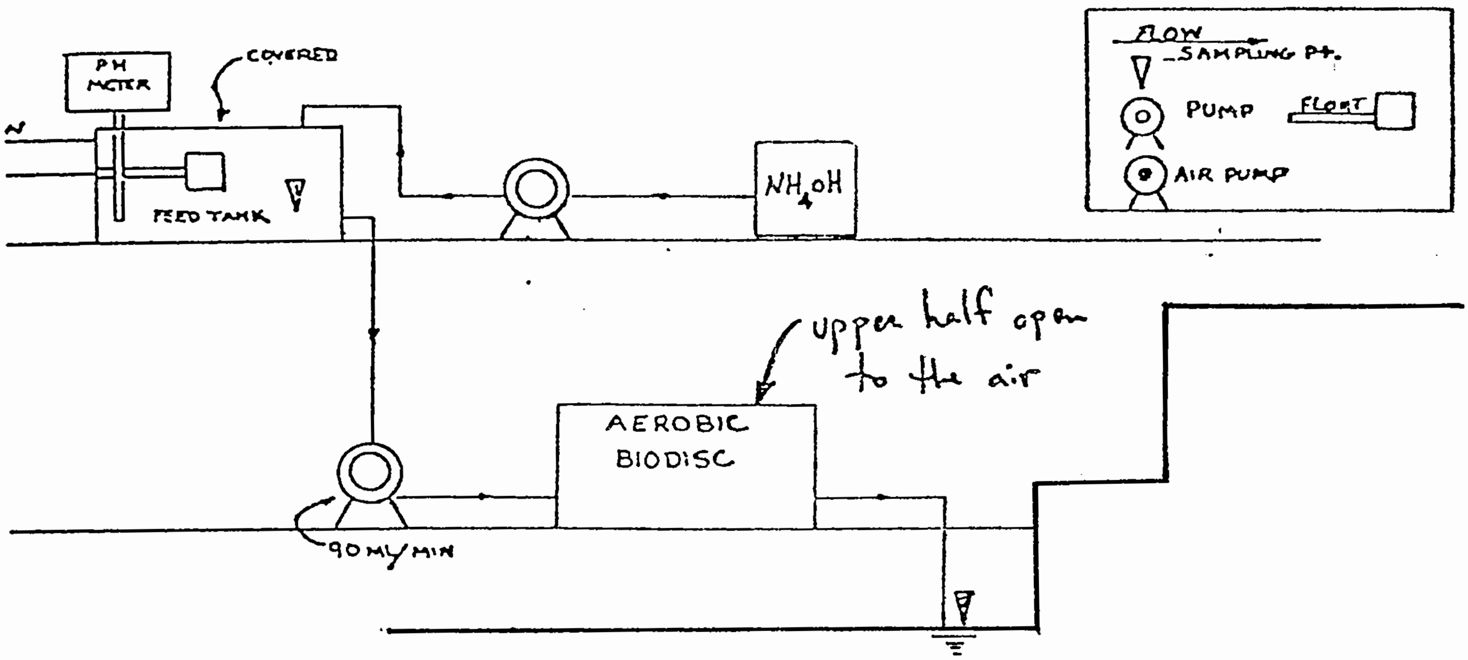

EFFLUENT
FROM PLANT
EQUALIZATION
POND
$NaOH/NH_4OH$
pH CONTROL
ROTATING
BIOLOGICAL
CONTACTORS
AERATION (?)
CLARIFIER
CHLORINE (?)
DUAL MEDIA
FILTRATION
ACTIVATED
CARBON
DISCHARGE

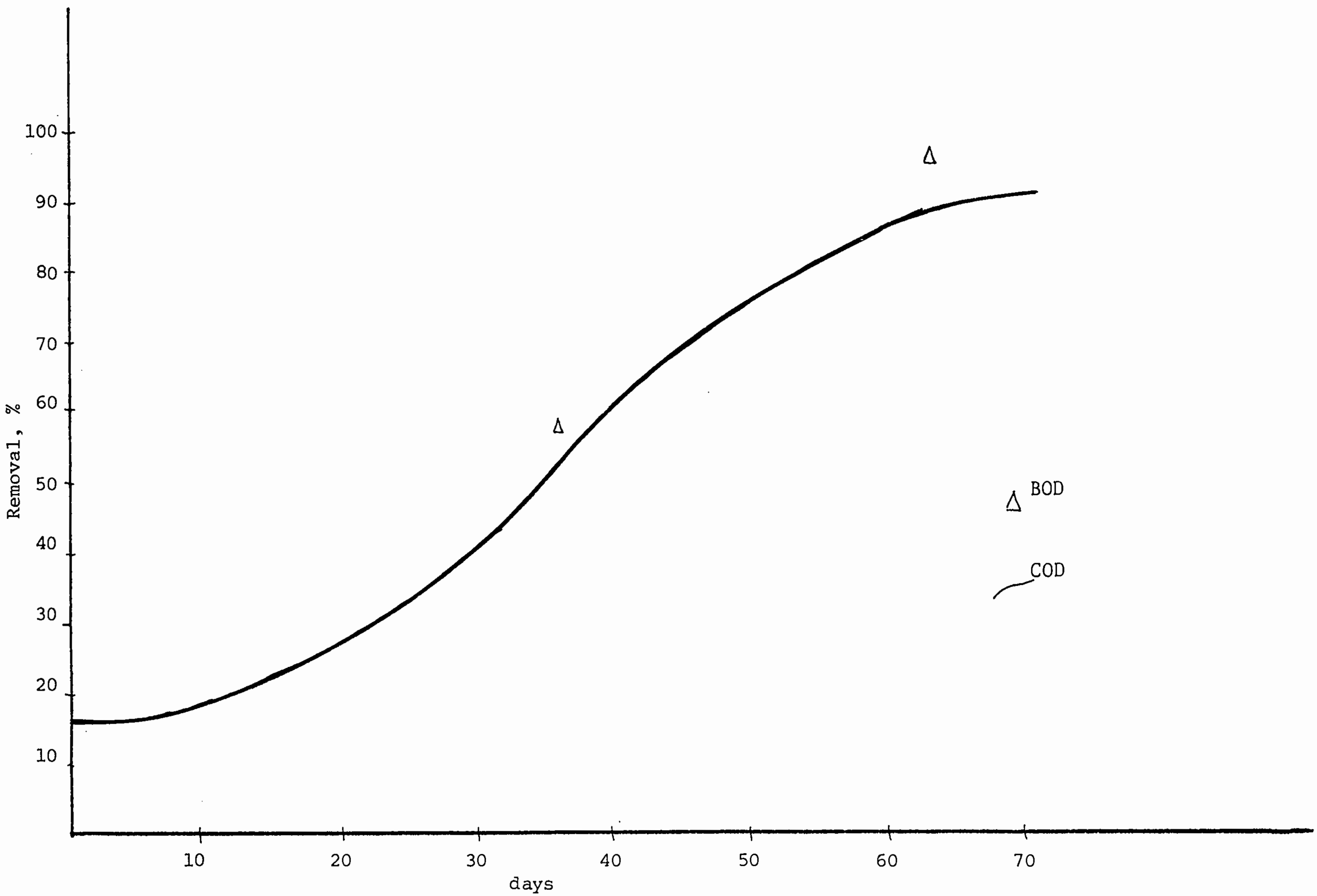
FIGURE 1





SCHEMATIC PHYSICAL LAYOUT
BENCH SCALE MODEL

FIGURE 3



Removal, %
100
90
80
70
60
50
40
30
20
10
10
20
30
40
50
60
70
days
BOD
COD

Figure 4.

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REMOVAL OF WASTE PETROLEUM DERIVED POLYNUCLEAR AROMATIC HYDROCARBONS BY ROTATING BIOLOGICAL DISCS

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Abstract

A staged, partially submerged rotating biological disc system was assessed to determine its performance in the reduction of concentrations of polynuclear (PNA) aromatic hydrocarbons attributable to waste crankcase oils (WCCO) in wastewater effluents. Removal of petroleum derived PNA hydrocarbons is important because of their known toxicity and carcinogenicity which pose potential public health risks. Samples of the influent, several successive stages of treatment by the disc system and the final effluent were collected, extracted with carbon tetrachloride (CCL4) and analyzed. Two UV-fluorescence spectroscopic techniques provide qualitative evidence of the presence of WCCO hydrocarbons. An IR-quantification method was utilized to determine the total extractable organics at each stage. The UV-fluorescence techniques rely upon the ability of aromatic hydrocarbons from WCCO to generate a "fluorescence profile" differentiable from other petroleum entities. Excitation of extracts at a specific wavelength produces WCCO fluorescence emission profiles which can be compared to known "standard oils." These fluorescence methods were successfully employed by us to identify weathered petroleum products

*The analyses for this investigation were performed under the auspices of the analytical research facility of the U.S. Environmental Protection Agency, Industrial and Environmental Research Laboratory, Oil and Hazardous Spills Branch, Edison, New Jersey, through an approved program that supports graduate level research study of the environment.

using exposed stock oils, and therefore they were used in this project for the detection of a gradient hydrocarbon response to the disc treatment system. Fluorescence maxima profiles (FMP's) showed a significant decrease in detectable PNA's attributable to WCCO through the successive stages of the disc system. Preliminary statistical analyses of IR quantification data revealed a direct relationship between the degree of treatment and the level of PNA concentration. These results indicate that such biological systems for the removal of WCCO aromatic hydrocarbons are viable alternatives to secondary treatment systems commonly being employed. Varying the flows, loadings and recycling may further improve removal efficiencies. Additional studies are warranted in light of the possible future need to reuse wastewaters.

Introduction

Considerable attention has been directed toward the use of rotating biological contactors (RBC's), and they have been shown to be an effective means of treating wastewaters.[1] Previous investigations[2] into the polynuclear aromatic hydrocarbon (PNH) character of wastewaters in New York City Water Pollution Control Facilities (WPCF), attributable to waste crankcase oils (WCCO) have exhibited a range of PNH's. Recent investigations[3] have also been directed at such compounds attributable to automotive WCCO not only because several constituents of these compounds (i.e. napthalenes) have been shown to "bioactivate" compounds into mutagens,[4] but also because of energy conservation needs.[5] The purpose of this preliminary work was by use of UV-fluorescence techniques, to qualitatively determine the performance of a pilot RBC system in the reduction of initial concentrations of detectable PNH's attributable to WCCO.

EQUIPMENT

The pilot plant at Newtown Creek is the same system operated by W. Torpey at the Jamaica WPCF during the period July to November 1969, and was comprised of three main component parts: (a) Ten stages of rotating disks (all discs were 3' in diameter) for the removal of organics and for oxidation of ammonia to nitrate, (b) six stages of illuminated rotating discs for the removal of nitrogen and phosphorus from the effluent of the preceding system by synthesis into attached algal cells, and (c) six packed beds of granular activated carbon columns for the adsorption of refractory organics from the preceding algal system. Sedimentation of 1.5 hours was interposed between the effluent from the ten stage unit and the algal unit. A mixed media filter preceded adsorption in carbon columns for removing the particulates, which were mainly algal cells generated on the illuminated disks.

The flow through the ten stages was 28.39 l/min. $\pm$ 10%, where as the algal unit rate was 11.35 l/min. Flow to the pressure downflow carbon columns was at a surface loading of 203.6 l/min/m^2. Disks rotated opposite to the direction of a flow through the successive stages. The theoretical detention time was six minutes in each stage, measured when the disks were devoid of slime. Actual time was somewhat less, depending upon the degree of displacement of fluid volumes by the slime. Addition information of the RBC system operating results can be obtained from Torpey, W., et. al. (1973).[6]

ANALYTICAL APPROACH

Because of the complex chemistry of petroleum, each petroleum sample lends itself to differentiation from others. This passive-tagging approach establishes specific qualitative parameters for oil samples in the form of "profiles" or "fingerprints" to be compared to a "reference standard profile." Thus, positive correlations for RBC pilot plant effluent samples are either established or not established with reference standards depending upon those portions of the petrochemical waste that exhibit themselves in fingerprints and remain stable under physiochemical processes and environmental conditions.

METHODOLOGY

Water samples were collected at successive stages along the system in 980 ml. wide-mouth, glass Mason jars with Teflon-lined caps. Samples were taken from the raw influent, stages 1, 3, 6, 8, 10, settling tank, 13, 16 and the final effluent. Each sample was adjusted for pH 3 and refrigerated throughout storage until analysis. Samples were extracted with 50 ml CCL_4 in seperatory funnels and the bottom layer collected. Solvent was stripped off and residue weight recorded. An infra-red (IR) quantification method was used to determine total extractable hydrocarbons (mg/l) from each sample.[7] CCL_4 extracts were jet-air evaporated, concentrated, and residues weighed and brought to volume in hexanes for UV fluorescence analysis.

Previous investigators have exhibited the ability of fluorescence spectroscopy to detect trace quantities of petroleum derived hydrocarbons in oceanic waters. Investigators[9] have been able to differentiate between a lubricating oil and a crude or fuel oil using fluorescence spectroscopic techniques. All petroleum products fluoresce when excited by UV light because of the presence of aromatic hydrocarbons with multi-ring configurations such as fused ring polynuclear aromatics.[10] A UV-fluorescence spectrophotometer with two independent monochromaters (150 watt xenon are light source), and a constant temperature cell bath maintained a 10mm path length quartz cell at 20° $\pm$ 0.5°C, was used for all fluorescence analyses. A synchronous excitation fluorescence spectroscopic technique was utilized for all analyses.[11] A standard reference WCCO was excited at 290 nm while scanning the emission spectrum from 240 to 540 nm, generating a maxima emission profile (MEP) for that excitation wavelength. This MEP was then used to correlate presence of WCCO in successive stages along the RBC system. In addition, each sampled stage was excited at successive excitation wavelengths from 240 nm to 440 nm (at 20 nm-intervals) while scanning for the maximum fluorescence emission at that excitation frequency. (Figure 1) Each maximum peak was utilized as a point to be plotted graphically, generating a "fluorescence maxima profile" (FMP) for each sample. Correlation was determined by visual comparison of maxima profile plots of the WCCO reference standard, to RBC stage maxima profile plots. (Figure 2).

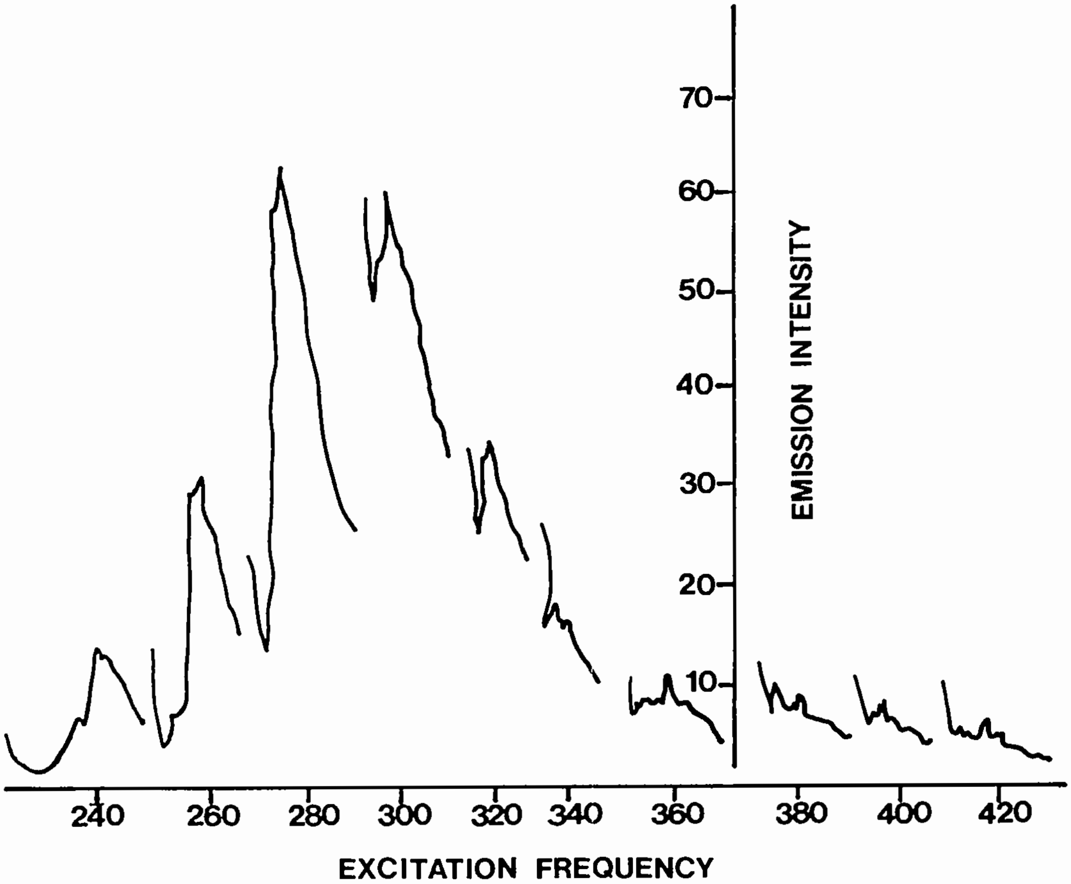

Figure 1.Fluoresence maxima profile (FMP) for WCCO.

It should be noted that the correlation criteria utilized for these analyses is essentially qualitative, in that source identification, (such as gas station or individuals dumping WCCO into sewers) of the detected waste petrochemical cannot be directly established by the technique. The Newtown Creek Sewage Treatment Facility (STF) however, is in an area of high petroleum hydrocarbon load.

There are several refineries and industrial facilities surrounding the treatment plant. Investigation by Mueller, J.A., et. al., [12] revealed the Newtown Creek STF to be the principal discharger of oils and grease in treated wastewaters to the New York Bight (12.6 metric tons/day).

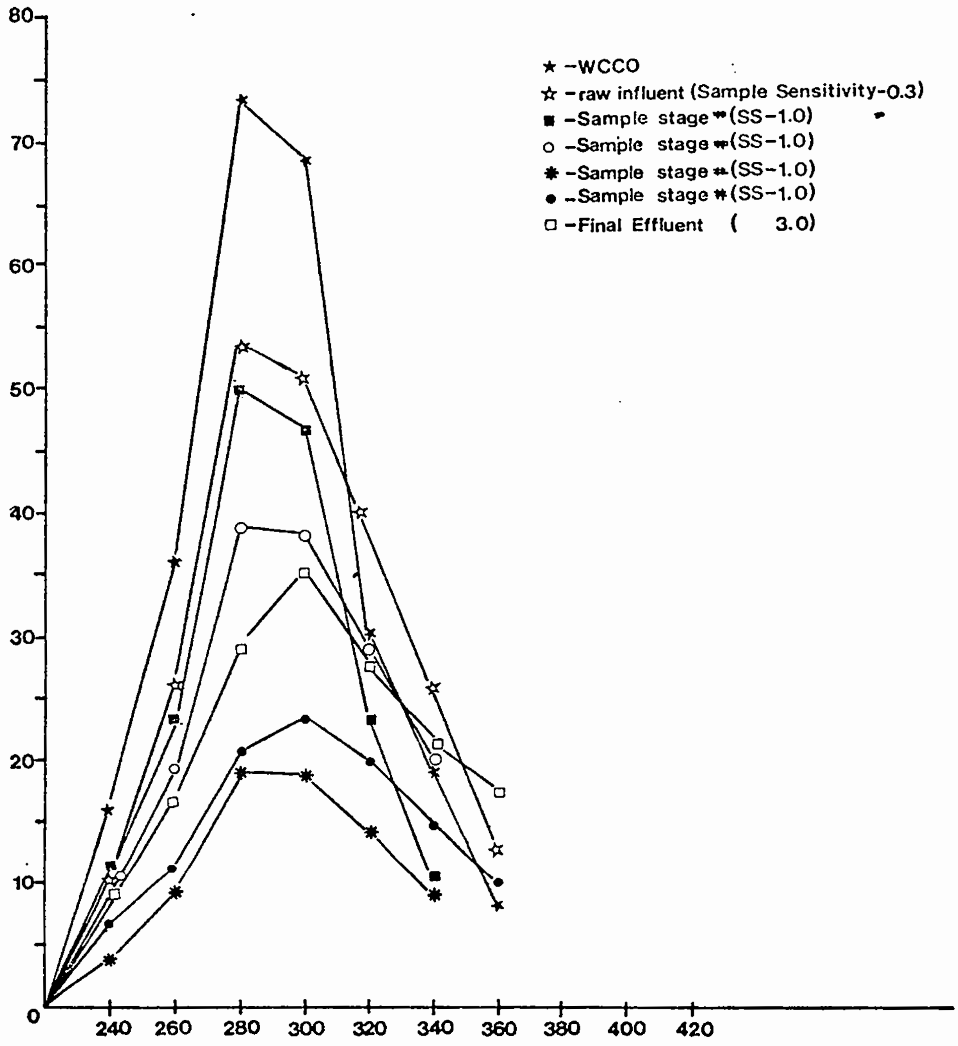

Figure 2. FMP's for RBC stages and reference standard.

RESULTS

Table 1 exhibits the total extractable organics (including petroleum hydrocarbons) obtained from the pilot RBC system at Newtown Creek. Preliminary statistical regression analysis appears to indicate that this data is a linear representation of RBC stage and PNH removal. Further detailed quantitative work is required here to substantiate a consistant linearity to PNH removal. In addition, fluorescence profiles of RBC stages RAW, to #16, correlate with WCCO FMP's. The final effluent sample and sample #16, at 10X over initial influent sample sensitivity setting, did not reveal profiles correlating to WCCO. (See Figure 2) Investigations into which PNH's are not being removed should be conducted. Results strongly indicate a considerable reduction of PNH's attributable to WCCO, in the final effluent. Future work in this area should establish an extensive sampling scheme, coupled with

TABLE 1

TOTAL EXTRACTABLE ORGANICS FROM RBC SYSTEM

Sample	Fluorescence Sample Sensitivity	Correlation with FMP of WCCO	MG/1 (TEO)
Raw	.3	(+)	16.7
1	.3	(+)	10.2
3	.3	(+)	9.9
6	.3	(+)	7.9
10	.3	(+)	4.4
Settling Tank	1.0	(+)	2.6
13	1.0	(+)	4.8
16	3.0	(-)	2.6
Final	3.0	(-)	0.12

variations in operational modes so as to clearly establish whether such petroleum derived hydrocarbon loadings are consistantly and effectively removed by RBC systems.

DISCUSSION AND CONCLUSION

A number of PNH's are potent carcinogens in animals and man. WCCO has been shown, along with other petroleum products to contribute significant quantities of detectable PNH's to aquatic environments. Traditional wastewater treatment facilities in major urban areas have been shown to be relatively ineffective in elimination, or providing a significant reduction of these compounds. RBC systems have on the other hand, been successfully employed in treatment of wastewaters,[13,14,15] and have exhibited here surprisingly effective removal of PNA's over suspended cultures.

With more investigation into variations in such operational parameters as detention times, slime build-up, influent loadings, effective disk surface area, submerged disc depth, wastewater flow rates and temperature, greater insight into PNH removal efficiencies may be provided.

This preliminary investigation appears to provide some credence to RBC systems as an effective means to treat sewage, and to prepare water for future re-use.

ACKNOWLEDGEMENTS

Special appreciation to W. Torpey for permission to sample the pilot RBC system at Newtown Creek, to M. Alavanja for the preliminary statistical analysis, and to M. Gruenfeld, U. Frank and Dr. G. Kupchik for their comments and support of this project. Discussion of results was provided by W.Torpey

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TREATMENT OF PHENOL-FORMALDEHYDE RESIN WASTEWATER USING ROTATING BIOLOGICAL CONTACTORS

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INTRODUCTION

Phenol-formaldehyde resin manufacture produces a very high strength waste (COD approximately 200,000 mg/l), containing 60,000 mg/l phenol, 20,000 mg/l formaldehyde and 15,000 mg/l methanol (PFM waste). Economical and environmentally acceptable disposal of this waste is a major problem.

The problem is emphasized because of recent regulatory activity which places phenols on the EPA list of priority pollutants. A survey of local sewer use ordinances for 40 municipalities (POTW's) [1] revealed phenol discharge limits of 0.005 - 3000 mg/l with a mode of 0.5 mg/l. Thus, to discharge PFM waste to most POTW's would require pretreatment and/or dilution to reduce phenol by 99.99992%. Because of such stringent requirements, PFM waste has been destroyed by incineration. While this method results in zero discharge, the economic attractiveness is decreased as the price of energy increases.

After researching the techniques available to treat and/or dispose of PFM waste including chemical oxidation [2,3,4], recovery systems [5,6,7], and biological treatment [8,9,10], a multi-staged process scheme was selected. Although all unit processes involved established technologies, the proposed combination of treating and disposing of the PFM waste was unique. A rotating biological contactor (RBC) was an essential part of the process train. This paper focuses on the use of the RBC in treating PFM waste; discussion of the other components in the process can only be covered briefly here.

PROCESS SELECTION

The selection of processes was largely influenced by the stringent local sewer ordinances for phenol (1.0 mg/l), by the location of the plant in Northern California, and by the low waste volume (less than 10,000 gpd). The stringent phenol limit meant that a treatment plant treating the PFM waste would have to attain greater that 99% phenol removal. The combination of Northern California location and climate, low waste volume and available land area made attractive a zero discharge approach, using terminal facultative evaporation ponds. Prior to discharge into the ponds, it was necessary to consider additional biological treatment so that pond area for treatment could be reduced and the need for pond aeration could be eliminated along with the possibility of stripping PFM into the atmosphere. Rotating biological contactors were selected for the pre-pond biological treatment step because of their ease of operation, their relative freedom from sludge settling problems compared to activated sludge, the ability to cover the system for odor control, and their ability, compared to trickling filters, to operate at high loadings without clogging.

A survey of the literature on phenol and formaldehyde removal by biological treatment revealed that up to 2000 mg/l of phenol is amenable to biological treatment with 91% phenol removal;formaldehyde concentrations of up to 670 mg/l have been treated biologically with removals generally in excess of 90%. The results of this review (Table I) indicate that the usual phenol and formaldehyde levels subjected to aerobic biological treatment are in the range of 100-500 mg/l phenol and 100-300 mg/l formaldehyde.

Organic removal rates have been reported for some phenol-containing wastes in fixed film biological systems. Porter and Dutch [15] report the use of a plastic media trickling filter tower with effluent recycle to dilute influent phenol to approximately 100 mg/l. Phenol removal was 98% when the filter was loaded at approx. 22 lb phenol/1000 ft^3 of media (approx. 2 lb COD/1000 ft^2/ day.) Jenkins [16] indicated that a phenol-formaldehyde resin waste was diluted 40-fold and treated at the rate of 5000 Imp. gal/yd^3/day on a rock media trickling filter (assumed specific area of 15 ft^2/ft^3) at a rate of 45 lb COD/ft^2/day with approximately 25 percent phenol removal.

TABLE I

BIOLOGICAL TREATMENT OF PHENOLIC/FORMALDEHYDE WASTES

Method	Reference	Initial Phenol Concentration (mg/l)	Phenol Reduction (%)	Initial Formaldehyde Concentration (mg/l)	Formaldehyde Reduction (%)
Activated Sludge	Eisenhauer, 1968 [11, 12]	100-750	90-100	-	-
Oxydation ditch	"	400-900	87.99	-	-
Lagoon	"	100-250	90-95	-	-
Trickling filter	"	200-500	85-98	-	-
Trickling filter	Wolnak, 1971 [13]	-	-	200	75-90
Trickling filter and lagoon	"	-	-	200	99
Activated Sludge	Biczysko, 1969 [14]	2000	91.0	670	93.2
Activated Sludge	"	1000	97.6	330	99.6
Activated Sludge	"	300	99.4	100	99.6
Trickling filter	Porter-Dutch, 1960 [15]	100	98	-	-
Trickling filter	Jenkins, 1957 [16]	1000	25	-	-

While these literature values for phenol waste loading and removal efficiency served as a guide for the the dilution requirements and loadings, as well as removals expected in an RBC plant, no specific information was available on the performance of RBC systems treating PFM wastes. It was also known that facultative fonds can act as clarifier, sludge digestor and a source of recycle water, and are able to efficiently treat low phenol concentrations (25-100 mg/l [11,12,17]). The RBC/ facultative pond recycle system was thus to be the basic treatment approach; however, we also discovered in laboratory scale testing that PFM waste, itself the distillate from a high temperature distillation at the resin manufacturing site, could be further distilled at low temperature (40-70°c) reducing phenol and formaldehyde concentrations significantly. The pot residue of the low temperature distillation could be incinerated at close to self-supporting combustion, as the residue had almost 60 percent of the energy content of #2 fuel oil.

A pilot study was needed to test the feasibility of this combined treatment process, to provide data that would establish the effectiveness of the approach, and to allow an attempt at optimizing the sizing of a distillation/RBC/facultative pond recycle system, or just an RBC/ pond system should the distillation step not prove cost-effective. The pilot study described in this paper had this objective.

MATERIALS AND METHODS

Pilot Plant Description

Two parallel but separate pilot plants were constructed during September and October 1977, and operated from November 1977 through March 1978. The biological treatment sections of both plants were identical (Figures 1 and 2), and consisted of two stage RBC units followed by a settling pond from which a recycle stream was drawn. The recycle stream passed through a small RBC that was installed to provide a source of seed organisms, in the event the larger two-stage RBC was inactivated by toxic shock loads of PFM waste.

Pilot Plant A. A 4-stage, 250 ft^2 surface area RBC pilot plant was leased from the Autotrol Corporation (Milwaukee, Wisconsin) and subdivided into two 2-stage, 125 ft^2 units by blocking the overflow weir between stages 2 and 3. RBC Unit A consisted of the first two stages of the Autotrol unit. It was fed by a calibrated scoop which rotated coaxially with the contactor. The discharge was taken from the second stage of this unit. The combination of raw waste and recycle flow rates,used in the pilot plant operation provided a hydraulic residence time of 3 hours in the RBC.

The settling pond was constructed from a 15-ft diameter, 4-ft deep plastic pool. With a water depth of 3 ft, the pond contained 4000 gal. Recycle flow was taken from a point 6 inches below the water surface by an 0.2 hp submersible pump, and discharged to a sump with an overflow

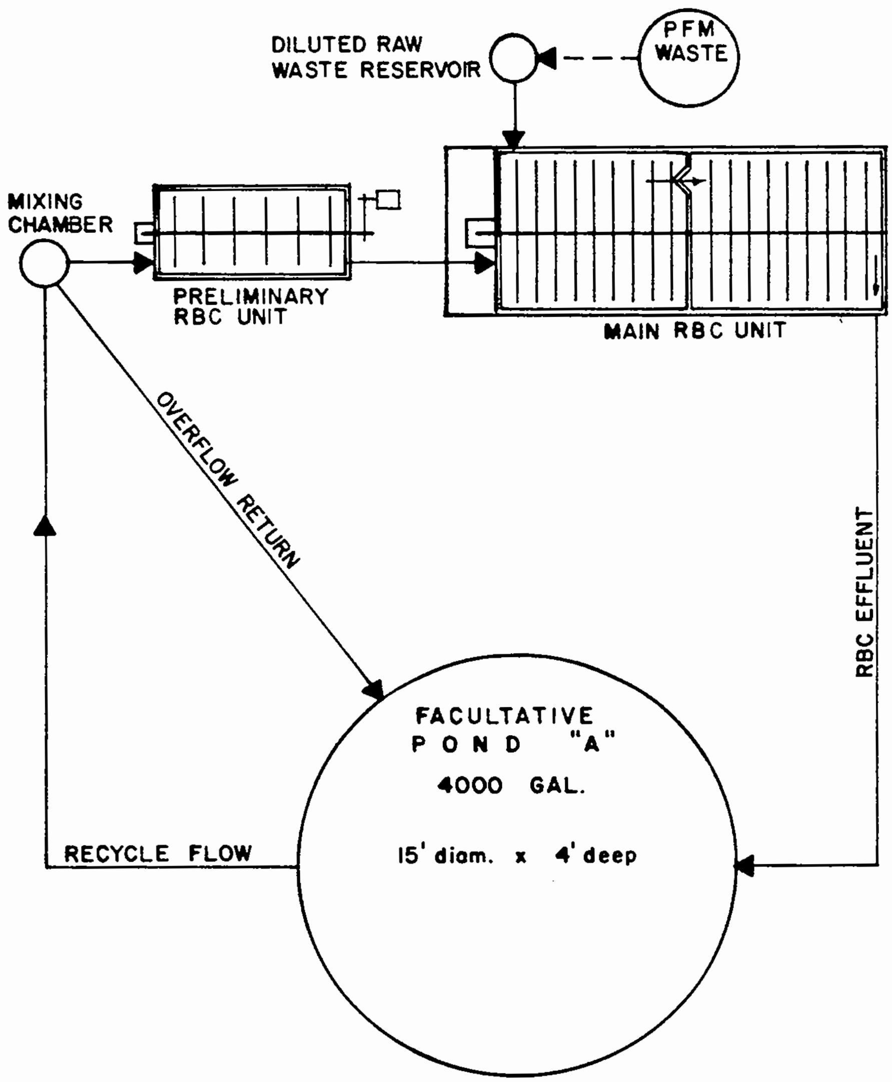

FIGURE I PILOT PLANT "A"

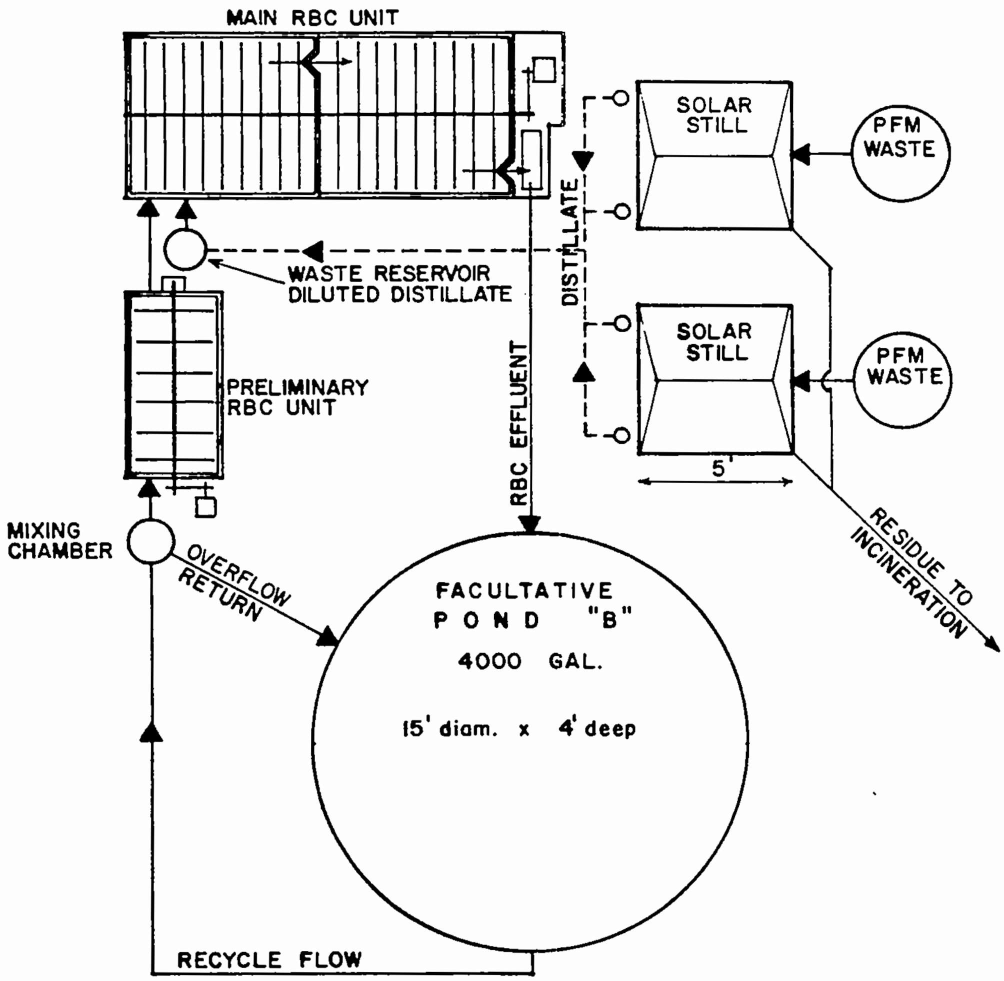

FIGURE 2 PILOT PLANT "B"

back to the pond. The recycle rate was controlled by a manually-operated valve on the sump outlet. The recycle flow then passed through a locally-fabricated RBC containing six 2-foot diameter roughened plastic discs, to provide 35 ft^2 of contact surface.

Raw PFM waste, diluted 4-fold with domestic primary treated sewage was fed into the RBC feed chamber, where it mixed with the recycle stream from the preliminary RBC. When required, nutrients (ammonium phosphate) were added directly to the diluted PFM waste reservoirs to provide a 90:3:1 ratio of C:N:P.

Pilot Plant B. In addition to the units described for Pilot Plant A, this plant had two 25-ft^2 solar stills prior to the biological system, fabricated from galvanized sheet metal, with glazing consisting of two 3-ft x 5-ft x 1/4" glass plates, set at 22° to the horizontal. The design of these solar stills was based upon those currently used for desalination of seawater [18,19].

Analytical and Sampling Methods

Analytical Methods. All samples were preserved and analyzed in accordance with Standard Methods for the Analysis of Water and Wastewater 14th Edition [20], except as noted. COD was by the method of Ryding and Forsberg [21]. The method was validated by running parallel samples by Standard Methods techniques and obtaining values which were 95-100% of those determined by the Standard Methods technique. Phenol was by gas chromatography with a Varian Aerograph, Model 1200, with a 3-ft x 1/8 in. SS, 10% FFAP on Chrom W column. Parallel and additional determinations were by the 4-amino-antipyrine method (Standard Methods). Formaldehyde and methanol were by gas chromatography, using a 5-ft x 1/8 in. SS Poropak N column. Parallel and additional analyses were attempted by the method of Jephcott [22]. Ortho-phosphate was by the stannous chloride method; Nitrate was by the zinc reduction method; Ammonia Nitrogen was by the method of Zadorogny, et al [23]; Dissolved Oxygen was *in-situ*, using a YSI specific oxygen electrode; pH was *in situ* using a Corning No. 6 pH meter.

Sampling and Data Collection. The pilot study was conducted during Fall and Winter of 1977-1978. Mean daytime temperature averaged 13°C.

In the pilot-scale distillation studies, distillation was discontinuous, with make-up volumes of raw PFM waste added as required to maintain a certain minimum volume in the still. The reported results of low temperature PFM distillation are weighted averages of fractions collected.

In the biological treatment studies, all samples were grab samples. At a minimum, a bi-weekly sample was collected and stored on ice until it was returned to the laboratory for analysis. Grab samples were deemed generally representative because of the controlled hydraulic and organic loading rates.

PROCEDURE

The pilot plant program is subdivided as follows:

Start-up and acclimatization
Phase I - Operation at moderate PFM waste loadings
Phase II - Operation at high PFM waste loadings
Phase III(a) - Treatability of pond recycle flow
Phase III(b) - Treatability of PFM solar distillate

Start-up and Acclimatization

Primary treated domestic sewage was fed into each pilot plant at the hydraulic loading rate recommended by the maufacturer (1.6 gal/ft^2/ day, or 200 gal/day) for a period of two weeks.When an observable biomass had developed on the RBC surfaces, PFM waste was introduced in steadily increasing concentrations, while maintaining the hydraulic loading constant at the above stated value.

Increases in PFM waste concentration were introduced first to Plant A and then one week later, to Plant B, as a precaution in case an overdose inactivated Plant A. This proved to be an unnecessary precaution, as no toxic effects ascribable to increasing PFM waste levels were observed during the study.

The PFM waste feed rates were increased to produce phenol increments of 50 to 75 mg phenol/l at approximately weekly intervals, until influent phenol concentrations of 300 to 400 mg/l were reached in each system. At this influent phenol level, the influent COD concentration was approximately 1500 mg/l. After 6 weeks of acclimatization the domestic sewage was replaced by recycle from the ponds.

Phase I - Moderate PFM Waste Loading

Influent COD levels greater than 1200 mg/l in RBC Unit A were obtained on December 1, 1977 (day 1 of the pilot program, and averaged 1543 mg COD/l (21 lb COD/1000 ft^2 media/day) over the next 48 days. RBC Unit B, the back up or control unit, did not reach influent COD levels over 1000 mg/l until December 23, 1977 (day 14) and averaged 1219 mg COD/l (16 lb COD/1000 ft^2 media/day) over the next 34 days. During this period it was attempted to keep the influent COD loadings as constant as possible because at these COD levels phenol concentrations averaged 350-400 mg/l, a value that from the literature was estimated to be readily treatable.

Dilution water was continually provided to both RBC units by recycling pond effluent. Nutrient sources present in the pond recycle stream left over from the RBC treatment of domestic primary effluent during start-up and acclimatization were sufficient to support biological growth during this phase. As a result additional sources of nutrients were not required.

Phase II - High PFM Waste Loading

The application of PFM waste was increased dramatically to RBC Unit A during the period from the forty-eighth to the fifty-eighth day and then gradually decreased over the next 19 days to test the stability of the RBC treatment process to shock loading. During this period influent COD and phenol averaged 3061 mg COD/l (41 lb COD/1000 ft^2/day) and 624 mg/l respectively with peak values on day 58 at 6228 mg COD/l (83 lb COD/1000 ft^2/day) and 1200 mg phenol/l. It was attempted to use RBC Unit B as a control during this phase and it was thus operated at a fairly consistent COD loading of 1442 mg/l (19 lb COD/1000 ft^2/day), almost the same influent loading as applied to RBC Unit A in Phase I.

Nutrients in the holding ponds were exhausted during this phase and nitrogen and phosphorous in the form of ammonium phosphate were added to both RBC waste feed systems in the C:N:P ratio of 90:3:1.

Phase III(a) - Treatability of Pond Recycle Flow

During the high PFM waste loading of Phase II, Pond A became overloaded with effluent COD and phenol concentrations over 600 mg/l and 100 mg/l respectively. To determine the comparative treatability of this COD and residual PFM waste in the holding pond, RBC Unit A continued to receive the recycle flow from Pond A but raw PFM waste was no longer added to the influent,i.e., RBC Unit A was used to treat the water in Pond A. This phase lasted 24 days.

Phase III(b) - Treatability of PFM Solar Distillate

The solar distillation rates of PFM waste during November through February for the size of stills used were inadequate to provide a sufficient distillate volume to adequately load an RBC Unit at 1200 mg COD/l on a daily basis. Therefore, solar distillate was collected and stored until March 1 and then fed for 24 days to RBC Unit B at the average influent COD loading level of 1235 mg/l (17 lb COD/1000 ft^2/day) and phenol level of 219 mg/l.

Solar distillation of PFM waste produces a distillate with a higher relative percentage of methanol compared to phenol and formaldehyde than in the raw PFM waste. This operating test run was undertaken to evaluate the comparative treatability of solar distilled PFM waste versus the previous runs treating raw PFM wastes.

PILOT PLANT RESULTS

Waste Streams

The chemical composition of the PFM waste stream to the pilot plants is summarized in Table II. Raw PFM waste was applied to RBC Units A and B during Phases I and II. Solar distillate from the PFM waste was applied to RBC Unit B during Phase III(b). Raw PFM waste has a pH of 4.6 and the chemical composition of its solar distillate is shown in Table

II. When the pH of the raw waste is elevated by the addition of NaOH or lime to pH 9 or 10.5, the phenol-formaldehyde removals increase substantially. Although this distillate was not used in the biological pilot system the chemical composition is presented in Table II for comparative purposes.

Pilot Plants A and B

Concentrations, loadings and removals of COD and phenol by each pilot plant unit are tabulated in Table III. Figures 3 and 4 present the time course of COD influent and effluent for each RBC unit. The effect of COD loading on COD removal rates by each unit is presented in Figures 5 and 6. The effect of COD loading on percent COD removal for both pilot plants is presented in Figure 7.

Results of COD, phenol, formaldehyde and methanol removal rates attained by the RBC units and pond system for Phases I through III are summarized in Table IV.

DISCUSSION OF RESULTS

It had been intended to conduct the pilot plant feasibility study during June through December; however, delays and time limitations required the program to be conducted from October through March during the coldest part of the year. Water temperatures in the small pilot plant averaged only 13-14 °C during the day and dropped to as low as 10 °C at night despite enclosing the RBC. Thus biological performance of the RBC was not optimal at these low temperatures and was reduced somewhat over what would be expected in a larger facility operating year-round. Despite the adverse weather conditions and organic loading extremes, the RBC pilot plant units showed operational stability and treatment reliability.

Biological Treatment of Waste

During Phase I when COD loadings were relatively steady and of moderate intensity RBC Unit A removed 61 percent of the influent phenol and 63 percent of the influent COD at average influent concentrations of 427 mg/l phenol and 1543 mg/l COD. On a media surface area basis COD loadings were approx. 21 lb/1000 ft^2/day with COD removals at 13.1 lb/1000 ft^2/day. The performance of RBC Unit B was slightly inferior - 54 percent phenol removal and 54 percent COD removal. Influent concentrations were 355 mg/l phenol and 1435 mg/l COD. The influent COD loading rate was 16 lb/1000 ft^2/day with COD removals of 8.8 lb/1000 ft^2/ day. The reason for this small difference in performance of the two RBC Units is not understood; in general, the results of the entire pilot program suggest that RBC Unit A may have performed slightly more efficiently than Unit B at the same organic loading rates. It should also be noted that the interpretation of the phenol results for Phase I of Pilot Plant B is limited as these are estimated values only consisting of two actual data points combined with extrapolations from the much more extensive COD data from Phase I.

TABLE II

PFM WASTE STREAM CHARACTERISTICS

PFM WASTE STREAM	COD mg/l	PHENOL, mg/l	FORMALDEHYDE, mg/l	METHANOL, mg/l
Raw Waste	180,000 - 200,000	45,000 - 65,000	20,000 - 24,000	15,000 - 22,000
Pilot Scale Solar Distillation pH 4.6	123,000	32,000	17,000	14,000
Laboratory Scale Distillation				
pH 9	31,300	9,000	5,000	6,900
pH 10.5	18,600	2,700	800	7,600

TABLE III

COD AND PHENOL REMOVALS BY RBC UNITS

COD		RBC UNIT A			RBC UNIT B		
		PHASES			PHASES		
		I	II	III	I	II	III
Influent[2] (mg/l)	min max ave	1220 1842 1543	1550 6228 3061	295 1070 694	138 2082 1219	1031 1830 1442	800 1659 1235
Effluent (mg/l)	min max ave	340 769 563	630 5568 2007	111 315 195	62 744 561	349 1572 812	310 973 455
Percent Removal	min max ave	46 76 63	11 59 34	47 84 72	30 78 54	14 66 44	44 72 63
Removal $\frac{lb}{10^3 ft^2 \cdot day}$	min max ave	8.6 20.0 13.1	4.4 18.7 11.1	2.4 10.0 6.7	1.0 19.3 8.8	3.4 13.6 9.0	6.5 14.8 9.8
PHENOL							
Influent[2]	min max ave	220 660 427	325 1200 624	<10 210 127	55 600[1] 355[1]	20 175[1] 165[1]	64 71 54
Effluent	min max ave	28 320 161	61 1181 414	<1 145 51	175 325 263	47 183 128	31 83 51
Percent Removal	min max ave	20 91 61	1.5 83 34	47 84 60	110 300 219	19 100 82	19 83 63

1. Estimate based on 2 data points and extrapolation from more extensive data.

2. Diluted with pond recycle.

TABLE IV

SUMMARY OF PILOT PLANT RESULTS

Parameter	PILOT PLANT A						PILOT PLANT B					
	Avg. Conc. (mg/l)[2]			Removal Percent			Avg. Conc. (mg/l)[2]			Removal Percent		
Phases	I	II	III	I	II	III	I	II	III	I	II	III
COD												
RBC Inf.	1543	3061	694	-	-	-	1219	1442	1235	-	-	-
RBC Eff.	563	2007	195	63	64	72	561	812	455	54	44	63
Pond Eff.	308	1112	298	46	45	(-)53	292	429	293	48	47	36
TOTAL:	-	-	-	79	64	57	-	-	-	76	70	76
PHENOL												
RBC Inf.	427	624	127	-	-	-	355[1]	263	219	-	-	-
RBC Eff.	161	414	51	61	34	60	165[1]	128	82	54	51	63
Pond Eff.	37[1]	139	22	73	66	57	78[1]	92	22	44	28	73
TOTAL:	-	-	-	90	78	83	-	-	-	76	65	90
FORMALDEHYDE												
RBC Inf.	147	112	47	-	-	-	94	70	83	-	-	-
RBC Eff.	25	56	14	83	50	48	37	36	42	61	49	49
Pond Eff.	20	12	10	20	79	29	10	18	15	73	50	64
TOTAL:	-	-	-	86	89	63	-	-	-	89	74	82
METHANOL												
RBC Inf.	100	323	72	-	-	-	118	181	318	-	-	-
RBC Eff.	43	205	41	57	37	43	38	144	170	68	20	47
Pond Eff.	na	124	26	-	40	37	na	115	39	-	20	77
TOTAL:	-	-	-	>57	62	64	-	-	-	>68	36	88

1. Estimate based on 2 data points and extrapolation from more extensive COD data.

2. Diluted with pond recycle.

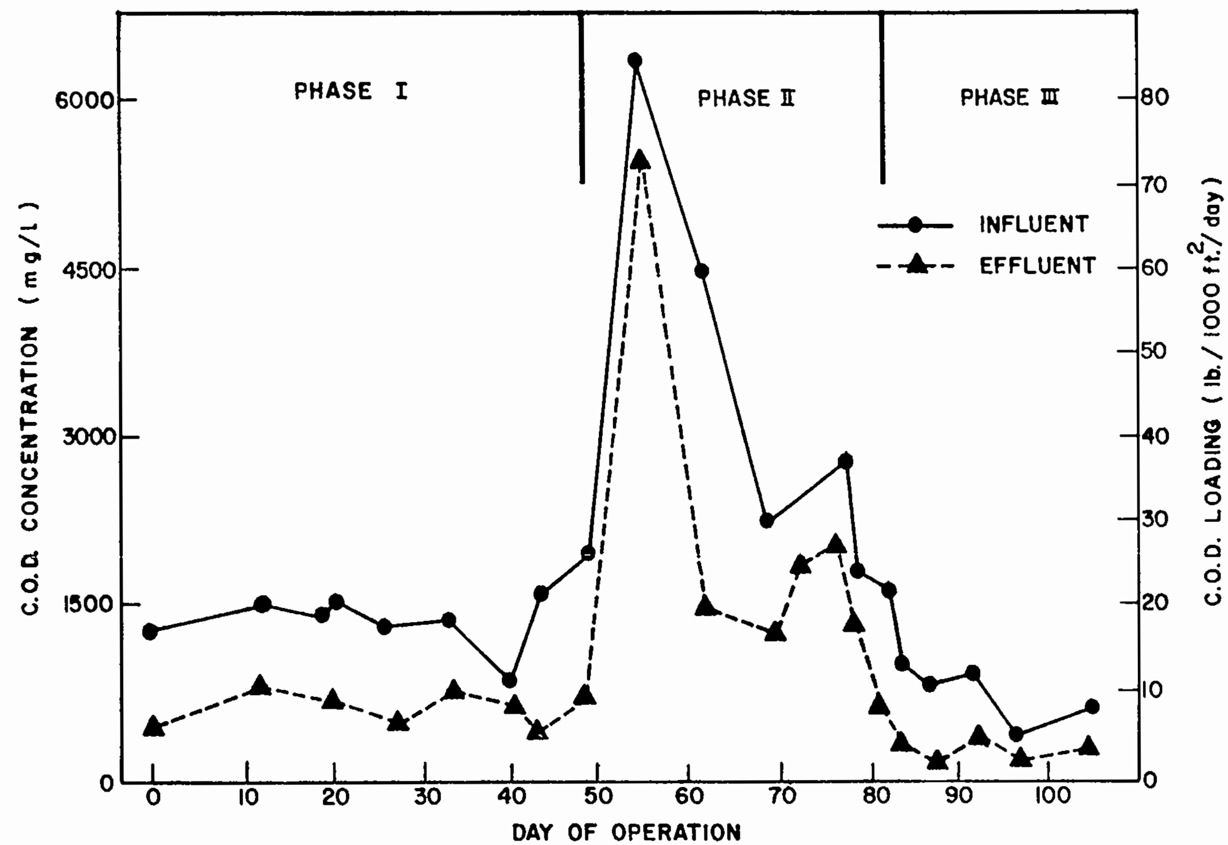

FIGURE 3 RBC UNIT A INFLUENT AND EFFLUENT C.O.D. CONCENTRATIONS

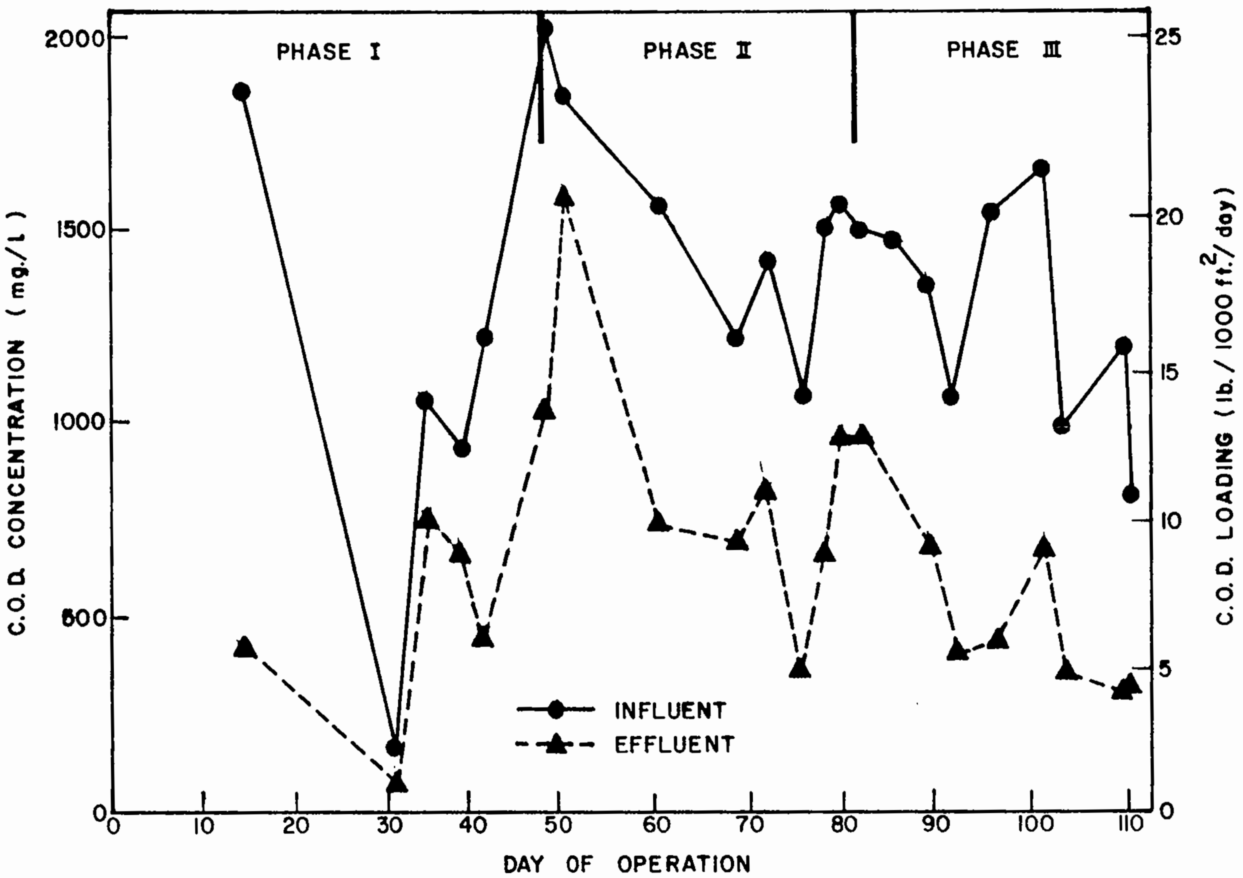

FIGURE 4 RBC UNIT B INFLUENT AND EFFLUENT C.O.D. CONCENTRATIONS

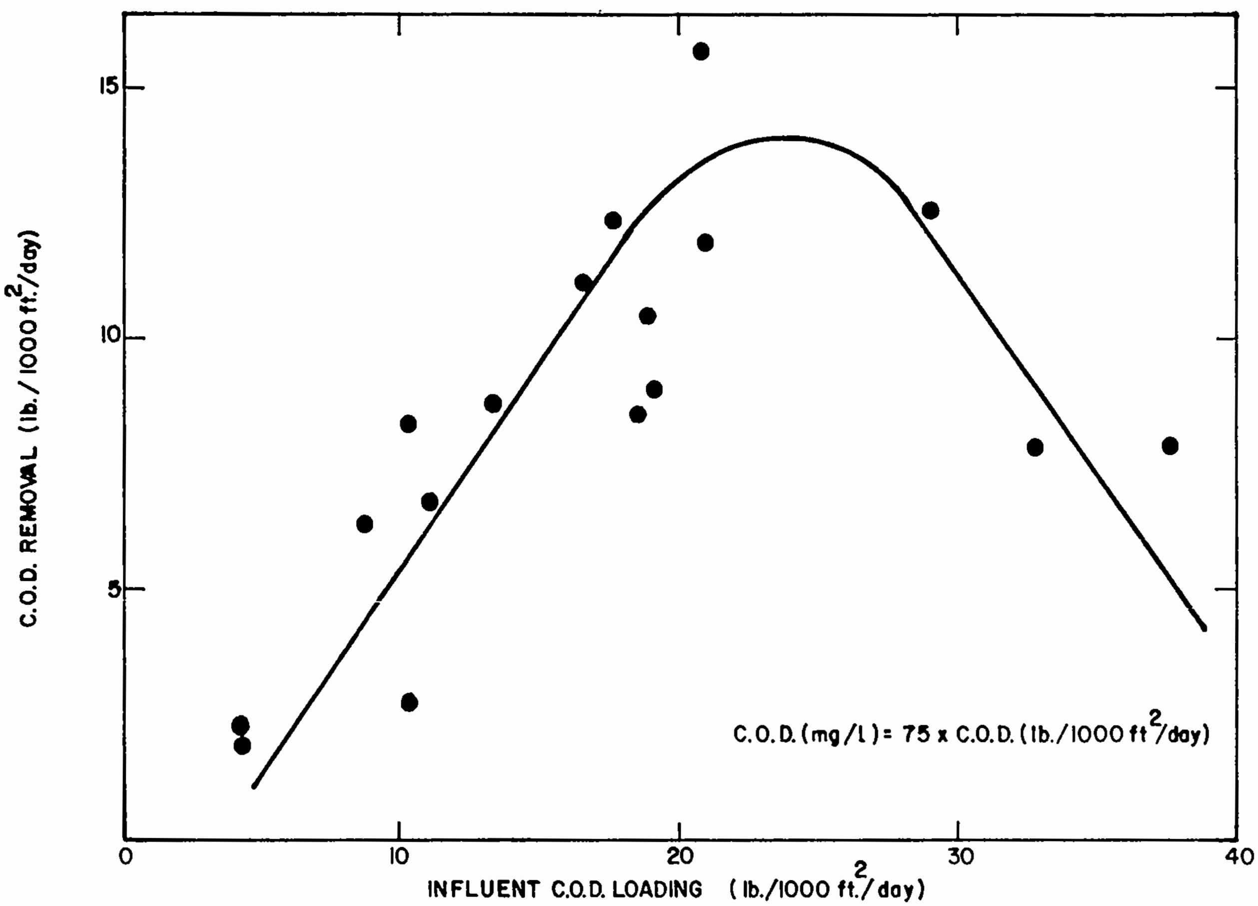

FIGURE 5 COMPARISON OF INFLUENT C.O.D. LOADING WITH REMOVAL RATES FOR RBC UNIT A

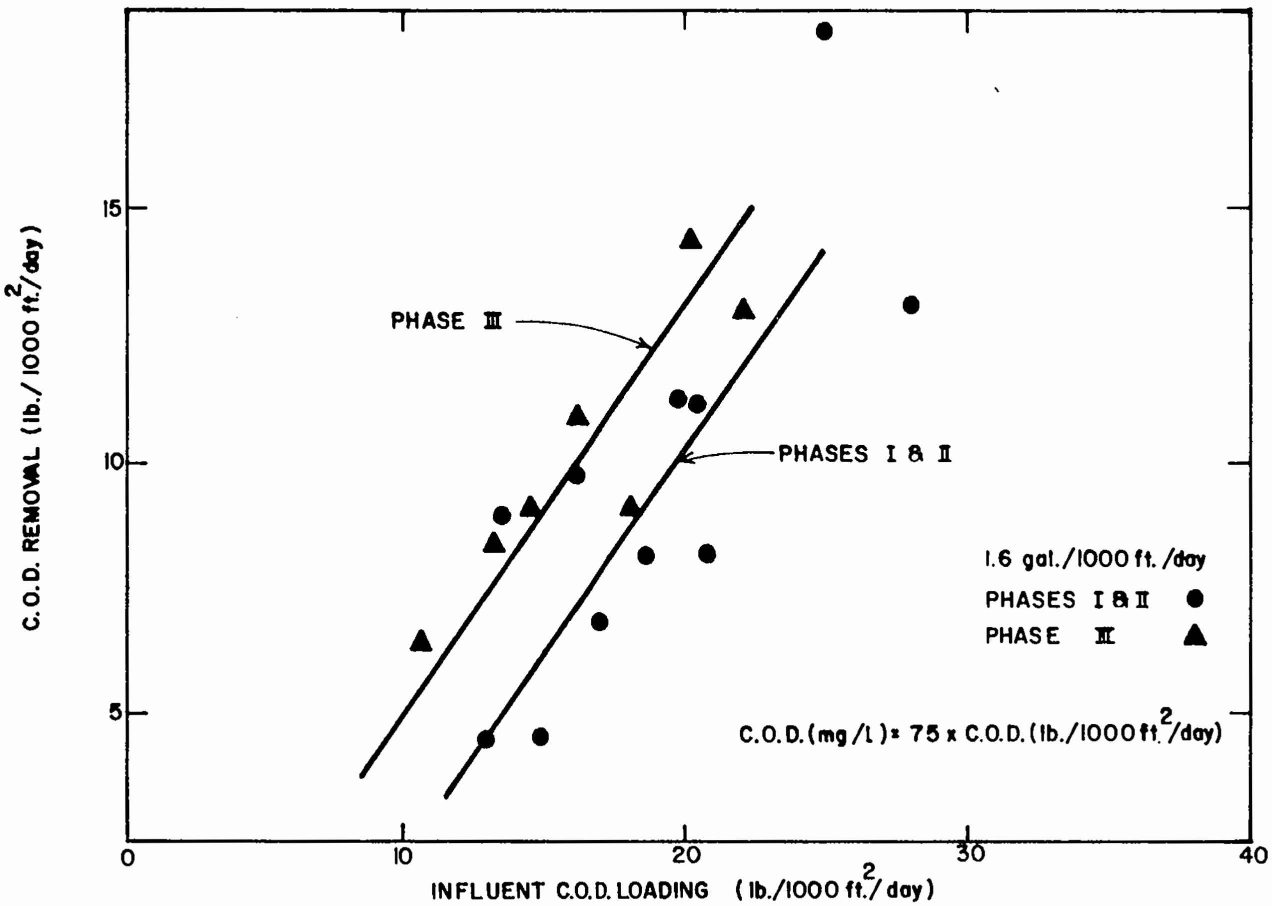

FIGURE 6 COMPARISON OF INFLUENT C.O.D. LOADING WITH REMOVAL RATES FOR RBC UNIT B

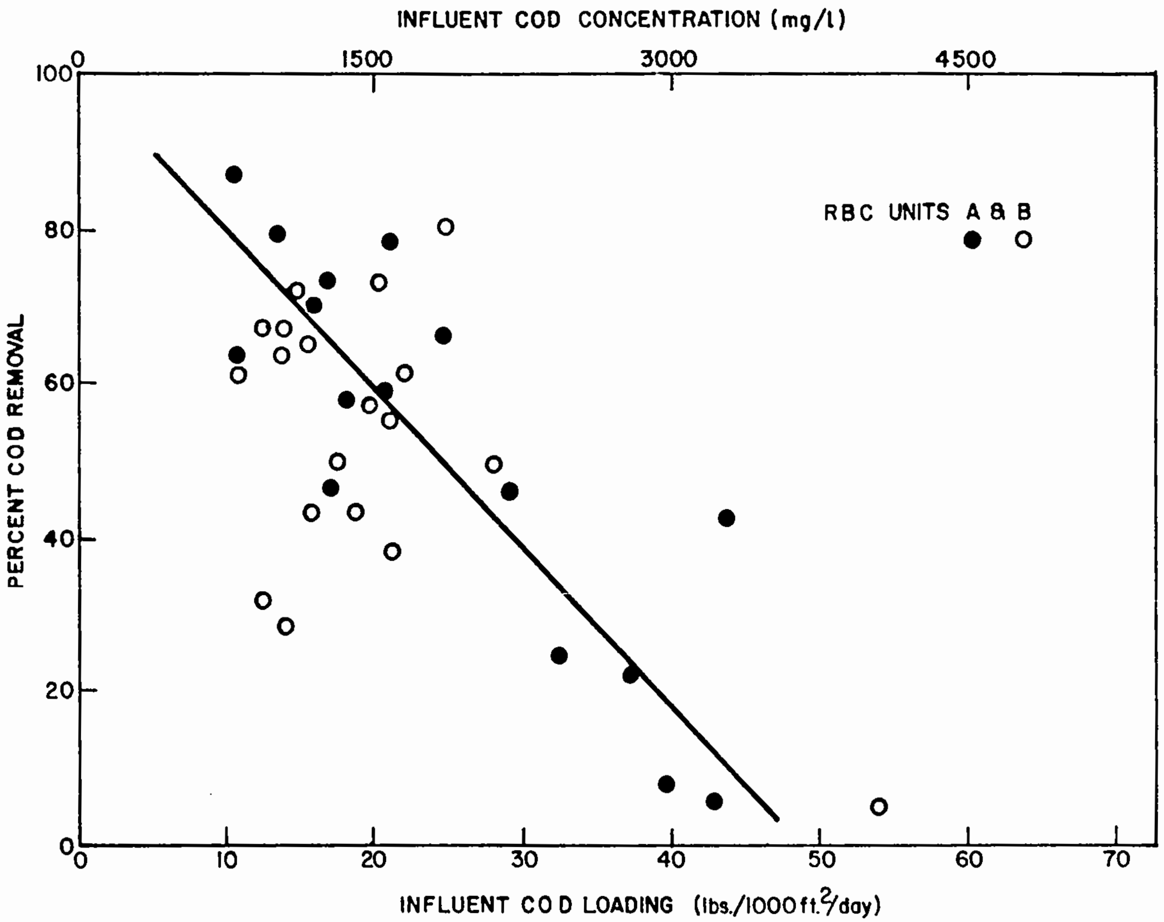

FIGURE 7 COD REMOVAL EFFICIENCY

During Phase II the influent COD concentration fed to RBC Unit A was increased to 3000 mg/l. The average percent COD removal dropped to 34 percent but the COD removal rate decreased only 15 percent from 13.1 lb/1000 ft^2/ day to 11.1 lb/1000 ft^2/day. At these high loadings phenol removals also decreased to 34 percent. During this period the biological film on the biodiscs which previously had a greyish-brown appearance, rapidly became white and filamentous. Microscopic examination of the biofilm showed that Beggiatoa spp. was predominant in the film to the extent of nearly excluding all other organisms. When the COD loadings to Unit A were again lowered in Phase III the amount of Beggiatoa spp. in the biofilm decreased as the biofilm regained its original grey-brown appearance concurrent with the increase in percent COD and phenol removals.

During this Phase II RBC Unit B continued to operate as a control to Unit A. The average COD loading to Unit B was increased 18 percent to 19 lb COD/1000 ft^2/day to approximate the COD loading applied to Unit A in Phase I. Comparing Unit B performance between Phases I and II phenol removal efficiency apparently decreased only 6 percent while COD removal efficiency decreased 19 percent; however, absolute COD removal rates were essentially unchanged at 9.0 lb COD/1000 ft^2/day. The comparison of phenol efficiency removal between Phases I and II is limited by the uncertainty level associated with the Phase I phenol data for Pilot Plant B, but the more extensive COD data indicate that the increase in loading was sufficient to significantly reduce COD removal efficiency. As a control RBC Unit B did not perform well in comparison to Unit A as there appeared to be consistent treatment efficiency differences between them.

In Phase III(a) RBC Unit A treated only the recycle stream from Pond A which had become overloaded during Phase II. As the pond contents were treated, the COD levels in the pond were reduced and the COD loadings to the RBC decreased gradually to an average of 9.3 lb/1000 ft^2/day with COD removals averaging 60 percent. It would appear that this recycle stream may have been somewhat more resistant to biological oxidation than the diluted raw PFM waste fed during Phase I because, although COD removal efficiency was 14 percent higher than in Phase I, the average COD loadings in Phase III were only 45 percent of those in Phase I.

When influent COD loading is plotted against COD removal for RBC Unit A (Figure 5) a maximum COD removal rate at approximately 14 lb/1000 ft^2/day is observed at a COD loading rate of 22-23 lb/1000 ft^2/day. For RBC Unit B (Figure 6) a maximum COD removal rate is not clearly defined because COD loadings were not increased to values high enough to establish the shape of the curve at COD loadings above 20 lb/1000 ft^2/day. Percentage COD removals for RBC Units A and B were then combined and plotted against influent COD loading as shown in Figure 7. The result is an almost linear relationship with 80 percent removals found at COD loadings of 11-12 lb/1000 ft^2/day and 20 percent removals found at COD loadings of 38 lb/1000 ft^2/day. The general relationships established between COD loadings and removals as shown in Figures 5,6, and 7 can now be used to assist in sizing an RBC to treat a PFM-waste. Various COD loadings and required removals can be plugged in to optimize the sizing and performance of the RBC unit in the overall process train.

Biological Treatment of Distilled PFM Waste

In Phases I and II it was clearly demonstrated that both RBC Units were able to remove 50-60 percent of the COD from diluted PFM waste with COD loadings of 16-20 lb/1000 ft^2/day. In Phase III(b) the impact of a distillation pretreatment step on the biodegradability of PFM waste was investigated. Here, RBC Unit B received solar distillate from raw PFM waste which had an undiluted COD of 123,000 mg/l but contained a much higher percentage of COD attributable to methanol than the raw PFM waste (Table II).

During biological treatment of the solar distillate COD removals in RBC Unit B were 63 percent at influent COD levels that were almost the same as in Phase I and only slightly lower (17 percent) than in Phase II. The higher percent COD removal in Phase III (63 percent) compared to Phases I and II (54 percent and 44 percent) indicate that the distilled PFM waste was somewhat easier to treat. Also a plot of COD removal versus influent COD loading in Figure 6 seems to indicate that absolute removal rates in Phase III were higher than in Phases I and II. Phenol removal in RCB Unit B was also higher in Phase III than in Phases I and II by an amount that would largely account for the higher COD removal. This result suggests that distillation may remove some poorly oxidizeable phenolic compounds with the more biodegradeable phenolics being more distillable. Thus, distillation of raw PFM waste besides reducing COD loading by up to 90 percent [if the pH is adjusted to 10.5 (Table II)] may also provide a distillate slightly more amenable to biological treatment.

Formaldehyde and Methanol Removal

Problems were encountered especially prior to Phase II with the analytical methods for formaldehyde and methanol. The colorimetric method of Jephcott was unsatisfactory and non-reproducible in our lab, even with standard solutions, so that a gas chromatographic method had to be adopted in the study. Data for formaldehyde and methanol are presented in Table IV together with the COD and phenol data for both pilot plants and all phases of operation.

Based on only 2 measurements, formaldehyde removal by RBC Unit A in Phase I was 83 percent; in Phases II and III formaldehyde removal decreased to 50 percent for both Units A and B. The only supportable conclusion that can be made is that formaldehyde removals of 50 percent were obtained independent of influent formaldehyde concentrations between 72 and 323 mg/l. This percentage removal rate is lower than some data reported in the literature for activated sludge. It is speculated that some removals of PFM in activated sludge are probably due to stripping by aeration.

In general, methanol removals in Phase I were 60 percent and about 40 percent in Phases II and III.

Pond Performance

Table IV presents COD and PFM results for Recycle Ponds A and B, and overall removals for the RBC units and the recycle ponds together. The pond data should not be considered representative of full-scale facultative pond performance because the pools were shallow and had a low hydraulic residence time. In particular during the heavy loading of RBC Unit A during Phase II, the Recycle Pond A was highly overloaded so that in Phase I only pond effluent was recycled to RBC Unit A. RBC Unit B was not so heavily loaded as RBC Unit A so that the performance of Recycle Pond B did not change markedly during the pilot program. The pond recycle retention times were 20 days so that changes in pond performance between various phases of operation were difficult to distinguish. On the average, it can be said that these shallow pools removed approximately 45 percent of the influent COD (largely through settling since very little algae grew during the winter months of operation) to produce an overall pilot plant removal of 70-75 percent. Pond phenol removalyle Pond A was 60-70 percent and 30-70 percent in Recycle Pond B to produce an overall plant phenol removal of 80-90 percent in Pilot Plant A and 70-90 percent in Pilot Plant B.

Formaldehyde removals in the Recycle Ponds varied substantially about an average removal of approximately 60 percent. Overall pilot plant removal for formaldehyde was generally greater than 80 percent. No methanol data was available for the Recycle Ponds in Phase I but in Phases II and III the Recycle Ponds averaged 40 percent methanol removal to produce an overall pilot plant methanol removal of greater than 60 percent.

Degradation of PFM Waste in RBC

Effluent COD values reported for both RBC units are for filtered samples in order to trace the removal of soluble COD. Unfiltered effluent COD values were also obtained throughout the pilot study to evaluate the relative percentage of influent COD biologically oxidized and converted to organic material. The unfiltered COD results for both RBC units showed that at average influent COD levels of 1500 mg/l, 32 percent of the influent COD was oxidized and 22 percent was converted to suspended COD (Table V). The suspended COD in the RBC effluent will then settle in the facultative pond and be anaerobically digested while the soluble COD will need to be treated aerobically in the pond. The pond soluble COD will be treated further by recycle of the pond effluent through the RBC.

TABLE V

SOLUABLE COD REMOVAL AND CONVERSION TO BIOMASS

	minimum	maximum	average
Influent COD, unfiltered (mg/l)	1030	1566	1445
Effluent COD, unfiltered (mg/l)	375	1906	982
Effluent COD, filtered (mg/l)	345	1818	715
COD removal (%)	1	76	32
COD conversion to biomass (%)	1	66	22

Air Quality

Considerations of State of California Air Resources Board and OSHA regulations made the emission of phenol and formaldehyde into the atmosphere from the RBC units a major concern. Air in the enclosed main contactor section of the RBC units was sampled for both phenol and formaldehyde utilizing a micro-impinger, when influent COD concentrations were at 1500 mg/l. No formaldehyde was detected in 1 m^3 of sampled air; phenol levels were 0.013 mg/m^3 of sampled air. By comparison, the TLV (Threshold Limit Value) for phenol established by OSHA is 10 (19 mg/m^3) and the NIOSH recommended limit for phenol is 20 mg/m^3.

RBC Performance

The RBC part of the pilot plant performed well overall removing 60 percent of the phenol and COD at moderate loadings i.e., influent phenol levels of 300-400 mg/l and COD levels of 1200-1500 mg/l. The RBC's demonstrated excellent treatment stability and showed no toxic effects even at very high influent phenol levels (over 1000 mg/l). By comparison to the literature values shown in Table I, however, the pilot plant RBC units did not achieve consistently the 85-95 percent phenol removals reported. It is possible that RBC's are not as efficient at treating PFM wastes as activated sludge or trickling filters; but, it should be noted that the aeration process in activated sludge and the cascading of sewage in the trickling filter dosing process certainly strips volatile organics from the water adding to the overall removal efficiencies. No data were found to indicate the level of phenol or formaldehyde in the air over activated sludge or trickling filter units treating high levels of these wastes and thus no relative weight can be given to the contribution of this process in the overall efficiency of treating PFM waste. In this study in an enclosed RBC unit no formaldehyde and less than 0.1 mg/m^3 of phenol could be detected in the enclosure, indicating that very little stripping of these organics occurred in the enclosed RBC pilot system.

The temperature of the pilot system also probably played a significant role in that optimum pilot plant performance can not be expected when the temperature is only 13 or 14°C. If the operation of the pilot plant had been continued through warmer temperatures biological activity would have increased and removal efficiences would no doubt have improved somewhat, perhaps to 80 percent but probably not higher unless loading rates were reduced significantly from the average values used. Despite the apparent inefficiencies of RBC's in treating PFM waste compared to activated sludge or trickling filters, it is likely that RBC's are actually equally efficient if all conditions were equal and air stripping of organics was included in an overall mass balance. This could only be proven though by comparing parallel pilot systems.

SUMMARY AND CONCLUSIONS

Economical and environmentally acceptable disposal of high strength phenol-formaldehyde waste (PFM waste) from resin manufacturing is a major problem. Phenol is on the EPA list of priority pollutants and local sewer use ordinances are generally stringent with respect to phenol, usually allowing less than 1 mg/l. With the phenol level at 60,000 mg/l in PFM waste, conventional biological treatment systems even designed for 99 percent removals can not achieve 1 mg/l effluents. Thus a no discharge approach was selected as the most appropriate solution. Currently PFM waste is disposed of by incineration and, while this method results in zero-discharge, its economic attractiveness decreases with the increasing cost of energy.

The selection of an alternative zero discharge treatment process was largely influenced by the local climate, availability of land, and the volume of waste to be treated - less than 10,000 gpd. The biological system selected consisted of a rotating biological contactor (RBC) followed by an evaporative facultative ponding system. A solar distillation unit was also evaluated as a pretreatment step for the PFM waste. The pilot program demonstrated that a strong phenol-formaldehyde-methanol waste from resin manufacturing or its solar distillate could be treated by a combination of rotating biological contactor and recycle ponding system.

The pilot system as a whole and the rotating biological contactors in particular performed well but less efficiently that expected when compared to literature values for activated sludge and trickling filters. However, the pilot program was carried out during winter conditions and the resulting low temperatures in the small pilot system undoubtedly reduced bacterial activity and thus prevented optimum removal rates from being attained. Furthermore, the extent to which air stripping contributed to the literature removal efficiencies quoted is not known. The RBC's did operate effectively throughout the study period under varying climatic and loading conditions and exhibited excellent stability in withstanding periodic shock loadings. Percentage and absolute COD removal rates for the RBC's were found to be a function of the influent COD loading and this data provided sufficient information to allow the sizing of full-scale rotating biological contactors for the treatment of the PFM waste involved.

The choice of treatment system and the relative sizing of the components would depend mainly upon the relative availability of land area, capital, solar insolation and process waste heat. The solar distillation of the PFM waste has the advantage of producing a distillate that is 15-20 percent more biodegradable and of reducing the COD loading from 40-90 percent depending upon the level of pH adjustment of the raw PFM waste. These benifits of solar distillation must be traded-off against the capital and operating costs of the solar still, as well as the decrease in sizing of the RBC and ponding system. Still size and hence cost is dependent most strongly upon solar insolation values and upon the amount of of process waste heat that could be made available to supplement solar energy.

The relative sizing of the RBC/pond system depends upon the land available for ponding and the waste stream volume if evaporative and zero-discharge are treatment goals. The overall sizing of these units depends on whether or not solar distillation is used and to what extent pH adjustment is made during solar distillation. After detailed cost-benifit analysis comparing all of the above variables in a capital and operating cost matrix and with the goal of an evaporative-zero discharge system, it was determined that the least cost system for capital as well as O & M was a 1.1 acre still, operated at pH 9, followed by 2 RBC units with 100,000 ft^2 of contactor surface each, and four 1-acre ponds. Two of these ponds would be deep facultative primary ponds, one would be a second pond and one just for evaporative and blowdown.

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ENERGY RECOVERY FROM ANAEROBIC ROTATING BIOLOGICAL CONTACTOR (AnRBC) TREATING HIGH STRENGTH CARBONACEOUS WASTEWATERS

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The continuing development of ever more stringent industrial effluent discharge standards coupled with rapidly escalating energy costs encourage the development of new approaches for the treatment of high strength carbonaceous wastewaters. Aerobic treatment designs for high strength wastewaters are usually constrained by energy intensive mixing, oxygen transfer or sludge handling processes. Previous experimental work with overloaded conventional rotating biological contactors (1, 2, 3) led to the concept of the anaerobic rotating biological contactor (AnRBC) for the low cost treatment of high strength wastewaters. Based on the bench scale pilot plant data presented below, it appears that the AnRBC process can attain high quality effluents with relatively low energy inputs and can even result in net energy yield under appropriate loading conditions.

AnRBC CONCEPT

Conceptually the AnRBC is similar to conventional aerobic rotating biological contactors in that microorganisms become attached to and grow on rotating discs that are partially submerged in the wastewater as shown schematically in Figure 1. Groups of discs, separated into sequential compartments called stages, are partially immersed in the wastewater and rotated continuously both to provide mixing within each stage and to facilitate product gas transfer to the anoxic atmosphere maintained above the water surface. Adjacent stages are separated by baffles to minimize short-circuiting. The flow is passed from stage to stage through holes in the baffles below the waterline.

However, unlike conventional RBC units, the AnRBC is enclosed in an airtight housing with an anoxic atmosphere maintained above the liquid. Also, the depth of disc submergence is substantially greater than in conventional aerobic units. Microorganism attachment to the rotating surfaces provides long mean cell retention times in the reactor, which in turn encourages the development of the slow growing methanogenic bacteria responsible for the conversion of low molecular weight compounds to methane. Sloughed excess biomass is carried from stage to stage through openings in the baffles and leaves the reactor in the effluent. Conceptually, the AnRBC process couples the advantages of the short hydraulic detention times typical for the fixed film horizontal flow RBC process with the high strength, carbonaceous degradation capabilities of anaerobic systems.

AnRBC PILOT PLANT AND EXPERIMENTS

The primary purpose of this study was to ascertain the conditions whereby anaerobic non-methanogenic and methanogenic microorganisms could successfully be grown on rotating disc surfaces. Secondary purposes included (a) the development of the models for predicting the removal of soluble organic substrate as a function of both wastewater organic strength and feed flow rate, and (b) the preliminary evaluation of process net energy requirements or yields.

To accomplish these goals, two identical four-stage reactors were constructed from plexiglass as shown in Figure 1. The inside diameter and disc diameters were 13.97 cm (5.50 in.) & 12.70 cm (5.00 in.), respectively. The 1.270 cm (0.500 in.) horizontal shaft was supported by external end bearings and was rotated by attachment through a pulley and belt arrangement with a variable speed drive motor. Successive stages were separated by fixed baffle plates with three 1.91 cm holes below the waterline for the passage of solids and the wastewater carrier stream from stage to stage. Ten 0.318 cm thick discs were contained in each stage, providing a total reactor disc surface area of 1.013 m^2 (10.908 $ft.^2$). The first stage of each reactor was preceded by a small mixing chamber that contained a flat bladed impeller to distribute the influent feed evenly through the numerous holes in the baffle separating the mixing chamber and the first stage. The water level in the reactor was controlled by a dynamic head tube resembling a vented inverted siphon on the effluent line. Valved liquid sampling ports permitted grab sampling from each stage as well as from the effluent. Gas collected in each stage was vented through a common manifold to a wet test gas meter.

The AnRBC pilot plants were operated with about 70 percent of the disc area submerged for the studies reported herein. The reactor liquid volume for these conditions was about 5.27 liters. Both reactor systems were housed in a controlled temperature room with the internal reactor temperatures maintained at 35 $\pm$ 0.5°C.

Experimental Parameters

The feed stock used for these experiments is similar to that used in earlier loading studies with aerobic RBC pilot plants (1, 2, 3). The synthetic wastewater contained dissolved sucrose as the sole organic carbon source and was mixed with the soluble inorganic constituents shown in Table 1. Ammonium salts served as the sole nitrogen source for these experiments to avoid denitrification conditions. These soluble wastewater constituents are biochemically similar to those that might be expected to result from food, bottling and some organic chemical processing industries.

Soluble organic carbon was measured with a Dohrman DC50 Organic Carbon Analyzer on samples that had been filtered through a 0.45 micrometer Millipore HA filter. Total alkalinity at a pH end-point of 3.7 was determined as suggested in Standard Methods (4). Volatile acid akalinity (VAA) was measured by the procedure developed by DiLallo and Albertson (5). Because the analytical results of this test are not exact, the volatile acid alkalinity results reported below should only be interpreted qualitatively. Since influent organic substrates were entirely soluble, effluent solids (measured as total residue, volatile residue, total nonfilterable residue, and volatile nonfilterable residue as suggested in Standard Methods) (4) represent either solids grown and sloughed in the reactor or solids precipitated in the reactor. Product gas generation rates were measured with a wet test gas meter and gas composition was determined with a gas chromatograph.

A total of eight experiments with variable loading conditions was conducted and they can be considered portions of a single factorial experiment as indicated in Table 2. Loading conditions for each steady state experiment are multiples of the influent flow rate (H) and the influent TOC concentration (F) used in Experiment 1. Thus, the 3F-2H loading notation shown in Table 2 for Experiment 8 indicates that the influent feed concentration (C_i) was about three times stronger than that used in Experiment 1 and the flow rate (Q) was twice as large.

Operations

Prior to start-up, tracer study experiments indicated that these four-stage reactors can be described hydraulically as four complete-mix reactors in series. The initial start-up procedures have been described elsewhere (6). The reactors were operated for a given loading condition until effluent soluble TOC concentrations and gas production rates varied by less than $\pm5\%$ for three successive days. Operation for a minimum of three weeks following a step change in loading was required to achieve these quasi-steady state operating conditions. Intensive sampling over the next day was used to obtain the data reported below. Continuous operation was extended for 218 days until an O-ring failed and reactor fluids began leaking from a bearing. Experimentation was discontinued at this time and both reactors were dismantled for inspection. Microbial solids were observed to coat all disc surfaces and decreased in thickness from the first through the fourth stages, with only a thin attached film layer present on the fourth stage discs.

RESULTS

Figures 2, 3, and 4 display the general data patterns seen as a result of these experiments. The influent volatile acid alkalinity (VAA) was nearly negligible relative to the values observed in the reactors for all feeding conditions. Even though stage data points have been connected with straight lines, it should be remembered that each stage is acting as a complete-mix reactor and that there are hydraulic and reaction discontinuities between adjacent stages.

Figure 2 illustrates that the VAA generated in the first stage was accompanied by simultaneous minor depressions of pH and carbonate alkalinity. About 80% of the TOC removal occurred in the first stage. Both TOC and VAA were further reduced in succeeding stages while the pH and alkalinity increased in the downstream direction. Note that the fourth stage did not contribute to TOC removal for this loading condition and apparently the hydraulic residence time in the reactor could have been reduced from 17.5 to about 13 hours with no loss in soluble TOC removal performance.

The results from Experiment 6, shown in Figure 3, are similar even though the organic loading rate has been increased by a factor of four and the reactor hydraulic detention time reduced to one-half that of Experiment 1 conditions. For this loading situation, the first stage removed about one-half the influent TOC. However, volatile acid production was more significant in the first stage. Note also that the fourth stage TOC and VAA concentrations were significantly higher, and that the overall soluble TOC removal was reduced from the 96% found in Experiment 1 to about 79% in Experiment 6. It is probable that additional stages would have resulted in more complete TOC removal.

Figure 4 illustrates the effects of a high loading rate and short hydraulic detention time on system performance. The organic loading rate in Experiment 4 was increased by a factor of about 8.5 compared to Experiment 1, with the detention time being reduced to one-eighth that of the original conditions. The sharp drop in pH and alkalinity along with the high production of VAA suggests that acid fermentation occurred at a rate that overwhelmed the slower-growing methanogenic bacteria in the system. Only 46% of the soluble TOC was removed under these loading conditions. The increase in VAA in the fourth stage may suggest that the fermentation process was inhibited by high concentrations of soluble substrate in the first three stages.

When the data from these experiments are arrayed as shown in Figures 5 and 6, the effects of influent feed concentration (C_i) and flow rate (Q) on performance can be observed. Figure 5 shows that increasing the flow rate (reducing the hydraulic detention time [Θ]) for constant C_i conditions results in higher TOC concentrations at most points within the reactor. Under relatively low mass loading conditions (Experiments 1 and 2), effluents contained relatively low volatile acid concentrations. At the higher mass loading rate (Experiments 3 and 4), the effluents contained relatively high volatile acid alkalinity concentrations and the product gas had a lower methane content. Thus, an increase in the influent flow rate has the apparent effect of moving methane fermentation downstream in the reactor. Alternately, these data suggest that there might be a critical reactor hydraulic detention time required to sustain effective methane fermentation. Based on these data, it appears that this critical time period is between 4.4 and 8.8 hours. Of special interest is the fact that the first stage substrate removal rates shown in Figures 5 and 6 are substantially higher than those reported for conventional aerobic RBC systems.

The data shown in Figure 6 can be used to compare TOC removal as a function of influent substrate concentration for constant flow rate conditions. As expected, the system requires more stages (area) to achieve a specific residual organic effluent concentration with increasing C_i values. Again, the first stage organic removal rates are substantially higher than those reported for conventional RBC treatment.

Mass loading and removal data are shown in Figure 7. Up to a loading of about 22 g TOC per day (21.7 g TOC per m^2 -day), soluble TOC removal appears to be independent of either influent concentration or flow rate and averages 95+ percent removal. Although a smooth curve could have been passed through all the data points, a sharp discontinuity in soluble TOC removal exists near this limiting mass loading and it is evident that both additional increments of TOC removal and the overall percent removal are highly dependent on mass loading conditions.

Due to the small flow rates, gas production was difficult to monitor with the available equipment. However, as shown by the dashed line in Figure 8, a good linear correlation exists between soluble TOC removed and total gas production. When data from Experiments 4 and 8 are omitted because high VAA effluent concentrations indicate incomplete reactions, an even better correlation exists (solid line). Hence, total gas production can be conservatively estimated as 1.76 m^3/kg TOC removed. Methane and carbon dioxide were present in about even quantities for Experiments 1, 2, 5 and 7. The carbon dioxide content increased from 54% in Experiments 3 and 6, to about 60% in Experiments 4 and 8. These, along with the VAA, pH, and TOC data, provide evidence that methanogenic activity was unable to proceed to completion under the higher loading conditions.

Effluent solids production is a critical consideration in the design of biological treatment systems. As expected, gross solids production was dependent on the mass of TOC removed as shown in Figure 9, with more net solids being generated for the higher loading conditions. The observed yield (Y_{obs}), defined as the mass of solids generated per mass of soluble TOC removed, appears to be related to the logarithm of the loading factor (ΘC_i) as shown by the solid line in Figure 10. Nearly as good a correlation and probably a more meaningful relationship was found when Y_{obs} was compared to the logarithm of $\Theta\Delta C$ as shown by the dashed line in Figure 10. It seems rational to expect lower observed cell yields under conditions of long detention times and more complete carbon conversion to stable end-products.

PREDICTIVE MODELS

When stage TOC data from the eight experiments are displayed on a semi-logarithmic plot as shown in Figure 11, apparent pseudo-first order relationships between soluble TOC remaining and time are observed with slopes and intercepts dependent on C_i. For a given influent substrate concentration C_i, TOC remaining can be approximated by

$$\ln C = \ln C_i - K_f T, \qquad (1)$$

where C is the soluble TOC (mg/l), K_f is the pseudo-first order rate constant (hr^1), and T is the hydraulic residence time (hr). When the slopes from the three lines are compared with C_i, a strong correlation between the pseudo-first order rate constant (K_f) and C_i is observed (Figure 12). Both linear and logarithmic relationships provide good predictions for K_f over the range of C_i data evaluated. Using the empirically determined linear relationship

$$K_f = 0.3744 - 7.96 \times 10^{-5}\, C_i \qquad (2)$$

for simplicity with Equation 1 yields

$$\ln C = \ln C_i - (0.3744 - 7.96 \times 10^{5} C_i)T. \qquad (3)$$

As shown in Figure 13, Equation 3 provides reasonably good predictions for C when compared against the measured data. The data scatter is suprisingly small when one considers that a simple pseudo-first order expression is being used to describe the results of a complex series of anaerobic reactions that were only partially complete for some of the loading conditions employed in these studies.

Of course, the empirically determined constants indicated in Equations 2 and 3 are only valid for the experimental parameters employed in this study. One would expect these constants to change as a function of substrate type, disc diameter, rotational speed, immersion depth, etc. Additional studies will be required to develop these relationships. However, a pseudo-first order approximation of the form indicated in Equation 3 may prove useful for some design situations.

A more useful model for applying the AnRBC concept to design situations is based on an empirical model originally developed by Schroeder (3) that considered mass flux concepts for the removal of organic material on the face of an RBC disc. Friedman, Woods, and Wilkey (2) and later Friedman, Robbins, and Woods (3) applied Schroeder's model to aerobic RBC systems by considering each RBC stage as a completely mixed reactor. They proposed the mass of organic material removed per stage could be described by

$$M_n = \frac{K\ C_i^2}{k' + C} \qquad (4)$$

where:

- M_n = mass or organics removed per stage per unit time, mass/time
- K = a removal rate constant, volume/mass-time per stage
- C_i = substrate concentration entering the stage, mass/volume
- k' = a constant.

If in the first stage $C_i >> k'$, then Equation 4 can be closely approximated by

$$M_n = KC_i\,. \qquad (5)$$

A K value was calculated for each set of loading conditions by using Equation 5 and first stage data for each of the eight experiments (Table 3). Using these K values, k' was calculated with Equation 4 for each experiment using the last stage data for which significant TOC removal occurred. This results in the average k' value of 0.0426 g/l. Despite significant scatter for individual k' values, this average value for k' was used for the succeeding calculations since k' has little effect in the early stages where most soluble TOC removal occurs.

Under identical mass loading conditions, first stage mass removal rates vary significantly as seen by comparing M_n and K values for Experiments 2 and 5 as well as Experiments 4 and 6 in Table 3. The observed differences are dependent on both the stage hydraulic detention time (Θ_s) and the influent organic concentration C_i. The natural logarithm of the product of these two parameters, $\ln(\Theta_s C_i)$, has been termed the "loading factor" and appears to be related to K as shown in Figure 14. A straight line provides a reasonably good fit of these data considering that pH and VAA values varied significantly for each first stage loading condition. Again, future studies will probably show K to be a function of disc diameter, rotational speed, substrate type, and other operating parameters as well as the loading factor. However, since a reasonable relationship does appear to exist between K and $\ln(\Theta_s C_i)$, trade-offs between hydraulic detention time (flow), influent concentration, and desired effluent concentration can be predicted with the use of an empirically determined relationship such as that shown in Figure 14.

In order to verify this empirical model, the measured soluble TOC data from the eight experiments were compared to calculated model values for each set of loading conditions. The linear relationship shown in Figure 14 was used to predict the K value for each set of loading conditions. Equation 4 was next used with the resulting K value and k' average to predict the mass removal for each stage. The C_i for the following stage was back calculated from the mass remaining in the previous stage. This iterative process was used to calculate the results shown in Figures 14, 15, and 16. The experimental and predicted values agree closely for the loading conditions investigated.

ENERGY CONSIDERATIONS AND PROCESS SCALE-UP

Estimating anticipated prototype performance based on bench scale data obtained under ideal laboratory conditions is subject to numerous errors and unforeseen problems. However, the following two examples will describe the net energy requirements and yield that may be obtained with the AnRBC designed to meet specific effluent requirements. It is assumed that conventional twelve foot diameter plastic discs will be used for the prototype installation. Based on torque measurements made with clean, full size discs with an additional 45 percent allowance for the mass of microorganisms adhering to the discs as well as motor and gear box losses, it is assumed that 4 HP will be more than sufficient to rotate a shaft carrying 9300 m^2 (100,000 ft^2) of media. As will be seen later, net energy requirements are relatively insensitive to motor horsepower requirements. The following analysis also ignores external pumping requirements, product gas compressor requirements and heat losses from the reactor. This latter loss is assumed to be offset by a heat exchanger operating between the AnRBC effluent and the influent streams. Because the direct scale-up of bench scale aerobic RBC data has been previously shown to lead to poor prototype design[3], the effective area used in three examples has been reduced by 25 percent from 9300 m^2 to 6975 m^2 per shaft.

Example 1 - High Quality Effluent Required (95 percent Removal)

Using the critical loading of 21.7 $g/m^2/day$ from Figure 7 as a basis for design, the mass of TOC that will be removed per day can conservatively be estimated as

$$21.7 \text{ g/m}^2\text{/d} \times 9300 \text{ m}^2 \times 0.75 \times 0.95 = 143{,}800 \text{ g TOC day.}$$

Using the gas production rate indicated by Figure 8 and a value of 50 percent methane in the product gas yields

$$V_{CH_4}\text{/day} = 143.8 \text{ Kg TOC/day} \times 1.76 \text{ m}^3\text{/Kg TOC} \times 0.5 = 126.5 \text{ m}^3\text{/day } CH_4.$$

Therefore, the potential energy, E_p, available from this methane is

$$E_p = 126.5 \text{ m}^3\text{/day} \times 35{,}800 \text{ KJ/m}^3 = 4.529 \times 10^6 \text{ KJ/day.}$$

Assuming that 79 percent of the methane energy content can be converted to useful heat, the net available energy becomes 3.57 x 10^6 KJ/day/reactor.

Energy input to the reactor will be required to turn both the discs and to heat the wastewater to a temperature of 35°C. The 4 HP will require an input of about 258,000 KJ/day, far below the energy equivalent of the methane produced by the reactor. However, the energy required to heat the influent wastewater to a reactor operating temperature of 35°C is dependent on both the influent temperature and the concentration of biodegradeable TOC. For this design example, the product of the flow rate and influent TOC concentration is

$$21.7 \text{ g/m}^2\text{/day} \times 9300 \text{ m}^2 \times 0.75 = 151{,}358 \text{ g/day.}$$

Thus, an influent concentration of 2000 mg/l yields a flow rate of 75,679 l/day. The minimum energy (Q) required to heat this flow for a 20° temperature increase is

$$Q = 75{,}679 \text{ Kg/day} \times 4200 \text{ J/Kg-°C} \times 20 \text{ °C} = 6..357 \times 10^6 \text{ KJ/day}$$

Figure 18 demonstrates the interactions between influent substrate concentration, required temperature change and net energy requirements. The solid portions of the curves represent the region where experimental evidence is available. The dashed portions of these curves represent projections for a well buffered wastewater. The vertical difference between any point on a curve and the 3.6 x 10^6 KJ/day line represents the net energy yield or input requirement for the reactor. The breakover points for energy equivalence for 5°, 10° and 20°C heating requirements appear to be around 800, 1700 and 3000 mg/l TOC, respectively. The region below the horizontal line and above the appropriate heating curve represents the potential energy recovery available from high strength wastewater treatment with the AnRBC process.

Example 2 - Lower Effluent Quality Required (68% Removal)

This example illustrates an AnRBC energy balance for conditions when a lower effluent quality is acceptable. Based on the data obtained from Experiment 6 (shown in Figures 7 and 8) and a 45 percent methane content in the product gas, it can be shown that a loading of 33.4 g TOC/m^2/day will yield about 123.8 m^3/day of methane with the same areal safety factor used in the previous example. Using the same procedure for estimating energy availability as in the previous example, gives a useable energy yield from product methane of about 3.32 x 10^6 KJ/day/reactor. Again, it is conservatively estimated that 4 HP of electrical energy will be required to rotate the shaft. Figure 19 illustrates substrate concentration and heating requirement effects on total energy requirements. This family of curves has been shifted to the right relative to those shown in Figure 18 due to the larger volumes of water that must be heated.

CONCLUSIONS

Based on the data and discussion presented above, it can be concluded that

1. The AnRBC treatment is both feasible and practical for high strength wastewaters.
2. Microorganisms, including the difficult to culture methanogenic bacteria, will readily adhere to and grow on rotating surfaces.
3. Because reaction phases or components can be separated by feed strength and flow rate adjustments, AnRBC pilot plant studies should prove useful for describing the complex interactions and removal kinetics for anaerobic systems. Similarly, AnRBC pilot plants should prove advantageous for the study of presumed toxicants on anaerobic processes.
4. The AnRBC process appears to be ideal for the pre-treatment of high strength wastewaters with energy recovery potentially available as a result of the production and utilization of methane in the product gas.

TABLE I

AnRBC FEED*

CONSTITUENT	CONCENTRATION
$C_{11}H_{22}O_{11}$	Variable
$MgSO_4 \cdot 7\ H_2O$	60.0 mg/l
KCl	120.0 mg/l
$MgCL_2 \cdot 6\ H_2O$	300.0 mg/l
$CaCl_2 \cdot 6\ H_2O$	100.0 mg/l
$CoCl_2 \cdot 6\ H_2O$	14.2 mg/l
$FeCl_2 \cdot 4\ H_2O$	90.0 mg/l
$(NH_4)Cl_2$	**
$(NH_4)_2HPO_4$	**
$NaHCO_3$	***

$$\frac{BOD_5}{TOC} = 1.72, \quad \frac{COD}{TOC} = 2.79, \quad \frac{COD}{BOD_5} = 1.62$$

* Tap water was used to provide trace nutrients.
** Amount added depended on influent carbon concentration. The N and P concentrations were maintained in excess of a C:N:P ratio of 150:15:1.
*** 3000 mg/l for Experiments 1-4, 6000 mg/l for Experiments 5-8.

TABLE 2

AnRBC LOADING CONDITIONS, 12 RPM

Flow Rate (Q), l/hr	Unit Hydraulic Detention Time (Θ), hr	Stage Hydraulic Detention Time (Θ_s), hr	Influent TOC Concentration, mg/l 1075	2320	3050
0.30	17.50	4.38	1F-1H (1)	2F-1H (5)	3F-1H (7)
0.60	8.75	2.19	1F-2H (2)	2F-2H (6)	3F-2H (8)
1.20	4.39	1.09	1F-4H (3)	--	--
2.40	2.19	0.55	1F-8H (4)	--	--

Experiment number indicated in parentheses.

TABLE 3

CALCULATED FIRST STAGE K VALUES

Experiment No.	Mn* g/day	Ci* g/1	K 1/g-day	LN (ΘCi)*
1	6.18	1.05	5.89	8.43
2	10.2	1.05	9.67	7.74
3	8.35	1.10	7.58	7.10
4	14.1	1.10	12.8	6.40
5	12.2	2.32	5.25	9.23
6	15.7	2.32	6.75	3.53
7	7.98	3.05	2.62	9.50
8	9.75	3.05	3.20	8.81

*Measured or the product of measured parameters.

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7. Schroeder, E.D., Water and Wastewater Treatment, McGraw-Hill Book Company, p. 307, 1977.

NOTE: At the time the research was conducted, S.J. Tait and A.A. Friedman were, respectively, graduate student and Associate Professor, Department of Civil Engineering, Syracuse University, Syracuse, New York. S.J. Tait is currently a Senior Research Engineer, International Paper Company, Tuxedo Park, New York. A patent is pending for the AnRBC process.

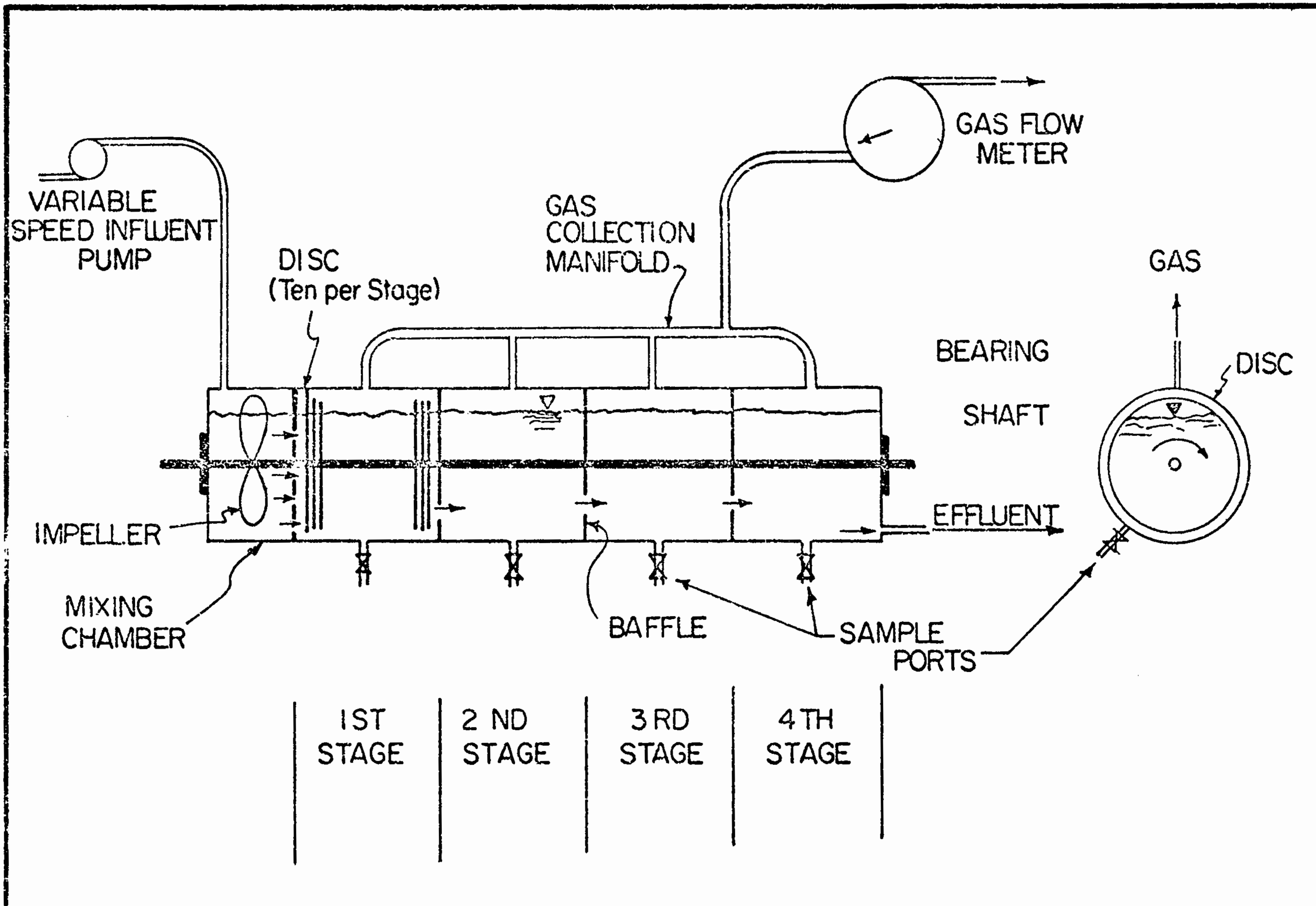

FIGURE I ANAEROBIC ROTATING BIOLOGICAL CONTACTOR

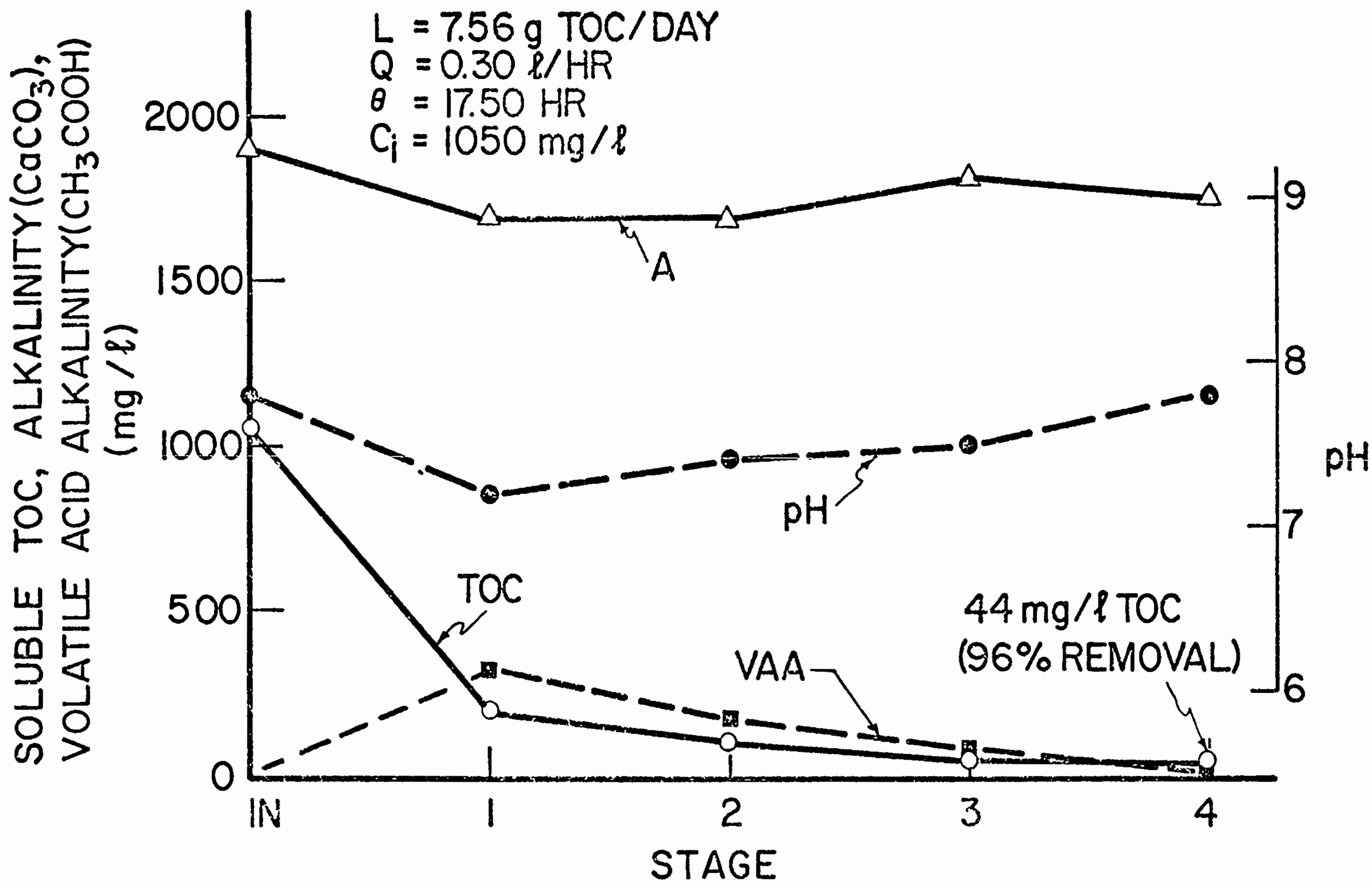

FIGURE 2 EXPERIMENT 1, IF-1H

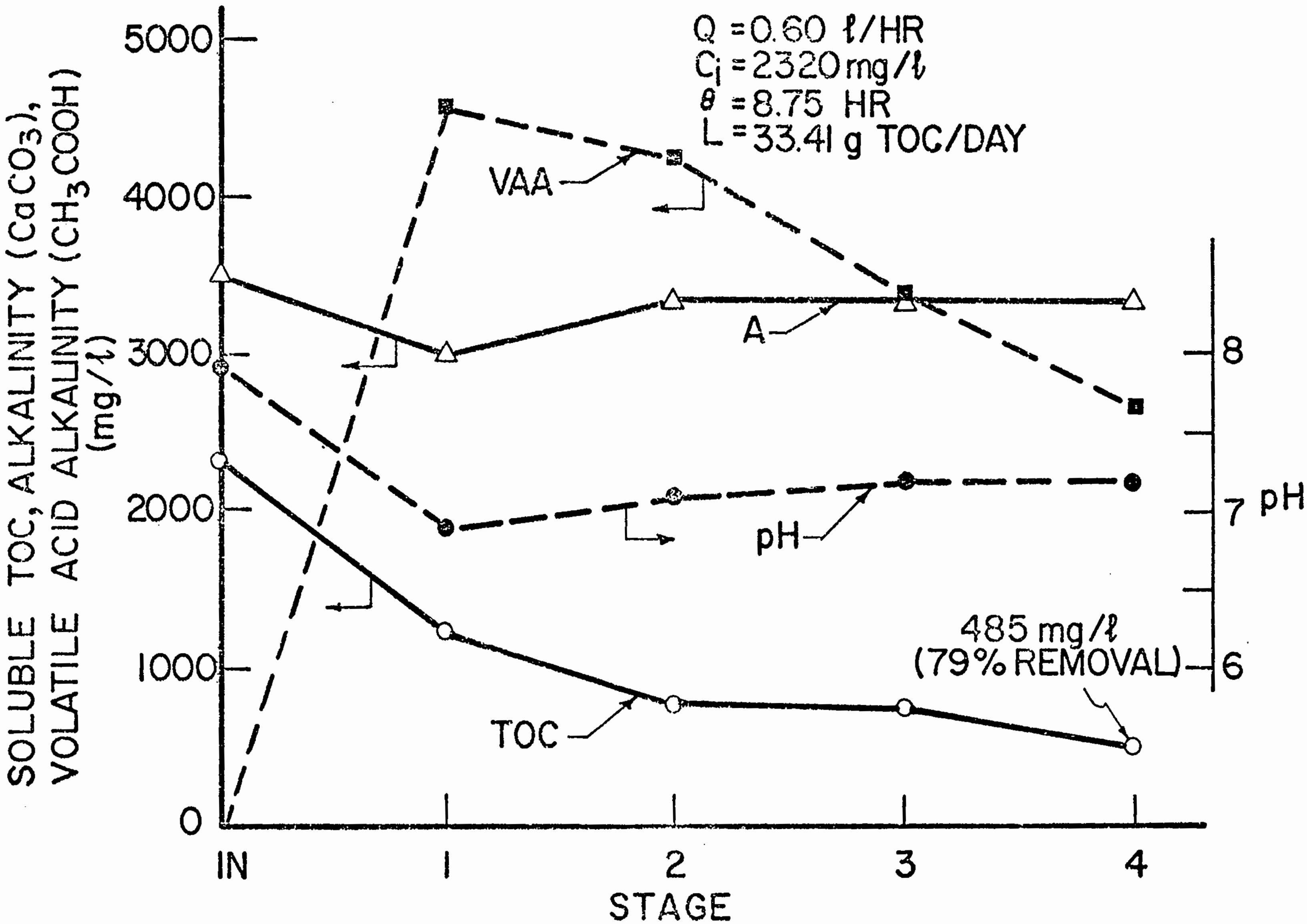

FIGURE 3 EXPERIMENT 6, 2F – 2H

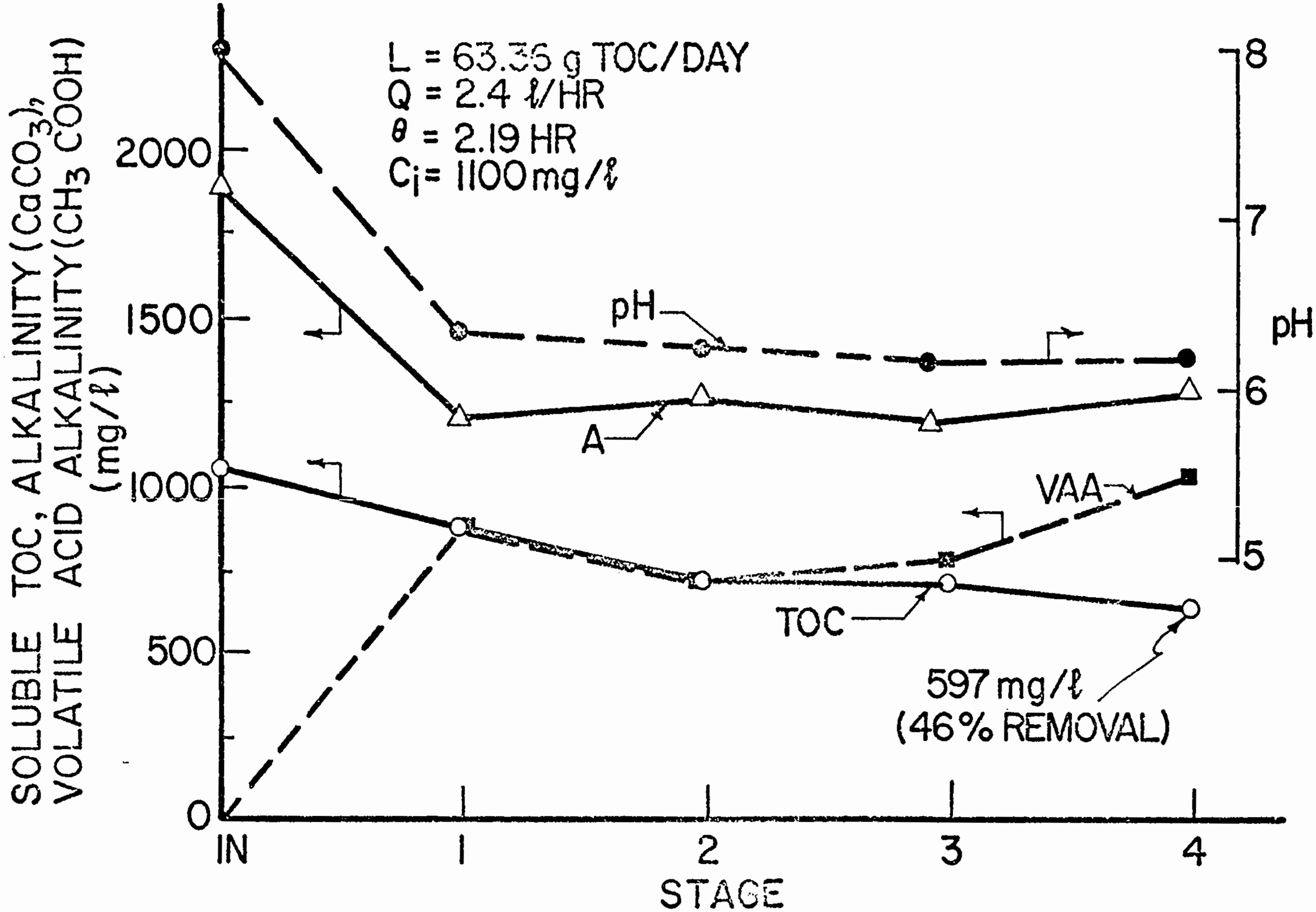

FIGURE 4 EXPERIMENT 4, IF-8H

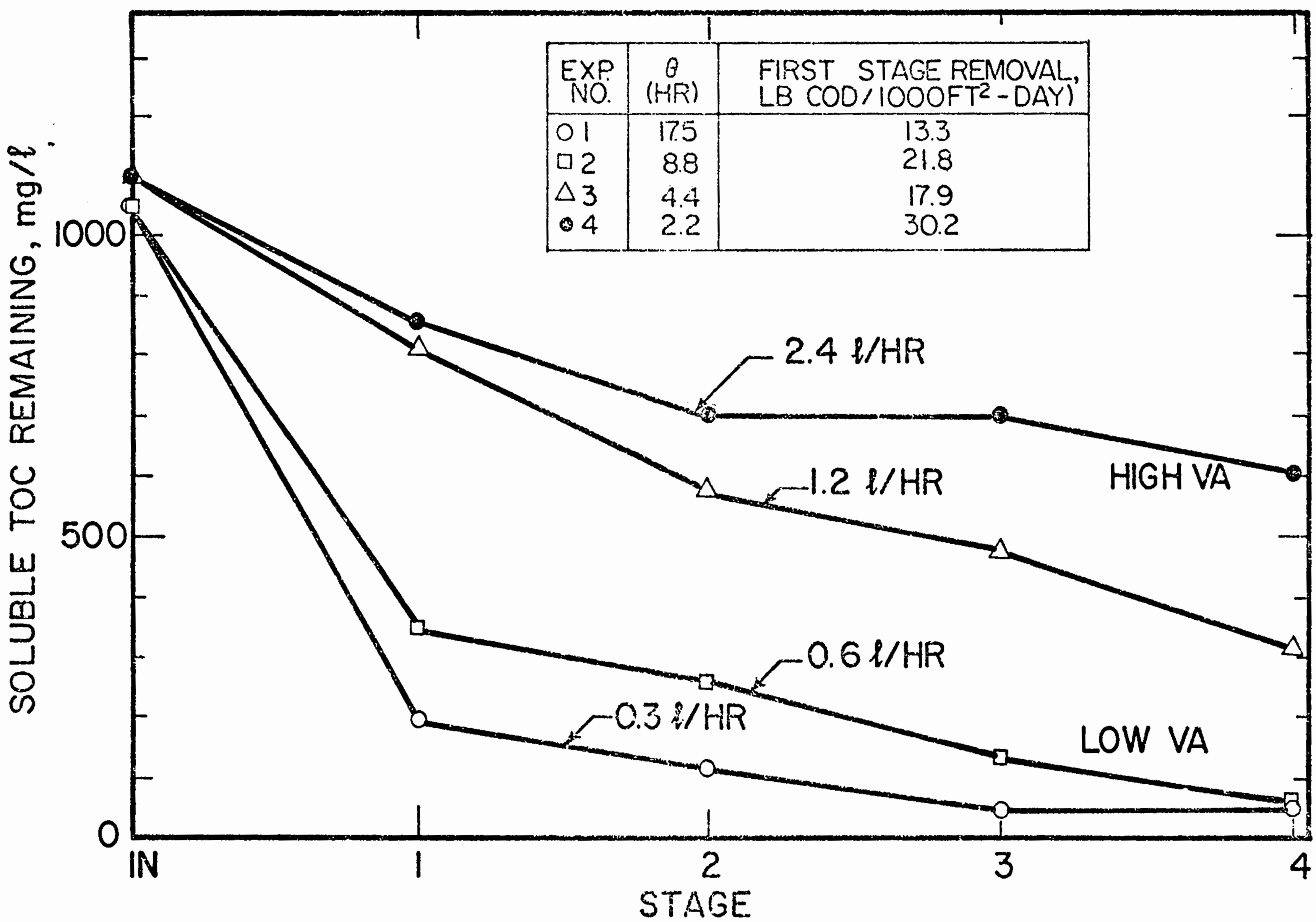

FIGURE 5 EFFECT OF FLOW RATE ON AnRBC TREATMENT

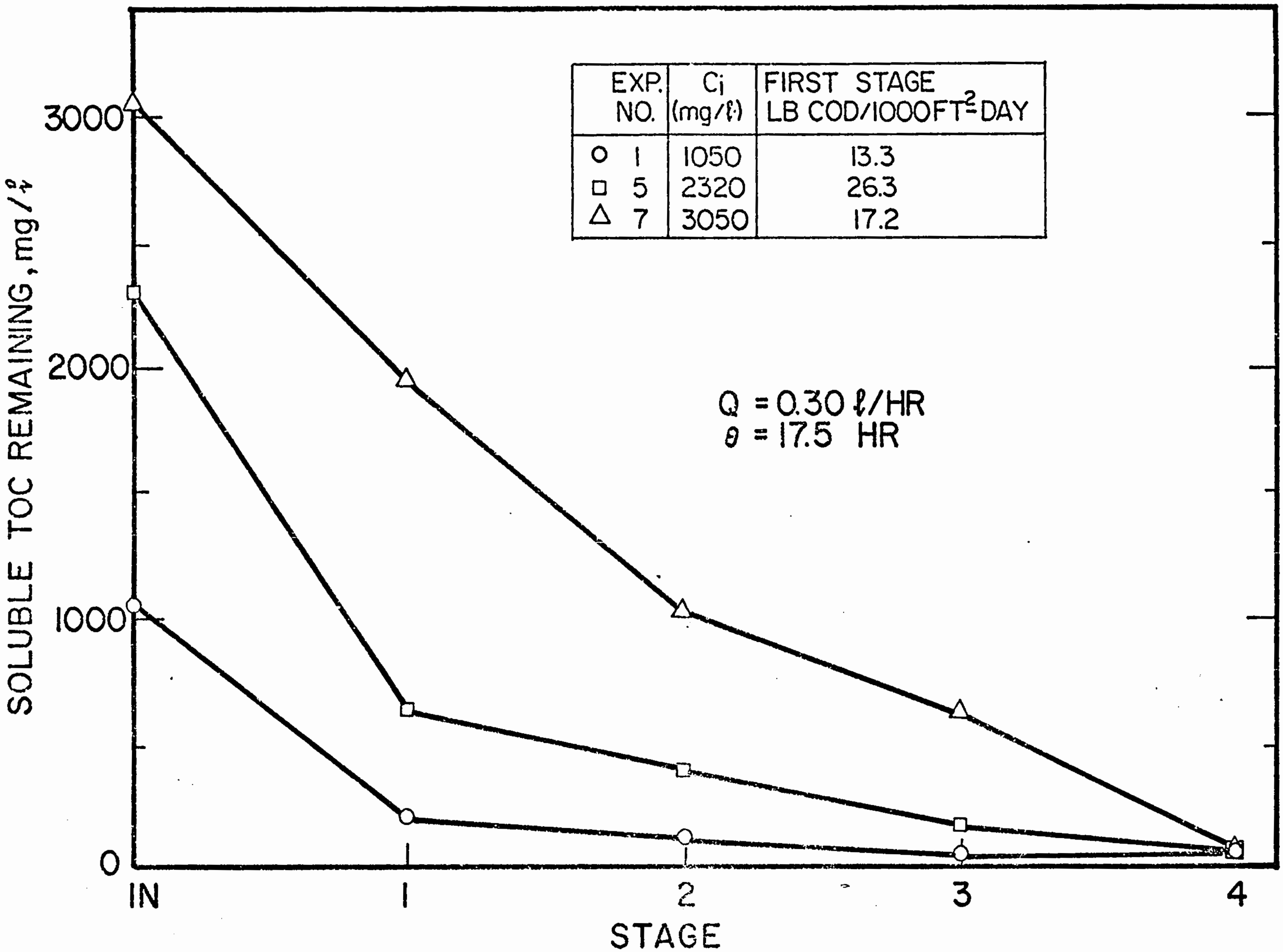

FIGURE 6 SUBSTRATE CONCENTRATION EFFECTS

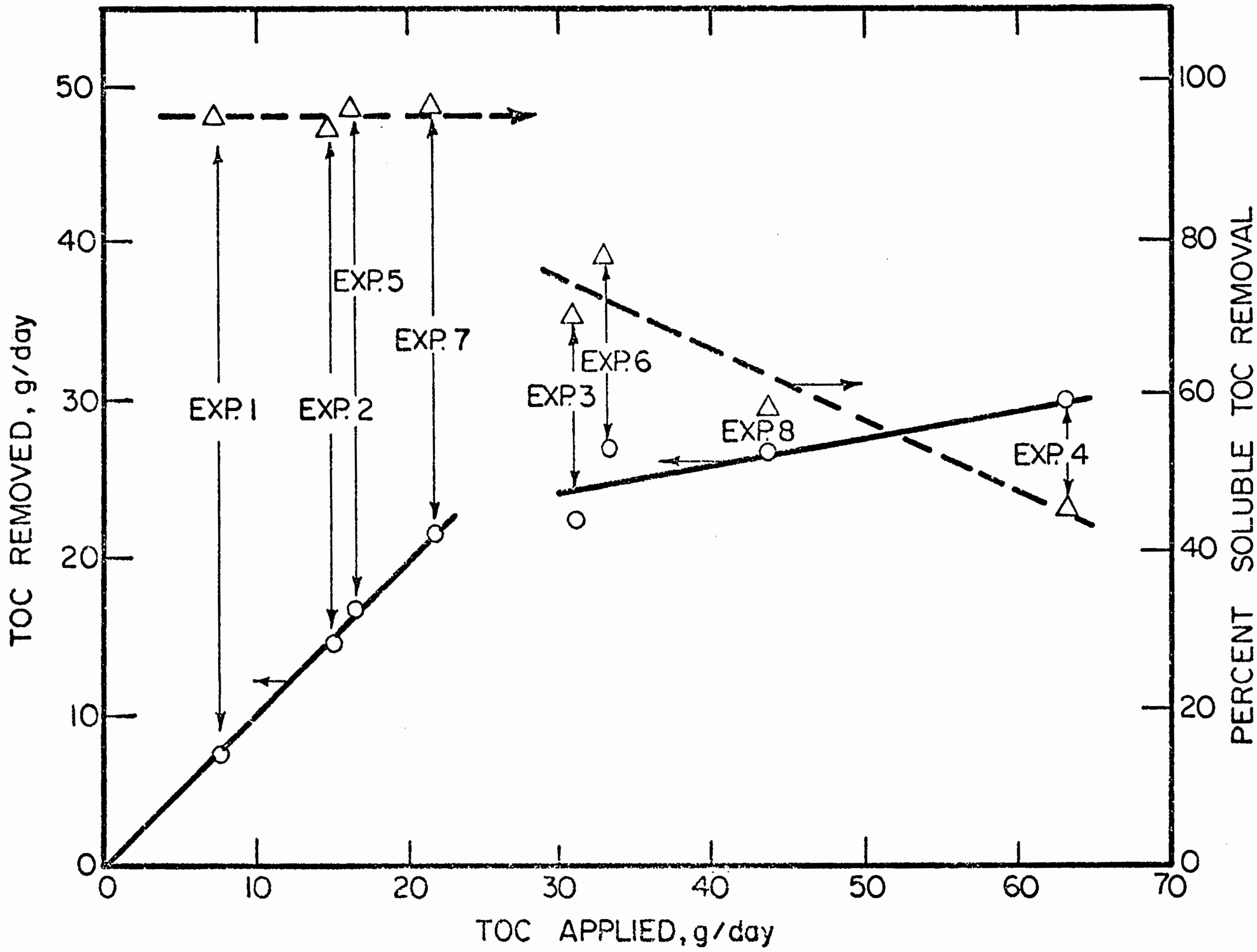

FIGURE 7 SOLUBLE TOC REMOVAL

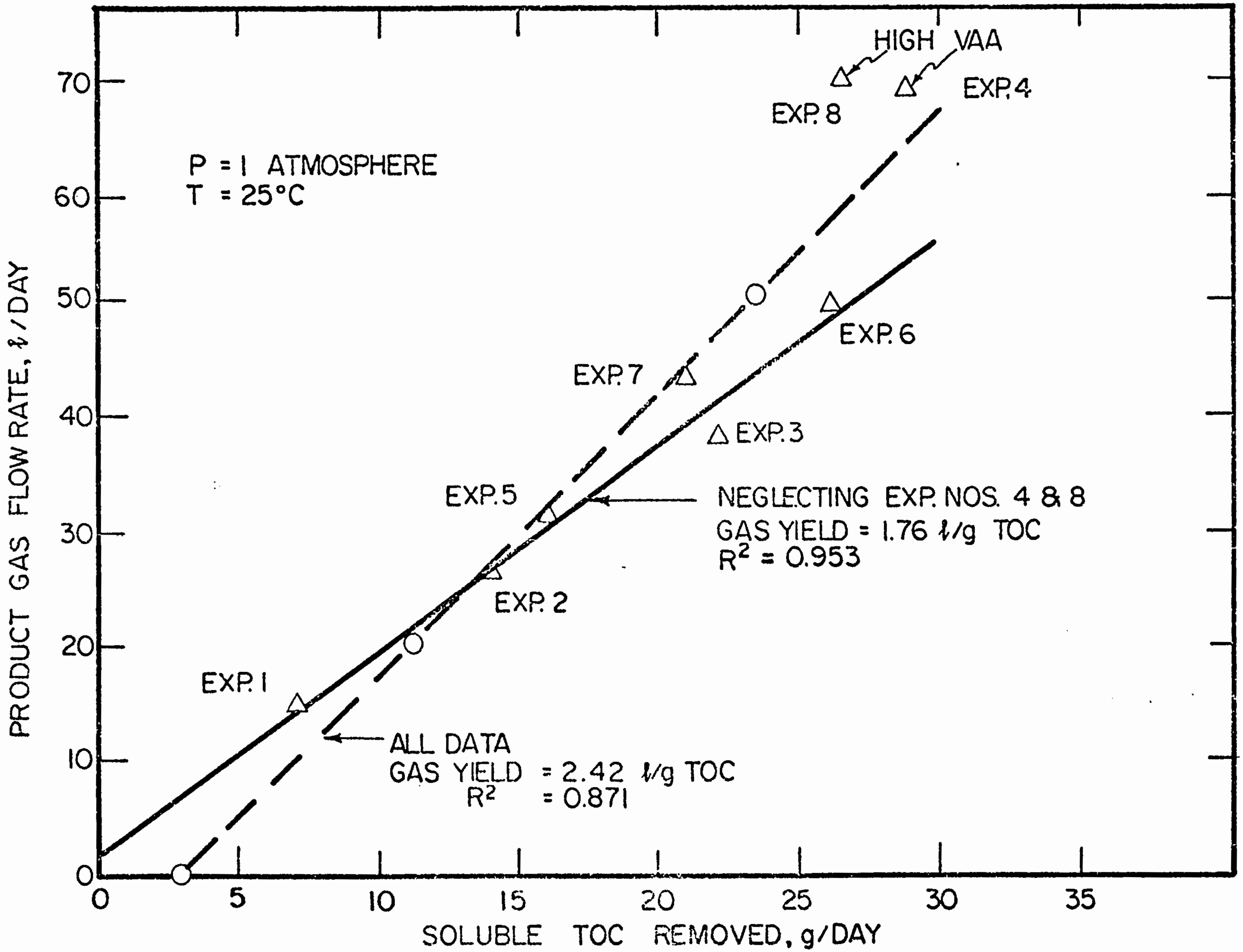

FIG. 8 TOTAL GAS PRODUCTION

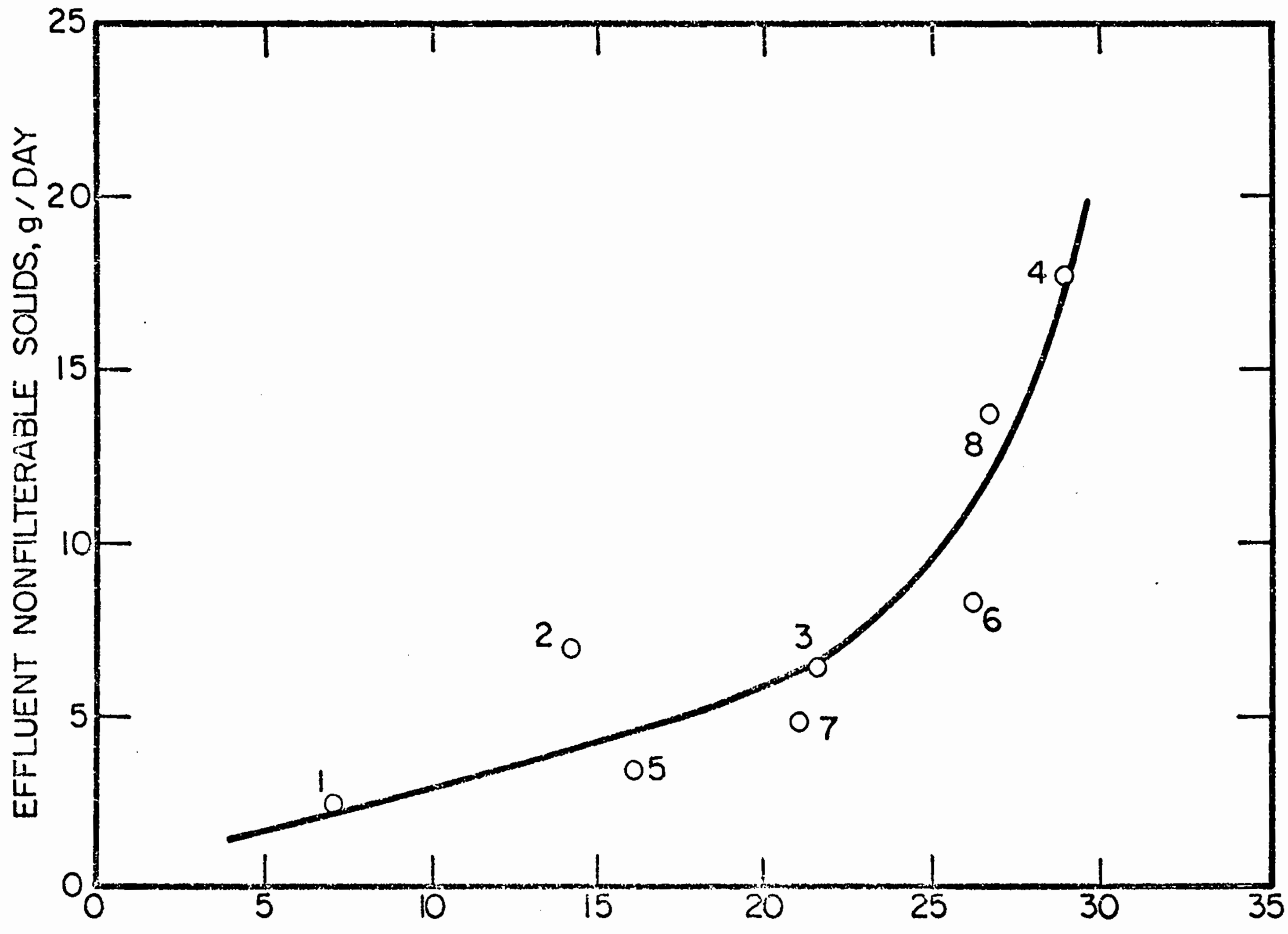

FIG. 9 SOLIDS PRODUCTION

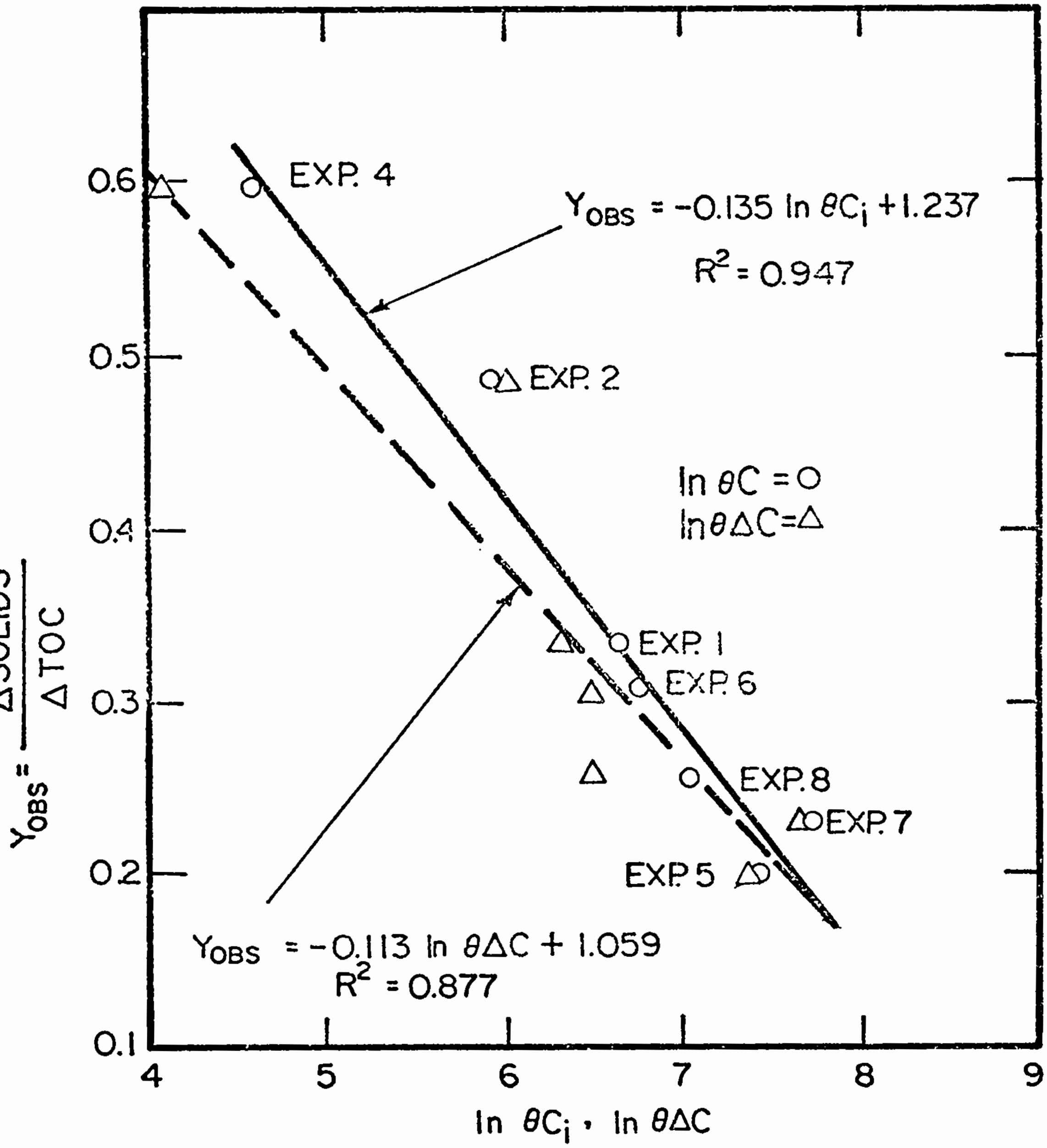

FIG. 10 OBSERVED YIELD AS A FUNCTION OF LOADING CONDITIONS

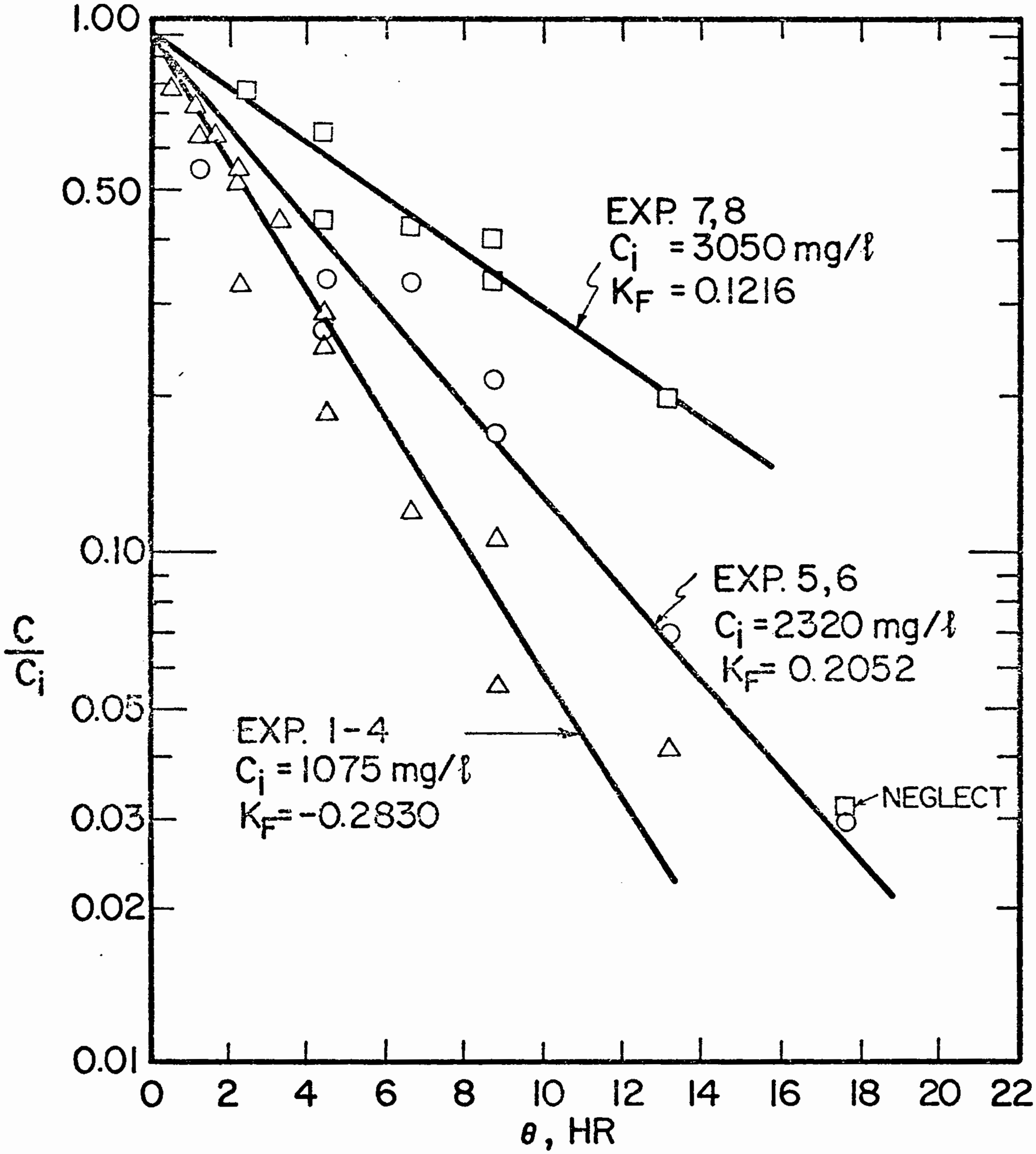

FIGURE 11 SOLUBLE TOC REMAINING

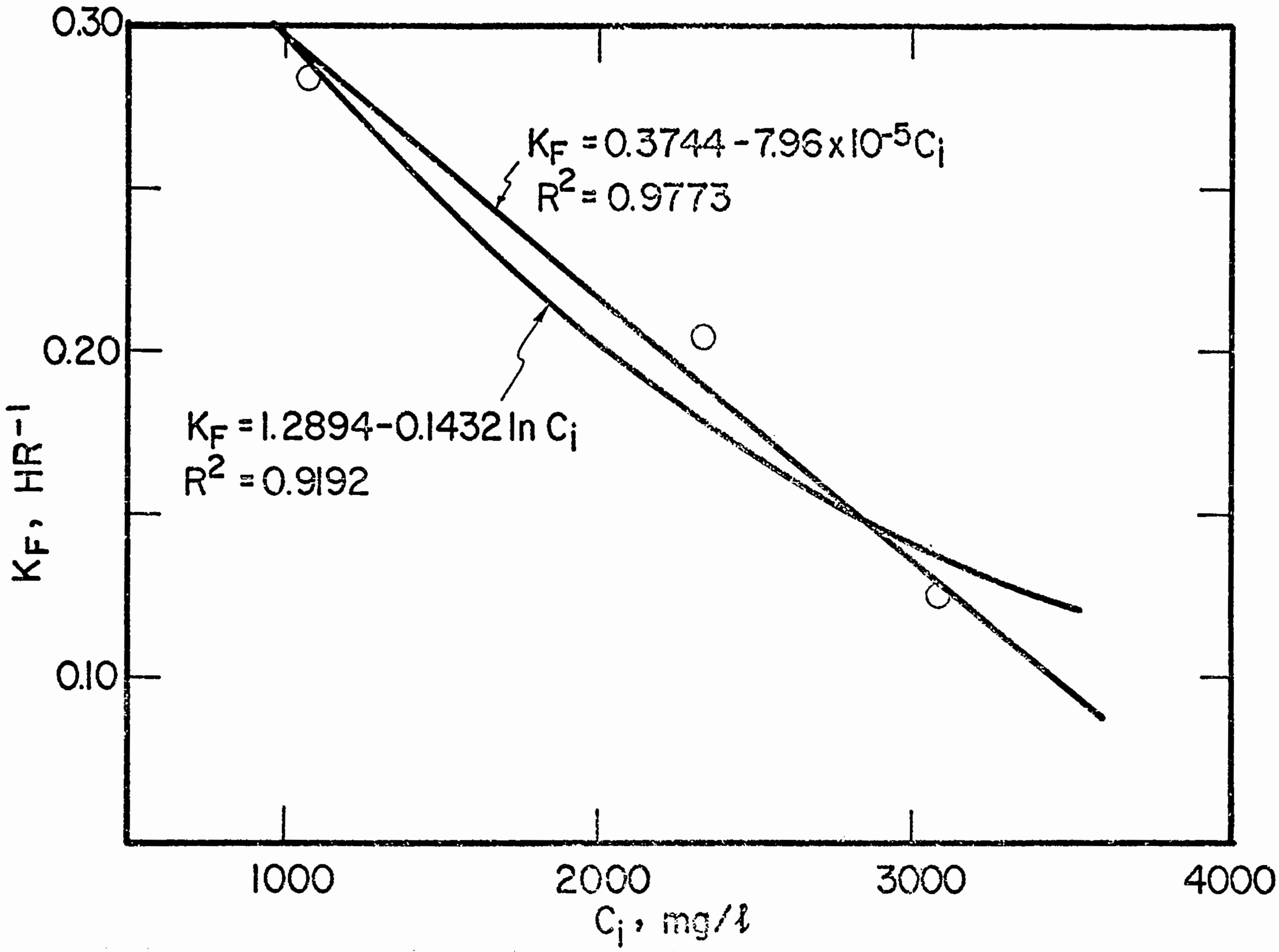

FIGURE 12 K_F DEPENDENCE ON C_i

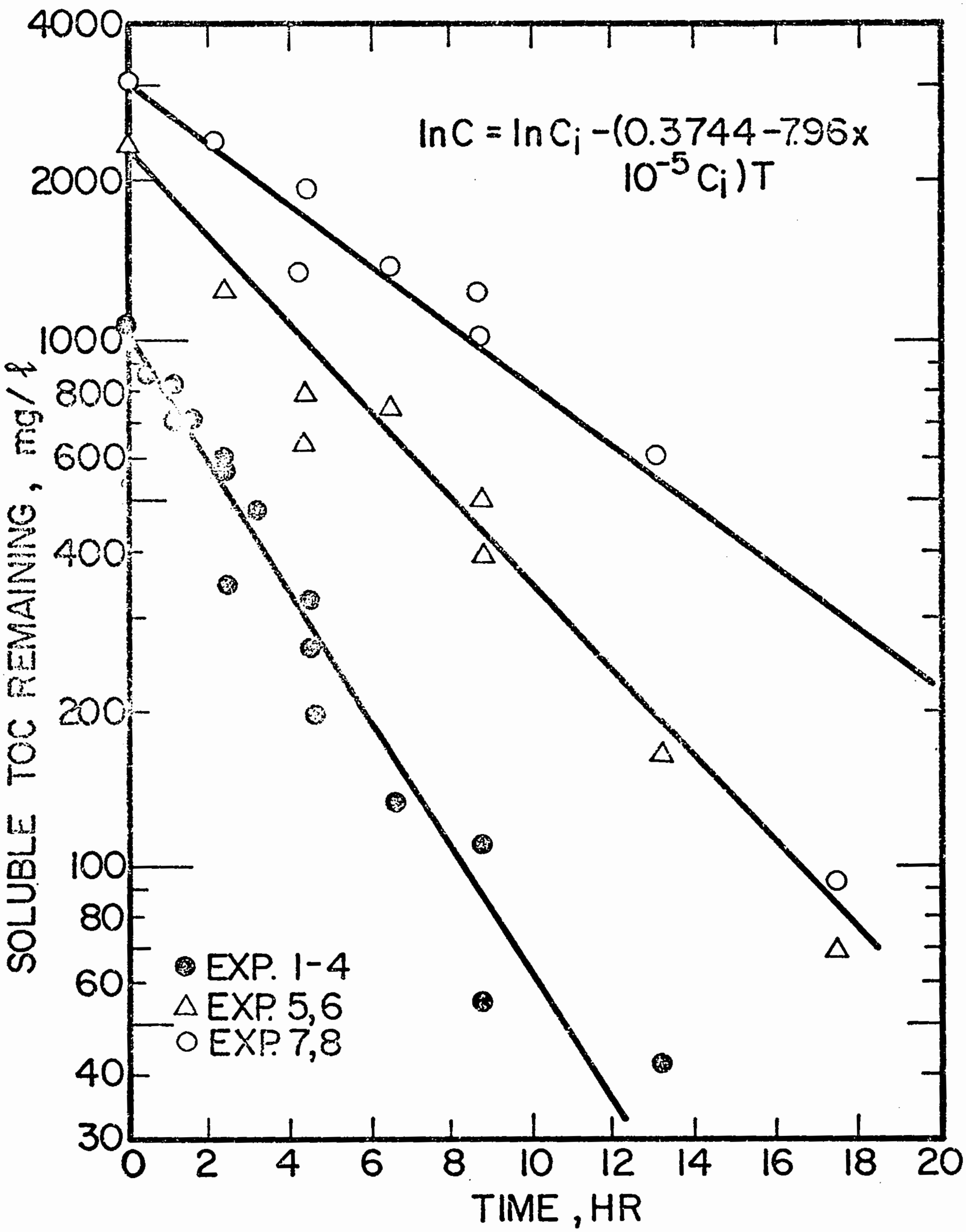

FIGURE 13 PREDICTED TOC VALUES

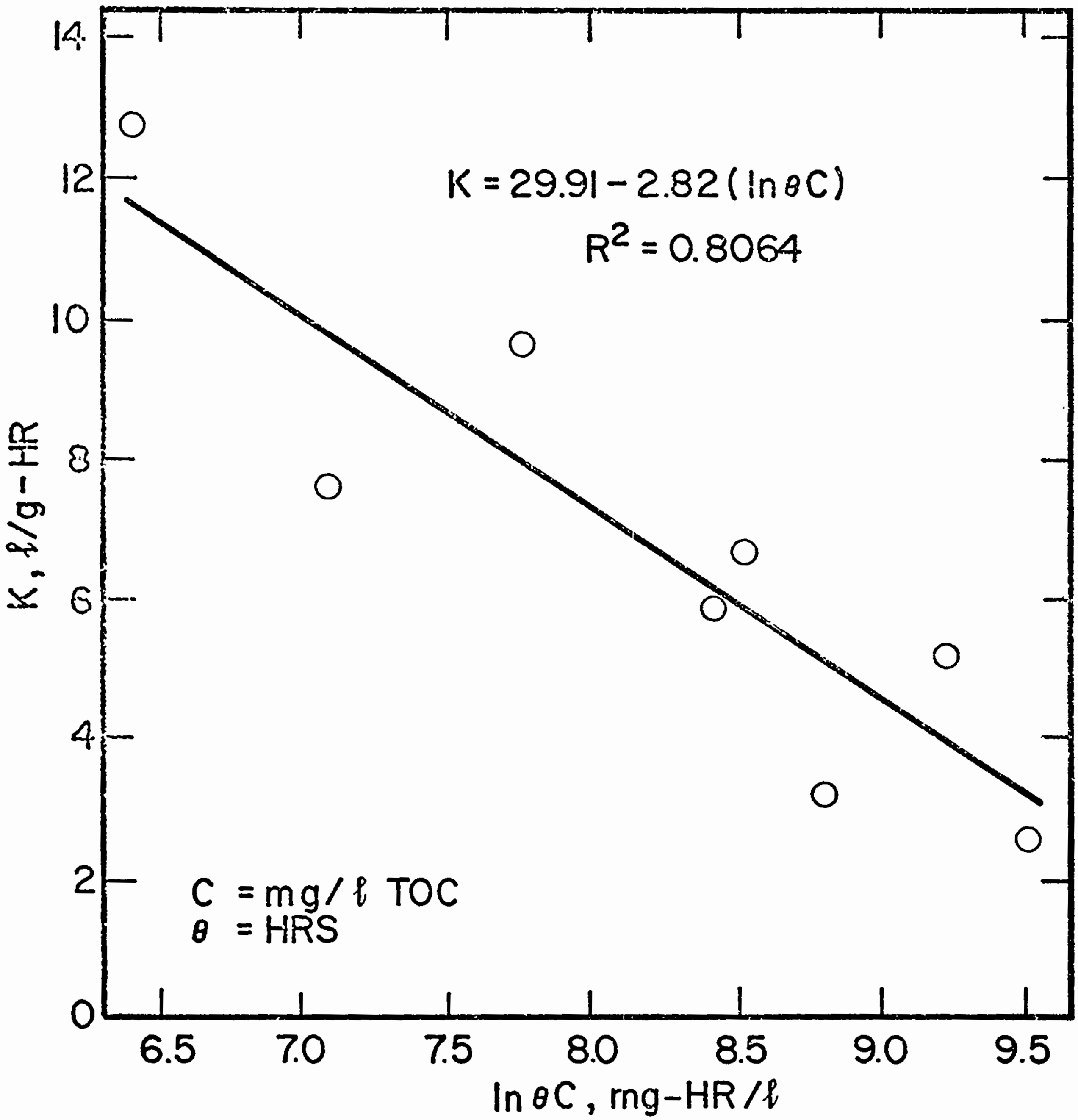

FIGURE 14 DEPENDENCE OF FIRST STAGE K VALUES ON THE LOADING FACTOR

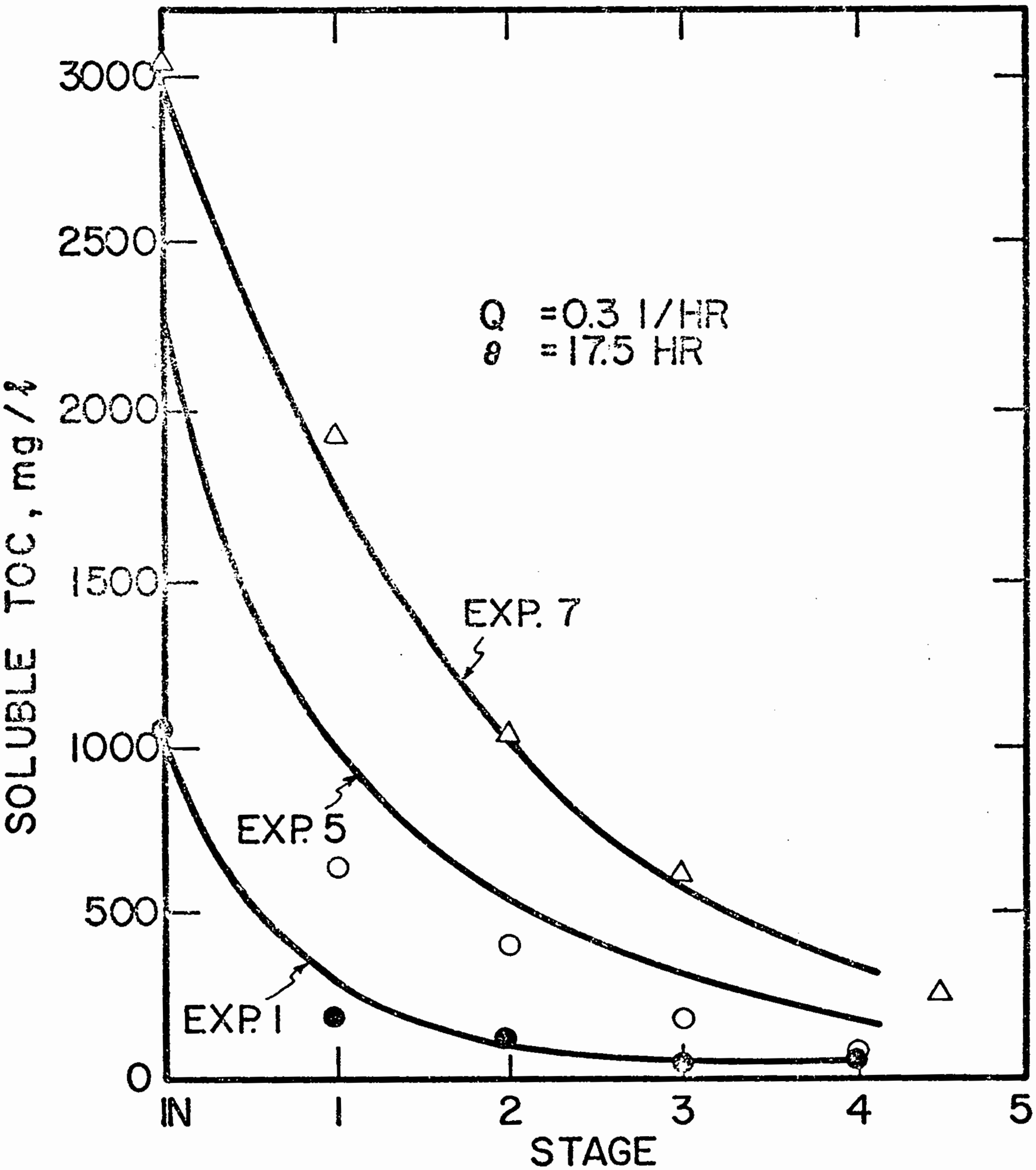

FIGURE 15 MASS TRANSPORT MODEL, EXPERIMENTS 1, 5, 7

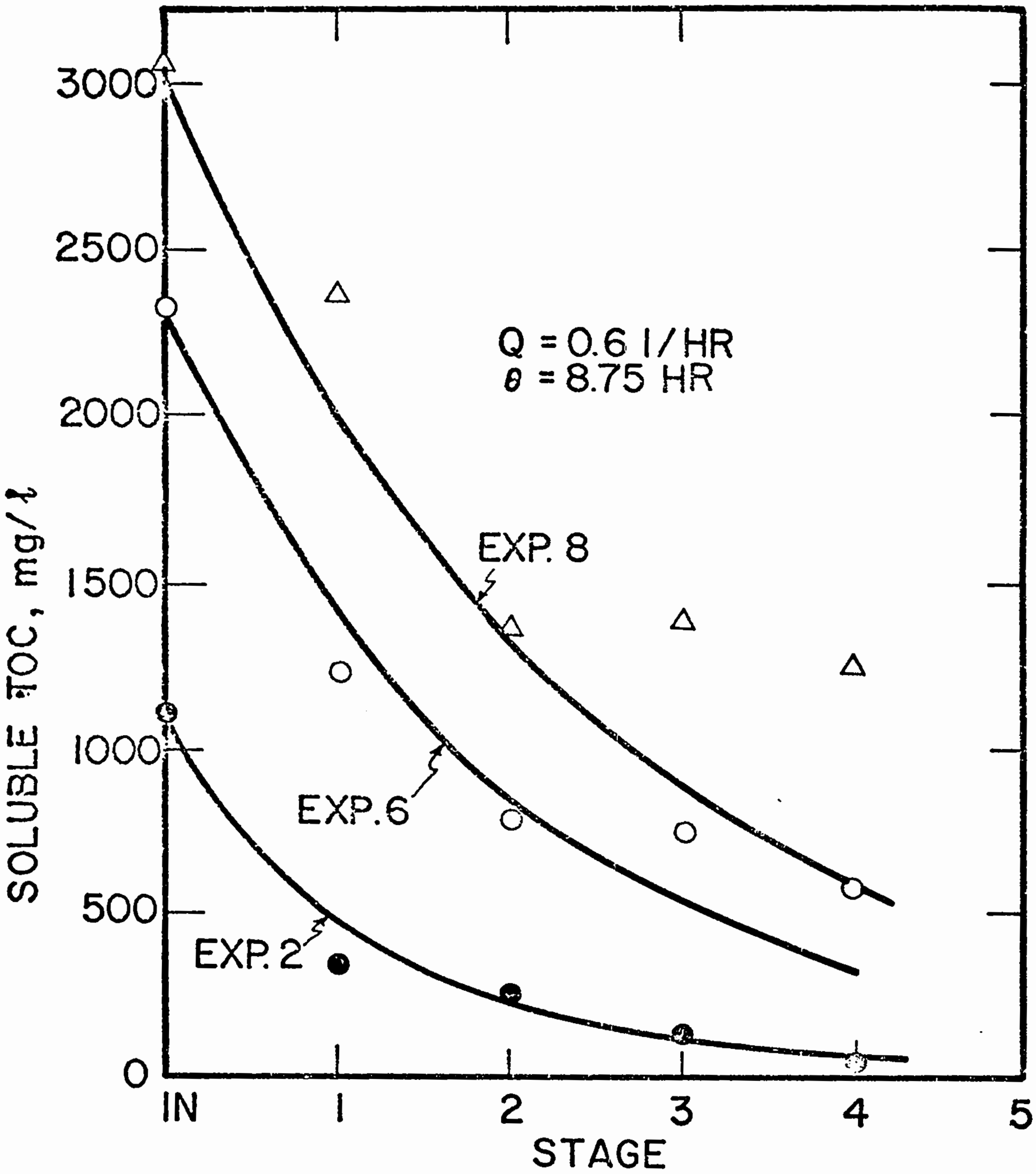

FIGURE 16 MASS TRANSPORT MODEL, EXPERIMENTS 2,6,8

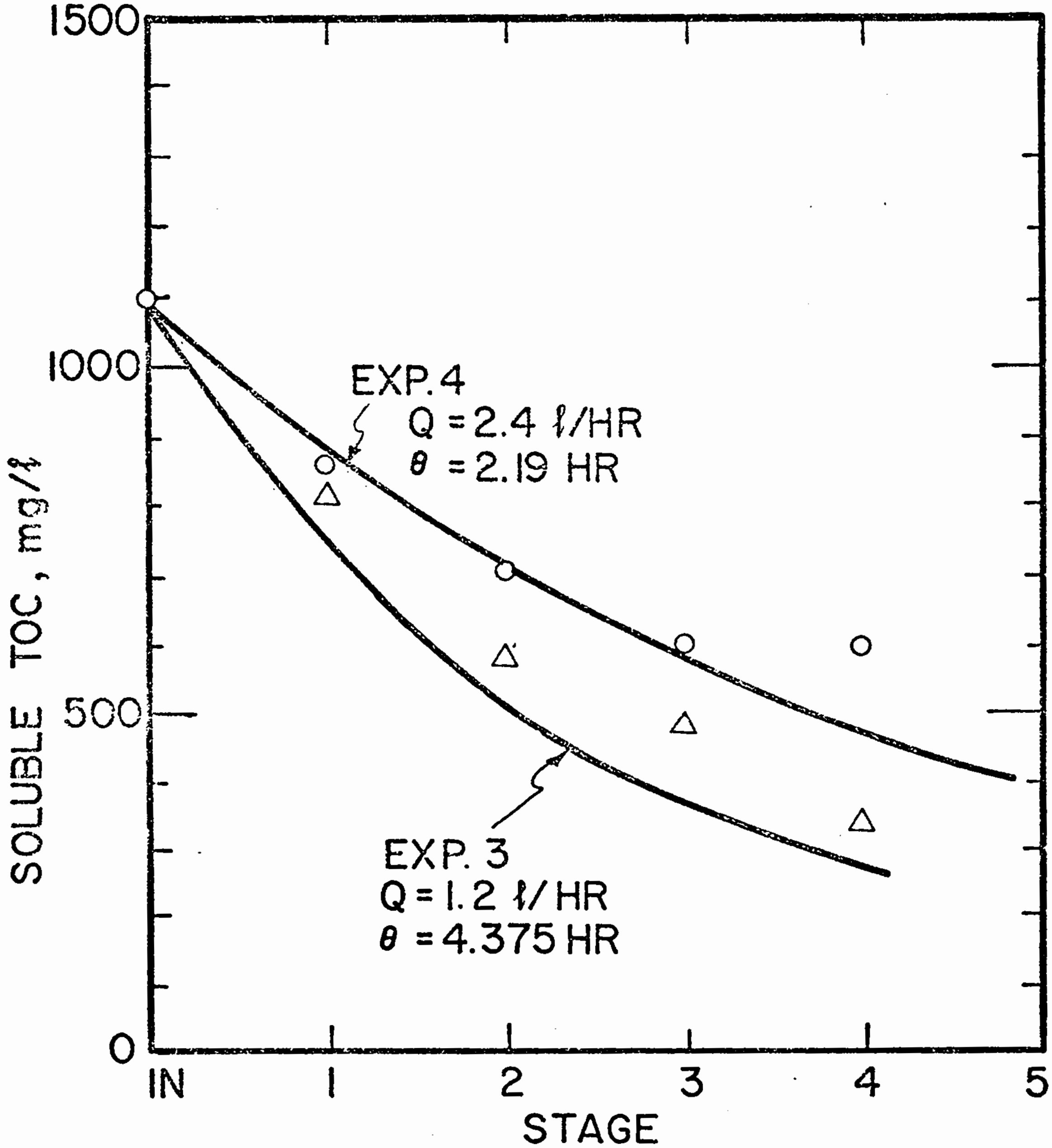

FIGURE 17 MASS TRANSPORT MODEL, EXPERIMENTS 3 AND 4

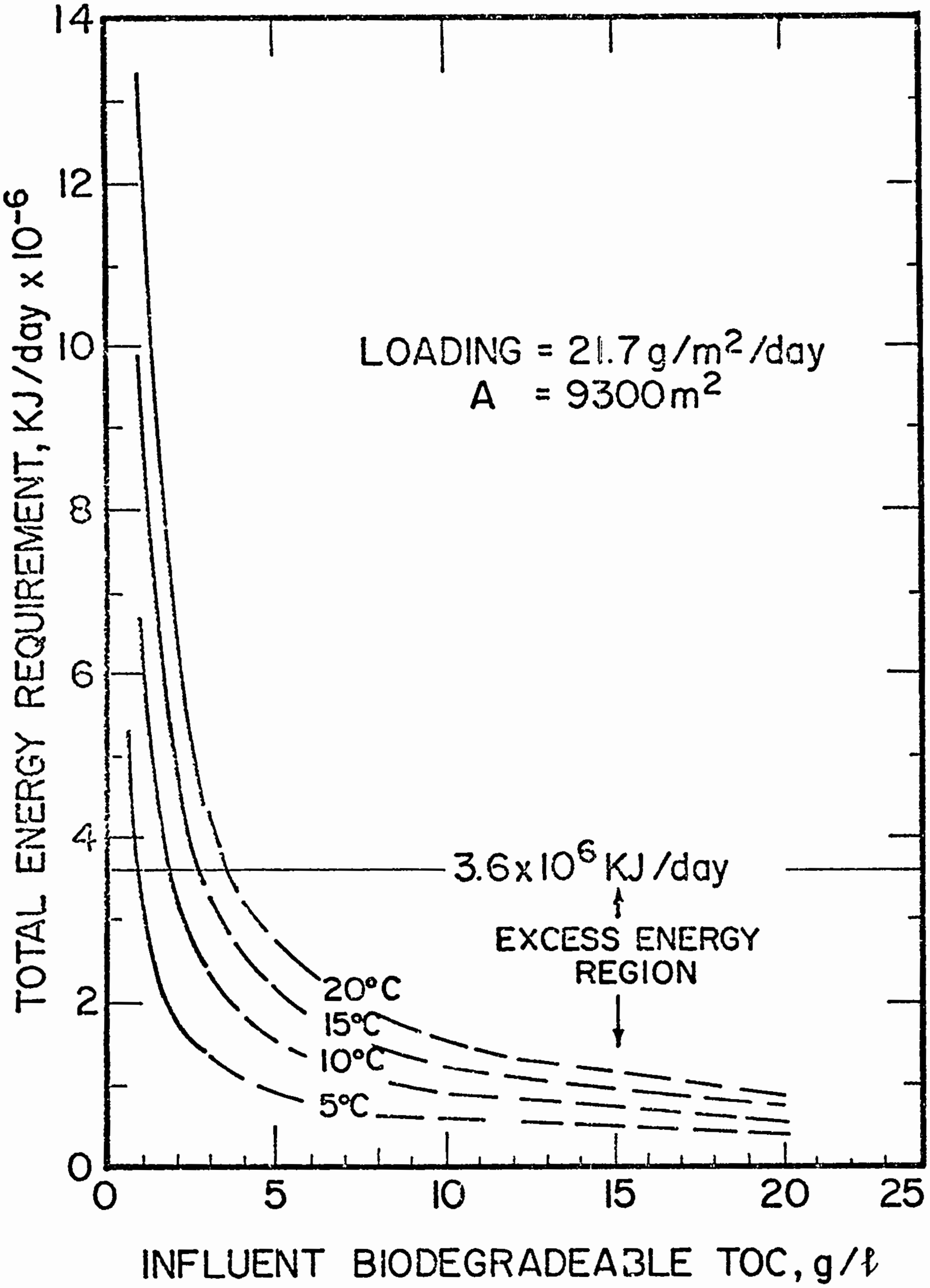

FIGURE 18 INFLUENT SUBSTRATE CONCENTRATION AND TEMPERATURE EFFECTS ON ENERGY REQUIREMENTS, 95% REMOVAL

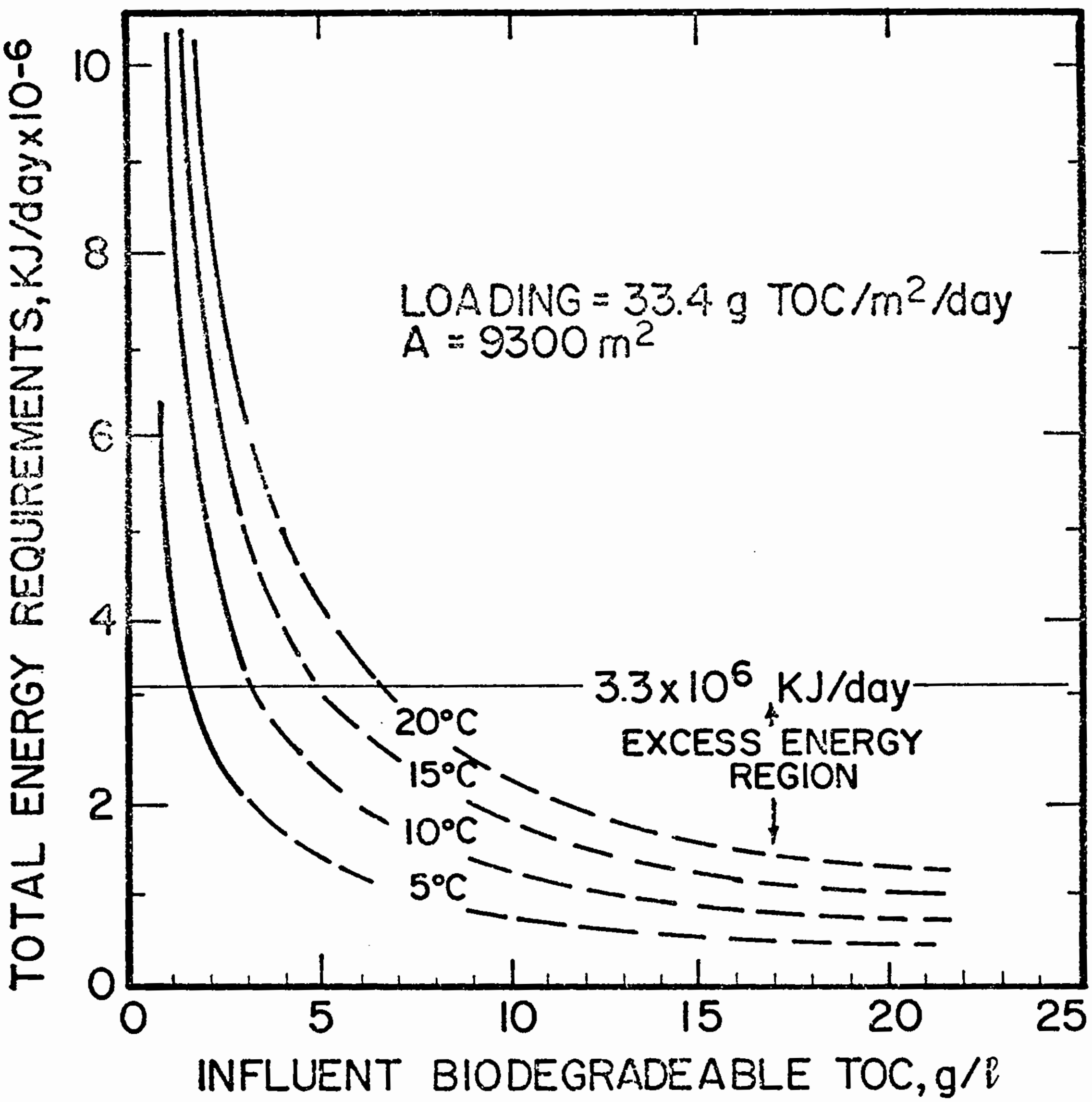

FIGURE 19 INFLUENT SUBSTRATE CONCENTRATION AND TEMPERATURE EFFECTS ON ENERGY REQUIREMENTS, 68% REMOVAL