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FIRST INTERNATIONAL CONFERENCE ON FIXED-FILM BIOLOGICAL PROCESSES

April 20-23, 1982
Kings Island, Ohio

Edited by Y.C. Wu, Ed D. Smith,
R.D. Miller, and E.J. Opatken

Vol. I

Members Of Organizing Committee:

Yeun C. Wu (Chiarman)
Department of Civil Engineering
University of Pittsburgh
Pittsburgh, Pa.

James V. Basilico
Office of Research and Development
U. S. Environmental Protection Agency
Washington, D.C.

Ed. J. Opatken
Wastewater Research Division
U.S. Environmental Protection Agency
Cincinnati, Ohio

Ed. D. Smith
Environmental Division
U.S. Army Construction Engineering
Research Laboratory
Champaign, Illinois

Ed. H. Bryan
Civil and Environmental Engineering
National Science Foundation
Washington, D. C.

Roy D. Miller
Environmental Health Engineering Branch
U.S. Army Environmental Hygiene Agency
Fort Meade, Maryland

Richard Dick
Department of Civil Engineering
Cornell University
Ithaca, New York

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FOREWORD

Biological wastewater treatment has been practiced in many forms since the early part of this century, but fixed-film biological processes have only recently been more intensively studied and applied by water pollution control researchers and engineers. Because of some inherent advantages over suspended growth processes, today there is a greater interest in fixed-film biological treatment processes than ever before. This Conference was designed to provide a forum for that interest and to help accelerate further development of this technology.

The objective of this Conference was to assess the State of Knowledge and identify the research needs regarding the full spectrum of fixed-film biological processes. The Conference addressed many new approaches to anaerobic as well as aerobic treatment. Many practical applications and new research findings were presented and many of the speakers expressed optimism for significant progress in the future. Because of their keen interest and the dedication of those who attended, this Conference was truly a professionally stimulating experience. There was much interaction and exchange between all participants.

The Conference consisted of 13 technical sessions with a total 80 presentations, one workshop on research needs for fixed-film biological wastewater treatment processes, and a field tour to the LeSourdsville Regional Rotating Bioloigcal Contactor Plant. The Conference Proceedings consisted of 77 of those papers presented by the authors. More than 300 participants representing a wide spectrum of researchers and practitioners attended the Conference. Worldwide interest was also evident from the 31 foreign participants who traveled from Canada, India, Saudi Arabia, Yugoslavia, Japan, Norway, Switzerland, Republic of China, Italy, South Korea, France, Belgium, England, West Germany, and Scotland.

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May 24, 1982

James V. Basilico
Yeun C. Wu

ABSTRACT

The First International Conference on Fixed-Film Biological Processes was held at the Kings Island Resort, Kings Island, Ohio on April 20-23, 1982. This Conference serves as an opportunity to assess the applicability of this advanced technology for the treatment of municipal and industrial wastewaters.

The proceedings are essentially the papers and discussion given by authors and participants. The papers are divided into 13 major topic areas:

1. Current Status and Future Trends
2. Biofilm and Biokinetics
3. Concepts and Models
4. Small Scale/On Site Systems
5. Municipal Wastewater Treatment-Case Histories
6. Nitrification and Denitrification
7. Industrial Wastewater Treatment Part I- Rotating Biological Contactors
8. Industrial Wastewater Treatment Part II- Biofiltration, Packed Bed Reactors
9. Innovative Research
10. Aerobic and Anaerobic Treatment-Submerged Media Reactors
11. Industrial Wastewater Treatment Part III- Submerged, Anaerobic Fixed-Film Reactors
12. Process Evaluation and Design
13. Experiences With Fixed-Film Treatment Facilities

The discussion occurred during the Research Needs Workshop was taped and printed as an appendix.

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Keynote Speakers:

Mark Williams
Dean, School of Engineering
University of Pittsburgh
Pittsburgh, Pa.

Ed. D. Smith
Environmental Division
U. S. Army Construction Engineering
Research Laboratory
Champaign, Illinois

William Jewell
Department of Agricultural Engineering
Cornell University
Ithaca, New York

Ed. D. Schoreder
Department of Civil Engineering
Univesrity of California
Davis, California

Conference Assistants:

John C. Kennedy
Chung C. Chen
Shen Y. Lien
Sin N. Hsieh
Jeff Greenfield
Li L. Lin

Department of Civil Engineering
University of Pittsburgh
Pittsburgh, Pa.

T. Casteel
J. Wingham

Kings Island Resort
Kings Island, Ohio

Session Chairman:

Joel I. Abrams
Department of Civil Engineering
Univesrity of Pittsburgh
Pittsburgh, Pa.

Roy D. Miller
Environmental Health Engineering Branch
U.S. Army Environmental Hygiene Agency
Fort Meade, Maryland

John A. Roth
Center for Environmental Quality Management
Vanderbilt University
Nashville, TN

Richard Dick
Department of Civil Engineering
Cornell University
Ithaca, New York

Hallvard Odegaard
Department of Sanitary Engineering
University of Trondeheim
Trondheim-NTH, Norway

Marvin E. Lambert
Columbus Gas utility
Columbus, Ohio

Dick Brenner
Municipal Environmental Research Lab.
U.S. Environmental Protection Agency
Cincinnati, Ohio

Michael Saunders
School of Civil Engineering
Georgia Institute of Technology
Atlanta, GA

Ed. D. Smith
Environmental Division
U.S. Army Construction Engineering Research Lab.
Champaign, IL

A. A. Friedman
Department of Civil Engineering
Syracuse University
Syracuse, New York

Michael Sweet
Engineering Science Ltd.
Cleveland, Ohio

James V. Basilico
Office of Research and and Development
U.S. Envrionmental Protection Agency
Washington, D.C.

Ed. J. Opatken
Municipal Environmental Research Lab.
U.S. Environmental Protection Agency
Cincinnati, Ohio

John Bandy
Envrionmental Division
U.S. Army Construction En gineering Research Lab.
Champaign, IL.

Workshop Organizers:

A. F. Gaudy, Jr.(Chairman)
Department of Civil Engineering
University of Delaware
Newark, Delaware

Ed. D. Opatken
Municipal Environmenatl Research Lab.
U. S. Environmental Protection Agency
Cincinnati, Ohio

A. A. Friedman
Department of Civil Engineering
Syracuse University
Syracuse, New York

W. W. Eckenfelder, Jr.
Department of Environmental Engineering
Vanderbilt University
Nashville, TN

C. P. Leslie Grady, Jr.
Department of Environmental Engineering
Clemson University
Clemson, South Carlonia

Yeun C. Wu
Department of Civil Engineering
University of Pittsburgh
Pittsburgh, Pa.

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PART I: GENERAL SESSION

KEYNOTE ADDRESS

STATE OF KNOWLEDGE FOR ROTATING BIOLOGICAL CONTACTOR TECHNOLOGY

E. D. Smith. Environmental Engineer and Leader of the Environmental Water Quality Management Team, U.S. Army Construction Engineering Research Laboratory, Champaign, IL 61820

J. T. Bandy. Environmental Engineer, U.S. Army Construction Engineering Research Laboratory, Champaign, IL 61820

INTRODUCTION

It is a real pleasure for me to be here this morning to discuss the state-of-knowledge on Rotating Biological Contactors (RBC's). When I made the Keynote address at the 1980 First National Symposium/Workshop on RBC's, held at Champion, PA in 1980, I had hoped that this type of conference might become a tradition. I believe that the 1980 conference was beneficial to the RBC industry. I expect this conference to be equally useful. Recently it has become more difficult to obtain funding for this type of symposium. I believe that the benefits of these meetings far outweigh their costs. I hope that the success of our conference will encourage the sponsoring of similar gatherings in the future.

I am happy to see the excellent turnout for this symposium. Most of the experience and competency in the field of

RBC technology are represented here today. The agenda indicates you will be quite busy in the next three days. I am confident that it will be a productive and pleasant experience for you. I am confident that the proceedings, which will be published from this meeting (and for which I am responsible), will provide very excellent technical guidance to those who could not attend.

Today, I plan to provide a state-of-knowledge definition of RBC technology. I plan to do this by discussing how the RBC scenario has changed from 1980 - the year that a state-of-knowledge definition was given at the 1st National Symposium/Workshop on RBC Technology.

PROGRESS AND PROBLEMS

It was reported at the 1980 conference that, in comparison with many other wastewater treatment technologies (e.g., activated sludge), few dollars and man-years of research had been devoted to RBC technology. The many excellent papers presented at the 1980 symposium significantly narrowed that disparity. Numerous additional research studies and field evaluations have been described in the literature since that time. Many new RBC installations have come on line since the First National Symposium. However, despite RBC technology's continued spread and despite the incorporation of field experience and research findings into RBC process and equipment design recommendations; one unfortunate characteristic of the technology has remained unchanged. Those who design and operate RBC facilities must largely rely upon the design recommendations and operation guidance of the vendors of RBC equipment.

The present state-of-knowledge is such that there is no single best design procedure or set of relationships that are universally applicable. No well-defined theory of RBC design and operation is 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. The situation is compounded by the fact that each manufacturer has a different approach to media fabrication, configuration, and shaft attachment and shaft design, and there exist many conflicting stories and

opinions as to the suitability of the alternative equipment for even conventional applications.

Although millions of dollars have been spent by American industries and municipalities for RBC process equipment, the latest wastewater treatment guidance documents still 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 process. 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 consideration either do not have RBC cost curves (capital, O&M, energy, etc.) or the curves are dated. Guidance remains scarce with regard to RBC applicability, design, O&M and economic considerations.

To make matters worse, the RBC industry has suffered a public relations problem because of numerous equipment failures. Premature shaft failures, stub end failures and media separation/degradation have been experienced at existing installations. The durability of the polyethylene is still uncertain because of the relatively short service record (most facilities with RBC systems were built during the past five years). Industry is attempting to rectify these problems.

To be fair to the RBC equipment vendors it must be noted that the RBC process has unique characteristics which almost guarantee that problems would occur during the early development of the technology. It would be very difficult to destroy or damage wastewater treatment technologies such as activated sludge or trickling-filters through improper design or operations. Improper design or operation of RBC units potentially could result in structural failure

problems. Even with proprietary assurances that current designs are much improved, the life expectancy of major components is not fully known. Consequently, choice of the RBC alternative should be accompanied by a negotiated performance/equipment warranty. This consideration is important when a pollution abatement engineer wants to be confident of the reliability of any wastewater treatment technology. However, one should keep in mind that if the manufacturers' assurances are accurate, current designs are much improved. Then RBC technology should be the technology of choice wherever it is applicable. It is significant that hundreds of RBC plants have been in operation for several years without experiencing media/shaft failure problems.

All manufacturers offer a warranty against defects in materials and workmanship after delivery or after plant start-up. The warranty period and conditions vary depending on system components and the manufacturer, and are often negotiable. For example, the Plainville Plant in Connecticut was given a warranty period of 30 years for the shafts, 10 years for the surface media, and 5 years for mechanical equipment.

Many RBC manufacturers offer performance guarantees that generally provide a specified effluent with the equipment installed and operating at design conditions. The guarantee usually obligates the manufacturer to provide new equipment or a partial refund if the design effluent standards are not met. This guarantee is predicated on the fact that influent characteristics are within the specified limit. Generally, the manufacturers are willing to negotiate a guarantee as long as they agree with the treatment design. During the maturational period of the RBC process, these guarantees and warranties will be especially important to the RBC user community.

STATUS OF RBC TECHNOLOGY

Even with these problems, the extent and magnitude of interest regarding RBC technology continues to increase. The participation at the conference session dedicated to RBC's is evidence of the interest of various sectors (private, academic, research, government agency, regulatory, A/E, professional organization, design engineer, industrial, and plant operators). All of the above and other profes-

sionals involved in wastewater treatment and mangement are represented at this symposium.

Two years ago it was reported that RBC technology was popular in Europe for both municipal and industrial applications and that it was being utilized ever more frequently in the U.S. Today, it can safely be reported that RBC technology has truly made the transition into a truly cosmopolitan treatment technology. There are more than 30 RBC manufacturers in Japan alone.

In the U.S., RBC's have been in operation treating municipal wastewater for more than 10 years. Over 250 installations are presently in operation with design flow rates ranging from less than 0.01 mgd to 54 mgd. Approximately 25 percent of existing RBC municipal facilities in the U.S. are package plants. RBC's are currently being evaluated for potential application to a 200 mgd plant which would have several hundred shafts. Approximately 70 percent of the RBC systems operating in the U.S.A and Canada are designed for organic carbon removal only, 25 percent for combined organic removal and notification and 5 percent for nitrification of secondary effluent.

Several significant developments in RBC technology are occurring. Some of these directly address the problems to which I alluded earlier. All are potentially important.

a. The U.S.E.P.A. has chosen RBC's as the topic of their first publication of a Design Information Series (DIS) document, the purpose of which is to provide selected design information. The document is currently under review. The DIS is not a manual specifying design criteria. It supplements commonly accepted RBC design procedures or approaches by providing appropriate qualifiers and/or information not readily available to the design community. The document seeks to address important design parameters and relationships (or lack of them) in order to provide a more rational RBC design approach. Topics considered are design loadings for carbonaceous removal, nitrification and denitrifiction, equipment reliability and service life, power requirements for air and mechanically driven units and structural design considerations such as flexibiity and hydraulics. The document attempts to provide practical usable design information rather than to emphasize theoretical considerations. The information in the document is intended to assist the design engineer by providing a more in-depth perspective on some of the key design considerations than is normally available in other design manuals.

b. The American Society of Civil Engineers (ASCE) has formed a "Rotating Biological Contactor Task Subcommittee." The subcommittee has prepared a report entitled "RBC for Secondary Treatment" which should be published in 1982.

c. The US Army Construction Engineering Research Laboratory (CERL) has prepared a report which provides assistance in determining when trickling filter plants can be effectively and economically upgraded using RBCs and which provides guidance in designing the RBCs.

d. USACERL will publish a lessons learned document based upon Dept. of Army and Corps of Engineers RBC applications at Fort Bragg, Fort Ritchie, Fort Knox, Jwalein Island, Korea, and Saudi Arabia.

e. The Corps of Engineers has sponsored a study dedicated to evaluating the potential of RBC's in recreational area applications.

f. Finally, this conference session devoted to RBC's is taking place. All the various sectors of the RBC community are represented in these meetings (academic/researchers, A/Es, manufacturers, government representatives, municipal and industrial engineers and operating personnel). My work in compiling these proceedings has convinced me that our meeting has already been productive. Much more will come of our personal interactions this week.

g. Rotating Biological Contactor related research reported in the technical literature is much more common during the last few years.

LITERATURE REVIEW

A literature search for 1980-1981 was performed which identified 126 studies. The following review provides information concerning various aspects of theory, design and operating experience associated with RBC systems.

The First National Symposium/Workshop on Rotating Biological Contractors held at Champion, PA, on February 4-6, 1980 more than doubled the literature of the technology(4). The proceedings are a compilation of the 68 papers delivered during the meeting and a transcription of the associated workshop. Eleven major topics are covered in the papers: perspective, overview, history, process variables and biofilm properties, municipal wastewater treatment, biokinetic studies, air drive and supplemental aeration, industrial wastewater treatment, concepts and models, upgrading

waste treatment systems with RBC's, design and operation, nitrification and dentrification, and selection and economics. Requirements for further research were identified at the workshop.

In September of 1980, Kneel and Godfrey announced in Civil Engineering a cooperative effort of the U.S. EPA and the ASCE to produce a new series of design books which would meet the twin goals of reducing the time required for new knowledge to be reflected in design manuals and of securing profession-wide peer review of design manuals as they are produced(5). One of the first manuals to be produced will cover rotating biological contactors.

Numerous papers have appeared since the First National Symposium in early 1980. Hitdlebaugh and Miller(6) discussed the operational problems of RBC's. Dehkordi(7) and Keihan:(8) described the effects of heavy metals upon RBC performance. Trinh(9), Allen(10) and Bauer et al.(11) assessed the applicability of RBC's for remote or on-site applications. Mueller et al.(12) discussed the impact of mass transfer considerations upon RBC and trickling filter design. Factors to be considered in scaling up were identified by Wilson et al.(13). Kinetics for domestic wastewater treatment were explored by Pano(14).

Reports of RBC applications to the secondary treatment of domestic wastewater continued to appear. Regent(15) reported several years of successful RBC operation in Yugoslavia. Interestingly, he described no mechanical failures. Spink(16) described the role of RBC's in the Province of Alberta, Canada. Rushbrook and Wilke(17) described an innovative treatment facility in Hillsborough, NH which will include RBC's, solar-heated anaerobic digestion and methane recovery. Shifts in sewage solids distribution across RBC installations were studied by Nunch et al.(18). Sapinsky(19) emphasized the importance of energy conservation in wastewater treatment and cited RBC plants at Hillsborough NH, Minneapolis and Chicago for their efficient use of energy.

Some interesting process modifications were explored as were some unusual applications of RBC technology at conventional wastewater treatment plants. Given(20) reported on the RBC treatment of dilute wastewater. Huang and Bates(21) compared RBC treatment of a synthetic milk waste using air and pure oxygen. RBC's were used in an innovative anaerobic treatment system for high strength carbonaceous wastes by Tait and Friedman(22). Cheung and Krauth(23) investigated

the feasibility of replacing conventional sedimentation by microstrainers in the RBC system. Hong(24) evaluated the use of RBC's in treating aluminum sulfate coagulated septage supernatant.

The use of RBCs for tertiary treatement continued to develop. Noss and Miller(25) described the use of an RBC for secondary treatment and recarbonation following low-level lime addition for phosphorus removal. The effects fo nitrate concentration and retention period upon RBC denitrification were investigated by Cheung and Krauth(26). Stephenson and Murphy(27) characterized the kinetics of denitrification in a biological fluidized bed. Buckingham(28) performed an engineering and marketing analysis of the rotating disk evaporator, a device physically similar to RBC which is designed to evaporate wastewater rather than biologically treat it.

Numerous nitrification studies have appeared since early 1980. Wu et al.(29) used data from many previous studies to derive and validate a model for the prediction of RBC nitrification performance. Mueller et al.(30) developed and verified a steady state model of nitrification and organic carbon oxidation in the RBC. Smith et al.(31) evaluated RBC's as an upgrading-retrofit process for BOD reduction and nitrification. Bridle(32) discussed RBC's in the context of biological processes for nitrogen conversion along with other processes capable of achieving the same ends. The kinetics of the nitrification process were modeled by Watanabe et al.(33) and by Margaritas et al.(34). Stratta(35) investigated the feasibility of enhancing nitrification by controlling the pH in RBC's. Marsh et al. described a coupled trickling filter - RBC nitrification process(36).

Additional nutrient removal work included the investigation by Knoetze et al.(37) into chemical inhibition of biological removal processes. Murphy and Wilson(38) performed pilot plant studies of BOD removal, nitrification and phosphorus removal. Singhal(39) described RBC nitrification at an advanced wastewater treatment plant in Cadillac, Michigan. An energy efficient extension to the Guelph, Ontario wastewater treatment plant was described(40). This plant effectively removes BOD, ammonium-nitrogen and phosphorus with RBC's followed by filtration.

The feasibility of using RBC's to upgrade existing plants was explored in several studies. Gutierrez et al.(41) evaluated upgrading primary tanks with RBC's. Smith

et al.(42) considered RBC's as an upgrade for existing trickling filter plants. Poon et al.(43,44) evaluated the effectiveness of RBC's in supplementing the BOD and ammonium nitrogen removals achieved at trickling filter plants. The Surfact process developed by the Philadelphia Water Department was described by Guarino et al.(45). The Surfact process which physically merges an RBC with a diffused aeration tank provides an inexpensive upgrade. Very little construction is required.

A final area of activity has involved industrial or primarily industrial wastewaters and rotating biological contactors. Chesler and Eskelund(46) evaluated RBC's for the treatment of explosives manufacturing wastes. Acid mine wastes were treated in pilot scale and prototype studies conducted by Olem and Unz(47) at Hollywood, PA. Dairy wastes were treated in an innovative process involving an aerated equalization tank and RBC's by Waggener et al.(48). Suria Pandian and Agarwal(49) also described RBC treatment of dairy wastes. The city of Monett, Missouri, overcame problems posed by industrial discharges to its sewage plant equivalent to 7 times its population by using RBC's(50). O'Shaughnessy et al.(51) applied RBC's to oil shale retort wastewater. Blanc et al.(52) evaluated RBC's for the treatment of beef slaughtering and processing wastewaters. The influence of the rotational speed of RBC's on the reaction rates observed was investigated by Odai et al.(53). Borghei(54) described treatment of the effluent of a glucose-production plant using a rotating biological packed bed.

The emphasis of a report by Chesner and Bender (55) was to review and compare current design procedures and performance capabilities of the RBC process. This was accomplished by a review of the literature, an evaluation of the process, O&M, equipment and power performance at RBC plants approaching design flow conditions and a comparison of current design guide information.

> Sixteen domestic RBC facilities providing carbonaceous BOD removal and approaching design flow conditions, supplied monitoring data that were used to evaluate process performance. The reliability of these systems in meeting effluent concentration and removal efficiency criteria, defined by NPDES as 30 mg/L BOD effluent concentration and 85 percent BOD removal efficiency,

respectively, was evaluated. The results indicated that the plants exceeded effluent criteria 12 percent of the time and failed to meet percent BOD removal 67 percent of the time. An analysis of performance data demonstrated that average values for both mass BOD removal rates (lbs BOD removed/day/1,000 sf of media) and BOD removal efficiencies increased with increasing influent waste strength. For the range of conditions found at the plants surveyed, RBC process performance followed design predictions for mass BOD removal rates and percent BOD removals for high wastewater influent strength (175 to 350 mg/L BOD), and progressively lagged behind those predictions as waste strength decreased below 175 mg/L.

Low labor requirements to operate and maintain an RBC secondary treatment unit are attractive features of an RBC system. Hourly labor requirements were reported in the range of 1 to 7 hours per week, averaging 2.6 hours per week for 23 plants, with an average design flow of 1.4 mgd. Power measurements were performed during the course of this investigatin to identify RBC energy consumption. The results established power consumption to rotate 100,000 sf of standard density media to be 3.46 kw for mechanical drive (1.6 rpm) and 2.93 kw for air drive (1.2 rpm). To rotate 150,000 sf/shaft of high density media at 1.6 rpm mechanical units used 3.77 kw of power.

Equipment performance is a severe problem in existing RBC systems. The nature of the problem centers on shaft failures and media degradation. Of the plants surveyed there were 12 shaft failures reported and the media in three plants had become brittle or failed due to shifting. As a result of this poor operating history it was concluded that design engineers should seek an RBC eqiupment warranty sufficient to protect the owner against equipment failures (55).

REFERENCES

1. U.S. EPA, "Process Control Manual for Aerobic Wastewater Treatment Facilities," EPA-403/9-77-006, (1977).

2. U.S. EPA, Office of Water Program Operations (WH-547), Washington, DC, 20460, EPA-430/9-78-004, MCD-42, (1978).

3. U.S. EPA Technology Transfer Publication, "Process Design Manual for Upgrading Existing Wastewater Treatment Plants," (1974).

4. Wu, Y. C.; Smith, E. D.; Opatken, E. J.; Miller, R. D.; Borchardt, J. A.; Proceedings: First National Symposium/Workshop on Rotating Biological Contactor Technology. Champion, PA; FEbruary 4-6, 1980. Sponsored by University of Pittsburgh, Municipal Environmental Research Laboratory, Cincinnati, Ohio and U.S. Army Construction Engineering Research Laboratory, Champaign, Illinois. June 1980. Volumes I and II.

5. Godfrey, Kneeland A. Jr., "New Developments in Wastewater Treatment -- EPA-ASCE Design Books Planned", Civil Engineering (NY) 50(9) 96-101 (1980).

6. Hitdlebaugh, J. A. and R. D. Miller, "Operational Problems with Rotating Biological Contactors", Journal Water Pollution Control Federation 53(8) 1283-1293 (1981).

7. Dehkordi, F. G., "The Effect of Heavy Metal on the Performance of Rotating Biological Contactors (RBC)", Master's thesis, Oklahoma State University, OWRT-A-087-OKLA(3), (1980).

8. Keihani, A., "Long Term Effects of Chromium and Copper on the Rotating Biological Contactor", Master's thesis, Oklahoma State University, OWRT-A-087-OKLA(2), (1980).

9. Trinh, D. T., "Exploration Camp Wastewater Characterization and Treatment Plant Assessment," Technol Dev. Rep. EPS No. 4 (1981).

10. Allen, G. A., "Methods Used to Provide Services to Outlying Mine Townsites", Water and Pollution Control 118(2) 18-19, (1980).

11. Bauer, D. H.; Conrad, E. T. and D. G. Sherman, "Evaluation of On-site Wastewater Treatment and Disposal Options", SCS ENGINEERS, EPA-600/2-81-178, (1981).

12. Mueller, J. A., Paquin, P. and J. Famularo, "Mass Transfer Impact on RBC and Trickling Filter Design", AIChE Symposium Series No. 197, Vol. 76, (1980).

13. Wilson, R. W.; Murphy, K. L., and J. P. Stephenson, "Scale Up In Rotating Biological Contactor Design", Journal Water Pollution Control Federation 52(3) 610-621, (1980).

14. Pano, Abraham, "The Kinetics of Rotating Biological Contactors Treating Domestic Wastewater", doctoral dissertation, Utah State University, (1981).

15. Regent, Aleksander, "Small RBC's Logging Hours in Yugoslavia", Water and Sewage Works 127(8) 42, (1980).

16. Spink, D., "Getting the Treatment", Env. Views 3(1)7, (1980).

17. Rushbrook, E. L. and D. A. Wilke, "Energy Conservation and Alternative Energy Sources in Wastewater Treatment", Journal Water Pollution Control Federation 52(10) 2477-2483, (1980).

18. Munch, R.; Hwang, C. P. and T. H. Lackie, "Wastewater Fractions Add to Total Treatment Picture", Water and Sewage Works 127(12) 49-54, (1980).

19. Sapinsky, C. P., "Energy Conservation is a Dire Necessity", Water and Wastes Engineering 17(8) 28-32, (1980).

20. Given, P. W., "RBC Treatment of Dilute Wastewater", Annual Conf. of the Water Pollution Control Federation, 53rd, Proceedings of the Research Symposium, Las Vegas, Nevada, WPCF, (1980).

21. Huang, J. C., and V. T. Bates, "Comparative Performance of Rotating Biological Contactors Using Air and Pure Oxygen", Journal Water Pollution Control Federation 52(11) 2686-2703, (1980).

22. Tait, S. J. and A. A. Friedman, "Anaerobic Rotating Biological Contactor for Carbonaceous Wastewaters", Journal Water Pollution Control Federation 52(8) 2257-2269, (1980).

23. Cheung, P. S. and K. Krauth, "Investigation to Replace the Conventional Sedimentation Tauk by a Microstrainer in the Rotating Disk System," Water Res 14(1) 67-75, (1980).

24. Hoag, G. E., "Rotating Biological Contactor Treatment of Aluminum Sulfate Coagulated Septage Supernatant", Master's thesis, University of Lowell, (1980).

25. Noss, C. I. and R. D. Miller, "Rotating Biological Contactor Process for Secondary Treatment and Recarbonation Following Low-Level Lime Addition for Phosphorus Removal," Final Report, Army Medical Bioengineering Research and Development Laboratory, Fort Deitrick, MD, USAMBRDL-TR-8007, (1980).

26. Cheung, P. S. and K. Krauth, "Effects of Nitrate Concentration and Roetention Period on Biological Denitrification in the Rotating-Disc System," Water Polution Control (London) 79(1) 99-105, (1980).

27. Stephenson, J. P. and K. L. Murphy, "Kinetics of Biological Fluidized Bed Wastewater Denitrification," Tenth Int. Conference International Association for Water Pollution Research, Toronto, Ontario, June 23-27, 1980 Prog. Water Tech 12(6).

28. Buckingham, P. L., "Production Engineering and Marketing Analysis of the Rotating Disk Evaporator", Municipal Environmental Research Laboratory, Cincinnati, Ohio, EPA-600/2-81-179, (1981).

29. Wu, Y. C.; Smith, E. D. and John Gratz, "Prediction of RBC Performance for Nitrification", Journal Environmental Engineering Division, ASCE 107(4) 635-652, (1981).

30. Mueller, J. A., Paquin, P., and J. Famularo, "Nitrification in Rotating Biological Contactor", Journal Water Pollution Control Federation 52(4) 688-710, (1980).

31. Smith, E. D.; Poon, C. P. C.; Mikucki, W. and J. T. Bandy, "Tertiary Treatment of Wastewater Using A Rotating Biological Contactor System", U.S. Army Construction Engineering Research Laboratory, Champaign, IL, CERL Tech. Report N-85, (1980).

32. Bridle, T. R., "Fundamentals of Biological Processes for Nitrogen Conversion," Presented at Env. Canada et al Nutrient Control Technology Conference, Calgary, February 7-8, 1980, P7-A(42).

33. Watanabe, Y., Ishiguro, M., and K. Nishidome, "Nitrification Kinetics in a Rotating Biological Disk Reactor," Tenth Int. Conference International Association for Water Pollution Research, Toronto, Ontario, June 23-27, 1980, Prog. Water Tech. 12(6).

34. Margaritas, A., Watanabe, Y., Ishiguro, M., Nishidome, K. and P. Harremoes, "Nitrification Kinetics in a Rotating Biological Disk Reactor", Water Science and Technology 13(4-5) 1219-1225, (1981).

35. Stratta, J. M., "Nitrification Enhancement Through PH Control With Rotating Biological Contactors", doctoral dissertation, Pennsylvania State University, (1981).

36. Marsh, D., Benefield, L., Bennett, E., Lindstedt, D. and R. Hartman, "Coupled Trickling Filter-Rotating Biological Contactor Nitrification Process," Journal Water Pollution Control Federation 53(10) 1469-1480, (1981).

37. Knoetze, C., Davies, T. R., and S. G. Weichers, "Chemical Inhibition of Biological Nutrient Removal Processes," Water SA 6(4) 171, (1980).

38. Murphy, K. L., and R. W. Wilson, "Pilot Plant Studies of Rotating Biological Contactors Treating Municipal Wastewater", Env. Canada Env. Protection Service Report SCAT-2, (1980).

39. Singhal, A. K., "Phosphorus and Nitrogen Removal at Cadillac, Michigan," Journal Water Pollution Control Federation 52(11) 2761, (1980).

40. Anonymous, "Innovative Nutrient Removal Process Attracts International Attention," Water and Pollution Control 118(10) 14-15, (1980).

41. Gutierrez, A., Bogart, I. L., Scheibe, D. K., and T. J. Mulligan, "Upgrading Primary Tanks With Rotating Biological Contactors", Municipal Environmental Research Laboratory, Cincinnati, Ohio, EPA-600/2-80-003, (1980).

42. Smith, E. D., Poon, C. P., and R. D. Miller, "Upgrading DA Trickling-Filter Sewage Treatment Plants," U.S. Army Construction Engineering Research Laboratory, Champaign, IL, CERL Tech. Report N-102, (1981).

43. Poon, C. P. C., Chin, H. K., Smith, E. D., and W. J. Mikucki, "Upgrading With Rotating Biological Contactors for BOD Removals," Journal Water Pollution Control Federation 53(4) 474-481, (1981).

44. Poon, C. P. C., Chin, H. K., Smith, E. D., and W. J. Mikucki, "Upgrading With Rotating Biological Contactors For Ammonia Nitrogen Removal," Journal Water Pollution Control Federation 53(7) 1158, (1981).

45. Guarino, C. F., Nelson, M. D., and T. E. Wilson, "Uprating Activated-Sludge Plants Using Rotary Biological Contactors," Water Pollution Control 79(2) 255, (1980).

46. Chesler, P. G. and G. R. Eskelund, "Rotating Biological Contactors for Munitions Wastewater Treatment," Final Technical Report, Army Mobility Equipment Research and Development Command, Fort Belvoir, VA, MERADCOM-2319, (1981).

47. Olem, H. and R. F. Unz, "Rotating Disc Biological Treatment of Acid Mine Drainage," Final Report, Industrial Environmental Research Laboratory, Cincinnati, Ohio, EPA-600/7-80-006, (1980).

48. Waggener, J. E., Fitzhugh, M. L., and G. E. Flann, "Innovative Approach to the Treatment of Dairy Wastewater with Rotating Biological Contactors," 53rd Annual Conference of the Water Pollution Control Federation, Proceedings of the Industrial Wastes Symposium, Las Vegas, Nevada, Sep 28-Oct 3, 1980, Published by WPCF, (1980).

49. Suria Pandian, P. S. and I. C. Agarwal, "Removal of Dairy Waste Organics by Rotating Biological Contactors," Indian Journal of Environmental Health 23(1) 27, (1981).

50. Riddle, W. G., "Small City Requires Large WWTP," Water and Wastes Engineering 17(7) 42-44, (1980).

51. O'Shaughnessy, J., Blanc, F. C., Wei, I. W., and J. Patinskas, "Biological Treatment of Oil-Shale Retort Wastewater Using Rotating Biological Contactors," Abstracts of Papers of the American Chemical Society, Vol. 181, p. 76, (1981).

52. Blanc, F. C., O'Shaughnessy, J. and S. H. Corr, "Treatment of Beef Slaughtering and Processing Wastewaters Using Rotating Biological Contactors," Abstracts of Papers of the American Chemical Society, Vol. 181, p. 74, (1981).

53. Odai, S., Fujie, K., and H. Kubota, "Effect of Rotation Speed on Reaction-Rate on a Rotating Biological Contactor," Journal of Fermentation Technology 59(3) 227-234, (1981).

54. Borghei, S. M., "Treatment of the Effluent of a Glucose-Production Plant Using a Rotating Biological Packed Bed," Process Biochemistry 16(2) 29, (1981).

55. Chesner, W. H. Iannone, J. and J. Bender, "Review of Current RBC Performance and Design Procedures," Report prepared for th Municipal Environmental Research Laboratory Office of Research and Development USEPA 22 W. St. Clair Street, Cincinnati, OH 45268 (contract No. 68-02-2775), March 1981.

ANAEROBIC ATTACHED FILM EXPANDED BED FUNDAMENTALS

William J. Jewell, Department of Agricultural Engineering, Cornell University, Ithaca, New York

ABSTRACT

The anaerobic attached film expanded bed process (AAFEB) has been shown to be capable of treating low strength wastewaters at low temperatures and at relatively short retention times. Such capability leads to the unexpected conclusion that the AAFEB is a municipal wastewater treatment alternative capable of meeting secondary effluent quality without producing a secondary waste sludge. In order to understand reasons for this extraordinary capability, recent research has focussed on reproducing the phenomena, evaluating the kinetics with soluble and insoluble substrate at varying temperatures, testing of the system under shock loads, and evaluation of the potential applications (algae harvested or operation under thermophilic temperatures, 50°C). This paper will relate the research data to process fundamentals (active biomass, solids retention time, substrate kinetics) and design requirements.

INTRODUCTION TO THE ANAEROBIC EXPANDED BED PROCESS

An anaerobic biological treatment process capable of treating dilute domestic sewage to secondary quality without the production of waste secondary sludges, and processing

capability superior to aerobic biological systems are among the characteristics suggested by the results of studies on the anaerobic attached microbial film expanded bed reactor (AAFEB) (1, 2). These surprising results have been obtained from nearly a decade of research and development efforts on a new process that has attempted to optimize the capability of the anaerobic fermentation process for wastewater treatment (3, 4, 5, 6).

The anaerobic methane fermentation process has been applied for many decades to waste management, as has been the expanded bed process. The definition of each unit process is well-known, but the combination of these two unit operations into one process has only recently been achieved. The application of the expanded bed physical process as a biological converter appears to be an optimum method of achieving fine solids separation and microbial conversions.

Goals and Objectives

The main goal of this paper is to synthesize the basic fundamentals of anaerobic biological processes and the characteristics of the physical expanded bed to illustrate the basis for the AAFEB process. The specific objectives are to review the physical considerations required in operating the expanded bed, to summarize the biological capability, to compare the resulting process to other attached film processes, and finally to briefly consider future research and development efforts required to clarify the process capabilities.

Background

The conceptual diagram of the expanded bed process is shown in Figure 1. It is a fine particle upflow filter in which the particles are slightly expanded but remain in close proximity to other particles. The basic mode of operation is similar to packed filters and fluidized processes. These similarities have led to some confusion between the processes and the terminology used to describe each.

A comparison of these various processes that use inert packing material is shown in Figure 2. The static filters, or packed filters, were developed in 1963 by Young and McCarty (7). By decreasing particle size and increasing upflow velocities, there will be a point at which the particles begin to be lifted in a slightly expanded form. The relationship between the porosity of the bed (the ratio of the void space to the total volume of the bed) and the head loss that occurs

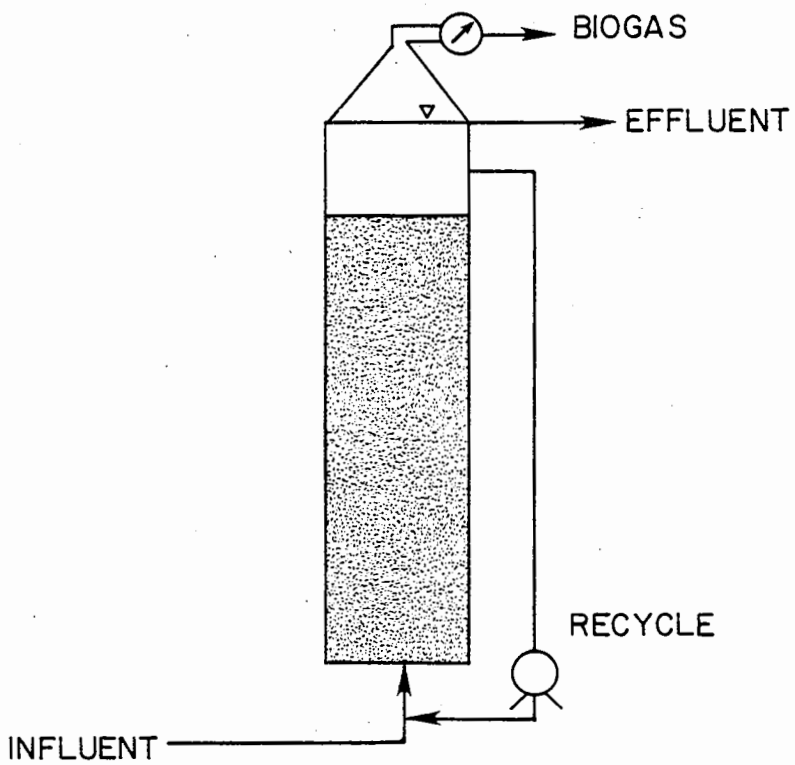

Figure 1. Schematic diagram of the attached microbial film expanded bed process.

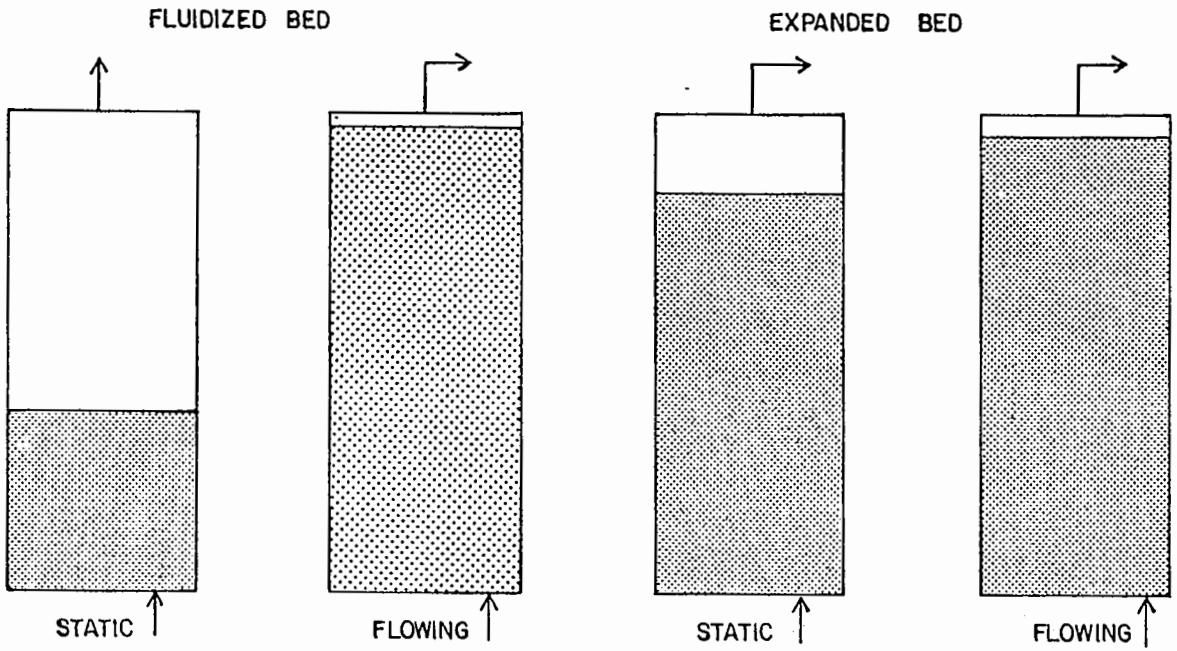

Figure 2. Qualitative comparison of reactor volumes occupied by media in typical fluidized and expanded beds.

in these units has been well defined for process applications such as water filter backwashing (8, 9).

The general relationships between the upflow rate and the bed characteristics are summarized in Figure 3. At low fluid velocities, the particles remain in contact or in packed form. As the velocity of the liquid is increased, the particles' resistance causes them to be slightly expanded or "fluidized." The operation of the expanded bed is most effective when the expansion is limited to a small fraction of the packed bed volume. This requirement enables the bed to inhibit the flow of fine solids through the filter but to avoid clogging, which occurs in a packed bed. Further increases in velocity cause further separation of the bed. As the velocity is further increased, the individual particles separate, and true fluidization begins. All particles are in motion, and the bed continues to expand; and particles move in more rapid and more independent motion. The bed continues to expand as the velocity is increased and maintains a uniform character. Particles move in random directions through all parts of the liquid at this state. Strong transient currents with many particles temporarily traveling in the same direction can be observed, but in general, particles move randomly as individuals. This phenomenon is known as particulate fluidization. It is the common state for fluidized processes in which the bed is fluidized up to 300 percent or more of its static, packed bed form.

Eventually, as the upflow of the velocities increases, the superficial velocity approaches the terminal settling velocities of particles, and the particles become entrained in the liquid and are carried out of the reactor. Thus, in relation to the diagram shown in Figure 3, the expanded bed operates as close to the fixed bed characteristics as possible, whereas the fluidized bed often operates at much higher superficial velocities further to the right of the abscissa on the diagram.

A comparison of anaerobic and aerobic microbial processes can be made if the capability of microorganisms is known under ideal conditions and these characteristics can be adjusted for the application to specific processes. Attached film processes complicate the comparison because of the increased influence of mass transfer limitations. The main parameters would be the temperature of operation, the concentration of substrate required to achieve a given removal rate, and microbial yield. Microbial yield is one of the most important criteria for comparison since it also is related to the minimum sludge retention time (SRT_{min}) that can be achieved in a biological

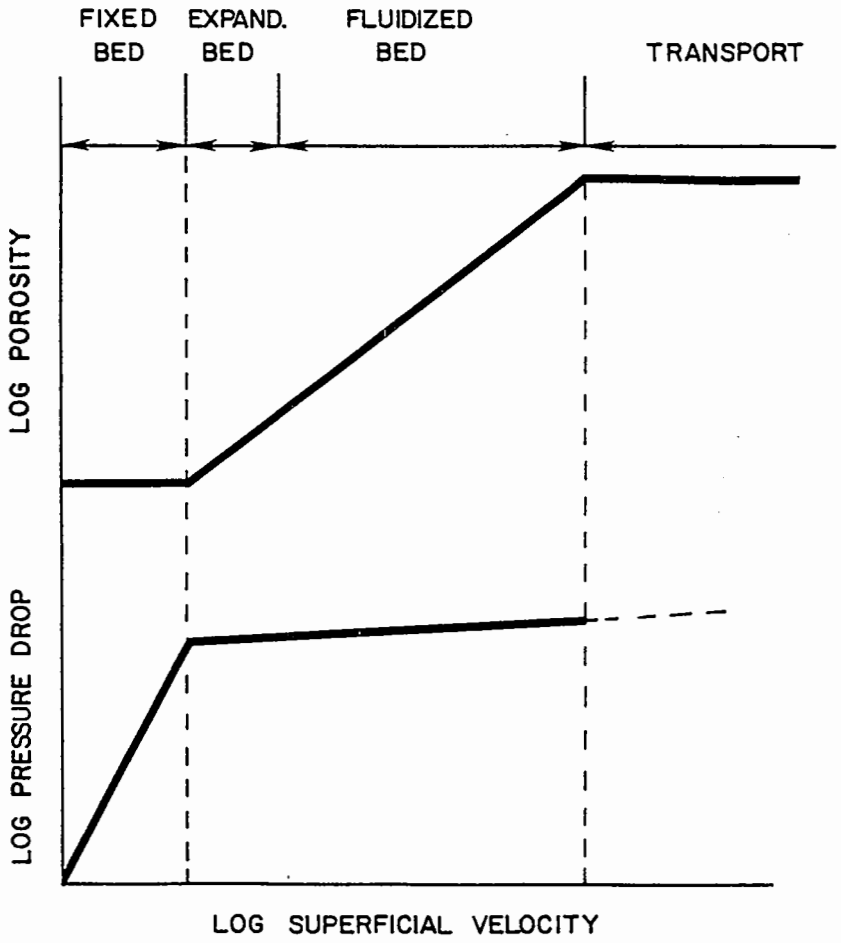

Figure 3. Effect of increasing upflow velocities with a filter of small particles on the friction losses and the void space or filter porosity.

process. The microbial solids retention time ultimately defines both the stability of the process as well as the safety factor under which the process is designed or operating. A summary of example values for anaerobic and aerobic treatment systems is given in Table 1. These data emphasize several well known process differences. The high reproduction rates of aerobic organisms lead one to conclude that they have significantly more capability for substrate removal and require a much smaller reactor volume than anaerobic processes. Minimum solids retention times in treating soluble substrates significantly greater than 10 days at an operating temperature of 25°C would not be unusual for a conventional anaerobic treatment process. The relationship of the solids retention time to the microbial mass in the system is given by the following equation:

$$SRT = \frac{X_o \cdot V}{X_e}$$

where SRT = the solids retention time
V = reactor volume
X_o = bacterial concentration in the reactor
X_e = biomass lost from the reactor in the effluent or intentionally wasted each day

Whenever a process operates at a sludge retention time less than the microorganism reproduction time, it is in the process of failing and/or going through a change to a situation where the process efficiency is changing as the microbial mass adjusts its concentration. A simple method of reviewing the process capability under a given set of operating conditions is to determine the solids retention time that can be achieved by various processes and compare it to the minimum acceptable with the system. This will be done in an example later.

EXPANDED BED PROCESS DEFINITION

The development of the expanded bed process is based on efforts to optimize the conditions required to achieve maximum microbial concentrations while good control over the microbial biomass in the fluid media is maintained. The goal of a biological process is to minimize the cost of the system that is designed to achieve a specific purpose. The maximum conversion rate per unit volume of reactor will lead to lower

Table I. Kinetic Coefficients for Anaerobic and Aerobic Treatment Systems

Biological Systems	Temp. °C	K_s Half Rate Coefficient mg/ℓ	Microbial Yield, mg VSS / mg Substrate	Minimum Cell Residence Time, Days	Coefficient Basis	Reference
Anaerobic Systems						
1. Acetic Acid	25	869	0.054	4.2	Acetic Acid	15
2. Acetic Acid	35	159	0.044	3.1	Acetic Acid	15
3. Milk Waste	25	24	0.370	---	COD	16
Aerobic Systems						
1. Domestic Waste	20	22	0.670	0.27	COD	17
2. Skim Milk	20	100	0.48	0.42	BOD_5	18

costs, and this should minimize the total treatment system cost. It follows that the achievement of maximum removal rates can be obtained by either using superior microorganisms or higher microbial mass concentrations. Since we have few opportunities to select microorganisms in waste management, the emphasis has focused on maximizing microbial concentrations. However, it is essential that the microbial mass be "active"; that is, that it be exposed to an available substrate in the bulk liquid. Therefore, the first part of the definition of the optimization of a process is that it must achieve a maximum active biomass concentration.

Of course, achievement of maximum biomass is only the first step in designing an optimized biological process. The second requirement for the system is to be able to achieve efficient, reliable management of microorganisms. A process that clogs or accumulates thick films through which substrate cannot penetrate is not acceptable. The approach that was taken with the development of the expanded bed was to first define a process that would achieve a maximum biomass per unit volume and then superimpose these requirements on the physical requirements that are necessary to operate the hydraulic flow regime.

Design Requirements for Maximum "Active" Microbial Concentrations

There are a number of factors to be considered in defining the maximum biomass accumulation potential of the expanded bed--mass diffusion characteristics of soluble and particulate organics, microbial growth rates, substrate requirements, and process kinetics. The growth kinetics and process requirements can be assumed to be similar to those shown in Table I in the absence of mass diffusion limiting processes. These emphasize the problem of low substrate removal rates whenever the substrate concentration is low and the requirement for long solids retention times.

It is well known that methane-producing bacteria are amongst the slower growing bacteria and at 35°C, under optimum conditions, have a maximum reproduction time of three to four days, as shown in Table I. Of course, temperatures less than 20°C are often experienced with domestic sewage, and the microorganism reproduction time may have to be significantly greater than 10 to 30 days under the cooler winter temperatures.

Two questions represent the challenge in the understanding of the process biomass requirements: (1) what are the

conditions required to accumulate a maximum active biomass in the system? and (2) how would we manage those bacteria such that they would be exposed to the substrate at maximum flow rates and still maintain control over the bacteria? Without the addition of inert particles and the growth of attached microbial film, it is obvious that there are a limited number of ways to try to increase the biomass concentration in the reactors. It would appear that reactors without inert media are limited to somewhere around 5 gm VS/ℓ unless we go to highly elaborate methods of maintaining bacteria within the system.

The depth of substrate penetration to a microbial floc or an attached film is well known for the lower substrate concentrations that are common in sewage (10, 11, 12). Substrate diffusion depths exceeding 60 microns occur at relatively low substrate concentrations. In an aerobic film LaMotta (10) showed that 5.2 mg/ℓ of glucose penetrated to greater than 10 microns. These diffusion-limited depths indicate that attached films thicker than 0.05 mm would result in some substrate-limited biomass, especially where the substrate in solution is low.

The substrate diffusion depth limitations for optimum microbial particle dimensions can be estimated as follows:

$$\Phi_M = 2\ (S_D) + \Phi_D$$

where Φ_M = maximum total particle diameter of the inert particle and the attached microbial film

S_D = substrate diffusion depth

Φ_D = inert particle diameter

The background development work at Cornell University has developed two surprising results in relation to methane-forming bacterial films. The thickness of the film is thin and usually around 0.020 mm. This unexpectedly thin film has a limited impact on particle management and indicates that all of the attached film will be active since the substrate diffusion depths, even at low bulk solution concentrations, will be no greater than 0.05 mm. Thus, the optimum inert particle diameter is exceptionally small, being around 0.02 mm. However, as will be seen, this particle is so small that it is difficult to manage with the practical hydraulic retention times.

The second major observation has been that the bulk density of these thin films is much higher than expected, being greater than 200 gm/ℓ of film in some cases. This compares to aerobic films that have densitites of about 34 gm VS/ℓ. A comparison of the particle size and the active biomass achieved with the particles, assuming that they all achieve a thin microbial film and bulk density as indicated above, is given in Table II. Maximum biomass concentrations that have been observed in expanded beds have exceeded 40 gm/ℓ. Data in Table II show that the goal for the expanded bed should be the achievement of as much as 100 gm/ℓ of active biomass.

Table II. Effect of Inert Particle Size On Maximum Microbial Mass Goal.

Particle Description	Size Particle mm	Area per Volume of Reactor cm^2/cm^3	Active Biomass As Volatile Solids gm/ℓ
None (activated sludge)	--	--	2
Large rocks	50 to 75	1.0	2
Plastic media	25	5.0	7
Coarse sand	0.2 to 2	21.0	16
Fine sand	0.02 to 0.2	210.0	150

Microbial Management Requirements

The velocities in the expanded bed and at the top of the bed should be less than or equal to those required for microbial management if we want to achieve maximum solids capture and separation with the process. Typical clarifier overflow rates are approximately 0.7 gal/min/ft^2. This is equivalent to an upflow velocity of approximately 6 ft/hr. Flocculated microbial particles or films that have been scraped off the inert particles would settle at velocities higher than these clarifier overflow rates. Thus, we would expect if the process could be designed at these lower velocities, any solids passing through the bed or escaping from the films would

collect at the top of the expanded bed. These solids could be recycled or removed at this point.

Bed Expansion Requirements

The liquid velocity required for bed expansion is a function of its viscosity and density and the particle size, shape, and specific gravity. Numerous attempts have been made to estimate expansion velocity and requirements. Most of these efforts have been directed at backwash requirements for various physical filters. The author is unaware of any specific work that has been completed with inert particles coated with mature microbial films. In the case of the expanded bed, it is possible to use the theory as developed for bed expansion for backwashing since the microbial films are thin and insignificant in most cases. Figure 4 summarizes example interactions between the physical factors controlling bed expansion at 25°C and the particle diameter. This figure also contrasts the unhindered terminal settling velocity to the superficial upflow velocity required for expansion for the specific gravity particles of 2.65. These data indicate that the lower densities (1.2 specific gravity) have upflow expansion velocities in the region where solids management is compatible with requirements. This is achieved with particle diameters between 0.1 and 0.4 mm. The operating zones for fluidized beds and expanded beds are qualitatively indicated on this diagram, indicating that the higher velocity requirements for the fluidized bed achieve a shorter hydraulic retention time. These reactors also require higher specific gravity particles. For example, an expansion velocity of 60 ft/hr is required for a particle diameter of 0.2 mm with sand. A 1 mm size sand particle requires velocities of 300 ft/hr or greater.

Head Loss Considerations

Head loss through an expanded bed or a fluidized bed is given by the general relationship equation as follows:

$$\Delta P = \frac{L\ (\sigma_s - \sigma)}{\sigma} \cdot (1 - \varepsilon)$$

where
- P = head loss through the filter
- L = bed height
- ε = porosity of the expanded bed
- σ_s = specific gravity of the particle
- σ = specific gravity of the fluid

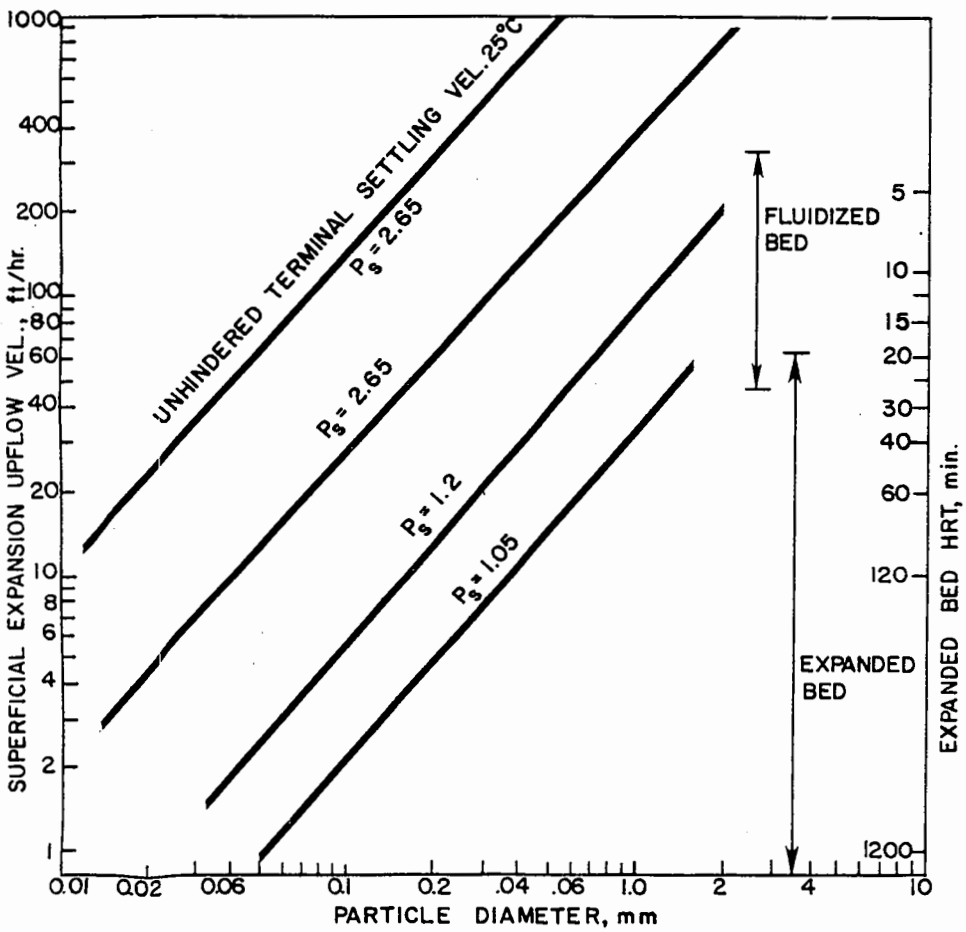

Figure 4. Relationship between spherical particle diameter, superficial upflow velocity required for bed expansion, particle specific gravity at 25°C.

In general, the friction pressure loss through media with a specific gravity similar to sand (2.65) results in head losses of 1 ft per ft of bed depth or greater, whereas the lower specific gravity particles required in an expanded bed results in much lower head losses, usually on the order of 1 in per ft of bed or less.

Expanded Bed Dimensions

Since no full scale expanded beds have been built to date, the information on the size of the system can be discussed in general terms only. Once the capability of the biological process is established and the particle management has been defined, the remaining concerns relate to volumetric requirements for flow and the dimensions of the unit. The relationship between substrate concentration, depth of the columns, the loading rate, and hydraulic retention time are illustrated in Figures 5 and 6. Since these are arithmetic relationships, they are only presented here as design guides for consideration of various processes. The height of the process is intimately involved in the particle selection and the bed management requirements. The shorter reactors would result in lower retention times at velocities that are acceptable for solids management. The typical range of upflow velocities in the fluidized bed tend to favor deeper beds. Of course, the relationship between velocity and depth can be changed by adding recycle to the system. The recycle requirements should be minimized and only utilized for bed management purposes. Note that at minimum flow requirements with the expanded bed it may be necessary to include a pumped recycle.

STATUS OF PROCESS DEVELOPMENT

Previous studies have focused on defining the characteristics of the attached film in relation to synthetic substrate concentration in sewage (1) and the effects of temperature (4), shock loadings (5), and particulates on the interaction (6). Ongoing studies are evaluating the thermophilic kinetics on both soluble and particulate substrates (13). Although a complete review of the kinetics of the process is beyond the scope of this paper, a review of selected data is included here to indicate the process capability. The relationship between temperature, substrate concentration, and process loading rate is summarized in Figure 7. These data show a wide scatter but indicate, as reported earlier by Switzenbaum (4), that the temperature effects are not as

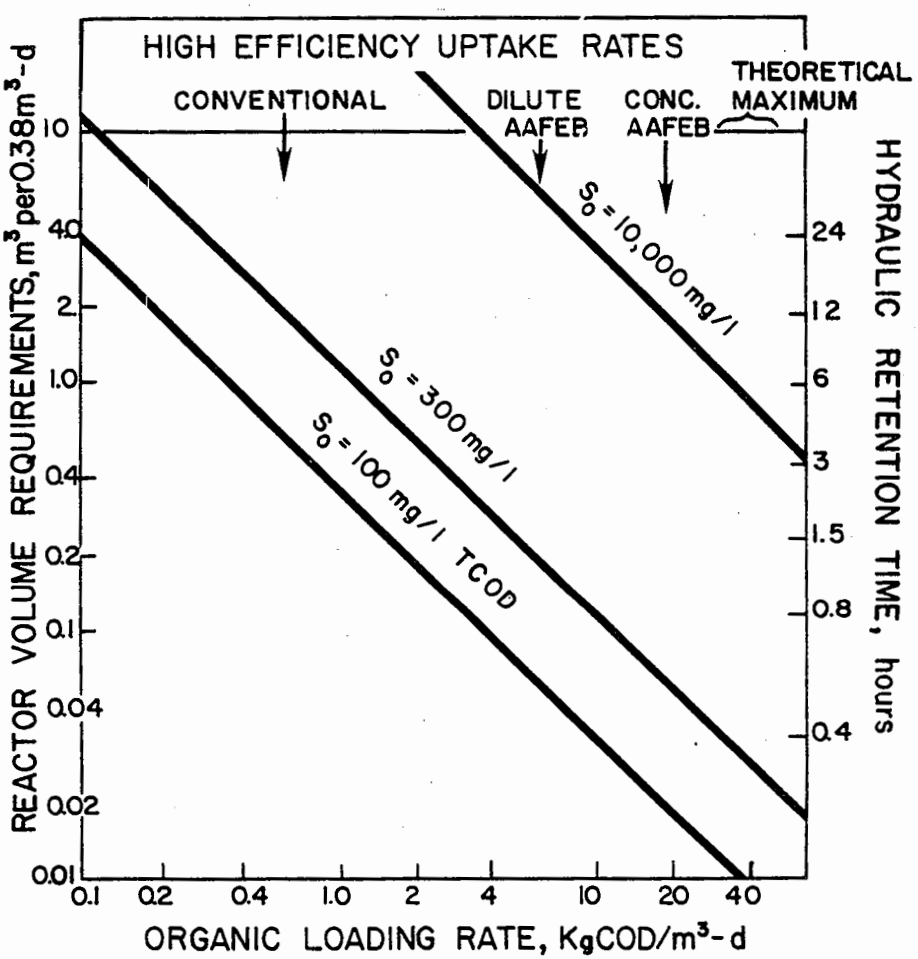

Figure 5. Relationship between process volumetric organic loading rate, reactor volume requirements, and hydraulic retention time for varying substrate concentrations.

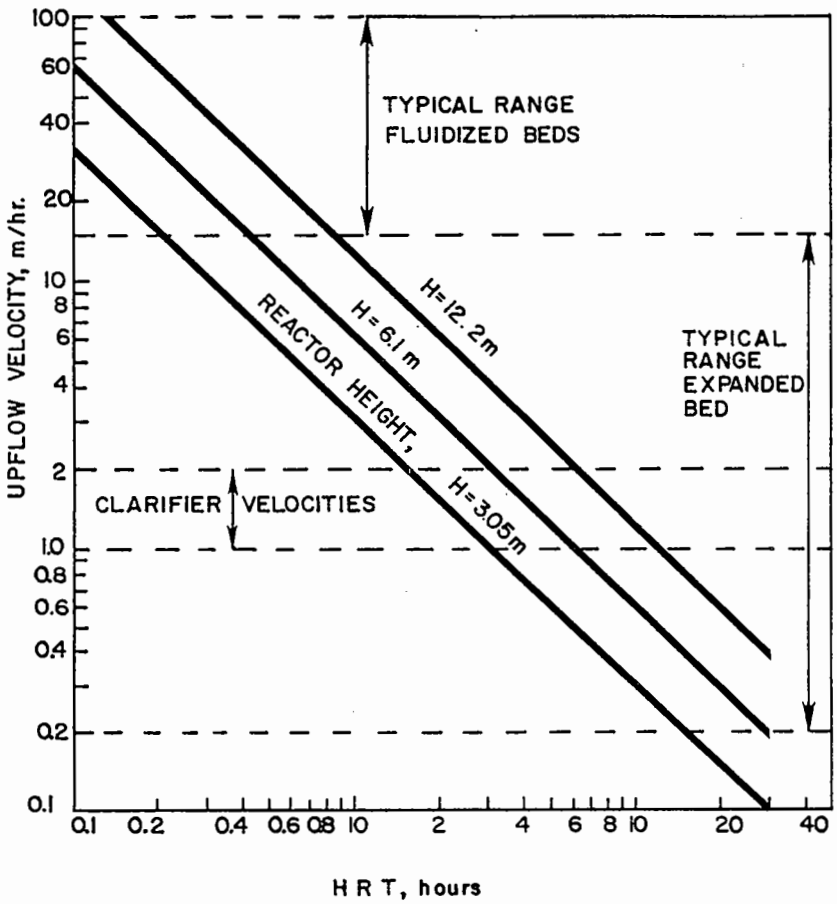

Figure 6. Relationship between reactor height, hydraulic retention time, and upflow velocity (superficial or empty bed basis).

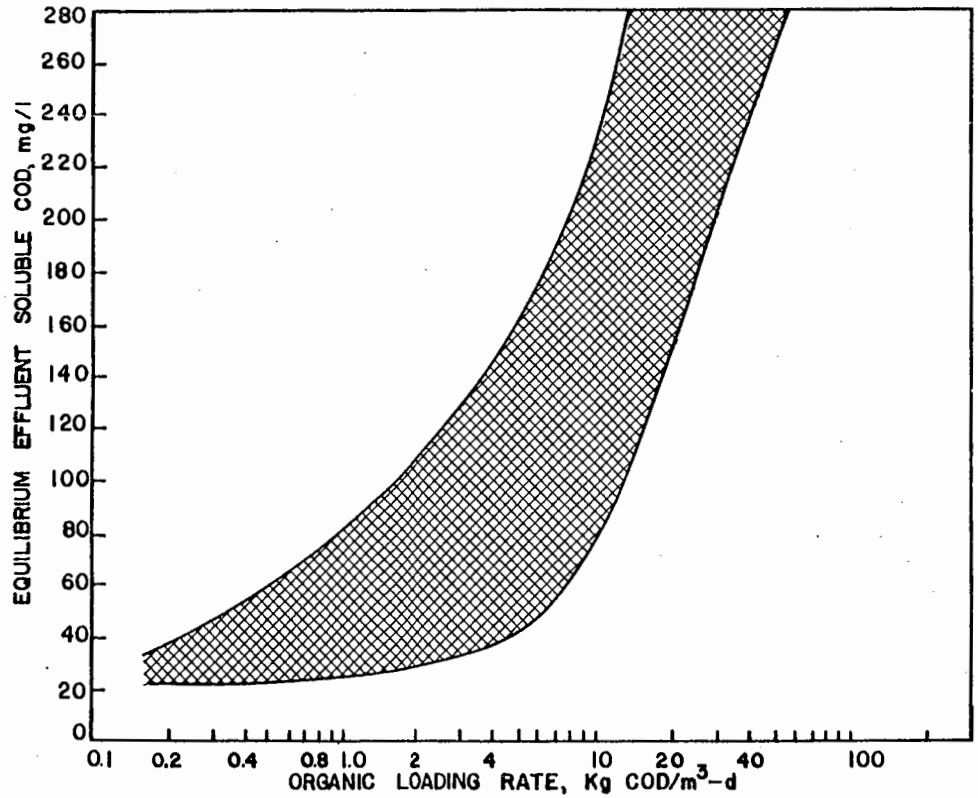

Figure 7. Organic loading rate effects on total effluent soluble COD for a wide range of steady state operating conditions (HRT values from 0.3 to 6 hr, 50 to 600 mg/ℓ influent TCOD, temperature 10°C to 30°C) from reference (4).

significant as one might expect with anaerobic processes applied to dilute wastewaters. Figure 8 contrasts the effluent when primary settled sewage was treated with the expanded bed to that reported in a pure oxygen fluidized bed study (14).

A study by Morris (6) focused on the interaction of particles in the expanded bed process. It was found that at 35°C pure cellulose particles loaded at a reactor loading rate of less than 7 kg/m^3/day resulted in a total effluent COD concentration of less than 60 mg/ℓ. Thus the loading rate and effluent quality relationship shown in Figure 7 for soluble organics also appears to hold true for particulate matter.

Although much additional work is required to define the detailed kinetics of the expanded bed process, Switzenbaum (4) showed that a highly simplified equation could describe the biological reaction rates. At low influent concentrations, the following equation was found to correlate the effects of substrate concentration on the process efficiency in the expanded bed:

$$\frac{S_b}{S_o} = K_2 \cdot A$$

where S_b = effluent COD concentration
S_o = influent COD concentration
K_2 = removal rate coefficient
A = specific film substrate utilization rate, day^{-1}

The coefficient, K_2, was closely correlated in a temperature relationship. This temperature relationship was used to extrapolate the reaction kinetics to 55°C, with the relationship between substrate removal efficiency, temperature, and removal rates and reactor volume shown in Figure 9. It is interesting to note that the early data developed by Schraa (13) on the thermophilic films' interaction with soluble substrates indicates that the rates achieved in Figure 9 will be supported.

DISCUSSION

The previous review of principles involved in controlling and defining the expanded bed process can be used to illustrate the AAFEB potential. As was indicated in the Introduction, the process appears to be capable of producing a secondary effluent quality without production of significant secondary microbial sludge. If the reactor upflow velocities are

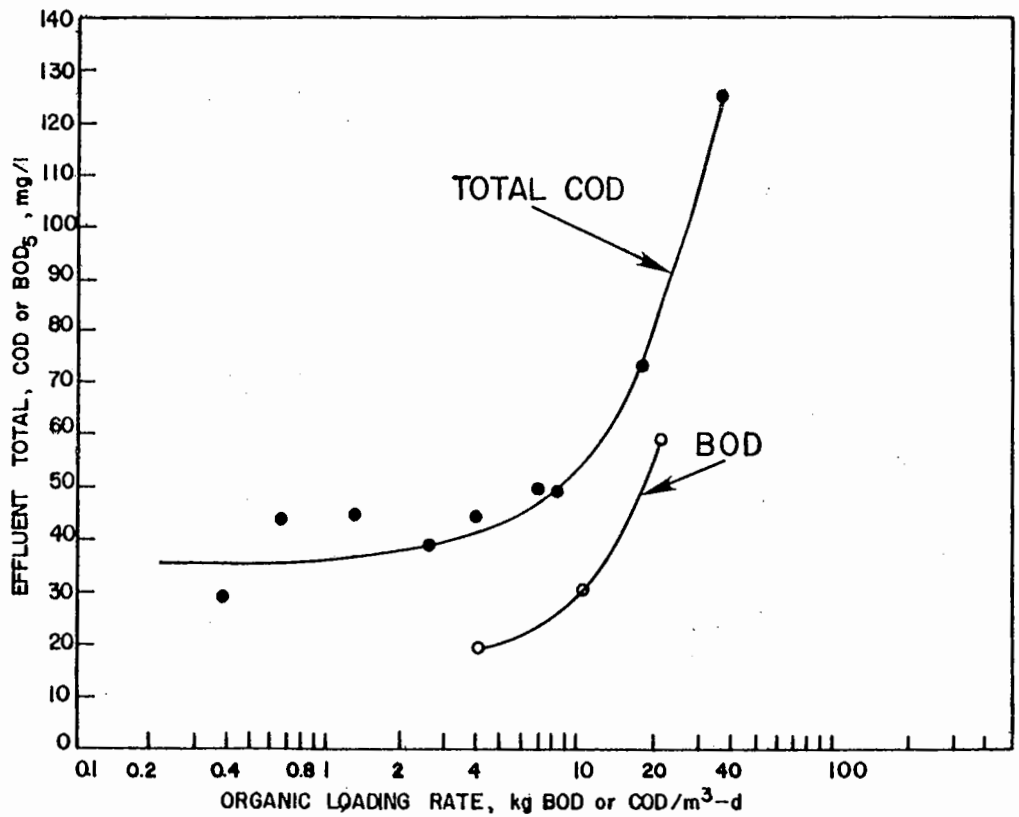

Figure 8. Comparison of process efficiency of the AAFEB (the COD data, reference [4]) and a pure oxygen fluidized bed (the BOD data, reference [14]) when applied to primary settled sewage.

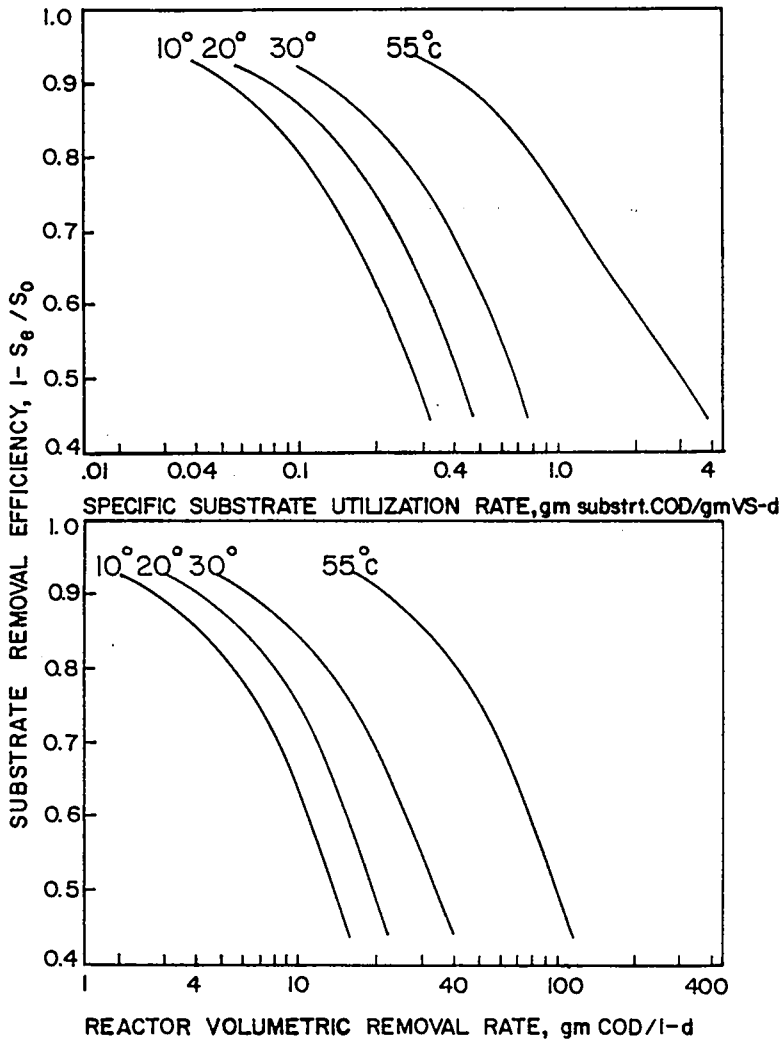

Figure 9. Comparison of AAFEB removal kinetics at varying temperatures and process efficiencies. The specific removal rate was calculated from the relationship $S_e/S_o = K_2A$ where K_2 = 1.77, 1.21, 0.75, and 0.25 ℓ/d for temperatures of 10, 20, 30, and 55°C, respectively. Volumetric removal rates were calculated assuming that the AAFEB microbial mass was 50 gm VS/ℓ. Values for A from reference (4) for 10, 20, and 30°C and the values for 55°C were estimated using temperature effect on K_2 from reference (4, 13).

compared to the clarifier velocities, it is clear that one can expect effluent suspended solids to be low. If the process is operating at a design organic loading rate of 7 kg/m^3/day, the reactor will have a hydraulic retention time of approximately one hour. Tests with primary settled sewage indicate that the effluent quality at this loading rate should be high and that one could expect much of the BOD to be converted. If 200 mg/ℓ of BOD is converted, this should result in a net yield of approximately 0.75 gm/ℓ/day. If the effluent suspended solids are lost equivalent to 30 mg/ℓ, the net change in volatile solids in the system is zero.

The long sludge retention times required to achieve an efficient anaerobic reaction and the high substrate concentrations required to drive the reaction combine to make the task of treating dilute, low temperature wastewaters amongst the most difficult challenges for anaerobic processes. It is essential that the solids management as well as the biological process be carefully controlled.

A comparison of various particle sizes and SRT values illustrates the problems that will occur if the fluidized bed process is used for sewage treatment, as compared to the expanded bed process. If it is assumed that sewage has an organic content of 230 mg/ℓ of BOD and a temperature of 20°C and effluent solids from the reactors are limited to 15 mg/ℓ, the following compares the solids retention time and therefore the capability of the processes to produce the required effluent. It is assumed that both reactors have equal efficiencies, even though this will probably not be the case. It will also be assumed that 200 mg/ℓ BOD is removed in each. Both units will be 20 ft deep. The expanded bed will have a 50 minute hydraulic retention time, whereas the fluidized bed will have 6.5 minutes. Due to the expansion and the small particle use in the expanded bed, the operating mass is estimated to be approximately 40 gm/ℓ of reactor. The fluidized bed will have an operating mass of approximately 8 gm/ℓ if 300 percent expansion is used. Based on the above assumptions, the net yield in the expanded bed is 0.4 gm/ℓ/day, whereas in the fluidized bed it would be 3.31 gm/ℓ/day because of the increased reaction rate per unit volume that is required. The resulting solids retention time is nearly 100 days in the expanded bed, as compared to 2.4 days in the fluidized bed. Clearly, the velocities and the solids management in the expanded bed result in the requirements for both the biological and physical processes to treat low strength wastewaters, whereas the fluidized bed is operating at very short solids retention times and will achieve a low quality

of effluent.

Finally, it is possible to make some gross comparisons between aerobic and anaerobic processes based on the data that are available. Figure 10 illustrates the relationship between the substrate removal rates and the resulting solids retention time in various aerobic and anaerobic processes that are able to achieve different reactor concentrations of biological solids. There are numerous assumptions included in this figure. For example, it is assumed that the efficiencies and the removal rates of the processes are compatible. The surprising results that this overview emphasizes is that the anaerobic process capability exceeds all aerobic processes. The high concentration of microorganisms in the AAFEB result in much longer solids retention times than the aerobic processes under comparable loading rates. This indicates that high organic loadings that are achieved either at high flow rates or organic concentrations with the aerobic processes can easily lead to unstable situations; whereas the anaerobic processes can still have an acceptably long solids retention time so that they can continue to operate successfully.

The AAFEB studies show significant promise for the application to a wide variety of wastewater purification problems. Areas that require further research and development are as follows:

- process scale-up to large pilot or full scale;
- impact of toxic substances;
- fundamental study of the biological reaction kinetics as affected by film thickness, substrate characteristics, and temperature;
- definition of the physical filtering capabilities of an expanded bed;
- definition and application of the thermophilic expanded bed;
- definition of impact on major practical problems such as: algae and eutrophication management, waste activated sludge treatment, and retrofit to existing processes;
- definition of physical process requirements for inexpensive inert particles coated with anaerobic microbial films, i.e., upflow velocity, bed management needs, recycle, solids, wasting;
- development of the specific application of series anaerobic-aerobic treatment to achieve high efficiency of carbon and nitrogen removal without any chemical additives.

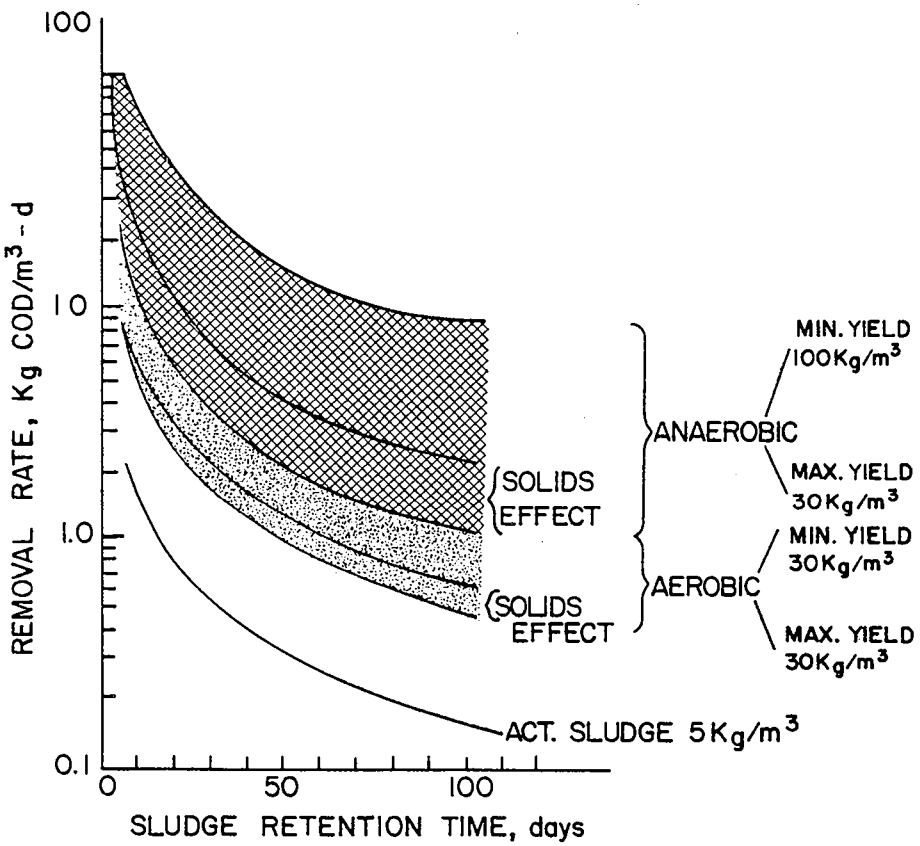

Figure 10. Comparison of resulting solids retention time and varying substrate removal rates for anaerobic and aerobic processes.

The title of this paper indicated that the topic was to be the fundamentals of a new process referred to as the expanded bed process. It is clear that the fundamentals that apply still remain to be defined in relation to many applications of the anaerobic expanded bed process. The work to date has focused on attempting to define the limits of the biological capability of a high biomass anaerobic system. Ongoing work indicates exciting possibilities for the applications of high temperature films, especially to concentrated waste stream management, excess waste activated sludge and substrates such as algae and weeds for energy production and pollution control.

SUMMARY AND CONCLUSIONS

The combination of process characteristics of the expanded bed filter with anaerobic microbial films has resulted in a process that provides the opportunity for maximum biomass concentration development while good control over the fluid forces required to retain solids is achieved. This enables the process to produce such surprising results as secondary treatment quality effluents from dilute wastes even at low temperatures, and substrate removal capability greater than any biological process, including all aerobic alternatives.

Two major unexpected results account for the capabilities of the AAFEB process. The anaerobic microbial films are exceptionally thin (around 0.020 mm) at low substrate concentrations, thus preventing mass diffusion limitations, and high bulk densities (calculated values in excess of 200 gm VS/ℓ for anaerobic films contrasted to 34 gm VS/ℓ for the thick aerobic films). Due to the combined characteristics of the expanded bed to achieve maximum biomass and suspended solids control at relatively high processing rates, it should be a desirable process for all microbial conversions. Additionally, the small particulate filtering capacities of the expanded bed appear to be significant but undefined.

Studies in progress on thermophilic films show promise of developing high rate processes for concentrated waste streams. Further research and development efforts should focus on both the fundamentals of the process and scale-up applications.

ACKNOWLEDGEMENTS

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REFERENCES

1. Jewell, W. J., Switzenbaum, M. S., and Morris, J. W., "Municipal Wastewater Treatment with the Anaerobic Attached Film Expanded Bed Process." *Journal Water Pollution Control Federation*, 53(4):482-490. 1981.
2. Jewell, W. J., "Development of the Attached Microbial Film Expanded Bed Process for Aerobic and Anaerobic Waste Treatment." Paper presented at the Biological Fluidised Bed Treatment of Water and Wastewater Conference, University of Manchester Institute of Science and Technology, England, April 14-17, 1980.
3. Jewell, W. J., "Future Trends in Digester Design." In: *Anaerobic Digestion*, Proceedings of the First International Symposium on Anaerobic Digestion, University College, Cardiff, Wales, September 17-21, 1979. 1980. pp. 467-491.
4. Switzenbaum, M. S. and Jewell, W. J., "Anaerobic Attached Film Expanded Bed Reactor Treatment. *Journal Water Pollution Control Federation*, 52(7):1953-1965. 1980.
5. Jewell, W. J. and Morris, J. W., "Influence of Varying Temperature, Flow Rate and Substrate Concentration on the Anaerobic Attached Film Expanded Bed Process." In: *Proceedings of the 36th Industrial Waste Conference, Purdue University*, May 12-14, 1981. pp. 655-664. 1982.
6. Morris, J. W. and Jewell, W. J., "Organic Particulate Removal with the Anaerobic Attached-Film Expanded-Bed Process." In: *Proceedings of the 36th Industrial Waste Conference, Purdue University*, May 12-14, 1981. pp. 621-630. 1982.

7. Young, J. C. and McCarty, P. L., "The Anaerobic Filter for Waste Treatment." *Journal Water Pollution Control Federation,* 41 5(2)R160. 1969.
8. Rich, L. G., *Unit Operations of Sanitary Engineering,* John Wiley and Sons, pp. 146-151. 1961.
9. Cleasby, J. L. and Fan, K., "Predicting Fluidization and Expansion of Filter Media." *Journal of the Env. Eng. Div.,* June, 1981, 107 (No. EE3), pp. 455-471. Paper No. 16321 in: *Proceedings of the American Society of Civil Engineers.*
10. LaMotta, F. J., "Internal Diffusion and Reaction in Biological Films." *Journal Environmental Science and Technology,* 10 (no. 8). August, 1976, pp. 765-769.
11. Kornegay, B. H. and Andres, J. F., "Characteristics and Kinetics of Biological Film Reactors." Federal Water Pollution Control Administration, Final Report, Research Grant WP-011811. Dept. of Environmental Systems Engineering. Clemson University, Clemson, SC. 1969.
12. Rittmann, B. E. and McCarty, P. L, "Substrate Flux Into Biofilms of Any Thickness." *Journal of the Env. Eng. Div.,* August, 1981, 107 (No. EE4), pp. 831-849 in: *Proceedings of the American Society of Civil Engineers.*
13. Schraa, G., "Thermophilic Anaerobic Attached Film Expanded Bed Treatment of Soluble Organics." Ph.D. Dissertation, Dept. of Agricultural Engineering, Cornell University, Ithaca, New York. In preparation.
14. Nutt, S. G., Stephenson, J. P. and Pries, "Aerobic Fluidized Bed Treatment of Municipal Wastewater for Organic Carbon Removal." Presented at the Water Pollution Control Federation Conference, Houston, Texas. 1979.
15. Lawrence, A. W. and McCarty, P. L, "Kinetics of Methane Fermentation in Anaerobic Treatment." *Journal Water Pollution Control Federation,* 41 (no. 2), part 2. 1969.
16. Gates, W. E., et al., "A Rational Model for the Anaerobic Contact Process." *Journal Water Pollution Control Federation,* 39 (no. 12). 1967.
17. Beneder, P. and Horvath, I., "A Practical Approach to Activated Sludge Kinetics." *Water Research,* 1 (no. 10), 1967.
18. Gram, A. L., "Reaction Kinetics of Aerobic Biological Processes." I.E.R. Series 90, Report 2, Sanitary Engineering Research Laboratory, University of California, Berkeley. 1956.

TRICKLING FILTERS: RELIABILITY, STABILITY AND POTENTIAL PERFORMANCE

E D Schroeder. Department of Civil Engineering, University of California, Davis, California

INTRODUCTION

Trickling filtration is probably the oldest and least understood of the modern systems for wastewater treatment. The process was developed shortly before the turn of the century and in its original form was an intermittant or periodic treatment system. Development resulted in two general types of operation; the standard or low rate process which is basically the original one, and the high rate process which incorporates effluent recirculation and higher hydraulic and organic loading rates. Some design parameters and operating characteristics of the two types of process are given in Table 1.

The basic operations difference between the standard and high rate operation is effluent recirculation, and as can be seen in Table 1 the loading characteristics are considerably different. Perhaps more interesting are the differences in operating characteristics. Effluent BOD_5 and suspended solids from standard rate trickling filters are usually comparable to activated sludge processes, while effluent from high rate systems is less satisfactory. In standard rate filters the biological film builds up for long periods of time. Large scale sloughing occurs periodically, most notably in the spring. In high rate filters sloughing occurs continuously.

TABLE 1
DESIGN AND OPERATING CHARACTERISTICS
OF TRICKLING FILTERS

CHARACTERISTIC	STANDARD RATE	HIGH RATE	
		ROCK	PLASTIC
Depth, m	1.8 -3	1 -2.5	4 -10
Specific Surface m^2/m^3	40 -65	40 -65	80 -100
Porosity	0.45 -0.55	0.45 -0.55	0.90 -0.97
Media size, mm			Dependent upon
Hydraulic Loading*	25 - 75	25 - 75	configuration
Rate, m^3/m^2-d	0.9 -2.8	9 -28	20 - 75
Organic Loading*			
Rate, kg BOD_5/m^3-d	0.11 - 0.37	0.37 - 1.8	up to 15
Recirculation Ratio	0	1 - 4	1 - 4
Sloughing	intermittant	continuous	continuous
Nitrification	yes	at lower loading rates	not in economic range of operation
Effluent BOD_5 g/m^3	< 25	> 30	> 30
Effluent SS, g/m^3	< 25	> 30	> 30

* Calculated using influent flow rate and Bod concentration

Because the loading conditions are quite different the actual effect of recirculation is difficult to determine. Obviously the actual hydraulic loading rates are increased over the nominal value found by dividing influent flow rate by cross-sectional area. Flow variation is damped because of the steady recycle component, and presumably distribution over the media is more uniform and complete. Larger organisms, such as fly larvae, that feed on the slime are washed out. Thus the microbial community should be different in high and standard rate systems. Because the predator/ grazing organism population is lower more overgrowth and plugging problems might be expected in high rate trickling filters. The opposite is the case, however. Two factors appear to be involved. First, the higher flow rates result in more complete distribution of the nutrients through the volume and the result is more uniform growth. More important are the higher shear rates associated with the larger flow rates. Bruce (1) reported that at higher hydraulic loading rates shear was the principle control mechanism and that at lower rates slime accumulation the most important control mechanism was grazing by invertebrates. Solbe and Roberts (1) performed an inventory of invertebrate organisms, in an experimental standard rate unit over a three year period and found both the total slime mass and the populations to be highly variable. The spring sloughing resulted in large decreases in invertebrates as well as bacterial slime. It is assumed that the large accumulation of film during the winter months is the result of decreased invertebrate activity at lower temperatures and the spring sloughing is caused by their renewed activity.

Recycle

The effect of recycle on process performance has always been controversial. Many workers have considered the recycle stream to provide additional passes through the reactor (3), and therefore improving process performance. This would actually be true only if the recycle stream remained segregated from the influent, a situation that is difficult to conceive. A similar result would be obtained if the higher flow rates of a recycle system caused a more complete wetting of the trickling filter surfaces. Over designed units and systems with high influent BOD concentrations where the organic loading rate controls the process design would be examples that might appear to follow the multiple pass concept.

Recycling the process effluent should have three physical consequences: 1) diluting the influent stream, 2) increasing the liquid film depth and 3) incorporating sloughed microbial culture into the liquid film 4, 5, 6. The first two factors will decrease process performance because liquid phase transport will be slower

at lower concentrations and greater distances. Recycling sloughed cells could result in significant quasi-homogeneous reaction rates in the liquid film. Oxygen transfer would be less of a problem because the portion of the reactions taking place in liquid film would be closer to the air-liquid-interface. The summary effects of these three factors is not clear. High rate trickling filters remove considerably more organic material than standard rate units but effluent quality is considerably lower also.

Media

Highly porous plastic media has been increasingly used in recent years. The greater porosity and regular shapes can be expected to result in more uniform flow distribution and improved oxygen transfer. The advantages of plastic media are realized only at high loading rates where conventional media would be quickly plugged (7). At lower rates rock media systems perform as well or better than plastic media units. This is not particularly surprising because the available surface area per unit volume is not greatly different.

Biofilm

The attached biological slime in trickling filters is highly variable. Mass distribution varies with time, season and flow (1,8,9).

Bacteria are the dominant types of organisms although fungi such as Geotrichium are often present in significant amounts. There has been very little study of the structure of trickling filter biofilms. They are obviously not uniform and vary greatly in depth. Pivetti (8) observed the accumulation of biomass as a function of depth in a pilot scale (0.25 m diameter, 2.4 m media depth) trickling filter using 5 cm plastic pall rings (Norton Actifil). The hydraulic loading rate was held constant at 9.9 m^3/m^2-d (10.6 mgad or 4.13 d^{-1}) but three organic loading rates of approximately 0.31, 0.81 and 1.05 kg BOD_5/m^3-d. Both the hydraulic and organic loading rates fall in the lower range for high rate trickling filters. Slime mass varied with depth for all three loading rates as shown in Figure 1.

Make up of the biofilm includes many filamentous.

Sloughing

Little information is available on the mass rate of sloughing because most workers report secondary clarifier effluent suspended solids concentrations rather than trickling filter effluent values. Eden et al (7) studied plastic media (Surffpac) at high loading rates (6.7-21.5 $m^3/m^2 \cdot d$ and 0.9-2.4 kg $BOD_5/m^3 \cdot d$) over a one year period and reported trickling filter effluent suspended solids concentrations ranging from 119 to 198 g/m^3. Influent suspended solids concentrations were approximately 25 percent higher and

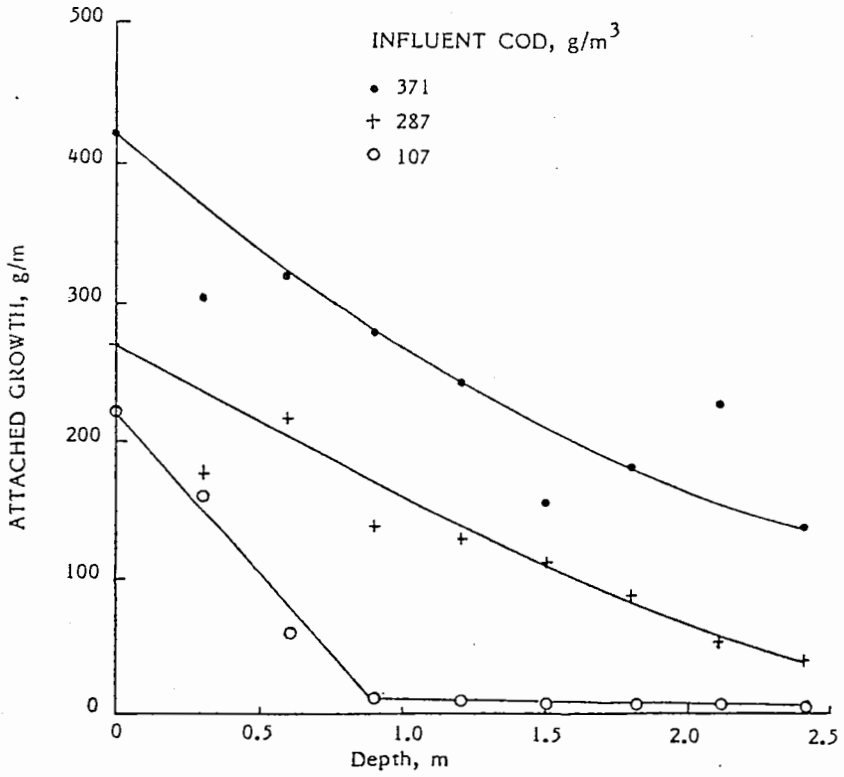


FIGURE 1 VARIATION IN ATTACHED SOLIDS WITH DEPTH AT HYDRAULIC LOADING RATE OF 10 m^3/m^2-d

thus a large fraction of the effluent solids could well have originated in the influent. Pivetti worked with a soluable fed and reported average effluent suspended solids values (unsettled) of 10, 44 and 72 g/m^3 for the three organic loading rates he used.

Aitken (10) used a hydraulic loading rate 15 $m^3/m^2 \cdot d$ and an organic loading rate of 2 $kg/m^3 \cdot d$ in studies with a 0.15 m diameter, 1.1 m deep model plastic media trickling filter. Like Pivetti the feed was soluble. Effluent suspended solids averaged 23 g/m^3 over a 103 day period, with a standard deviation of 7.7 g/m^3. The media used by Aitken (10) was 1.3 cm plastic pall rings. Quite possibly the small media size was a factor in the low effluent suspended solids values. Periodically Aitken pulsed the hydraulic loading rate to determine the effect on effluent suspended solids. Pulses consisted of increasing the liquid application rate by a factor of 4 or 8 for a 30 or 60 minute duration period. Large increases in sloughed solids during the pulse resulted. For a period of approximately one day after the pulse effluent suspended solids were less than normal steady state value (Figure) 2, but after this relatively brief period effluent suspended solids approached the steady state values.

Nitrification

Control of nitrification in biological wastewater treatment is still a somewhat elusive objective. The fact that extensive nitrification is normal in slow rate trickling filters and relatively insignificant in high rate systems is a consistant observation. There is no reason to suspect that nitrifying organisms are washed out of the high rate systems and another cause must be considered. A reasonable conjecture is that competition for oxygen is too great in high rate systems for nitrification to occur. If this is correct nitrifiers activity should be limited to the lower depth of low rate units but this has not been demonstrated. The actual mechanisms of nitrification have not been established. Quite possibly adsorption of NH_y^+ on slime surfaces is the ammonia removal mechanism, with oxidation of the adsorbed NH_y^+ following. An interesting set of experiments could be developed to test this hypothesis.

CONCEPTUALIZED AND REAL TRICKLING FILTERS

A considerable amount of mathematical modeling of trickling filters has been done. Examples include the early work of Velz (11), Howland (12) and Eckenfelder (13) that utilized first order reaction models, the work of the mid-1960's characterized by Swilley and Atkinson (4), Kehrberger and Bush (5), Meir et al (14) and Kornegay and Andrews[15] and the more recent studies by Atkinson and his co workers (16,17,18), Williamson and McCarty (19), Rittman and McCarty (20) and Harromoes (21). All of these

workers were forced to make idealizing assumptions about the fluid flow conditions. Generally these assumptions are uniform steady flow over the entire surface, a smooth microbial slime of uniform depth throughout the trickling filter and all reactions in the slime. Such assumptions are very unrealistic. All trickling filters are loaded periodically if one considers a particular point or flow path. Thus flow would be expected to occur in a rippling pattern with mixing occuring at points were sections of media intersect. The microbial slime can be seen to vary in characteristics throughout trickling filters and sloughing would result in a patchy surface of variable friction characteristics. Under such conditions it would be surprising to find the wastewater running smoothly over the entire media surface. Quite likely the flow runs in riveluts for short distances before joining with other streams which are then separated out into smaller flows at media interfaces or junctions. As growth and sloughing occurs the pathways of the flow would be expected to change resulting in an extremely dynamic system.

The third assumption, all reactions occur in the microbial film, has varying validity. Sloughed film would be biologically active. Recycling from a point prior to the secondary clarifies would enrich the liquid film with micro organisms and result in what Swilley (22) termed a pseudo-homogeneous reaction system. Swilley concluded from theoretical studies that pseudo-homogeneous systems would have better performance than heterogeneous systems and that recycle would improve process performance if suspended cells were included and decrease efficiency if suspended cells were excluded. Kehrberger and Busch experimentally validated Swilleys conclusion for an inclined plate system. Unfortunately the idealized flow conditions of an inclined plate are quite different from real trickling filters and there is substantial evidence that removal rates are increased when recycle is used, regardless of the configureation (ie before or after the secondary clarifes).

Most of the recent attached growth models (14-22) are based on mass transport concepts. None consider the possibility of more than one limiting nutrient. In their most easily applied forms there is an assumption that the same mechanism is rate limiting throughout the depth, but models such as Atkinsons (16,17) are based on spatially varying conditions. These models are useful in delineating the interaction of system variables and parameters but are far too simplified to predict process performance without extensive, system specific calibration. This was demonstrated by Atkinson and Ali (28) and Pivetti (5) in their studies with simplified systems.

ACTUAL PROCESS PERFORMANCE

A number of reports on actual performance of trickling filter systems over extended periods of time are in the literature. Particularly notable in the historical sense is the NRC report (23) the used annual average values to develop loading/performance relationships. Although the results were at best shakey (24) the NRC formula is still used in design. Gallen and Gotaas (25) developed a performance relationship based on the best fit of data from trickling filters and Fairall (26) and Rankin (27) worked with average data from a number of systems also.

More recently Haugh et al (28) and Niku et al (29) reported the results of the analysis of one years daily composite data from 11 high rate trickling filters systems located through the midwest. Average daily flows ranged from 900 m^3/d (0.5 mgd) to 130,000 m^3/d (34 mgd). In these studies summary statistics (mean, standard deviation, show etc) were examined to determine general process characteristics. Five common probability density functions were tested with the effluent BOD_5 and suspended solids data. Three of the two parameter empirical, gamma and log normal were found to adequately describe the distributions. The log normal distribution provides the best fit for activated sludge effluent data and is the easiest of the three to apply and was chosen as the model of choice.

Multiple regression analysis was used to estimate effects of selected process parameters on effluent quality. In general primary effluent BOD and suspended solids concentration and temperature had the greatest effect with flow rate also being significant. Weekly cycles were found in several plants and monthy variation was established in all 11 plants. As would be expected effluent quality was lower in the Winter than in the Summer.

The data were non-random as noted above. This means that effluent quality depends on the day of the week and month and that the chosen distributions can only be used in a limited manner. Prediction of annual frequencies and other long term behavior is possible, however. It is quite likely that the non-randomness is due to somewhat predictable loading and temperature variations and this would justify use of expressions such as Eckenfelder's (13) or the modified Atkinson model (6)/

EFFLUENT QUALITY, RELIABILILTY AND STABILITY

Effluent Quality for the 11 plants is summarized on a daily, seven day and 30 day average basis in Tables 2 and 3. Seven of the 11 plants met the 30 day 30 g/m^3 standard for BOD_5 and suspended solids. The daily, 7 day and 30 day average were virtually identical in each case, and therefore there is no advantage in reporting more than one value.

TABLE 2

EFFLUENT BOD_5 DATA FROM 11 MIDWESTERN TRICKLING FILTERS

Plant	Daily Average g/m^3	7 day Running Average g/m^3	30 day Running Average g/m^3
1	33.3	33.1	32.1
2	10.7	10.6	10.0
3	10.1	10.0	9.8
4	58.4	58.3	57.7
5	29.2	29.2	29.0
6	27.0	27.0	26.8
7	23.2	22.9	22.6
8	43.1	43.0	42.8
9	51.1	51.5	52.9
10	21.0	21.0	21.1
11	18.3	18.0	17.1

TABLE 3

EFFLUENT SUSPENDED SOLIDS DATA

FROM 11 MIDWESTERN TRICKLING FILTERS

Plant	Daily Average g/m^3	7 day Running Average g/m^3	30 day Running Average g/m^3
1	52.5	52.5	51.6
2	21.5	21.2	20.9
3	21.0	21.1	21.1
4	54.9	54.8	54.3
5	18.3	18.4	18.6
6	15.1	15.1	14.9
7	24.1	23.6	23.2
8	34.0	34.0	34.0
9	41.1	41.2	41.8
10	23.6	23.9	24.0
11	16.2	16.1	15.8

Process reliability has been defined by Niku, Samaniego and Schroeder (29) as the probability that a given system will meet a chosen standard. The proposed a coefficient of reliability (COR) based on the log normal distribution that uses and a processes coefficient of variation, V_x

$$COR = (V_x^2 + 1)^{1/2} \exp \left\{ - z_{1-\alpha} [\ln(V_x^2 + 1)]^{1/2} \right\}$$

where $z_{1-\alpha}$ = standard normal variate for the distribution.

Process reliability can be plotted as a function of V_x and the nomalized mean, m_x/X_s (m_x = actual or design value and X_s = standard value) as shown in Figure 2. Based on the data from the 11 plants V_x values for effluent BOD_5 and SS should be approximately 0.50 and 0.55, respectively. Using Figure 2 and X_s = 30 g/m^3 we can conclude that 95 percent reliability would require an average effluent BOD_5 concentration of slightly less than 15 g/m^3.

Evaluating process stability is a more qualitative procedure than evaluating reliability. Defining stability is somewhat of a problem in itself. Normalized parameters generally do not differentiate between systems with good and poor effluents. For example of the ratio of the standard deviation to the mean were used a plant with a ratio of one could have standard deviation and mean values of 5 and 5 or 100 and 100. The standard deviation is a useful value in estimating process stability, but does not provide information on what caused a particular value. For example a given standard deviation might be the result of a number of small deviations or one or two colosal failures. The range provides information on maximum values experienced but not on their frequency. A plant with one failure per year would not be called unstable by most people, but might be identified as such if range were the sole criteria. For these reasons an ideal stability measure does not exist, but a somewhat qualitative estimate can be developed by plotting range vs standard deviation (Figures 3 and 4). As can be seen the range tends to increase with standard deviation for both effluent BOD_5 and suspended solids concentration. A standard deviation value of 10 g/m^3 was taken as the stability cut off point for both variables. The decision was based to a large extent on a similar analysis for activated sludge processes where the differences are considerably more clearcut.

SUMMARY AND CONCLUSIONS

Trickling filters can be best designed using pilot scale studies. Estimates of process performance can be made using simple models incorporating loading rate parameters and the reliability of process can be estimated for a given effluent standard using Figure 2.

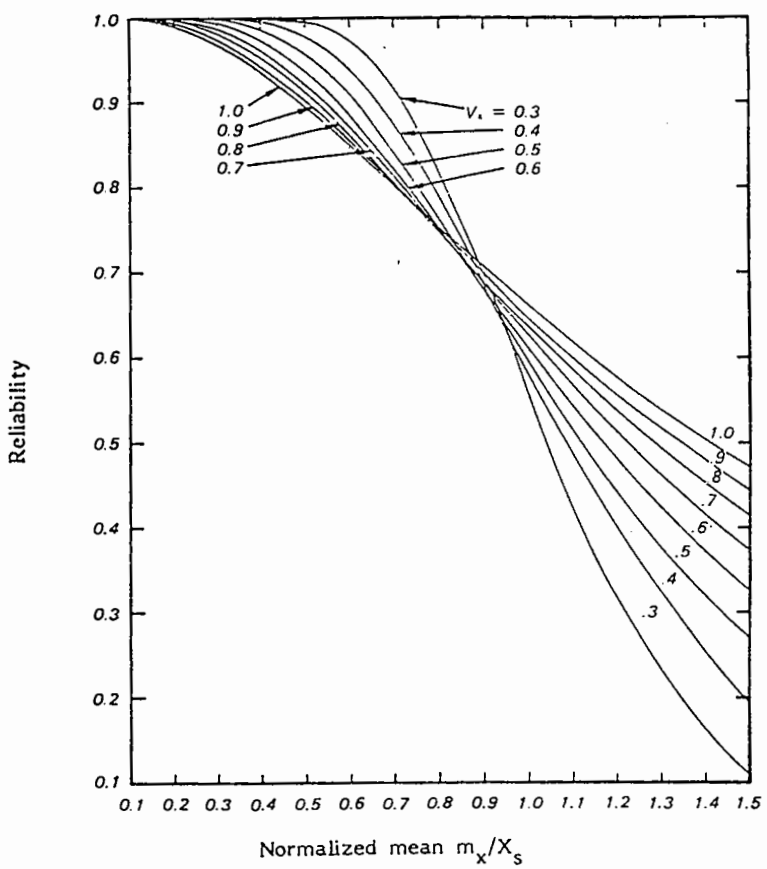


FIGURE 2 RELIABILITY AS A FUNCTION OF COEFFICIENT OF VARIATION AND NORMALIZED MEAN

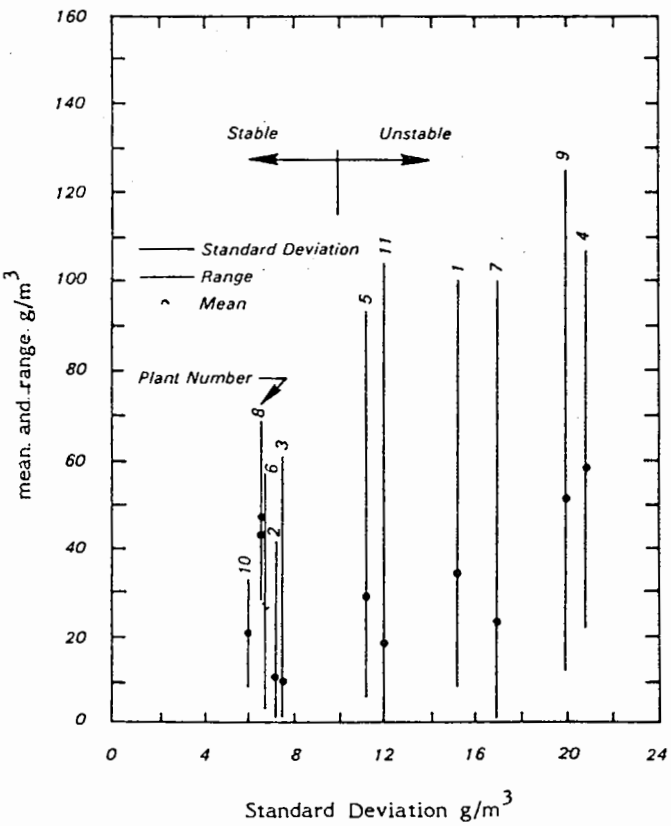


FIGURE 3 VARIABILITY OF EFFLUENT BOD AS A FUNCTION OF STANDARD DEVIATION AND RANGE

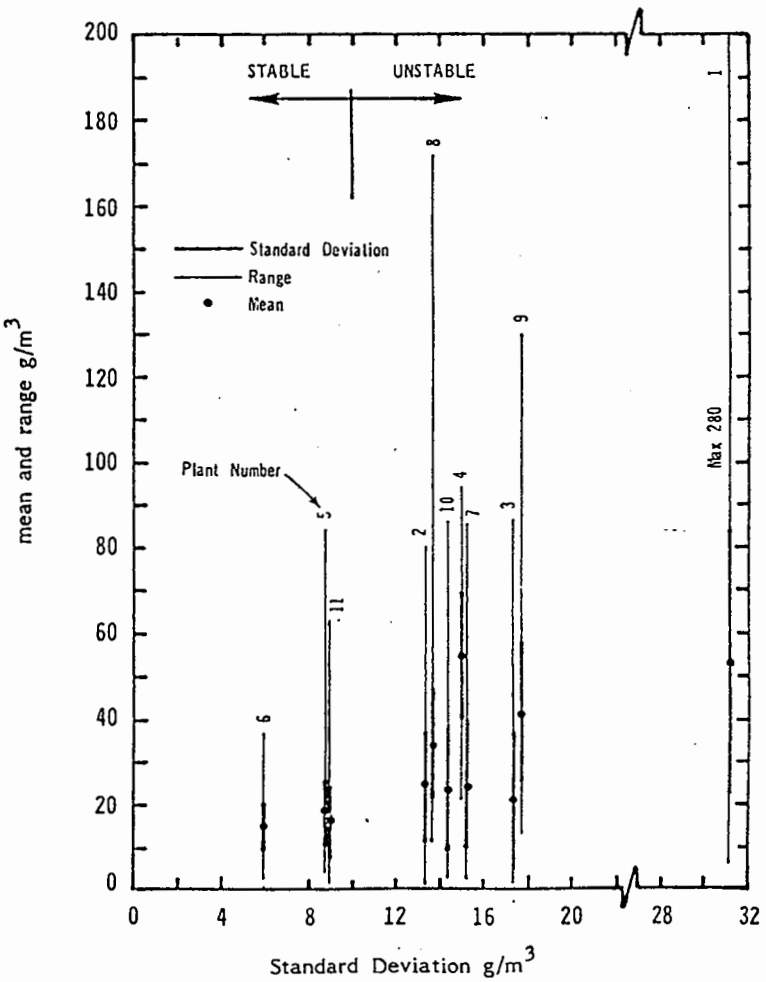


FIGURE 4 VARIABILITY OF EFFLUENT SUSPENDED SOLIDS AS A FUNCTION OF STANDARD DEVIATION AND RANGE

Process stability cannot be predicted, but it is clear that the lower the effluent BOD and suspended solids concentrations are the more stable a plant will be.

High rate trickling filters can produce good quality effluent as shown in Tables 2 and 3. In general the lower the organic and hydraulic loading rates the better the effluent quality will be. The performance history of low rate systems is quite good, but capital and land requirements are high and this will probably restrict their use to smaller communities even under current energy restrictions. High rate trickling filters can be competitive with activated sludge processes in terms of land and cost, but effluent quality is definitly lower than activated sludge processes. Application of 30-30 standards makes selection of high rate trickling filters risky unless further treatment is included. If standards are allowed to reflect a particular set of discharge conditions trickling filters would be a more widely used alternative.

REFERENCES

1. Bruce, A.M. "Percolating Filters," Process Biochemistry, 4, April, 1969.
2. Solbe, J.F. and H. Roberts "The colinization of a Percolating Filter by Invertebrates and Their Effect on Settlement of Humus Solids", Water Pollution.
3. Fair, G.M. and J.C. Geyer Water Supply and Wastewater Disposal, John Wiley and Sons, New York, 1956.
4. Swilley, E.L. and B. Atkinson "A Mathematical Model For The Trickling Filter" Proceedings 18th Industrial Waste Conference, Purdue University, 1963.
5. Kehrberger, G.J. and A.W. Bush "Mass Transfer Effects in Maintaining Aerobic Conditions in Film Flow Reactors" Journal Water Pollution Control Federation, 43, 1514, 1971.
6. Schroeder, E.D. Water and Wastewater Treatment, McGraw Hall Book Company, New York, 1977.
7. Eden, G.E., G.A. Truesdale and H.T. Mann "Biological Filtration using Plastic Filter Medium" J. Institute of Sewage Purification, 65, 562, 1966.
8. Pivetti, D.A. "Influent Concentration, Slime Mass Distribution and Substate Removal In A Trickling Filter" MS Thesis, University of California, Davis, 1976.
9. Howell, J.A. and B. Atkinson "Sloughing of Microbial Film In Trickling Filters", Water Research, 10, 307, 1976.
10. Aitken, M.D. "Hydraulic Modeling of Suspended Solids In Trickling Filter Effluents" MS Thesis, University of California, Davis, 1980.
11. Velz, C.J. "A Basic Law For The Performance of Biological Beds", Sewage Works Journal, 20, No. 4, 1948.
12. Howland, W.E. "Flow Over Porous Media as In A Trickling Filter", Proceedings 12th Industrial Wastes Conference, Purdue University, 1958.
13. Eckenfeder, W.W. "Trickling Filter Design and Performance", J. Sanitary Engineering Division, ASCE, 87, SA3, 33, 1968.
14. Maier, W.J., V.C. Behn and G.D. Gates "Simulation of the Trickling Filter Process", J. Sanitary Engineering Division, ASCE, 93, SA4, 91, 1967.
15. Kornegay, B.H. and J.F. Andrews "Kinetics of Fixed Film Biological Reactors" Journal Water Pollution Control Federation, 40, 460, 1968.
16. Atkinson, B. Biochemical Reactors, Pios Press, London, 1974.
17. Atkinson B. and I.S. Daovd "Diffusion Effects Within Microbial Films" Trans. Inst. Chemical Engineers, 48, 245, 1970.

18. Atkinson, B. and M.E. Abdel Rahman Ali "The Effectiveness of Biomass Hold-Up and Packing Surface in Trickling Filters" Water Research, 12, 147, 1978.
19. Williamson, K. and P. McCarty "A Model of Substrate Utilization By Bacterial Films" J. Water Pollution Control Federation, 48, 9, 1976.
20. Rittmann, B.E. and P.L. McCarty "Substrate Flux Into Biofilms Of Any Thickness" J. Environmental Engineering Division, ASCE, 107, 831, 1981.
21. Harromoes, P. "Significance of Pore Diffusion To Filter Denitrification" Half Order Reactions in Biofilm and Filter Kinetics", VATTEN, 2, 1977.
22. Swilley, E.L. Ph.D. Thesis, Rice University, 1964.
23. National Research Council Subcommittee on Sewage Treatment "Sewage Treatment at Military Installations, Sewage Works Journal, 18, May 1946.
24. Schroeder, E.D. and G. Tchobanoglous "Another Look at the NRC Formula" Water and Sewage Works, 122, 58, 1978.
25. Gallen, W.S. and H.B. Gotaas "Analysis of Biological Filter Variables" J. Sanitary Engineering Division, ASCE, 90, SA6, 59, 1964.
26. Fairall, J.M. "Correlation of Trickling Filter Data" Sewage Works Journal, 28, 1069, 1956.
27. Rankin, R.S. "Evaluation of The Performance of Biofiltration Plants" Transactions ASCE, 120, 1955.
28. Haugh, R. , S. Niku, E.D. Schroeder and G. Tchobanoglous Performance of Trickling Filter Plants, PB 82-108 143, National Technical Information Service, 1981.
29. Niku, S., E.D. Schroeder and F. Samaniego "Performance of Activated Sludge Processes and Reliability Based Design" J. Water Pollution Control Federation, 51, 284, 1979.

PART II: CURRENT STATUS AND FUTURE TRENDS

THE HISTORY OF FIXED-FILM WASTEWATER TREATMENT SYSTEMS

Robert W. Peters. Department of Civil Engineering, Purdue University, West Lafayette, Indiana.

James E. Alleman. Department of Civil Engineering, University of Maryland, College Park, Maryland.

INTRODUCTION

The science and 'art' of wastewater engineering stretches only slightly beyond one hundred years. Within this period, the applied technology has certainly made significant strides in promoting disease control and environmental protection. Fixed-film treatment unquestionably plays an important role in this history, particularly since it represented the original biological mechanism. Beginning with options like the trickling filter, intermittent filter and contact bed, fixed-film systems dominated the technology of wastewater treatment for several decades. And although this status has subsequently been assumed by suspended growth process, there is unquestionably a resurgence of interest in fixed-film applications.

Given the relative historical significance, and projected future of fixed-film systems, a chronological review of the associated progressive developments should be both interesting and informative. This paper will therefore explore the genealogy behind our current fixed-film technology, condensing the relevant yesteryear literature into twenty-five year increments. While attempting to limit this synopsis to a reasonable length, every effort has been made to facilitate a thorough documentation of the associated literature.

1850 - 1875

As described by the classic Dickens tale in 1859, "It was the best of times, it was the worst of times..." (1) This literary image poignantly portrays a mid-nineteenth century era freshly endowed with the blessings of an Industrial Revolution, yet virtually helpless in the face of rampant, epidemic disease. Cholera, alone, flared through the British Isles in four deadly outbreaks within one terrifying ten year period. (2)

Without question, these problems with communicable disease provide a sad reflection on the existing deficiencies in environmental sanitation. However, the concurrent infancy of bacteriology yielded only vague clues regarding the dangerous correlation between fecal contamination and disease transmission. Existing efforts towards sewage disposal, let alone treatment, were virtually non-existent. (3) Certainly it was fortuitous, then, that legislation (i.e. the Nuisance Removal Act) was enacted in 1858 to control sewage discharge, albeit more so a function of safeguarding asthetics rather than a perceived health hazard. (4) This emphasis quickly shifted towards disease control, though, following Dr. John Snow's monumental publication on epidemiology within the same year. (2,4,5)

England shortly organized a series of Royal Committees (6,7,8) charged with the study of problems relating to sewage disposal and treatment. Their initial findings categorized the existing state-of-the-art according to chemical precipitation, filtration and irrigation, with the latter two procedures generally associated with land treatment. While land systems carried a traditional background extending several centuries,(4,9) some of the other available options were rather curious. One such precipitation procedure, the ABC process, employed a bizarre mixture of alum, blood and clay. (4,10)

None of the available treatment mechanisms were, however,

recognized as biologically-related systems. Hence, Dr. Alexander Mueller's demonstration in 1865 that sewage could be purified by living organisms in a filtration column provided a major revelation. (11) Dr. Mueller, a prominent City Chemist of Berlin, subsequently patented his biological purification process several years later. Unquestionably avant-garde, neither the patent nor the fundamental concept attracted much attention, though.

In 1868, one of the Commission members, Sir Edward Frankland, began an epic study of filtration performance on raw London sewage in laboratory columns packed with media ranging from coarse gravel to peaty soil. Using a twice daily dosing pattern, Sir Frankland maintained successful filtration performance for over four months. (11,12,13) Although the filter's treatment capability was solely credited to physical-chemical means, the associated establishment of the intermittent filtration concept had notably introduced a necessity for resting or aeration periods between sewage applications.

Based on these results, the Royal Commision began to place considerable emphasis on the use of intermittent land filtration. (14) In 1871, J. Bailey-Denton initiated the first full-scale operation at Merthyr Tydvil, Wales. (14) Success at this facility, and others subsequently developed by Bailey-Denton, soon promoted several engineers to apply Frankland's concept. (4,11,14) Unfortunately, these engineers oftentimes neglected critical factors such as soil permeability and/or the necessity for intermittent dosing, such that failures became commonplace. And with subsequent documentation of 38 such failures, (4,11,14) technical interest in the intermittent concept quickly faded.

1875 - 1900

Following upon the singular work by Mueller over a decade earlier, several researchers successively explored the microbial aspect of sewage purification. Schloesing and Müntz (15) first demonstrated soil nitrification in 1877. Five years later, Warrington (16) confirmed that sterilized solutions lost their nitrifying ability until inoculated by fresh soil. And in 1890, Winogradsky (17) succeeded in identifying Nitrosomonas bacteria. These pioneers were, however, still uncertain as to the pragmatic application of these bacterial mechanisms to effective treatment.

Up to this point, Europe had dominated the developments in wastewater treatment technology. Within the United States,

though, comparable concern for pollution control resulted in the establishment of the Lawrence Experimental Station by the Massachusetts State Board of Health. (4,18) Organization of the Lawrence facility was handled by Hiram F. Mills, a distinguished hydrologist, and Professors Sedwick and Drown from the Massachusetts Institute of Technology. (4,19) Under the direction of Allen Hazen, the Lawrence group began a series of filtration experiments in 1887 which were comparable to the prior Frankland tests on intermittent dosing. In this case, however, the filters were significantly larger, at 1/200th acre per unit. While their results subsequently verified the treatment potential afforded by an intermittent filtration mechanism, the Lawrence group's first publication in 1890 provided a monumental analysis of the associated microbial activity. (18) Indeed, their findings truly furnished the hallmark demonstration that microorganisms carried within the filter media could degrade sewage in an aerobic environment facilitated by intermittent dosing.

Given the success of the Lawrence experiments, biological treatment systems rapidly expanded in terms of application and sophistication. Considerable controversy had arisen in the 1890's over patent rights obtained by Donald Cameron for septic tanks, (4) such that most municipalities were anxious to find suitable treatment alteratives. Several full-scale intermittent filtration systems were therefore constructed in the New England area, most of which were successfully maintained for several decades.

In Europe, though, sanitary engineers were still hesitant to accept the intermittent filtration concept. This opinion likely stemmed either from a lingering dissatisfaction with the Frankland-era facilities, or because of the widespread unsuitability of European soil. (4) Instead, they chose to intensify filtration rates using coarser media such as coke breeze, gravel, burnt clay and coarse chalk. Scott-Moncrief (9) probably began the first such tests, using sewage percolation through sequential trays of 1 inch diameter coke media. In 1893, J. Corbett (20) also employed a serial filter scheme, with an additional wooden trough to continuously distribute influent sewage across the bed. And in the same year, F. Wallis Stoddart (21) reported on the use of a coarse media filter receiving a continuous, trickling flow. Of these two latter researchers, Corbett acknowledged the impetus and direction provided by the previous Lawrence findings. Stoddart, however, insisted that his work stemmed from Frankland's principles and that his continuously percolated units were the first of their

kind. In either case, the trickling filter had been conceived.

Another classic European option which developed at much the same time was the contact bed. Acting along the lines of the Lawrence experiments, W. Santo Crimp and W. J. Dibdin decided in 1891 to experiment with a dosing pattern which flooded a coarse media filtration bed for 8 hours, followed by 16 hours in a drained state. (4,9) Of the coarse media materials tested on chemically-treated London sewage, Dibdin found that the coke breeze provided satisfactory treatment, while sand clogged extensively. In subsequent tests, Dibdin experimented with a double-contact approach, using primary and secondary beds respectively containing successively smaller media. (4) The success of this operation quickly led to several full-scale installations, all of which maintained the cyclic fill, drain and react periods. And in their fifth report (6), the Royal Commission provided extensive technical support for the installation and operation of such contact beds.

Dosing strategies for both the trickling filters and contact bed systems received intensive study in the years immediately following their development. For uniform loading of intermittent filter units, Waring and Lowcock devised a simplistic technique in 1892 based on an overlying fine gravel layer to promote equivalent flow distribution. (4,14,23) However, this procedure retarded desired bed aeration. Perhaps as a consequence, Waring also devised and patented a trickling filter system which employed forced aeration. (4,14)

Stoddart's (4,21) approach to flow distribution was that of corrugated sheet-metal plates with symmetrical discharge ports. Although considered satisfactory, leveling of these horizontal plates required tedious adjustment. Corbett (4,20) initially used slotted wooden troughs and then switched to a variety of fixed-spray jets. In 1896, Carfield (4,14) improved the fixed distributor concept by adding a siphoned dosing tank. The siphon action insured an intermittent dosing procedure which prevented localized flooding at the media.

Rotary flow distributors were originally tested in 1889, with additional refinement by Corbett in 1894. (20) Two years later, Whittaker and Bryant (4) introduced a rotary sprinkler equipped with a pulsometer. This latter addition not only produced a pulsed, intermittent flow, but also warmed the influent sewage. However, their model employed perforated pipe distribution arms prone to clogging. Rotary wooden troughs were then introduced by Mather and Platt to avoid this plugging problem. (4,14)

1900 - 1925

Given the classic technical advancements made by Hazen, Stoddart, Corbett and Dibdin in the past quarter century, the next twenty-five years could be viewed as an era of practical application and refinement. Of the available biological treatment systems (i.e. intermittent filtration, trickling filters and contact beds), it is interesting to note that each comprised a fixed-film process. Rudimental experiments in sewage aeration were underway at the time, but suspended growth systems did not originate for several years. (4)

Trickling filters were first introduced to the U.S. in 1901 at Madison, Wisconsin. (4) By 1910, several additions in mid-west and eastern cities brought the total to ten. (9) Monumental in size alone, the 31 acre Baltimore trickling filter system is remarkably still in operation some seventy-five years after its initial development. (24)

Amongst these early U.S. trickling filter units, and for several decades, fixed spray jets served as the norm for flow distribution. (4) Contemporary sewage treatment texts typically carried several pages devoted to spray jet design and installation. (4,9,14) In most cases, these distributions were also equipped with siphon dosing tanks. While rotating distributors were only randomly tested in the United States (i.e. Springfield, MO in 1912 and Pontiac, MI in 1920),(4) European trickling filter designs favored the rotary or travelling sprinkler approach. (11)

With the advent of trickling filter applications, interest in intermittent-filtration began to fade. Experimentation continued on both options at Lawrence, (19) demonstrating that the higher loading rates provided by coarse media design could significantly reduce the requisite land area. Mathematical modeling of these biological filters was also initiated in 1916 by Tatham. (25) In using a mass-balance derivation based on first-order kinetics, this study classically sought to define the purification process according to precise chemical engineering principles.

As for contact bed design, several full-scale applications were recorded. (4,26) Although a few large scale units were built in the United States, (4) contact beds did not receive much interest outside Europe. Because of the involved flooding routine, anaerobic conditions tended to lower final effluent quality. (26) This circumstance, combined with frequent clogging of the bed media by entrained sludge, (4,26) certainly began to cast doubts on the usefulness of contact bed treat-

ment.

Recognizing the desirability of an aerobic biofilm, Dibdin decided in 1904 to experiment with forced bed aeration. (27) And to facilitate flushing solid matter from the bed, the coarse media was replaced with slate slabs packed in horizontal layers. Operation of the modified unit still followed the phased fill-and-draw routine. (4,28) After 12 months of laboratory study, Dibdin successfully progressed to a full-scale demonstration of his slate bed design at Devizes in 1905. (29) However, in their fifth report, the Royal Commission indicated that the slate bed approach should only be considered as a primary sedimentation mechanism. (6)

Within the U.S., Dibdin's slate bed technique drew immediate interest. Experimental testing was initiated in Plainfield, New Jersey in 1905. (30) Historically important experimentation on slate bed treatment was also begun at Lawrence under the direction of H. W. Clark and S. Gage. (19, 31) In comparing aerated slate bed units and aerated bottles containing algal suspensions, these investigators reported in 1913 that the bottles provided better treatment efficiency. (31) This variance was attributed to a failure by the previously scrapped slate plates to accumulate a suitable biofilm during the short period of study.

Shortly thereafter, Gilbert John Fowler, a British Professor of Chemistry at Victoria University, visited the Lawrence labs and witnessed these same experiments. (31) Upon returning to England, Dr. Fowler's students Edward Ardern and William Lockett began the historic study of suspended growth treatment. In 1914, these two students then published the first account of an activated sludge process; sticking with the accepted intermittent (i.e. fill-and-draw) pattern, but distinctively switching to a suspended biomass. (32) Speaking on behalf of his students, Fowler did acknowledge the contributing and inspiration provided by Clark and Gage, referring to Lawrence as "the Mecca of sewage purification..." (32)

In much the same vein as Dibdin's slate bed, Dr. William Owen Travis also sought to improve upon the contact bed procedure. (22) As the local health officer in charge of a contact filter at Hampton, England, Dr. Travis was quite familiar with the problem of bed clogging. (4) His solution, introduced in 1904 as the Travis Hydrolytic or Colloider Tank, was essentially configured as a multi-stage septic tank. Successively divided into detritus, hydrolytic and finishing tanks, the latter two zones contained wooden colloider baffles or laths placed in a parallel array. These baffles were intended to

attract fine particulates for subsequent degradation. Only one such plant was ever built, at Norwich, England in 1909. (4) The construction at another Travis facility by the Emsher Drainage District Board was discontinued after the death of the involved design engineer, Wattenberg. (4) His replacement, Karl Imhoff, subsequently convinced the Board to switch to his personal design, known thereafter as the Emscher or Imhoff Tank. (9,14)

As a footnote to this era, mention should also be made of two unique patents obtained for rotating support media. (33,34) The first, conceived by Weigand in 1900, (33)comprised a moving cylinder with wooden slats. Poujoulat's patent in 1916 (34) employed agglomerated slag or porous brick fashioned as a hollow cylinder and rotated about its horizontal axis. Flow distribution was provided using a perforated pipe placed over the cylinder. Although neither option attracted much attention at the time, these designs could well be considered vintage predecessors to rotating biological contactor technology.

1925 - 1950

Over the next twenty-five years, intermittent filtration and contact bed systems were effectively discarded in favor of trickling filter design. Within the U.S., extensive efforts were made to improve and upgrade trickling filter performance, including the development and adoption of technical standards for design loading, bed construction and system operation. (35) High-rate designs, developed to increase hydraulic capacity, were marketed by several companies, including: Lakeside Engineering (Aero-filter), Dorr/Link-Belt Comp. (Bio-filter) and Infilco (Accelo-filter). (35) In most cases, fixed-spray jets were also discarded in favor of rotating distributor systems.

Much of the popularity of these trickling filter units could certainly be attributed to their relative simplicity, ease of operation and cost-effective performance capabilities. Activated sludge was still a somewhat innovative process, and one which prompted considerable concern regarding its intensive energy demand for aeration. (31,36,37) Legal problems also plagued the activated sludge process, with costly patent infringement suits filed against several major cities by Activated Sludge, Ltd. (38) Many municipalities consequently turned away from suspended growth systems in favor of the more conservative trickling filter option.

There were, however, several tangential developments in

fixed-film technology which deserve considerations. The application of one such option, the Hays process, actually rivaled the installation of trickling filters for the period of 1930 to 1940. (39) Developed in 1930 by Clifford Hays, a chemist from Waco, Texas, this procedure employed large asbestos-concrete sheets vertically stacked with a 1" to 2" spacing. (39) This design approach was physically analogous to the Dibdin slate bed (although vertically arrayed, rather than horizontal) or the Travis colloider system (with the added feature of a diffused aeration system). By 1942, there were 63 such units in operation throughout the U.S., many of which were located at military installations. (40) However, the limited availability of corrugated asbestos-concrete sheets during wartime conditions necessitated the use of flat sheets. (41) Lacking surface rigidity, these latter sheets frequently buckled and collapsed, resulting in process failures which doomed its future consideration.

Another such resurrected concept was that of the Nidus Rack. (42) Developed by A.M. Buswell in 1929, the Nidus Rack was intended to advance the Travis Colloider principle by significantly increasing the surface area for colloid/particulate attraction. Numerous woven lattice units constructed of veneer or basket wood were placed into a contact tank and mechanically agitated to promote deposition into an underlying settling compartment. Buswell's article also mentions a number of related studies incorporating straw and corncob filter arrays. (42)

Following along the research line established by Weigand and Poujoulat, a number of investigators independently studied the use of rotating support media. J. Doman (43) reported in 1929 on the development of a contact filter using partially submerged rotating plates constructed from galvanized steel. The schematic overview provided with this report (43) bears a striking resemblance to modern RBC designs.

One further option on rotating media, the Biological Wheel, was patented by A. T. Maltby shortly before 1930. (44) The unit consisted of a series of paddle wheels partially submerged in, and rotated by, sewage flowing through a surrounding channel. Biofilm attached to these wheels consequently rotated in alternating fashion through the sewage and into the atmosphere.

1950 - PRESENT

Mohlman's Sewage Works Journal (45) editorial entitled,

"Revival of the Trickling Filter," provides an excellant commentary on the mid-twentieth century state-of-the-art for fixed-film systems. Despite referencing the relative advantages of system reliability and economy, this editorial acknowledged that trickling filters, "were almost relegated to limbo." (45) Indeed, over the next few years, conventional trickling filter construction using rock media was unquestionably surpassed by activated sludge. Mohlman also provided a timely reference to the related technologies recently developed by Buswell, Maltby, Doman and others. In essence, he collectively defended fixed-film treatment as a worthy alternative to the rapidly advancing suspended-growth concept.

At much the same time, significant developments were occuring with the incorporation of plastic-based support media into various fixed-film treatment systems. These synthesized media forms offered several advantages over naturally available materials particularly in terms of surface contact area, voidage fraction, packing density, and construction flexibility.

Research and development on plastic media proceeded along two distinct lines during the early 1950's. In America, bundled plastic units were being proposed and tested as innovative packing for stationary filter applications. (46) Investigators in Europe, though, began testing rotating plastic discs in much the same manner as Doman's rotating cast iron system. (47) These latter researchers at Stuggart University, West Germany, conducted extensive testing on wooden and plastic discs, 1 meter in diameter. (47) Further improvement by Popel and Hartman (48,49) led to the use of expanded polystyrene media which then opened the door for commercial application.

By 1957, the J. Conrad Stengelin Company in Tuttlingen, West Germany had begun manufacturing expanded polystyrene discs 2 and 3 meters in diameter for use in wastewater treatment plants. The first commercial installation went into operation in 1960, (44,45) and soon thereafter the process began to attract considerable interest through Europe.

During the early 1960's, the research division of Allis Chalmers Corporation also investigated the use of rotating discs in various chemical processing applications. Their disc was called a two-phase contactor (TPC), and was tested for applications of gas absorption and stripping, liquid-liquid extraction, liquid-liquid heat transfer, and other mass and energy transfer applications. Eventually, the device was considered for oxygen transfer. In the summer of 1965, three-foot diameter metal discs were evaluated at the Jones Island treatment plant in Milwaukee, Wisconsin. These units were

initially employed for oxygen transfer in an extended aeration process, and then tested without sludge recycle and with an attached biomass (i.e. as a biological contactor). [In retrospect, the Jones Island site was an ironic location, as it represents the original application of activated sludge on a large commercial basis]. To confirm the favorable results of these initial tests and to learn more about the treatment process, laboratory tests were subsequently conducted using a synthetic dairy waste and 3-foot diameter aluminum discs. (49)

After learning of the European activities, Allis-Chalmers reached a licensing agreement in 1968 with the German manufacturer for production and sales distribution in the U.S. The treatment process was marketed under the trade name Bio-Disc. The first commercial installation in the U.S. went into operation at a small cheese factory in 1969. (50)

In 1970, Allis-Chalmers sold its rotating biological contactor technology to Autotrol Corporation. At that time, polystyrene discs were still not competitive with the activated sludge process, primarily due to the high capital cost of the polystyrene discs. However, in 1972, Autotrol announced the development of new rotating contactor media constructed from corrugated sheets of polyethylene. Until then, (51) the RBC unit consisted of a series of parallel, flat 0.5 inch thick expanded polystyrene sheets, each separated by a 0.75 inch space. The new arrangement used 1/16 inch thick polyethylene sheets with a 1.2 inch space.

Numerous terms are used throughout the wastewater treatment literature to describe RBC's. Among the terms in current use are the following: rotating biological contactors, rotating biological discs, rotating biological surfaces, RBS, biodisks, bio-discs, biological rotating discs, rotating filters, rotating biological filters, as well as trade names such as Bio-Surf, Aero-Surf, Surfact, and BioSpiral.

Several proprietary RBC options are currently available, including the following variations on media construction: parallel disc media attached perpendicular to the rotational shift, media sheets spiral wound about the shaft, and segmented media bundles placed as pie-shaped wedges about the shaft circumference. Another recent development amongst the field of rotating media units is that of providing supplemental aeration, either for enhanced oxygen transport and/or to provide for shaft rotation. In one instance, a full-scale system employing mechanical shaft rotation will shortly be retrofitted with such aeration capabilities in an effort to enhance system performance. (52) Numerous additional research, pilot-scale and full-

scale (including commercial and industrial) investigations have also been reported in the literature, notably including the Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, held at Champion, Pennsylvania, February 4-6, 1980.

RBC's have a number of characteristics which make them an attractive process for the design engineer. They can provide a high degree of treatment and, like trickling filters, have lower energy and maintenance requirements than activated sludge units. RBC's require less land area than most other comparable processes. A large microbial population in the form of mixtures of filamentous and non-filamentous bacteria and fungi grow on the contactor surface. (53) A large active surface area is obtained by the filamentous character of the growth. RBC's can provide a highly nitrified effluent, since different biological communities can be developed and maintained in separate stages. Because the biofilm is exposed to air roughly 50% of the time, concentrated industrial wastes can be treated without becoming anaerobic. RBC's systems can be designed to handle a wide range of flows, from less than 1 MGD to over 100 MGD. (54) No recycle is required. The sloughed biomass generally has good settling characteristics and can easily be separated from waste streams.

Rotating biological contactors show high efficiency in oxygen transfer. Organic overloads are handled well due to the large biomass on the discs. (51) Since they involve attached growth, they are less likely to fail through washout when conditions adverse to biological growth occur. No bulking, foaming, or floating of sludge occurs to interfere with a plant's overall efficiency. Short circuiting in the RBC cannot occur, due to the effect of staging in this plug flow system. Shock loads are dampened. (55)

In designing a plant, RBC's have advantages beyond their low area requirements. Most RBC's operate with nominal hydraulic head, so that pumping which otherwise might be required may be avoided. The change in head through the disc sections is less than 1.0 ft. Less excavation is required for RBC's than for activated sludge aeration tanks. RBC's are versatile both in the functions they perform and in the flexibility with which they can be configured. The discs can either be rotated by mechanical drive (such as the Bio-Surf process) or use an air drive mechanism (such as the Aero-Surf process) which has fewer moving parts and uses less energy. (56) For the mechanical drive systems, a 25 ft by 12 ft diameter module which contains 104,000 ft^2 of total surface area, can be driven by a

5-hp motor. (54) The Bio-Surf process can be designed to produce an effluent BOD_5 of 10 mg/l. The composition of effluents between 10 and 20 mg/l BOD_5 generally consists of approximately 1/3 soluble and 2/3 insoluble BOD_5. (54) However the discs are rotated, RBC technology use up to 50% less energy than activated sludge units. The low speed of the mechanical drive units reduces maintenance requirements and prolongs their lives. Air driven RBC's allow the rotational speed to be adjusted by turning a few valves.

The power requirements are low because the buoyancy of the plastic discs offsets their weight, the weight of the biomass, and the weight of their support structure so that the shaft structure half submerged, has almost no resultant downward force. (53) The process is virtually absent of nuisances: no clogging of the disc surface, no flies present, and no objectionable odors or noise. A high treatment capacity exists because of the large microbial population which is contacted with wastewater and aerated. BOD removal of 90% or more are obtained on domestic or industrial wastewaters for retention times of 60 minutes or less. Toxic shock loads affect only the more completely exposed organisms so recovery is rapid and complete. Cyclical fluctuations in wastewater flowrate are absorbed with no loss in overall treatment efficiency. The time required in introducing waste to the discs to steady state operation is usually one week.

RBC's are simple to operate. Nominal skill is required in plant operation. Since the sloughed biomass settles well and can be removed more reliably than solids from activated sludge tanks, clarifier design and operation is far less critical in RBC installations.

The RBC process also 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. Reliable winter performance is obtained when the discs are sheltered by a modest enclosure.

RBC technology is not without its share of problems, however, the structural integrity of RBC units is untested by time. Plastic media has torn loose from its drive shaft in one instance. (51) Tie rods can loosen and cause uneven rotation and need for realignment. Oil leaks from drive units are common. Friedman (57) has discussed some of the failure modes for RBC's. Failure can be defined as any situation where the process does not effluent goals, or does so in an objectionable manner. Situations such as process instability to meet

effluent BOD and/or ammonia standards, or production of solids that won't settle or cannot be separated readily from the carrier stream, or production of objectionable odors are examples of process failure modes. Media separation, shaft, bearing and mechanical drive train problems are precursors of process failure. The authors of this paper know of at least 15 process failures. (58) The reasons for failure were: shaft failure, bearing failures, plastic weld failure, structural support failure, steel shaft failure, and failure of the media. Smith and Bandy (51) point out that although maintenance costs are cited as an advantage, the costs are proportional to plant capacity, exhibiting none of the economies of scale observed with other non-modular technologies. Area requirements are also proportional to plant capacity. Mechanically driven RBC's are not able to vary the rotational speed easily; each drive unit must be modified.

Enclosures are necessary where low air and wastewater temperatures occur to achieve acceptable performance. RBC systems must ordinarily be protected by a roof since heavy rains may strip off the slime growth and hail may damage the plastic discs. (59) In northern climates, an enclosed heated building may be necessary to prevent freezing during the winter. Provision for enclosures increases an RBC installation's initial cost, which is a disadvantage. However, protected RBC's probably operate more stably in winter.

With inadequate grit and primary solids removal, suspended solids may accumulate in RBC reactors, resulting in lower process efficiency and possible foul odors. This can be avoided by providing adequate primary treatment. The RBC operation is subject to influent fluctuations which upset other processes; although RBC's handle organic and hydraulic shock loadings comparatively well, but with some loss in process efficiency. Toxic substances can cause a catastrophic loss of biomass from the discs, although recovery is more rapid than that of trickling filters under similar toxic loadings. Extremes of pH have an adverse effect on RBC performance. Overloadings on the first stage of RBC'c can cause an odor problem and loss of efficiency.

The conclusions on the advantages and disadvantages of the RBC process are varied. Antonie and Hynek (60) concluded the RBC processes are stable, versatile, and competitive with activated sludge. Their studies included a wide variety of municipal and industrial wastewaters. Thomas and Koehrsen (61) worked with distillery wastewaters, concluding that the activated sludge process was more stable when subjected to shock

loads, provided better removals, and was less expensive on both capital outlay and annual cost basis. Some disadvantages charged to the RBC provess will probably disappear as the technology matures. Controversy exists regarding design criteria, matrix design, surface-to-volume ratio for the reaction chambers, optimum rotational speeds, appropriate scale up procedures, recirculation requirements, and media configuration. Antonie (48) further compares the rotating biological contactor with the trickling filter process. Further operational experience, additional research, and symposia such as this one can be expected to remedy these shortcomings.

At much the same time (i.e. early 1950's) that the West German researchers began exploring plastic RBC's, American investigators at Dow Chemical Company were initiating their experiments with the production and use at plastic packing media. (46) Two initial plastic units were devised at Dow including a modified 'berl-saddle' (trademarked as Dowpac FN-90) and bundled arrays of nested, corrugated sheets (trademarked as Dowpac HCS). (46) Dow subsequently reassigned the Dowpac term, substituting it with 'Surfpac.' Further detailed review of the genealogy for these synthetic media is provided in the following paper by Bryan. (62)

Pilot-scale tests were conducted on both Dow packing materials using various types of industrial wastes. Both performed acceptably well, but future emphasis was given to the bundled form (i.e. Dowpac HCS) because of its perceived cost-effectiveness and operational flexibility. This material was designed to distribute falling liquid wastes in thin films over large surface areas so that maximum efficiency of contact with aerobic micro-organisms was attained. It provided a high percentage of void space for unimpeded draft circulation and waste flow. It provided large surface area adherence of biological slimes. The material produced by Dow Chemical Company consisted of individual sheets of polystyrene or Saran plastic material, (63,64) corrugated in two directions, having dimensions of 3 ft by 1.75 ft. The individual sheets were typically shipped stacked in bundles, and then assembled into structurally self-supporting modules at the point of use. In assembly, the sheets provided approximately 1 inch of free space. These modules were laid in the filter structure in a layered grid pattern to provide good distribution of flow of liquid, and to assist in structural stability. Void space within the assembled filter bed was about 94 percent. Assembled weight of the individual modules was 4 to 6 lb/ft^3. This enables the modules to be stacked to depths of 30 to 40

feet, conserving the use of land space.

Because of the variable character of different wastes, Dow developed two types of plastic, each suitable for certain waste streams:

1. Dowpac 10, which has good resistance to alkalies, salts, dilute mineral acids, and water, and is stated not to be suitable for some hydrocarbons, ketones, oxidizing acids, vegetable fats, and oils.
2. Saran, originally known as Dowpac 20, which is extremely chemically resistant to all common acids and alkalies, with the exception of strong ammonium hydroxide. It is suitable for most alcohols, esters, ketones, nitroparaffins, benzene, xylene, and toluene which diffuse slowly through the interstices between the modules, and have little effect on the material itself.

The sheets of Dowpac 10 are assembled with a solvent adhesive supplied by the manufacturer. Dowpac 20 is heat welded by special assembly machines supplied by Dow. In estimating the cost of plastic media for trickling filters, the cost of assembly must be included. The use of heat welding caused some modules to literally go up in smoke, which was a common failure (64).

Because of the light weight of this new material and its available void space, the development of small diameter towers with great height has occurred. This has incorporated important savings in the use of the filter since it materially reduced the amount of underdrain required. The enclosing structure for the trickling filter may be made of aluminum or other light metal or wood, since no structural containment walls are necessary. In place of the vitrified underdrain tile used in ordinary trickling filters, these under drains may be made of pressure-treated lumber concrete partition blocks, subway grating, etc. (63) Since the assembled modules are rectangular in shape, to avoid expensive cutting and shaping of the material, the tower structures are usually rectangular or hexagonal in shape.

The advantages of this lightweight and resistance substance generated the interest of other manufacturers. Since that time, similar plastic materials have been developed. ICI offers a polyvinyl chloride (PVC) packing named Flocor, which was formerly available from Ethyl Corporation. This was developed in England by the Imperial Chemical Industries, Ltd, and consists of flat and corrugated sheets bonded into a module 2 feet in width and depth, and 4 feet in length. The configu-

ration of the sheets is a patented feature. It offers such advantages as large practical surface area with the lowest bulk density.

Another development in the field is the use of a polyvinyl chloride plastic called Koroseal developed in 1963, produced by B.F. Goodrich Industrial Products Company. (63,64) The material is shipped, packaged, and assembled into modules in the field. The most outstanding demonstration of the use of this material was at the Rome, Georgia mill of a manufacturer of kraft paper. (63) This filter handles a flow of 16 MGD and is 80 feet in diameter, with a medium depth of 20 feet, and a total medium volume of 100,000 ft^3. The material is supported on epoxy-coated steel gratings in the tower, which has concrete block walls, with a total height of 30 ft. To fit the rectangular module shape, the tower is octagonal in shape. B. F. Goodrich next changed to a lower density medium (4) but contained less surface area (27 ft^2/ft^3). This material, derived from polyvinylidine chloride, required thicker sheets. The sine wave corrugations had a wave length of 4 inches and an amplitude of 2 inches. This compares with the Koroseal, having 37 ft^2/ft^3, of a sine wave corrugation with wavelength 3 inches and 1.5 inch amplitude. A 1.5 inch amplitude is generally the smallest amplitude put into use commercially, otherwise bridging and plugging problems occur, especially for high BOD wastewaters. B. F. Goodrich (4) has developed a cooling tower media, which can be used for nitrification-denitrification operations. This material has a surface area to volume ratio of 44 ft^2/ft^3. The corrugations are of wavelength 1.5 inches and amplitude 1 inch. Another recent development was Vinyl Core.(65)

An additional development in the plastic line was provided by American-Standard, New York. A cellulose-fiber sheet impregnated with plastic resin was made in a honeycomb design and was suitable for stacking in a column. Other varieties of plastic material for trickling filters are offered by Tex-Vit Company at Texas and Norton Chemical Process Products Division. (66) The structural engineering aspects of the plastic media has been addressed by Mabbott.(67)

Table I. Comparison of Plastic and other Trickling Filter Media (63)

Source	Brand Name	Density lb/ft^3	Surface area, ft^2/ft^3	Void Space %
Dow Chemical Co.	Surfpac	3.6	25	94
B. F. Goodrich	Koroseal	2.7-3.5	40	94
ICI	Flocor	4.06	--	95
Raschig Rings	--	30.3	22.7	74.9
Blast furnace slag	--	68.0	20	49
Stone, granite	--	90.5	30	45

Table II. Available Synthetic Media (48)

Supplier	Trade Name	Construction	Specific Surface Area ft^2/ft^3
Envirotech Corp. Brisbane, CA	Surfpac[a]	Flat and Corrugated PVC sheets	27
B.F. Goodrich Marietta, OH	Koroseal Vinyl Core	Flat and Corrugated PVC sheets	30.5
ICI Great Britain	Flocor[b]	Flat and Corrugated PVC sheets	29
Neptune-Microfloc Corvallis, OR	Del-Pak[c]	Horizontal wooden slats	14
Koch Eng. Co. New York, NY	Flexirings	Plastic pall rings	28
Norton Chemical Co. Akron, OH	Actifil	Plastic pall rings	29
Institute de Reserche Chimique Applique, France	Cloisonyle	PVC tubes	68.5

[a]Formerly available from Dow Chemical Co., Midland, MI

[b]Formerly available from Ethyl Corp., Baton Rouge, LA

[c]Formerly available from Del-Pak Corp., Corvallis, OR

Table I illustrates the weight and surface area advantages of these synthetic materials. Diversity within the current market of proprietery plastic packing media is demonstrated by Table II.

The primary merits associated with trickling filters stem from their simplicity low operating cost and ease of operation, which makes them ideal for remote sites or small communities (68). Because large masses of organisms must be present to achieve high quality effluents, they possess substantial reserve capacities making them robust and tolerant to changes in the influent. The dense nature of the microbial film which slough from the media produces sludges of relatively constant character which can be removed by sedimentation. Trickling filters have an ability to survive shock loads of toxic wastes (69) due to the relatively short retention time of the wastewater in the reactor (70) and/or because only organisms on the surface may be killed. If a shock load of long duration is applied or of a type which will be adsorbed onto the biofilm, then the trickling filter can be severaly affected (71,72).

Problems of clogging by excess biomass have been experienced when using a trickling filter, due to having too small an interstitial volume within the stones. The clogged areas become anaerobic, generating objectionable odors, and are difficult to clear once clogged. Filter flies often breed in a trickling filter to cause a further nuisance. The major operational problems of trickling filters are associated with cold weather operation, producing excessive cooling of the wastewater and ice formation on the surface of the stones. Efficiency in high rate filters is reduced with decreased temperature by approximately 30 percent per 10°C. Freezing may cause partial plugging of the filter medium and resulting over load in the open area. In northern climates, fiberglass covers or windbreaks have been employed to prevent ice formation. Covers also help contain odors which may be produced in the filter.

The main reason for the gradual loss of popularity of the trickling filter is the limited degree of treatment achieveable. Some of the largest plants have been built in recent years, but the use of the trickling filter is steadily decreasing, due to its inability to consistently achieve high degrees of soluble BOD removal. The short wastewater retention time limits the soluble BOD removal to the extent that it cannot meet the levels of treatment possible in an activated sludge system with a much longer retention time. With effluent discharge requirements becoming more stringent, the trickling

filter could no longer compete economically with the activated sludge process. The popularity of the trickling filter has also lost some of its popularity in favor of the rotating biological contactor. (49,73)

Generally operated as aerobic systems, these latter packed bed units typically receive a trickling flow which facilitates desired tower ventilation. Submerged contact has been recently tested, though, both for aerobic and anaerobic treatment. Tunick et. al (74) and Hines and Weeter (75) have accordingly reported on the behaviour of upflow anaerobic contact systems packed with selected media materials. A down-flow submerged contact process has also been marketed by Cytox (76), incorporating a parallel array of vertically stacked plastic sheets. Continuous fluid recycle within the vessel is directed towards a splash pad above the tank which then promotes oxygen transport. Aside from this latter aeration mechanism, the Cytox system could well be considered a resurrected Hays process. Another option for submerged media will be presented by a subsequent author, Li and Whang (77). This unique approach employs a synthetic ribbon media design which is then unfurled and weighted to maintain extension.

SUMMARY

This paper has described the important historical developments of fixed-film wastewater treatment systems. Beginning in the 1860's with filtration columns, various methodologies have been developed for wastewater treatment. This paper addressed the development of such fixed-film systems like trickling filters, intermittent filtration, contact beds, hydrolytic tanks, and rotating biological contactors. This paper can not possibly include all the relevant references on fixed-film processes. Rather, the goal of this paper is to highlight the technological advances which have occurred within the field. Fluidized bed systems have not been included in this discussion. They were intentionally omitted since they are semi-suspended growth cum fixed growth systems. Figure 1 highlights the important chronological developments of fixed-film wastewater treatment systems. This figure provides a quick synopsis of the involved genealogy described in this paper.

With the resurgence of interest in fixed-film applications, these processes are indeed consistent with the current federal policy regarding "trickle down theory." (78)

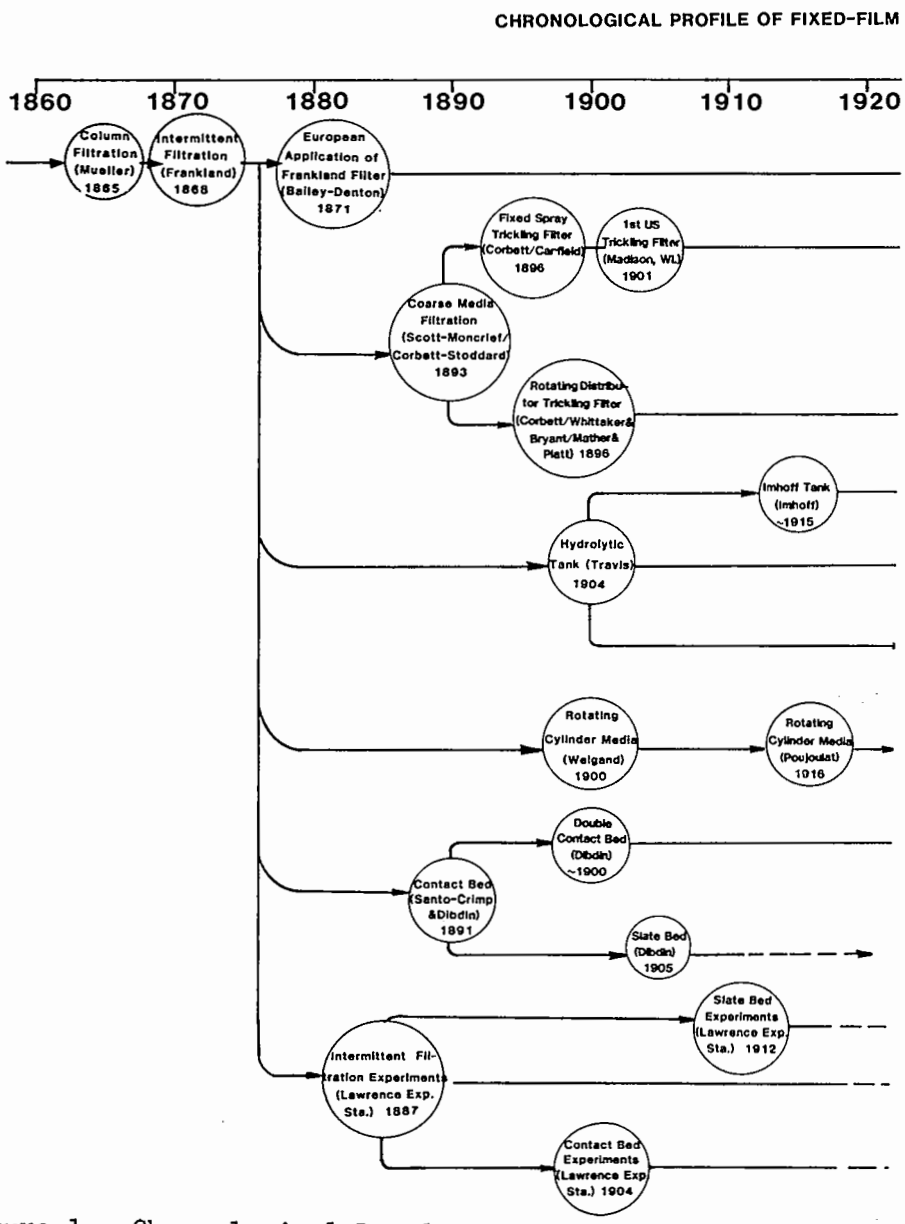
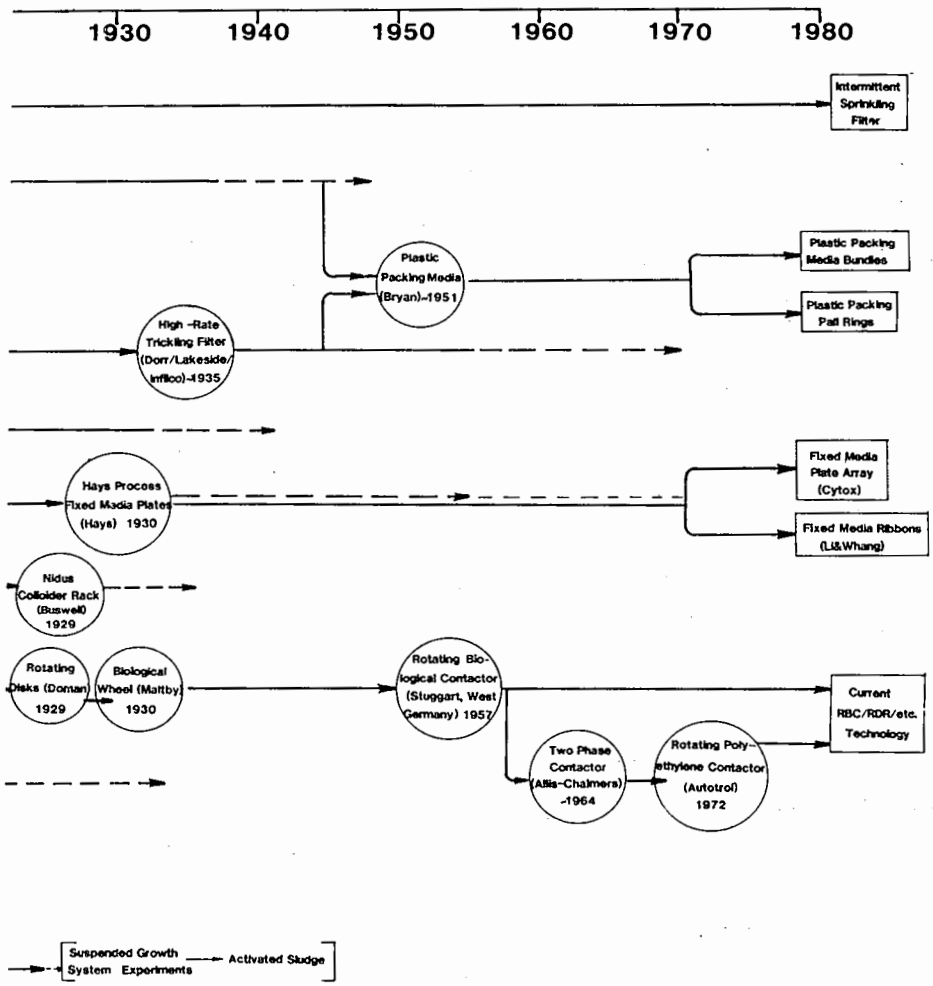


Figure 1. Chronological Development of Fixed-Film Wastewater Treatment Systems.

WASTEWATER TREATMENT SYSTEMS



REFERENCES

1. Dickens, C., A Tale of Two Cities, Oxford University Press, London (1949).
2. Charnock, A., "When Beer Was Safer Than Water." New Civil Engineer, 18-19, 21/28 December (1978).
3. ASCE Committee on History and Heritage, Pure and Wholesome, Amer. Soc. Civil Engin. (1982).
4. Metcalf, L. and Eddy, H. P., American Sewerage Practice, Vol. III - Disposal of Sewage, McGraw-Hill Book Company, Inc., New York (1916).
5. Chanlett, E. T., Environmental Protection. McGraw-Hill Book Company, New York, NY (1979).
6. Thompson, J. T., "The Fifth Report of the Royal Commission on Sewage Disposal." Jour. Soc. Chemical Industry, 33 (2) 72-76 (1909).
7. Calvert, H. T., "The Eighth Report of the Royal Commission on Sewage Disposal." Jour. Soc. Chemical Industry, 32 (6), 265-274 (1913).
8. Johnston, J. H., "Final Report of Royal Commission on Treating and Disposal of Sewage." Jour. Soc. Chemical Industry, 34, 507-508 (1915).
9. Kinnicut, L. P., Winslow, C. E. A., and Pratt, R. W., Sewage Disposal, John Wiley and Sons, Inc., London (1919).
10. Anonymous, "Sewage Treatment and Disposal of Town Refuse." Engineering, 30, 519 (1884).
11. Pearse, L., Modern Sewage Disposal, Lancaster Press, Inc. Lancaster, PA (1938).
12. Kiersted, W., Sewage Disposal, John Wiley and Sons, New York, NY (1894)
13. Hutchinson, C. C., "On the Disposal of Sewage Sludge." Jour. Soc. Chemical Industry, 3, 41-51 (1889).
14. Buswell, A. M., "The Chemistry of Water and Sewage Treatment, Chemical Catalog Company, Inc., New York, NY (1928).
15. Schloesing, T. and Müntz, A., "Sur la Nitrification Par les Ferments Organisés," C. R. L'Acadamie Des Sciences, 84, 301 (1977).
16. Warrington, R., "On Nitrification," Jour. Chemical Society, 59, 484 (1891).
17. Winogradsky, S., "Sur les Organismes de la Nitrification." C. R. L'Acadamie Des Science, 110, 1013 (1890).
18. Hazen, A., "(Abstract) The Lawrence Experiments on the Purification of Sewage in 1890 and 1891." Engineering News-Record, 28, 559-560 (1892).

19. Clark, H. W., "Past and Present Developments in Sewage Disposal and Purification." Sewage Works Journal, 2, 561-571 (1930).
20. Corbett, J., "A Dozen Years of Sewage Purification Experiments on a Large Scale at Salford, England." Engineering News-Record, 49, 191-192 (1903).
21. Stoddart, F. W., "Nitrification and the Absorption Theory." Jour. Soc. Chemical Industry, 30, 236-237, (1911).
22. Folwell, A. P., Sewerage, John Wiley and Sons, Inc., New York, NY (1918).
23. Lowcock, S. R., "Experiments on the Filtration of Sewage." Jour. Inst. Civil Engin., 115, 297 (1893).
24. Kratz, H., "Operation of Trickling Filters at Baltimore." Sewage Works Journal, 4, 135-141, (1932).
25. Tatham, G. T. P., "Some Theoretical Considerations Bearing on the Performance of Biological Sewage Purification Plant." Jour. Soc. Chemical Industry, 35, 711-715, (1916).
26. Hurley, J., "An English View on Contact Aeration." Engineering News-Record, 130, 895 (1943).
27. Keefer, C. E., Sewage-Treatment Works, McGraw-Hill Book Company, Inc., New York, NY (1940).
28. Babbitt, H. E., Sewerage and Sewage Treatment, 3rd edition, John Wiley and Sons, Inc., New York, NY (1928).
29. Dibdin, W. J., "The Slate Bed Treatment of Sewage." Jour. Soc. Chemical Industry, 32, 55-61 (1913).
30. Lanphear, R. S., "Sewage Purification Results at Plainfield During 1910-1911." Engineering Record, 64, 29 (1911).
31. Martin, A. J., The Activated Sludge Process, MacDonald and Evans, London (1927).
32. Ardern, E. and Lockett, W. T., "Experiments on the Oxidation of Sewage Without the Aid of Filters," Jour. Soc. Chemical Industry, 33, 523-539 (1914).
33. Weigand, Ph., "Varfahren zur Biologischen Reinigung von Abwassern," German Patent No. 135755, 1900.
34. Poujoulat, G., "Apparatus for the Purification of Sewage." French Patent No. 478,669; 11 May 1915, Jour. Soc. Chemical Industry, 35, 1129 (1916).
35. Hardenbergh, W. A., Sewerage and Sewage Treatment, International High Rate Filters Textbook Company, Scranton, PA (1936).
36. Ardern, E., "A Resume of the Present Position of the Activated Sludge Process of Sewage Purification." Jour. Soc. Chemical Industry, 36, 822-830 (1917).

37. Hammond, G. T., "Some Remarks on the Biological Treatment of Sewage, With Especial Reference to the Activated Sludge Method." Jour. Ind. Engin. Chemistry, 9, 399-403, (1917).
38. Anonymous, "The Activated Sludge Patent." Engineering News-Record, 113, 507 (1934).
39. Griffith, L. B., "Contact Aeration for Sewage Treatment." Engineering News-Record, 130, 138-142 (1943).
40. Kessler, L. H. and Norgaard, J. T., "Operation of Army Sewage Treatment Plants." Sewage Works Journal, 14, 1228-1261 (1942).
41. Otts, L. S.,Personal Communication, University of Maryland, College Park, MD (1982).
42. Buswell, A. M. and Pearson, E. L., "The Nidus (Nest) Rack, A Modern Development of the Travis Colloider." Sewage Works Journal, 1, 187-195 (1929).
43. Doman, J., "Results of Operation of Experimental Contact Filter with Partially Submerged Rotating Plates." Sewage Works Journal, 1, 555-560 (1929).
44. Allen, K., "The Biological Wheel." Sewage Works Journal, 1, 560 (1929).
45. Mohlmann, F. W., "Editorial: Revival of the Trickling Filter." Sewage Works Journal, 8, 647-648 (1936).
46. Bryan, E. H., "Molded Polystyrene Media for Trickling Filters." Proceedings of the 10th Purdue Industrial Waste Conference, Purdue University, Engineering Bulletin, 164-172 (1955).
47. Daillaire, G., "Behind the Rapid Rise of the Rotating Biological Contactor." Civil Engineering-ASCE, 49, 70-73 (1979).
48. Antonie, R. L., Fixed Biological Surfaces - Wastewater Treatment, Chemical Rubber Company Press, Cleveland, OH 1976.
49. Antonie, R. L., Private Communication, Autotrol Corporation, Milwaukee, WI, 1982.
50. Birks, C. W., and R. J. Hynek, "Treatment of Cheese Processing Wastes by the Bio-Disc Process," Proceedings of the 26th Purdue Industrial Waste Conference, 26:89-105, Purdue University, West Lafayette, IN (1971).
51. Smith, E. D., and J. T. Bandy, "A History of the Rotating Contactor Process," Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, Vol. I, pp. 11-23, 1980.
52. Canaday, J., Personal Communication, University of Maryland, College Park, MD (1982).

53. Antonie, R. and K. Van Aacken, "Rotating Discs Fulfill Dual Wastewater Role," Water and Wastes Engineering, 8 (1): 37-38, 1971.
54. Benefield, L. D., and C. W. Randall, Biological Process Design for Wastewater Treatment, Prentice Hall, Englewood Cliffs, NJ, 1980.
55. Metcalf and Eddy, Inc., Wastewater Engineering: Treatment, Disposal, Reuse, 2nd edition, McGraw-Hill, New York, NY, 1979.
56. Autotrol Wastewater Treatment Systems: Design Manual, Autotrol Corp., Bio-Systems Div., Miluwakee, WI (1978).
57. Friedman, A. A., "RBC Design Considerations for Industrial Wastewaters." Paper presented at the 74th Annual Meeting of the American Institute of Chemical Engineers, New Orleans, Louisiana, November 8-12, 1981.
58. Anderson, B. L., Private Communication, B. L. Anderson Co., West Lafayette, IN, 1982.
59. Steel,E. W., and T. J. McGhee, Water Supply and Sewerage, Fifth edition, McGraw-Hill, New York, NY, 1979.
60. Antonie, R. L., and R. J. Hynek, "Operating Experience with Bio Surf Process Treatment of Food Processing Wastes," Proceedings of the 28th Purdue Industrial Waste Conference, 28: 849-860, 1973.
61. Thomas, J. L., and L. G. Koehrsen, "Activated Sludge-Bio Disc Treatment of Distillery Wastewater," EPA-6601 2-74-014, Superintendent of Documents, Washington, DC, 1974.
62. Bryan, E. H., "Development of Synthetic Media for Biological Treatment of Municipal and Industrial Wastewaters." Proc. 1st Internat. Conf. Fixed-Film Biol. Processes, Kings Island, OH, 22 April (1982) [in press].
63. Besselievre, E. B., and M. Schwartz, The Treatment of Industrial Wastes, 2nd edition, McGraw-Hill, New York, NY, 1976.
64. Grieves, C. R., "Dynamic and Steady State Models for the Rotating Biological Disc Reactor," Ph.D. Dissertation, Clemson University, Clemson, SC, 1972.
65. B. F. Goodrich, "Vinyl Core and Koro-Z: PVC Filter Media for Wastewater Treatment," Bulletin FPD-81-EP-5027, B. F. Goodrich, Akron, OH, 1981.
66. Norton Chemical Process Products Division, "Actifil Media for Packed Biological Reactors," Bulletin ACT-30, Norton Chemical Process Products Division, Akron, OH.
67. Mabbott, J. W., "Structural Engineering of Plastic Media for Wastewater Treatment by Fixed-Film Reactors." Paper presented at the First International Conference on Fixed-

Film Biological Processes, King's Island, OH, April 20-23, 1982.
68. Chipperfield, P.N.J., M. W. Askew, and J. H. Benton, "Multistage, Plastic-Media Treatment Plants," JWPCF, 44, 1955-1967, 1972.
69. Dow Chemical Co., Functional Products and Systems Department, "A Literature Search and Critical Analysis of Biological Trickling Filter Studies," Vol. I and II, Environmental Protection Agency Water Pollution Control Research Series, Report No. 17050 DDY 12/71, December 1971.
70. Cook, E. E., and L. P. Herning, "Shock Load Attenuation, Trickling Filter," J. Environ. Engr. Div., ASCE, 104: 461-469, 1978.
71. R. E. McKinney, "Biological Treatment Systems for Refinery Wastes," JWPCF, 39: 346-359, 1967.
72. Cillie, G. G., "The Effect of Toxic Loads on Effluent Purification Systems," Water SA, 3: 83-89, 1977.
73. Gray, K. L., Private Communication, B.F. Goodrich Co., Engineered Products Division, Akron, OH, 1982.
74. Tunick, M. H., Friedman, A. A., and Bailey, D.G., "Treatment of Tannery Beamhouse Waste with a Bench-Scale Anaerobic Reactor." Proc. 13th Mid-Atlantic Ind. Waste Conf., Ann Arbor Science Publishers, Inc., 197-207 (1981).
75. Hines, D. and Weeter, D. W., "Production of Ammonia in a Packed Bed Anaerobic (ANFLOW) Bioreactor." Presented at the 1980 Annual AIChE Meeting, Chicago, IL, 19 November (1980).
76. CTX™ BIOX Treatment System, Cytox Equipment Corp., Dayton, OH, U.S. Patent No. 5. 3,966,608; 3,972,965; 3,966,599.
77. Li, C. T. and Whang, J. S., "Treatment of Domestic Sewage by Aquatic Ribbon System," Proc. 1st Internat. Conf. Fixed-Film Biol. Processes, King's Island, OH, 22 April, 1982 [in press].
78. Hao, Oliver. Personal Communication, University of California-Berkeley, Berkeley, CA, 1982.

DEVELOPMENT OF SYNTHETIC MEDIA FOR BIOLOGICAL TREATMENT OF MUNICIPAL AND INDUSTRIAL WASTEWATERS

Edward H. Bryan. Division of Civil and Environmental Engineering, National Science Foundation, Washington, D.C.

ABSTRACT

In Midland, Michigan, during June of 1954, a pilot-scale experimental trickling filter ten-feet in diameter and ten-feet deep began receiving the unsettled effluent from The Dow Chemical Company's four conventional trickling filters, the first of three stages for biological treatment of its strong phenolic wastewater. Half of the experimental unit was filled with crushed blast furnace slag identical to that used in the four large filters. The other half was packed with a fabricated plastic medium trademarked Dowpac HCS (since re-named Surfpac). With biological activity evident after eleven days, the feed was changed to a synthetic wastewater containing pure phenol and ammonium phosphate dissolved in Midland tapwater.

A paper presenting results of the direct comparison between performance of the two media in the experimental unit was presented in May of 1955 at the Tenth Purdue Industrial Wastes Conference and subsequently published in its Proceedings. From 1954 through 1960, an extensive research and development program was conducted by The Dow Chemical Company with cooperation of potential industrial users, municipalities, consulting engineers, educators and government personnel at local, state and federal levels. During this period, results from design, construction and/or operation of approximately 35 units provided guidance for decisions made during the development period.

This paper presents aspects of the critical early stages in the development of plastic media, experiences with relevance and potential applicability to current implementation of

innovative and alternative solutions to problems of wastewater treatment and management. Previously unpublished results from operation of several experimental units during the period from 1954 through 1958 are presented. Included are data and results from units that were packed to depths of 42 feet and which were constructed to make intermediate depth-sampling possible. One unit, which was constructed to permit measurement of air-flow through the packing, provided data confirming the previously known but sparsely documented potential for stagnation in trickling filters, a factor potentially affecting performance.

INTRODUCTION

During 1953, The Dow Chemical Company's effort to produce a tower packing for its own internal needs resulted in the successful development of two types of media that could be produced from synthetic plastics. Then trademarked "Dowpac FN-90" and "Dowpac HCS"*, efforts were initiated early in 1954 to investigate broadening their potential application to cooling of water and biological treatment of wastewaters.

The earliest public disclosure of Dow's pioneering work in development of synthetic media for biological treatment of municipal and industrial wastewaters was by Griess in a paper presented at a meeting of the American Chemical Society in 1954 (1). This paper contained initial, preliminary data from operation of a pilot-scale experimental trickling filter, ten-feet in diameter and ten-feet deep, half-filled with crushed blast-furnace slag and the other half containing Dowpac HCS (Figure 1).

In contrast to Dowpac FN-90, a unique modification of the conventional "berl-saddle" type of packing, which was injection-molded; Dowpac HCS was vacuum-formed from flat sheets of plastic. The forming process produced corrugations at right-angles to each other, and ribs that served to stiffen the individual sheets, produce an average spacing of one-inch between sheets when assembled into packs, and as positions of additional contact for joining sheets into packs.

*The Dow trademark "Dowpac" was reassigned to other products and replaced by "Surfpac" after the period of time during which the author of this paper was responsible for conduct of the research and development program described in this paper. To avoid any misunderstanding, the trademark designation used in this paper coincides with that in use when the work was conducted that led to results cited.

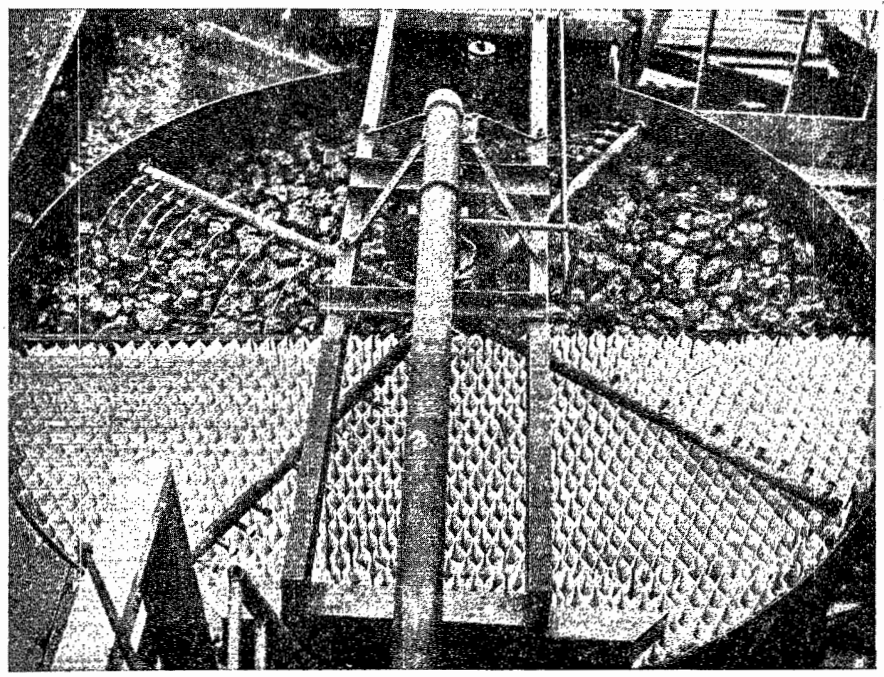


Figure 1. Pilot-scale experimental trickling filter, 10-feet in diameter, 10-feet deep filled with conventional crushed stone and Dowpac HCS media at The Dow Chemical Company in Midland, Michigan (1954).

The unique design of Dowpac HCS permitted individual sheets to "nest" in one position, but when alternate sheets were rotated 180° in the plane of each sheet, the pack expanded to produce a structure with the appearance of a "honeycomb" when viewed from either end. The combination of edge-loading, rib-stiffening and composite-sheet action produced modules of remarkable strength when subjected to compressive loading (Figure 2).

Experimental operation of the original pilot-scale unit which began in June of 1954 continued until September of 1955. With the exception of the initial eleven days during which the unit was inoculated by passing through it the effluent (unsettled) from the full-scale, conventionally packed Dow phenolic wastewater treatment plant trickling filters, the unit was operated until June of 1955 using a synthetic waste-

water consisting of pure phenol and a proportional amount of ammonium phosphate dissolved in City of Midland tapwater (treated Lake Huron water). In June of 1955, the unit was put on-line in parallel operation with the full-scale Dow trickling filters and was used to evaluate other potential packing shapes, materials and configurations as part of the materials/fabrication component of the development program. The range of phenol concentrations to which the pilot unit was subjected during the initial phase of its operation (on pure phenol) was from 10 to 536 mg/1. Results of this pilot plant study were presented by Bryan at the Tenth Purdue Industrial Waste Conference in May of 1955 (2) and were subsequently also published in Industrial Wastes magazine (3).

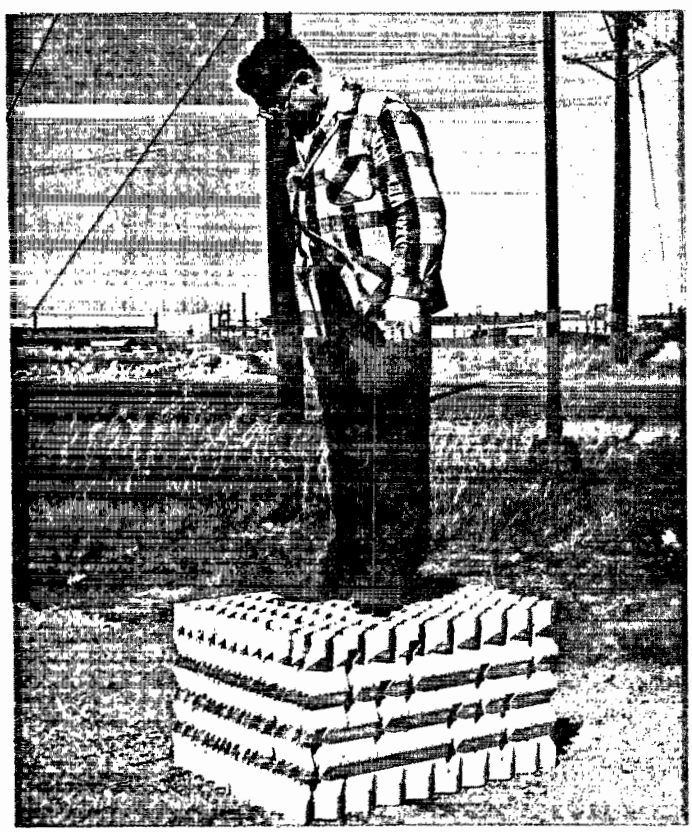


Figure 2. Stanley Mogelnicki, Supervisor of Waste Treatment Operations, The Dow Chemical Company standing on a module of Dowpac HCS illustrating its ability to support weight of treatment plant personnel.

A unit of identical dimensions to the initial pilot plant was constructed and packed with Dowpac FN-90. It was operated in series with the original unit. Despite favorable results, it was concluded that Dowpac FN-90 was likely to be more expensive and less likely to provide the flexibility in design of full-scale units for biological treatment of wastewaters when compared with Dowpac HCS.

During the preparations for operating the initial pilot-scale unit, it became evident that while polystyrene resin used in fabrication of the media would be satisfactory for process evaluation, it would not be satisfactory for the wide range of conditions to which full-scale units would be subjected. Test coupons of alternative materials were placed on and buried within the full-scale Dow trickling filters. While there was some variance in the length of time it took to form the initial films, all plastics tested responded favorably. Process studies to assist in identifying and characterizing the potential market were given priority over further research on materials of fabrication. The Dow Plastics Technical Service Bulletin issued in October of 1955 (4) announced availability of the two packings, suggested some potential applications, listed their physical properties, contained results of research to date, and contained a note of caution regarding limitations of polystyrene with regard to its chemical resistance.

TECHNICAL PROCESS EVALUATION PROGRAM

During 1954, it became evident that patent protection was likely for the unique designs of both packings but process patent protection in conventional applications for biological treatment of wastewaters was not. Accordingly, the decision was reached to utilize the technique of full public disclosure and offers of cooperative assistance to industries, municipalities and consulting engineers who expressed interest in assessing the potential applicability of the packings to meet their needs as the principal component of the Dow development strategy.

The previously cited paper presented at the Purdue Industrial Waste Conference was followed by technical papers which were essentially reports of progress on the Dow research and development program at the Michigan (June 1955), West Virginia (October 1955), Kansas (April 1956), Central States (June 1956), and Pennsylvania (August 1956) Sewage and Industrial Waste Association annual meetings, and at the Texas A & M Short Schools in March of 1956 and 1957 by Bryan (5)(6).

Kountz of the Pennsylvania State University referred to the results of his Dow-funded studies using catalyzed sodium sulfite to measure capacity for oxygen-transfer in a "philosophical" paper on "total oxidation treatment" at the Purdue (7) and Honey Harbour, Ontario Industrial Waste Conferences in May and June of 1956, respectively. In May of 1956, Towne and Becher of the U. S. Public Health Service's Robert A. Taft Sanitary Engineering Center presented a brief report on a Dowpac HCS research project that was in progress at the Battle Creek, Michigan wastewater treatment plant to the annual meeting of the Michigan Sewage and Industrial Wastes Association meeting in Benton Harbor.

All personnel who were cooperating with Dow in this development program were encouraged to present their findings in technical papers at conferences and meeting that were appropriate to their content. Stack presented results of a pilot plant study conducted at the Union Carbide Chemicals Company's South Charlestown, West Virginia plant at a meeting of the Manufacturing Chemists Association's Air and Water Pollution Abatement Committee's Joint Conference in Washington, D. C. on April 4, 1957. Trepanier (8) presented results of his research that was conducted at the Ford Motor Company's coke production plant in Dearborn, Michigan at a conference in Pittsburgh, Pennsylvania on April 8, 1957. Mills of the Naugatuck Chemicals Company in Elmira, Ontario discussed his research at the Ontario Industrial Waste Conference in Honey Harbour, Ontario on June 10, 1957.

Results from this expanding external evaluation program continued to be encouraging, equalling or exceeding the original process-related expectations and confirming results of a continued, parallel internal research and development program. While providing gradually increasing encouragement for its process-potential, the program was equally effective in disclosing weaknesses that would need to be addressed before marketing Dowpac HCS. Problems disclosed included confirmation of the already well-documented property of polystyrene to sustain combustion, its already well-established solubility in gasoline, and its tendency to absorb some organic compounds from wastewater which weakened its structural integrity to the point where it would no longer support the combined dead and live loads imposed on it in packed towers.

Increasing confidence in its technical promise was instrumental in increasing attention to alternative plastics for fabrication of Dowpac HCS early in 1956. A number of

approaches were tried centering around potential use of polyethylenes and polyvinyl chloride resins. By May of 1957, two test packs fabricated from Saran were sent to the Great Northern Oil Company's petroleum refinery in Pine Bend, Minnesota for preliminary testing in their trickling filter that had been originally packed with Dowpac HCS produced from polystyrene. In December of that same year, the unit was completely re-packed with 13,300 cubic feet of Dowpace HCS fabricated from Saran. The design and preliminary operation of this first, full-scale installation of a plastic-media packed trickling filter was described by Anderegg (9) in 1959 and by Bryan (10) in 1962.

During the initial, critical years while the product was in Dow's "development stage", it was necessary to simultaneously excite the interest of potential users, establish and maintain credibility regarding the relationship between its promise and proven performance, and maintain Dow internal interest to sustain the research and development program. Efforts to utilize technical forums in pursuit of public disclosure sometimes led to misunderstandings of intent. This is evident from the following abstract of a letter received from a consulting engineer in October of 1956:

> "Is Dowpac HCS available for purchase by my clients? In the plant designs I am not commiting your company as to its effectiveness nor as to claims for its use...and...if your answers are negative then I am confused. You never should have disclosed your information in technical society meetings and their journals if you did not want the engineering profession to be interested and to help you develop the ideas applications. It would seem that Dow takes the attitude of giving supreme and final approval to the engineering profession when Dow is ready. This is neither a scientific approach nor enticing to engineers interested in process development"

In response, the engineer was informed that:

> "...Dowpac HCS is not presently available for purchase except for experimental use. The magnitude of our existing program precludes duplication of experimental installations. All installations at the present time would be regarded by us as experimental...we have been pleased to participate in many technical programs by presenting 'progress reports' dealing with our work in this field. We have never

expressed nor attempted to imply that Dowpac HCS is currently available as a product at such meetings. Our desire to proceed to full-scale usage through the experimental pilot plant stage can hardly be considered by the profession as a 'neither scientific approach nor (one) enticing to engineers interested in product development'. We are sorry that your impression of our effort to provide the profession with a new and perhaps better tool for the solution of waste treatment problems is summarized by the preceding extract from your letter."

In response to another consulting engineer in October of 1957, it was necessary to emphasize again that:

"Dowpac HCS is still considered by us to be a developmental product. We feel that Dowpac HCS offers to the potential user a number of unique properties which will result under many circumstances in performance and economic advantages over conventional technology. We have strongly urged the prospective user to recognize the essential uniqueness of his particular waste treatment problem attacking it through pilot plant experimentation."

Even development of a product with such limited public appeal as a packing for wastewater treatment processes had its moments of difficulty with "the press". The March 1957 issue of Chemical Engineering contained a statement that:

"Entry of a big chemical company like Dow, with its technical and promotional skills, should produce results in a field long dominated by sanitary engineers."

In the conventional wisdom of public relations that it doesn't matter what is said about one in the media just as long as one's name is spelled right, a decision was made to not request a printed correction of this misinterpretation of the "Dow" approach which was to work through rather than around the traditional methods of obtaining product acceptability.

A somewhat more intriguing error occurred in the article by Egan and Sandlin in the August 1960 issue of Industrial Wastes (11). In their article, while correctly identifying Mead-Core, a plastic packing being developed by the Mead Corporation, Dowpac HCS and Polygrid (a plastic packing being developed by the Fluor Corporation) were reversed as to their identity in a set of four pictures and their captions.

Both Mead-Core and Polygrid bore historical ties to Dowpac HCS and The Dow Chemical Company's development strategy. The Fluor Corporation has the unique distinction of being the first customer for a full-scale installation of plastic media in their design of the previously cited Great Northern Oil Company refinery in Pine Bend, Minnesota. Recognizing promise of the basic concept inherent in its design, Fluor, in cooperation with Dow and Great Northern Oil Company personnel worked together to resolve the technical issues associated with that initial, full-scale installation while simultaneously beginning its development of the Polygrid packing, primarily for application in cooling of water. Almost forgotten "heroes" in the risk that was inherent in that initial installation were the personnel of the Minnesota Department of Health who approved the initial plan and who were patient during subsequent efforts to functionally integrate the Dowpac HCS unit into routine operation.

The Mead Corporation's interest which led to development of Mead-Core was directly related to its comparative studies of Dowpac HCS and Polygrid packings at pilot-scale (11). Cawley, who reported subsequently on full-scale use of Mead-Core at the Rome Kraft Company (12) himself conducted a Dowpac HCS pilot-scale study while with the Rayonier Corporation in Jessup, Georgia during the late 1950's.

Entry of other potentially competitive plastic media into the "arena" was an important factor in maintaining internal interest within The Dow Chemical Company, where assessment of its continued development program seemed to be subject to re-evaluation every other week. Equally important to the emergence of competitive packings was the continued evidence of technical superiority that Dowpac HCS was exhibiting over conventional media, emerging competitive shapes, and alternative processes.

During the period from May of 1955 to January of 1957, an average of one pilot plant study was initiated each month over a wide spectrum of potential applications, as summarized in Table I. The general arrangement was that The Dow Chemical Company would provide the packing and technical assistance in planning, conduct of the study, and evaluation of the results in return for a technical report of performance. While emphasis was on the external effort during this period, a complementary internal program was maintained and modestly expanded. By August of 1957, 28 pilot-plant studies had been conducted or were underway and an additional 6 were at an advanced stage in planning, design or construction.

Table I. Dowpac Development Program Summary from 1954 - 1958

Date of Internal Dow-Report	Number of Pilot Studies Ongoing and/or Complete (Summation) Internal	External	Types of Wastewater and/or Application (Items are additive)
May 1955	2	-	Synthetic phenol, Cooling water, Brine settling,
November 1955	4	2	Construction prototype, Semi-chemical boxboard, Kraft pulping
May 1956	4	8	Domestic wastewater, Coke oven, De-inking, Glycol, hydrolyzer, Sulfite oxidation, Dehumidification
January 1957	6	20	Ammonia removal, Corn steep-Vegetable oil refinery, Oil gas processing, Chlorinated phenols, Water treatment to remove carbon dioxide and hydrogen sulfide, Contact aeration, Milk waste, Sour-water scrubber, Solids flotation
August 1957	7	21 plus plans for 6	Alternative materials for media fabrication

Shortly after initiating its initial Dowpac HCS and conventional stone-packed unit, a construction prototype was designed and constructed at the Dow plant in Midland, Michigan (Figure 3). Another pilot plant containing a packed depth of 42 feet was constructed and operated at the City of Midland, Michigan Sewage Treatment Plant (Figure 4), initial results of which were presented by Bryan at the Michigan Sewage and Industrial Wastes Association meeting in June of 1955 and at the Texas Water and Sewage Works Short School in March of 1956 (6). Both units provided breadth to the development program not possible by response to external interests.

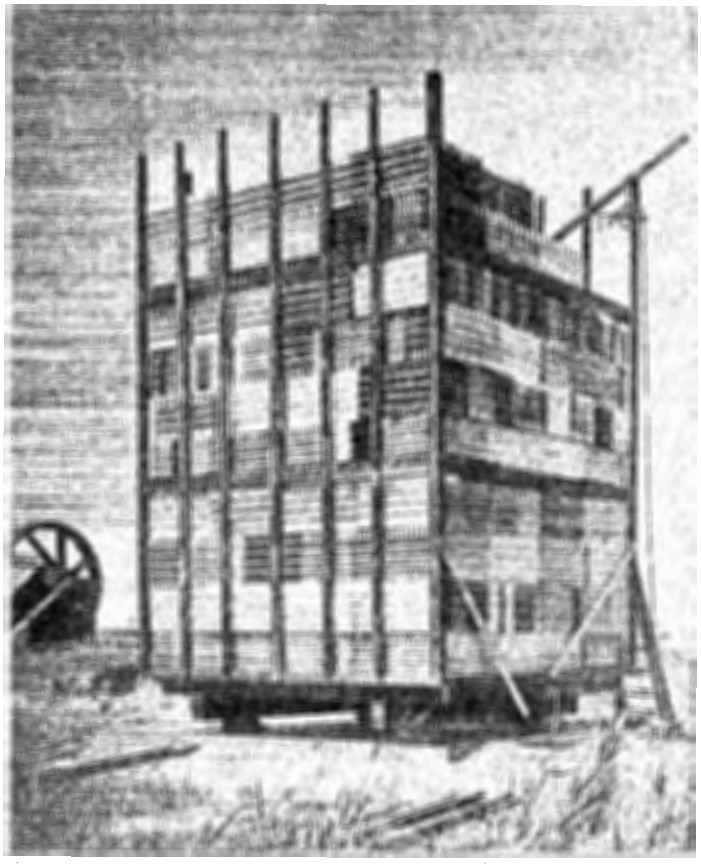


Figure 3. Dowpac HCS Construction Prototype, The Dow Chemical Company, Midland, Michigan (1955).

With the Dowpac HCS Construction Prototype, Handt (13) observed an efficiency of phenol removal of 96% for the 20-foot packed depth in comparison with 82% for the original unit containing a packed depth of 10 feet at the same hydraulic and organic loading rates. Brelsford (14) continued to operate this unit with the objective of determining the "protein-value" of harvested slimes, concluding their protein-equivalent based upon their organic nitrogen content was between 31.6 and 34.4 percent. In a subsequent study, Froman (15) found the unit to remove between 86.4 and 88.4 percent of the acrylonitrile in a synthetic wastewater using ammonium

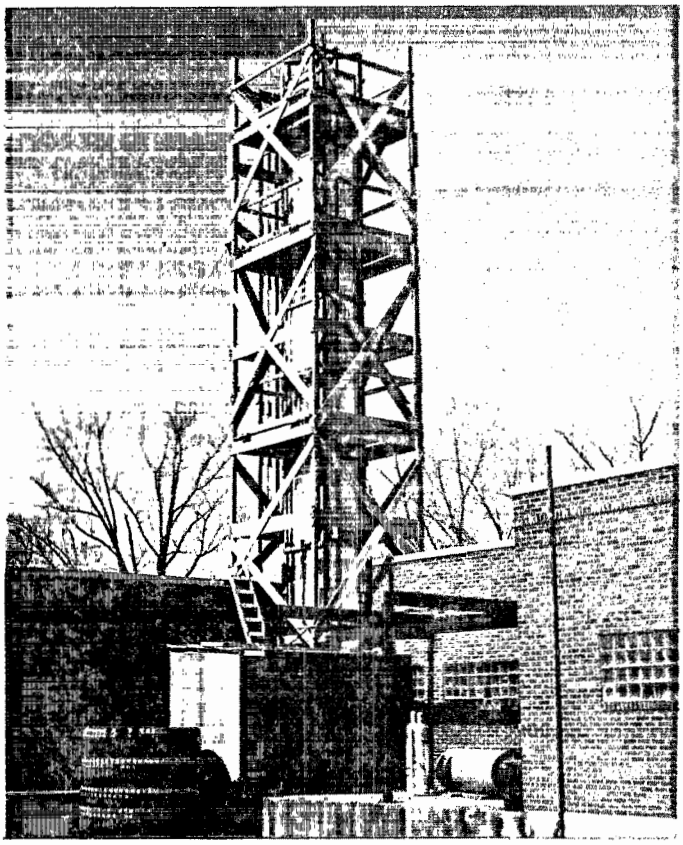


Figure 4. Dowpac HCS pilot plant at the City of Midland, Michigan Sewage Treatment Plant, containing a packed depth of 42 feet (1954).

phosphate as a supplemental source of nutrients, at loading rates of 83 to 162 pounds of oxygen-equivalent per 1000 cubic feet per day. The data from this study was used by Roy F. Weston, Inc. in the design of the full-scale plant for the treatment of wastewater at The Dow Chemical Company's acrylic fiber production facility near Williamsburg, Virginia which was placed into operation during 1958 (16).

The experimental operation of the pilot plant at the City of Midland Sewage Treatment Plant (Figure 4) included observations of air-flow by Heckeroth (17) and Greene (18). Their data (Figure 5) provided clear evidence of the potential

for reversal of air-flow through trickling filters and therefore the potential for stagnation and its consequences in affecting the performance characteristics as suggested by Bryan (19.

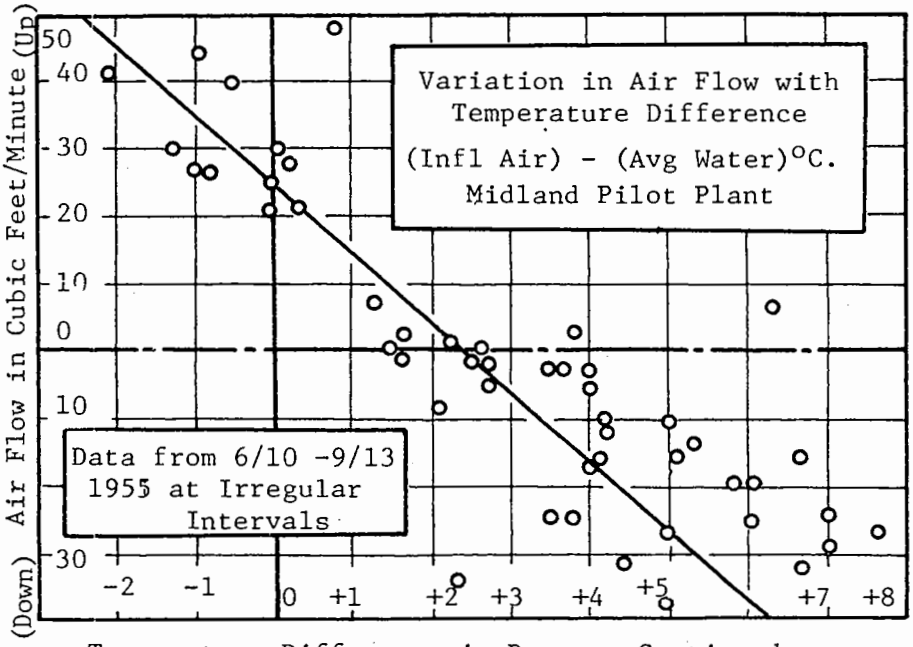


Figure 5. Relationship between temperature difference (Air - Wastewater) and air-flow through the 42-foot Dowpac HCS experimental unit at the City of Midland, Michigan Sewage Treatment Plant.

Greene conducted a four-week study in which solids from the settling tank at the City of Midland, Michigan experimental unit were returned to the Dowpac HCS tower as a "test" of the "total oxidation" concept of Kountz (7). Greene found that the loss of solids over the settling-tank weir was approximately equal to solids produced which were, in turn, produced in direct proportion to the reduction in chemical oxygen demand of the wastewater treated. His brief, preliminary study of the relationship between air-flow and performance suggested that theories of trickling filter performance and consequent "formulations" that ignore the effect that potential stagnation may have on availability of oxygen to the biologically active films may poorly represent the performance of actual trickling filters. These observations clearly indicated the

superiority of Dowpac HCS over conventional media in the freedom with which air could move under natural conditions and the relative ease with which forced-ventilation could be implemented in design, construction and operation of full-scale units.

Within the range of organic and hydraulic loadings used in studies with the City of Midland unit, its performance was found to be dependent only on the hydraulic rate of application. Results of the two rates most comparable to those used in trickling filters studied by the National Research Council were compared with the empirical formula resulting from those studies and found to be in essential agreement with those findings (Figure 6).

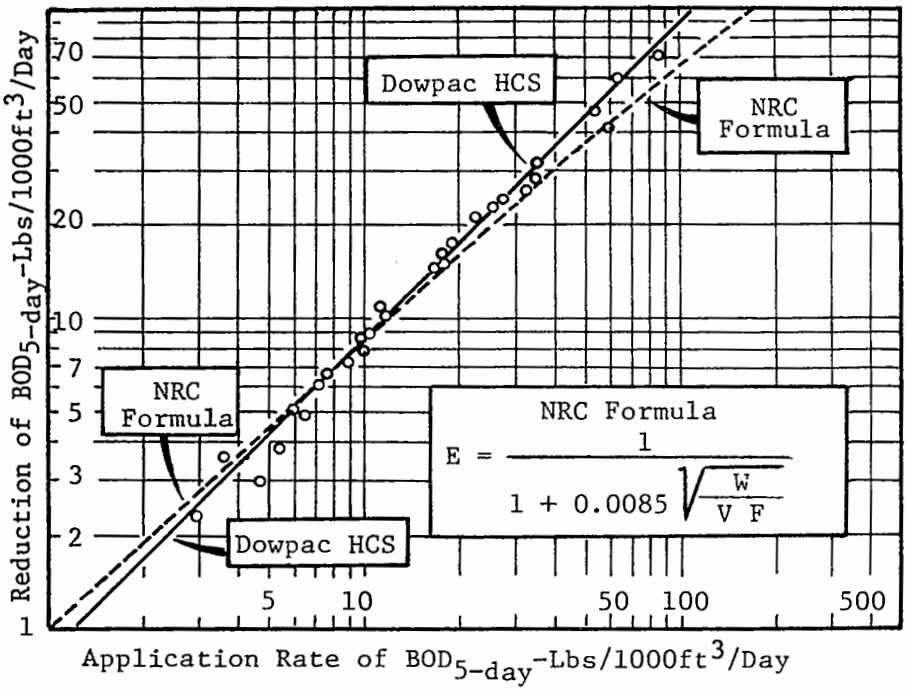


Figure 6. Performance of the Dowpac HCS unit at the City of Midland Municipal Sewage Treatment Plant which contained a packed depth of 42 feet. Hydraulic application rates for results compared to those of the National Research Council (1946) were 18 and 36 million gallons per acre per day.

The concept of returning solids to the influent with the packing "functioning as an aeration device for mixed liquors in addition to supporting bacterial slimes" as suggested by Bryan (5) was tested initially by Bauer (20) who found that the strength of the City of Midland wastewater was insufficient to build-up enough activated sludge for a good test of this concept. In a subsequent study, Ellis (21) used activated sludge from the Dow general wastewater treatment plant (10), whey from a local dairy and ammonium phosphate as a source of supplemental nutrients in tests ranging from two to nine hours in duration. He found the oxidation rate to be in a range of from 2.4 to 6.2 pounds per cubic foot per day (Chemical Oxygen Demand).

Since slimes had been chemically cleaned from the packing prior to his tests, the reduction in Chemical Oxygen Demand was solely attributed to the packing acting as an aerator. However, Ellis felt those rates were "exaggerated" by his assumptions in sampling, but after accounting for potential error, he concluded that:

> "...removal rates of greater than 1,000 pounds of Chemical Oxygen Demand per day per 1000 cubic feet were obtained."

This rate was in the mid-range of those plotted by Bryan (Figure 7) from data obtained by Kountz (22) using the cobalt-catalyzed, sodium sulfite technique in studies he conducted at the Pennsylvania State University.

Late in 1955, while studies were in progress at the City of Midland Sewage Treatment Plant, an opportunity arose to conduct a similar study in Battle Creek, Michigan. Following some preliminary discussions between personnel of The Dow Chemical Company, City of Battle Creek, and the firm of Jones, Henry and Williams (consultants to Battle Creek), a meeting was held in Battle Creek on January 4, 1956. The eleven persons present included personnel from the State of Michigan Department of Health and the Water Resources Commission and the U. S. Public Health Service's Robert A. Taft Sanitary Engineering Center in Cincinnati. A decision was reached to conduct a pilot-scale evaluation of Dowpac HCS at Battle Creek in a unit analogous to the unit in operation at the City of Midland. Financial support, estimated at $10,000, was agreed would be equally shared by the City of Battle Creek, The Dow Chemical Company, General Foods and Kellogg Corporations. A Steering Committee was appointed to include representation from all participants in the proposed study.

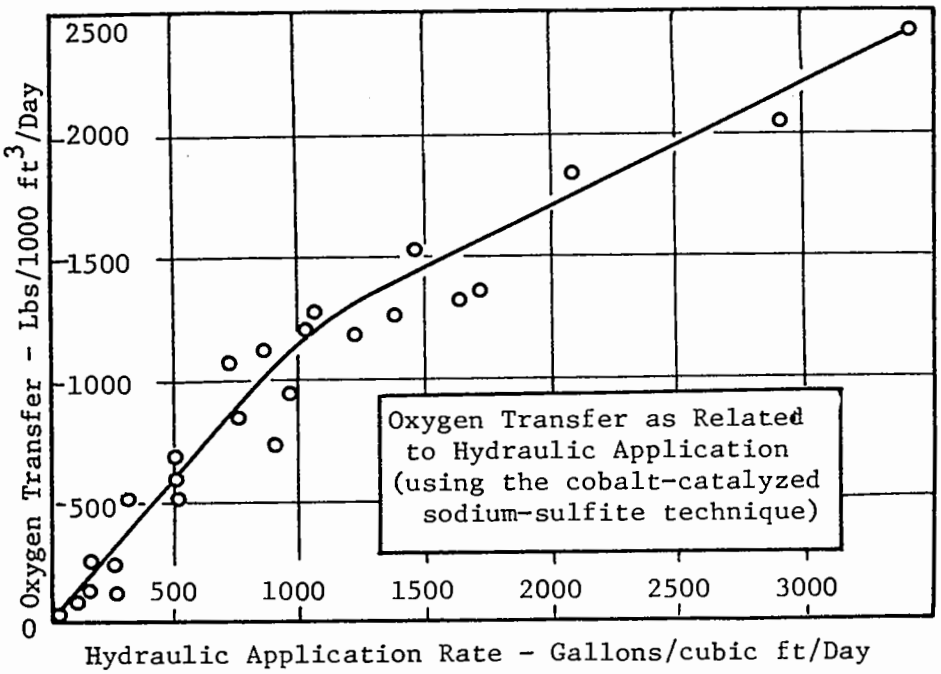


Figure 7. Relationship between oxygen transfer rate and hydraulic application rate for Dowpac HCS using cobalt-catalyzed sodium sulfite (Kountz data).

On January 13, 1956 - only nine days after the initial meeting at which the Battle Creek Study was formulated, the pilot plant was placed into operation. Activities during the intervening nine days between the initial meeting and the start of operation included construction of the pilot plant (Figure 8), fabrication of a settling tank, construction of a large BOD-incubator, augmentation of the City of Battle Creek Treatment Plant's laboratory for conduct of Chemical Oxygen Demand, and correlation of the 5-day Biochemical Oxygen Demand and the Chemical Oxygen Demand for the City's primary effluent. This pilot unit was operated continuously through April 28, 1956 while it was intensively studied. It was on "stand-by" operation until June 11, 1956 when it was operated at a low dosing rate to provide data for extending the range of operation to include the the highest hydraulic dosing rate then in general use for design of conventionally packed trickling filters. During the entire period of operation, the Steering Committee provided guidance to the study.

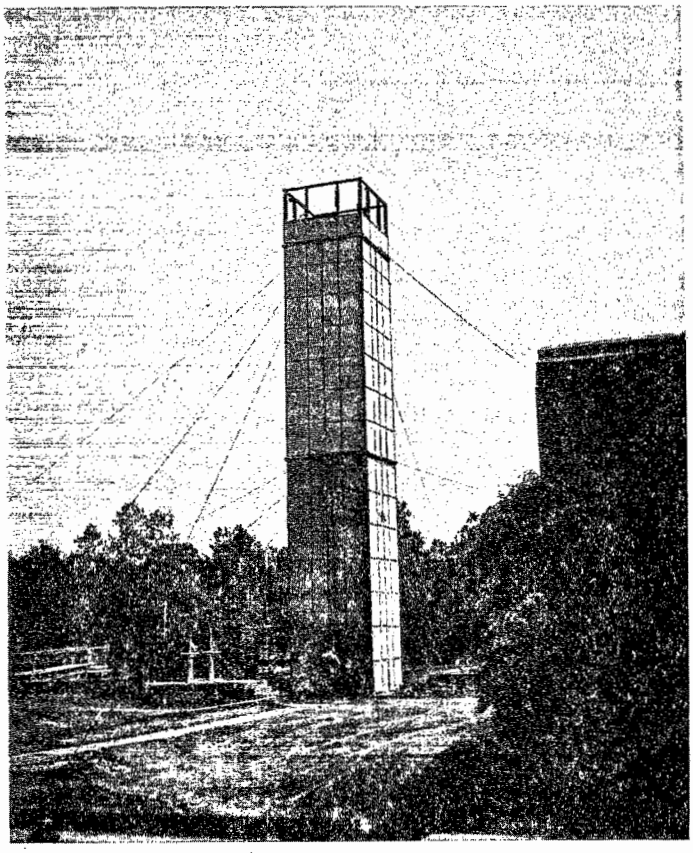


Figure 8. Dowpac HCS pilot plant constructed at the City of Battle Creek. The unit contained a packed-depth of 42 feet with provision for intermediate sampling.

Details regarding the construction and operation of the Battle Creek pilot plant, guidance provided by the Steering Committee, results of operation and their analysis were contained in a Report of the Steering Committee authored by Becher and Bryan (23) with a statistical analysis by Busch. Table II contains a summary of results. An Appendix to the Report (23) contains all observed data obtained during the reported study. Stack (24) commended the Steering Committee for: "accomplishing an excellent study...the most thorough study of trickling filtration treatment of sewage that I have seen."

Table II. Summary of Results From the Dowpac HCS Pilot Plant at Battle Creek, Michigan (1956)

Time Period (1956)	Hydraulic Rate (gpm/ft^2)	5-Day Biochemical Oxygen Demand				Chemical Oxygen Demand				Temperature-(Degrees F)	
		Infl C (mg/l)	lb/1000-ft^3/day	Efficiency-% Tank	Efficiency-% Lab*	Infl C (mg/l)	lb/1000-ft^3/day	Efficiency-% Tank	Efficiency-% Lab*	Air	Water**
2/3-17	3.39	249	241	34.0	48.1	453	440	29.2	44.9	26.2	57.9
2/17-3/14	1.63	222	83.2	56.0	71.9	389	156	52.4	60.3	29.0	53.0
3/20-4/7	1.6 + 1.6(R)***	241	101	62.2	66.7	420	176	49.3	51.1	31.8	51.0
4/10-28	0.820	227	53.0	76.0	80.3	397	92.9	63.5	69.2	45.6	54.2
4/28-6/11	Standby operation of unit while data were being analyzed - No data were obtained										
6/11-27	0.318	218	19.9	-	86.7	393	35.8	-	72.2	75.2	66.5**

Notes: *Tower effluent was given 60-minutes of quiescent sedimentation in a laboratory graduated cylinder.

**Average of influent and effluent. Passage through unit reduced temperature by 3.4°F.

***1.6 gpm direct (primary effluent) plus 1.6 gpm recirculation (unsettled filter effl).

**Influent temperature only, effluent temperature was not obtained in this period.

Tower depth was 42 feet. In calculation of loadings, it was assumed sewage applied by a 3-foot diameter rotary distributor contacted all packing in the 37-1/2"square section. No adjustment was made for packing-equivalent of sidewalls which provided a maximum of an additional 5% surface area in the packed tower.

As likewise determined from operation of the pilot plant in Midland, Michigan, the efficiency of operation at Battle Creek was linearly and inversely proportional to the hydraulic application rate (Figure 9). However, with respect to removal of oxygen demand, within the limits of hydraulic application rates studied, removal of both biochemical and chemical oxygen demand was linear (in two "regimes") proportional to the hydraulic application rate. The particular advantage of Dowpac HCS as a "roughing" unit was obvious from this study as it was from all other prior studies.

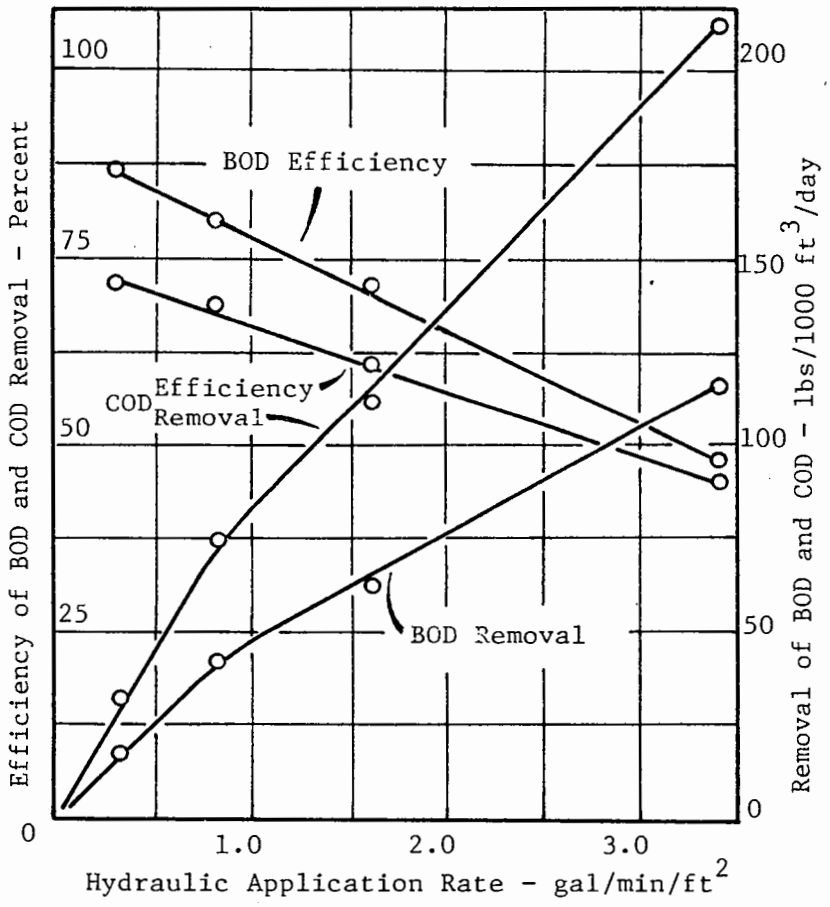


Figure 9. Efficiency and total removal of biochemical and chemical oxygen demand as affected by the hydraulic application rate of the Dowpac HCS pilot unit at Battle Creek, Michigan.

It was concluded that the Battle Creek pilot plant data were best represented by the following empirical equation:

$$R = \frac{1}{1 + 0.0148\ D^{1.748}\ 10^{-0.127\ Q^{2/3}}}$$

where: R is the "fraction" of 5-day BOD remaining at depth D
D is the depth of the Dowpac unit in feet
Q is the hydraulic application rate in gallons/minute
(Note - the unit was 9.77 square feet in area)

The exponent of "Q" in the above equation was noted to be in accord with Howland's theoretical development (25) and with the results of studies of laboratory trickling filters conducted by Bloodgood, Teletzke and Pohland (26).

SUMMARY

Although the general principle of trickling filtration had been previously well established and prior attempts had been made with little success to introduce synthetic media, the process by which synthetic media fabricated from plastic resins were developed was without precedent. In 1960, Zwick and Benstock (27), in a draft of their "Study Group Report on Water Pollution," attributed the origin of plastic media to an undocumented source - a person who had suggested replacing conventional media with wooden planks mounted in a box. Correspondence in which they were provided with a copy of the Battle Creek Report (28) resulted in some modification of the draft to provide a more balanced and accurate description of the origin of the concept and the role of personnel from The Dow Chemical Company, the U.S. Public Health Service, and others in its development.

The period during which The Dow Chemical Company's effort took place was one in which plastics were emerging to take the place of other materials in applications that went beyond the production of toys and novelties. Its own internal needs were the initiating cause for action taken by Dow in development of plastic media. The initial step is most accurately attributed to R. S. Chamberlin, D. E. Lake and F. E. Dulmage of The Dow Chemical Company who conceived the basic design of Dowpac FN-90 and related shapes. Dowpac HCS was a product of the joint efforts of D. E. Lake and Thomas J. Powers, Sr. The distinction of recognizing their potential for treatment of wastewaters belongs to Powers who provided the initial context within which

the development effort was initiated and nurtured. His seemingly unlimited capacity for seeking simpler and more direct ways of solving problems was coupled with a gift of almost infinite patience up to a point where action was both necessary and wise...attributes which, in the complex process of product innovation and development, are essential if not indispensible to balance potential risk with reward.

ACKNOWLEDGMENTS

Initial stages in the development of plastic media that are described in this paper were under the joint responsibility of James A. Struthers, Plastics Technical Service and Edward H. Bryan, Waste Disposal, The Dow Chemical Company, Midland, Mich. Other Dow personnel not specifically noted in the text or in the References section who made significant contributions during the early stages of development described in this paper included: E. E. Chamberlin, G. F. Dressell, John Hoy, William C. Goggin, A. A. Gunkler, Frank H. Justin, Earl Kropscott, Paul H. Lipke, Frank J. MacRae, Del H. Moeller, Stanley Mogelnicki, W. L. Nelson and Gordon B. Thayer.

Dr. Edward H. Bryan is currently Program Director, Water Resources and Environmental Engineering in the Division of Civil and Environmental Engineering, National Science Foundation, Washington, D. C. He was responsible for the technical and process related research and development activity, subject of this paper from 1954-58 while employed by The Dow Chemical Company. Responsibility for development of alternative plastics and methods of fabrication was initially that of James A. Struthers who was ably succeeded in this portion of the development program by Del H. Moeller, who assumed full responsibility for the program in 1958.

The content of this paper and any opinions expressed are solely those of the author and do not reflect a position by either The Dow Chemical Company or The National Science Foundation. The author expresses his appreciation to The Dow Chemical Company for permission to use photographs contained in this paper.

REFERENCES

1. Griess, G. A., "Plastics in Plants Manufacturing Heavy Chemicals", Industrial and Engineering Chemistry, Vol 47, pp 1343-1349 (July 1955).

2. Bryan, Edward H., "Molded Polystyrene Media for Trickling Filters," Proceedings of the Tenth Purdue Industrial Waste Conference, pp 164-172 (May 1955).
3. Bryan, Edward H., "Molded Polystyrene Media for Trickling Filters," Industrial Wastes, pp 80-84 (November-December 1955).
4. Dowpac FN-90 and Dowpac HCS, Bulletin of The Dow Chemical Company, Plastics Technical Service, 16 pp, (October 1955).
5. Bryan, Edward H., "Dowpac Tower Packing for Gas-Liquid Contact Systems," Proceedings of the 38th Texas Water and Sewage Works Short School, pp 121-122 (March 1956).
6. Bryan, Edward H., "The Role of Oxygen in Sewage Treatment," Proceedings of the 39th Texas Water and Sewage Works Short School, pp 88-89 (March 1957).
7. Kountz, R. Rupert, "Total Oxidation Treatment," Proceedings of the Eleventh Purdue Industrial Waste Conference, pp 157-159 (May 1956).
8. Trepanier, Norman W., "Biological Treatment of By-product Coke Plant Phenolic Wastes," Blast Furnace, Coke Oven, and Raw Materials Conference, American Institute of Mining, Metallurgical and Petroleum Engineers, Discussion by: Edward H. Bryan, Charles Drake and Hayes H. Black, pp 204-210 (1957).
9. Anderegg, Fred C. "Biological Disposal of Refinery Wastes," Proceedings of the 14th Purdue Industrial Waste Conference, (May 1959).
10. Bryan, Edward H., "Two-Stage Biological Treatment - Industrial Experience," Proceedings of the Eleventh Southern Municipal and Industrial Waste Conference, pp 136-153 (April 1962).
11. Egan, John T. and McDewain Sandlin, "Evaluation of Plastic Trickling Filter Media," Industrial Wastes, Vol 5, No 4, pp 71-77 (August 1960).
12. Cawley, William A., "Polyvinyl Chloride for Trickling Filter Media," Industrial Water & Wastes, pp 111- (July-August 1962).
13. Handt, Paul R., "Progress Report on Evaluation of Dowpac HCS as used in Trickling Filters," Student Trainee Report, The Dow Chemical Company (1956).
14. Brelsford, Donald L., "Harvesting Protein-Rich Bacterial Material From a Dowpac HCS Type Biological Trickling Filter," Summer Technical Employee Report, The Dow Chemical Company (1956).
15. Froman, Charles O., "Biological Oxidation of Acrylonitrile on Dowpac HCS Pilot Plant," Internal Report, The Dow Chemical Company (March 1957).

16. Sadow, Ronald D., "Acrylonitrile and Zinc Wastes Treatment," Parts 1 and 2, Industrial Water & Wastes, pp 42-45 and 66-70, (March-April 1961) and (May-June 1961), respectively.
17. Heckeroth, Earl T., "Progress Report on the Evaluation of Dowpac HCS as a Trickling Filter Medium," Student Trainee Report, The Dow Chemical Company (September 1955).
18. Greene, Robert E., "Progress Report on the Evaluation of Dowpac HCS Trickling Filter at the Midland Municipal Sewage Treatment Plant," Chemical Engineer Trainee Report, The Dow Chemical Company (September 1955).
19. Bryan, Edward H. and D. H. Moeller, "Aerobic Biological Oxidation Using Dowpac," in Advances in Biological Treatment, Proceedings of the Third Conference on Biological Waste Treatment, Manhattan College, edited by: W. W. Eckenfelder and Joseph McCabe, pp 341-346, Pergamon Press (1963).
20. Bauer, K. C., "Progress Report on Dowpac HCS as a Trickling Filter Medium - Midland Sewage Plant," Summer Technical Employee Report, The Dow Chemical Company (1956).
21. Ellis, William J., "Progress Report on the Evaluation of Dowpac HCS as a Trickling Filter Medium," Student Trainee Report, The Dow Chemical Company (1956).
22. Kountz, R. Rupert, "Oxygen Solution Capacity of Wetted Dowpac HCS Towers," Report to The Dow Chemical Company (1956).
23. Becher, A. E. and Edward H. Bryan, "A Study of the Performance of Dowpac HCS When Applied to the Treatment of Settled Sewage from the City of Battle Creek, Michigan," Report of the Project Steering Committee, Statistical Analysis by K. A. Busch, 149 pp, The Dow Chemical Company (June 1958).
24. Stack, Vernon T., Personal Communication (February 1959).
25. Howland, W. E., "Flow Over Porous Media as in a Trickling Filter," Proceedings of the 12th Purdue Industrial Wastes Conference, pp 435-465 (May 1958).
26. Bloodgood, Don E., G. H. Teletzke and F. G. Pohland, Fundamental Hydraulic Principles of Trickling Filters," Sewage and Industrial Wastes, Vol 31, pp 243-253 (1959).
27. Water Wasteland, Nader Task Force Report on Water Pollution, Edited by David R. Zwick and Marcy Benstock, Vol 2, pp XIX-28 and 29, Preliminary Draft, Center for Study of Responsive Law, Washington, D. C. (1971).
28. Bryan, Edward H., Personal Letter to David Zwick (July 1971).
29. Water Wasteland, Edited by David R. Zwick and Marcy Benstock, pp 381-382, Grossman Publishers (1971).

CURRENT STATUS AND FUTURE TRENDS OF ROTATING BIOLOGICAL CONTACTOR IN JAPAN

Masayoshi Ishiguro. Professor of Civil Engineering, Miyazaki University, Kirishima 1-1-1, Miyazaki, Japan.

INTRODUCTION

This paper presents the current status and future trends for the use of Rotating Biological Contactors in Japan. It includes a historical survey. Since 1966, the number of waste-water treatment plants using RBCs has risen to over 1,323. The total flow was 443,000 m^3/day in June 1981. Over 300 additional plants are now under construction. Most of these are utilized for secondary waste-water treatment, but 42 plants have been installed for nitrification and 17 other plants for BOD and nitrogen removal. The first denitrification RBC plant has been in operation since 1976. In 1981 the first objective RBC nitrification plant was built for treating surface water prior to water purification. Another special nitrification plant is under construction for nitrification of rice field irrigation water and has a design flow of 84,000 m^3/day. At the present time there are 22 RBC manufacturers in Japan. Seven of them have technical tie-ups with foreign enterprises, the other 15 manufacturers have developed their own technology. Investigation of the RBC is very active. About 100 papers were presented at Annual conference and published in the Journal of the Japan Society of Civil Engineers, and the Journal

of the Japan Sewage Works Association, etc. last year.

1. HISTORICAL REVIEW OF THE RBC PROCESS IN JAPAN

K. Kohyama, Department of Sanitary Engineering, Hokkaido University conducted the first experimental research of the RBC in 1960, for the treatment of Potato starch wastewater (1). In 1964, M.Ishiguro,Department of Civil Engineering, Miyazaki University, began studies on the RBC for treatment of Sweet Potato starch wastewater. As a result of this research, the first full scale RBC process in Japan for the treatment of Sweet Potato starch wastewater was installed at the end of 1966 in Miyazaki prefecture. This plant was constructed with five stages, a disk diameter of 2.0 m, and a surface area of 1,500 m^2, Polystyrene was used for the discs. The concentration of influent BOD_5 is 10,000 mg/l, and the flow rate is 600 m^3/day. The design for BOD loading is 900 $gBOD/m^2day$ which achieves a 70% BOD reduction (2). Not many more RBC process plants were installed until 1971, but investigation of the RBC continued steadily at both the above mentioned Universities and at other places (3, 4, 5, 6).

Table - 1 summerizes the number of operating RBC plants from 1972 up to June 30, 1981.

Table 1. Number of RBC Plants in Japan

Year	1972	1973	1974	1975	1976	1977
No. of Plants	4	25	60	96	252	469
Year	1978	1979	1980	1981 (June. 30)		
No. of Plants	701	948	1206	1323		

In 1972, there were only 4 wastewater treatment facilities utilizing the RBC. Since 1973, the number has increased from year to year to more than 1,323, with another 300 now under construction (7).

2. CURRENT STATUS

2-1 Exsisting RBC Plants

Table 2. summerizes the number of locations and the quantity of flow for the six major wastewater categories identified by source; pre-purification surface water (tap water sources), domestic, food processing, industrial (e.g. the pulp and chemical industries), waste treatment and disposal (e.g. landfills and wastematerial treatment plants), and animal breeding (7).

Table 2. Summary of RBC Plants in Japan (June 30, 1981)

Wastewater	Flow (m^3/d)	Flow(%)	Site	Site(%)
Tapwater sources	14,200	3	1	0.1
Domestic	224,321	53	654	50
Food processing	35,844	8	243	18
Industrial	121,454	29	261	20
Waste treatment and disposal	23,675	6	135	10
Animal breeding	3,381	1	29	2
Total	422,875	100	1323	100

There were over 1,323 RBC plants with a total flow of 442,875 m^3/day by June 1981. The tap water sources in Table 2 reflects the fact that in Japan the largest volume of water for municipal use is taken from surface water. The water sources have become polluted with organic wastes and nitrogen. Therefore, an RBC nitrification process has been installed for surface water prior to the water's treatment in the water purification plant. Further details of the plant are given in section 3.

There are approximately 654 RBC plants currently treating municipal wastwater. The largest operating RBC facility in Tokushima City has 32 shafts, and a flow of 31,600 m^3/day (Design flow: 63,200 m^3/day)(8). Two hundred forty three instalations treat food processing wastewater, two hundred sixty one installations treat industrial wastewater. The largest operating RBC facility has 40 shafts, a flow of 12,000

m^3/day for the treatment of water from the manufacture of pulp. One hundred thirty five installations treat landfills (garbage dump) for BOD removal, nitrification and denitrification. The first such RBC plant has been in operation in Miyazaki City since 1976 (9,10). There are twenty-nine plants treating wastewater from animal breeding (7,11).

Table 3. lists the distribution of Table 2. summary treatment facilities by flow range. Approximately 53% of the existing facilities are package plants treating a discharge flow below 100 m^3/day (0.03 MGD.).

Table 3. Total Number of Operating RBC Installations (June 30, 1981)

Flow range (m^3/day)	Total No.	Sub.total	%
0 - 99	702		53
100 - 299	405	1,170	84
300 - 499	101	1,208	91
500 - 999	57	1,265	96
1,000 - 2,999	37	1,302	98
3,000 - 4,999	8	1,310	99
5,000 - 9,999	5	1,315	99.4
10,000 - 19,999	6	1,321	99.8
20,000 - 29,999	1	1,322	99.9
30,000 - 39,999	1	1,323	100.0

Approval and financing by the Japanese Ministry of Construction for the RBC process for municipal wastewater is about ten years behind Europe and the U.S.A. The first RBC plant for public sewerage treatment was constructed in 1978. For that reason, in the early years after the RBC process was introduced into Japan, almost none were installed for municipal sewerage; therefore Japanese RBC engineers concentrated their efforts on the most difficult wastewater treatment for industrial etc. They have achieved to success with that wastewater treatment.

Table 4. summarizes the design criteria for surface loading of BOD in order to achieve the concentration effluents of BOD below 20 mg/l except for domestic wastewater (11).

Table 4. Design surface loading rates for all types of wastewater excluding domestic

No.	Wastewater	Influent BOD_5 (mg/l)	BOD loading (g/m^2d)
1	Marine product process	400 - 1000	30 - 90
2	Fish meat process	150 - 450	25 - 60
3	Fish market place	100 - 600	15 - 20
4	Meat process	100 - 1500	10 - 20
5	Eatable bird process	300 - 1500	15 - 20
6	Bean paste (Miso) Soy manf.process	150 - 600	5 - 25
7	Eatable food oil manf. process	400 - 600	20 - 25
8	Pickles manf.	500 - 1500	30 - 50
9	Sake brewing (brewery)	700 - 2000	15 - 20
10	Dairy	300 - 400	30 - 60
11	Fruit Canning	1000 - 1600	20 - 60
12	Orange Canning	200 - 1400	30 - 40
13	A taro Canning	100 - 200	15 - 25
14	Center of feeding	200 - 500	10 - 30
15	Silk yarn manf.	1400 - 6000	10 - 20
16	Dyeing manf.	120 - 200	20 - 40
17	Paint material product	70 - 140	5 - 10
18	Woolmil manf.	150 - 200	20 - 25
19	Wood pulp manf.	1000 - 2300	10 - 80
20	Refinery bleaching	800 - 1000	50 - 65
21	Old paper reproduct	300 - 800	10 - 20
22	Bleaching paper manf.	50 - 100	10 - 15
23	Petrochemistry manf.	100 - 800	5 - 80
24	Cleaning (wet)	80 - 140	8 - 10
25	Cleaning (dry)	300 - 500	10 - 20
26	Medicine manf.	600 - 1000	5 - 25
27	Hospital wastewater	120 - 450	10 - 15
28	Slaughter-house	750 - 2500	80 -100
29	Hog yard	200 - 1300	5 - 50
30	Diluted night soil	1500 - 2000	5 - 30
31	Waste material treatment plant	300 - 1000	10 - 20
32	Garbage dump	10 - 200	2 - 20

2-2 RBC Manufacture

At the present time there are 22 RBC manufactures in Japan. Seven of them have technical tie-ups with foreign enterprises: Autotrol (U.S.A.) with Nippon Autotrol (1972), Schuler-Stengelin (West Germany) with Pacific Engineering and also with Mitsuitoatsu (1973), Ames Croster (U.K.) with

Niigata Tekko (1973), Mecana (Switzerland) with Takuma (1974), Clow Envirodisc (U.S.A.) with Sinko-Pfaudra (1978) and Bio-Shaft (U.S.A.) with Maezawa Kogyo (1980). The other 15 manufactures have developed their own technology: Kurita-Kogyo, Shinmeiwa Kogyo, Dengyosha, Tore-Engineering, Sekisui Kagaku Kogyo, Asahi Engineering, Unichica, Matsushita Seiko, Showa Koji, Sanki-Kogyo, Meiden-sha, Kyushu Denko, Organo, Sekine Sangyo and Tsutsunaka Plastic (11).

2-3 Existing Facilities

Table 5. shows the nominal parameters associated with the media and mechanical components for the 22 RBC manufactures in Japan. Each equipment manufacturer offers variations of the media and drive components. The media material, support, shaft strength, tank shape, and clearance are some of the items which have affected RBC performance. The maximum values of disc diameter, surface area, and shaft length are 5.0 m, 19,170 m^2, and 8.8 m, respectively. There are many shapes for the disc surface, e.g. flat, combined flat and corrugated, waved, double-waved, two flat plates combined, flat-netted, etc.

Table 5. RBC Equipment Dimentions

Media	: Disc	
	Shape	: Circular, Octagonal
	Material	: High density Polystyrene, Polyethylene, Hard Polyvinyl Chloride, FRP (Fiber glass Reinforced Plastic)
	Diameter	: Standard: 3.6 m, Range: 1.0 - 5.0 m
	Thickness	: 0.7 - 7.0 mm
	Surface area	: 300 - 19,170 m^2/shaft
	Spacing	: 1.0 - 3.2 cm
	Construction	: Segmented (12, 8 or 5 pieces) : Steel supported. Unitized : Heat welded self-supported.
Mechanical	: Shaft	
	Shape	: Cross-section : Circular, Round Square Octagonal
	Material	: Steel
	Thickness	: 1.90 - 3.80 cm
	Length	: Standard : 7.5 m, Range : 1.0 - 8.8 m

: Motors
Horsepower : 0.5 - 15.0
: Drive Units
: Multi-V Belts, Chain and sprocket, Enclosed cartridge, Air Driven, Water Driven.

Recentry, a new drive unit process has been developed by Kurita-Kogyo : Water is introduced to aid in rotating the discs. Plastic water cups are welded onto the periphery of the media over the entire length of the contactor. The waste water is dropped from a height of about 1.0 m above the top periphery of the media and is captured by the plastic cups. The falling wastewater causes the 3.6 m-diameter RBC disc to rotate. The process could be combined with Activated Sludge process, e.g. Northeast sewage treatment plant in Philadelphia, U.S.A.(12, 13, 46).

Another more highly technique, developed by Meiden-sha, is the automated control of the rotational speed of the discs depending on the quality of the influent water. It is well recognized that when the BOD concentration in the influent increases, the additional BOD removal can be achieved by increase the rotational speed of the disc. Self-variation of rotational speed by the newly developed equipment could match the variation of influent flow rate, temperature, and concentration of organics. In the operation of the RBC process for variable loading such as industrial wastewater, this new technique and equipments will improve the maintenance and treatment efficiency (14, 15).

2-4 RBC System Study Mission to Foreign Countries

RBC system study mission have been organised six times since 1975 and have visited foreign RBC manufacturers and RBC plants under construction or operating : (1) August 1975 (U.S.A.), (2) November 1975 (U.S.A.), (3) September 1977 (50-th annual conference of the Water Pollution Control Federation in Philadelphia), (4) June 1978 (Europe-Denmark, Sweden, West Germany, Austria, France, Switzerland, and the U.K. including the 9th International Conference of the International Association on Water Pollution Research in Stockholm, Sweden), (5) June 1980 (Canada and the U.S.A., the 10th International Conference of the IAWPR, Toronto, Canada), (6) April 1982 (U.S.A. attended the 1st International Conference on Fixed - Film Biological Processes, Kings Island, Ohio, U.S.A.) (11, 16, 17).

3. SPECIAL APPLICATION OF THE RBC

As already mentioned, most of the RBC plants are utilized for secondary wastewater treatment. About 5% of the total number are utilized for nitrification and denitrification. Tth following discussion will be concerned with two specisl applications of RBC plants for the removal of low concentrations of ammonia-nitrogen.

3-1 Nitrification Prior to Water Purification

In Japan, the largest volume of water for municipal use has been taken from surface water. The surface water has became polluted with organics and nitrogen, so that the cost for prechlorination (addition of chlorine at the mixing basin) and for other chemicals have greatly increased at water purification plants. Therefore, it has caused a rise in the cost of water supply and plant maintenance. Trihalomethane (THM: a cancerous growth matter) is produced in the reaction between organics and chlorine, is becoming a world-wide problem. In addition, the rejection of high amounts of ammonia-nitrogen in raw water requires a large amount of chlorine, which might cause the production of THM.

Several studies have been made to find a process which could be installed prior to the water purification process in order to solve the problem. The unit processes evaluated were activated sludge, trickling filters, submerged biological filters, stripping, and the RBC process. The RBC process was selected because of its simplicity of construction, operation, maintenance, and low energy requirements.

Field tests using a RBC pilot plant were carried out from April 1976 to October 1980 in order to examine the effect of the reduction of organics and the oxidation of ammonia-nitrogen in low concentrations in river surface water. Based on the results of the field test, the first RBC nitrification plant for use prior to water purification plant was constructed in april 1981 with the approval by the Ministry of Health and Welfare (18). The plant is installed in Nakama-City, Fukuoka-Prefecture, in the island of Kyushu and treats most of the downstream water of the Onga River, running into the Genkai Sea.

Figure 1. is a diagramatic sketch of a rapid sand filter. It shows the path of the water through the various units. Conventional types of water purification plants are shown by the broken line and RBC nitrification unit by solid lines.

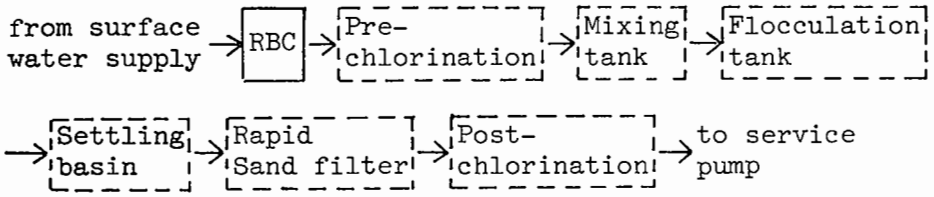


Fig 1. Schematic flow diagram of an upgraded Water purification plant with RBC nitrification

Table 6. shows the characteristics of the Onga River water and the percent reductions in the listed items by the RBC pilot plant. It indicates that the concentration of NH_3-N is higher in winter than in summer due to the small discharge rate of the river in winter.

Table 6. Characteristics of River water and RBC test

	Concentration			RBC test % reduction
	Max.	Min.	Mean	
Water temperature (℃)	30.4	4.0	15.9	
DO (mg/l)	10.7	4.5	7.7	
pH	8.3	7.3	7.8	
COD (mg/l)	13.6	9.8	11.6	32
BOD (mg/l)	5.3	1.2	3.3	70
SS (mg/l)	47.2	7.6	15.2	32
NH_3-N (mg/l)	3.0	0.02	0.67	90
Degree of turbidity (mg/l)	11.0	4.5	8.2	58
Color (mg/l)	46.0	24	36	30
Threshold odor number (TO)	50.0	8.0	25.0	60
Chlorine requirement (mg/l)	15.5	8.8	11.5	59
Total iron (mg/l)	0.34	0.18	0.29	70
Manganese (mg/l)	0.22	0.09	0.13	80

Design criteria for the flow rate, loading rate and equipment are summarized in Table 7. for the total installation.

Table 7. Nitrification Criteria

A. Design Flow Rate (Water Consumption)

Maximum - Summer	:	14,200 m^3/D (high water temperature)
Yearly Average	:	12,000 m^3/D (mid-term water temp.)
Maximum - Winter	:	11,000 m^3/D (low water temp.)

B. Loading Rates

Hydraulics	:	200 l/m^2D (5 gpd/ft^2D)	(winter)
Hydraulics	:	259 l/m^2D	(summer)
Hydraulics	:	219 l/m^2D	(mid-term)
NH_3-N	:	0.256 g/m^2D	(winter)

C. RBC Equipment Dimensions

Total Surface area	:	55,000 m^2 (592,020 ft^2)
One shaft surface area	:	9,150 m^2/shaft
No. of shafts	:	6
No. of trains	:	3
Length of shafts	:	7.4 m
Diameter of discs	:	3.6 m
Material of discs	:	high-density Polyethylene
Type of disc surface	:	composed of flat and corruguted sheet
Peripheral rotational speed	:	18 m/min. (1.6 rpm)
Electric power consumption	:	5.5 KW/shaft
Detention time	:	43.4 min. (winter)
Detention time	:	40.0 min. (average)
Detention time	:	33.6 min. (summer)
RBC manufacturer	:	Nippon Autotrol

In winter, the average values of the NH_3-N concentration in the river water and the required dosage of chlorine are 1.28 mg/l and 13.5 mg/l, respectively. However, the new purification plant with RBC nitrification unit has achieved the effluent NH_3-N concentration of 0.18 mg/l (86% reduction) and the chlorine dosage of 3.3 mg/l (76% reduction). Yearly averages of the feeding ratios for prechlorination have decreased from 11.3 mg/l to 4.7 mg/l (a 58% reduction). Moreover, the yearly average feeding ratios of activated carbon for the elimination of odor, etc. of 11.2 mg/l has decreased to 4.7 mg/l (a 57% decrease). The reduction in expense for chemicals is about ¥ 10 million (U.S. $ 50,000) a year.

The plant was started-up of April 1, 1981 at a flow rate of 5,220 m3/day (maximum) and 4,219 m3/day (average), which corresponded to 37 and 35% of the design flows, respectively. The performance of the RBC treatment has almost coincided with the design criteria from start-up to the present day.

A similar RBC nitrification plant is also under consideration in Nakama-City and will have a design flow of 19,700 m3/day.

3-2 Nitrification of River Water for Rice Field Irrigation

Rice is the staple food for the Japanese. The 3,081,000 hectares (7,700,000 acres) of rice field comprise 56% of the total farm land in Japan. The largest volume of water for rice field irrigation has been taken from the surface water of natural rivers and from irrigation reservoirs. Poor rice yields have been traced to high NH_3-N which causes excessive stalk growth compared to desired kernel growth.

The RBC process was selected to solve this problem. The field test of an RBC pilot plant with disc diameters 2.0 m and a flat and waved media surface was carried out from September 1977 to October 1979. The tested discs peripheral rotational speeds were 10.0, 13.5, 18.0, 24.0, 27.0, 30.0, and 36.0 m/min. The hydraulic loadings were 200, 300, 400, 450, 600, and 800 l/m2day.

Table 8. summarizes the water quality of the river water and the performance of the RBC pilot plant. Hydraulic loading is 600 l/m2day and peripheral rotation speed is 27 m/min. which are the optimum conditions for the removal of ammonia-nitrogen (19, 20).

Table 8. Characteristics of river water and RBC test

Items	Concentration of influent	Concentration of RBC effluent	Percent of reduction
Water temp. (°C)	19.2 - 22.5	18.8 - 20.0	
DO (mg/l)	5.2 - 6.5	7.7 - 8.7	
COD (mg/l)	7.8 - 14.0	4.6 - 5.9	24 - 43
BOD (mg/l)	3.7 - 18.2	1.5 - 5.5	46 - 70
SS (mg/l)	17.7 - 48.3	7.0 - 15.9	48 - 67
NH_3-N (mg/l)	0.9 - 1.4	0.0 - 0.3	79 -100
NO_3-N (mg/l)	0.6 - 0.7	1.2 - 1.9	200 -283
Org-N (mg/l)	0.5 - 1.0	0.3 - 0.8	20 - 50
Kej-N (mg/l)	1.4 - 2.2	0.4 - 0.8	64 - 80

The first RBC nitrification plant for rice field irrigation water with a design flow of 70,000 m^3/day (18.5 MGD) is under construction with the approval of the Ministry of the Agriculture and Forestry. The RBC nitrification plant is being installed in Ibaragi-City, Osaka Prefecture in Central Japan on the Yodo River, which into the Seto Inland Sea (Setonaikai).

Irrigation water for rice fields is required from June to September, therefore, the RBC plant is operated only four months a year. Water quality standards for rice field irrigation water is as follows: pH (6.0 - 7.5), COD (6 mg/l), SS (100 mg/l), DO (5.0 mg/l), and TN:Kej-N (1.0mg/l).

The characteristics of the RBC influents are summarized in Table 9.

Table 9. Characteristics of RBC influents

Items	NH_3-N, NO_3-N, Org-N, Kej-N,	T-N	DO	COD	BOD
Concentration (mg/l)	1.2 1.3 1.0 2.2	3.5	3.7	8.5	10.6

Final effluent values from the RBC plant of Kej-N, COD and DO were defined as 1.0, 6.0, and 5.0 mg/l respectively. Design flow rate, loading rate, and equipment are summarized in Table 10.

Table 10. Nitrification criteria for rice field irrigation water

A. Design flow rate

Average flow	: 70,000 m^3/D (18.5 MGD)

B. Loading rate

Hydraulics	: 600 l/m^2D
NH_3-N	: 0.507 g/m^2D

C. Design RBC equipment dimensions

Total surface area	: 164,160 m^2
One shaft surface area	: 13,680 m^2/shaft
Number of shafts	: 12
Number of trains	: 6

Length of shafts	: 8.85 m
Diameter of discs	: 5.0 m
Spacing of discs	: 17.5 mm
Thickness of discs	: 1.3 mm
Material of discs	: high-density Polyethlene
Type of disc surface	: flat
Peripheral rotational speed	: 27 m/min
Electric power consumption	: 9.2 KW/shaft
Detention time	: 20 min.
RBC manufacturer	: Dengyo - sha Machine Works

This RBC nitrification plant has six trains mechanically driven. There are two shafts per train. All twelve shafts have been installed in one building with 91.2 m in length, 15.6 m in width, and 5.9 m in height.

An additional RBC nitrification plant for field irrigation water with a design flow of 14,000 m^3/day (3.7 MGD) will be constracted within a few years in the same area.

4. CURRENT STATUS OF RESEARCH

Research of the RBC is very active. Last year 100 papers were presented at Annual Conference and published in the Journal of the Japan Society of Civil Engineers (JSCE), the Journal of the Japan Sewage Works Association (JSWA), the National Symposium on RBC Technology of the Environmental Conservation Engineering Association (ECEA), and other journal of wastewater treatment, etc. (7). The first special edition on the RBC process appeared in Journal of the Environmental Conservation Engineering (ECE), Vol.4, No. 7, July, 1975 (4, 5, 6). Another Journal of Engineering has edited a special issue on RBC every year.

The first Seminar on the RBC was held in September 1975 and sponsored by the ECEA. Since then, the Seminar was held annualy until 1979. In November, 1977, the RBC Wastewater Treatment Div. of the Association was established. As a result of the first National Symposium on RBC Technology with the ECEA, November 13 - 15, 1979 (21), an RBC Symposium has been held every year. At the end of the Symposium, field tours observe operating RBC plants. In October, 1979, the first Annual Conference of the Fixed-Film Biological process Research Association was held in Tokyo (22). The conference presents research in every year that is on the Rotating Bio-

logical Contactor, Trickling filter, Submerged Biofilter,etc.

To date, research on the RBC has been conducted at the following institutions : Hokkaido University, Kitami Institute of Technology, Tohoku University (23), Tokyo University (24), Tokyo Institute of Technology (25,26), Tokyo Metropolitan University, Miyazaki University (27 - 30), Kagoshima Technical College, the National Institute for Environmental Studies (31, 32), the Japan Sewerage Works Agency (12), the Consulting Engineers Co. (33, 34), and among RBC manufacturers.

The most foundamental research on the RBC was conducted by: M. Ishiguro, Y. Watanabe, S. Masuda, K. Yamaguchi, and H. Uchida, "Advanced Wastewater Treatment by RBC Unit (I-IV)" and published in the Journal of the Japan Sewage Works Association, Vol. 12-16, from 1975 to 1979. These papers were awarded the 1980 thesis prize of the JSWA (35). At the 9th, 10th, and 11th International Conferences of the IAWPR in 1978, '80, and '82, Y. Watanabe, M. Ishiguro, and K. Nishidome presented papers on Denitrification Kinetics, Nitrification Kinetics and Simulation of Nitrification in an RBC Unit (36, 37, 38). At the 1st National Symposium on RBC Technology in Feb. 1980, Pa. U.S.A., K. Ito and T. Matsuo presented a paper on "The Effect of Organic loading on Nitrification in RBC Wastewater Treatment Processes" (24), and H.Iemura and R.J.Hynek presented a paper on "Nitrogen and Phosphorus Removal with RBC" (10).

Many books have been published concerning RBCs: 1) The Newest Technique of wastewater treatment by Biochemical Processes (39), 2) Guide book for night soil treatment (40), 3) Wastewater treatment by the RBC (41), 4) Compilation book of Wastewater treatment technique by RBC (42), 5) Guide book for sewage, industrial wastewater, and sludge treatment (43), 6) The Fixed-Film Biological Process (11), 7) Guide book for Domestic wastewater treatment (44), etc. In particular the literature 6) includes the newest theory and the design procedures for Trickling filter, Rotating Biological Contactor, and Submerged Biofilter systems.

5. FUTURE TRENDS

To date, the pace of development of sewerage facilities in Japan has been slow. There are many reasons for this. Until recently, night soil was plowed back into farmland. Because water pollution problems were rare in the past, the importance of sewerage tended to be minimized. After World War II, farmers began to use chemical fertilizers instead of

night soil (septage). Night soil was disposed of in other ways, primarily being discharged into rivers and other bodies of water, eventually polluting them (45).

The 7.4% of the population was served by sewerage in 1963. The spread of sewer system in Japan lags far behind that of other developed countries. The systematic construction of sewerage facilities began with the First Five-year plan of Sewerage Construction (1963 - 67) in 1963. Although the population served by sewerage has increased along with increased Five-year plan, it was still only 30% at the end of fiscal year 1980 (the end of the Fourth Five-year plan). Coverage was 70% in 10 major cities having populations of more than 1 million, whereas that in other cities was under 20%, showing that wastewater treatment works in smaller cities lags far behind that in large cities. By about the year 2,000, 80% of the population should be served by sewerage. According to the Fifth Five-year plan, coverage should increase to about 44% by about 1985. Because the RBC process is characterized by low maintenance costs and low energy consumption, Sewerage facilities using the RBC process would be constructed especeally in smaller cities in Japan (46). Over twenty cities are constructing or planning sewage treatment plants with the RBC process.

In Japan, approximately 75% of the night soil collected by vaccum trucks. The collected night soil is treated at public wastewater treatment plants, home night soil purification tanks or collected night soil treatment plants. There are 1,186 night soil treatment plants, with a total planned processing capacity of 94,126 kl/day in 1980. These plants are viewed as a transitional measure until public sewerage could be established. Thus, Japan will continue to depend heavily on the collected night soil treatment plants for some time. RBC design criteria for night soil purification tanks (including domestic wastewater) have been authorized by the Ministry of Health and Welfare since July, 1980 (47). It seems likely that the number of RBC plants should increase in smaller cities, towns, rural communities (farm, fishing and mountain villages) etc.

Finally, the RBC process should increase rapidly in the following field in Japan :

a) public sewage treatment plants
b) community plants
c) home night soil purification tanks
d) treatment plants for
    i) food processing

ii) waste material treatment and disposal,
iii) animal breading, and other industrial wastewater.

e) special applications
i) pre-nitrification of surface water for water purification plant,
ii) nitrification plants treating river water for rice field irrigation.

In addition, research on RBC will continue actively in Universities, Government Institute, Consulting Engineering Companies, and RBC manufacturers in Japan.

REFERENCES

1. Kohyama K., Inoue I., and Takayasu M.,"Studies on the Biochemical Treatment for Wastewater in the Manufacture of Potato Starch", Journal of Water and Wastewater, Vol. 3, No.12, Dec, 1961, pp.1-10. (in Japanese)
2. Ishiguro M., Takahata S., and Nakamura I.,"Studies on the Sweet Potato Wastewater Treatment by the Rotating Biological Contactor", Proc. 21st Annual Conference of Japan Society of Civil Engineers, May, 1966, pp.146-149. (in Japanese)
3. Ishiguro M.,"Wastewater Treatment by the Rotating Biological Contactor", Journal of Japan Sewage Works Association, Vol.10, No.111, Aug.1972, pp.18-29. (in Japanese)
4. Ishiguro M.,"The Secondary and Tertiary Treatment of Municipal and Industrial Wastewater by the RBC", Environmental Conservation Engineering (ECE),Vol.4, No.7, July,1975, pp.1-21. (in Japanese)
5. Ishiguro M.,"Current Status of RBC in Abroad", ECE., Vol.4, No.7, July.1975, pp.42-52. (in Japanese)
6. Kohyama K. and Kato Y.,"Purification Mechanism of the RBC", ECE., Vol.4, No.7, July.1975, pp.31-41. (in Japanese)
7. Ishiguro M.,"Current Status and Future Trends of RBC", Proc., The Third National Symposium on RBC Technology. Oct.1981, pp.61-71. and ECE., Vol.10,No.12, Dec.1981, pp.37-46. (in Japanese)
8. Yamashiro Y.,"RBC Process in Tokushima City Central Sewage Treatment Plant", Proc. The Fifth Seminar of the RBC., May.1979, pp.57-70. (in Japanese)
9. Ishiguro M., Watanabe Y., and Masuda S.,"Treatment of Leachate from Sanitary Landfill" ECE., Vol.7,

No.6, June.1978, pp.3-11. (in Japanese)
10. Hynek R.J., and Iemura H.,"Nitrification and Phosphorus Removal with RBC", First National Symposium on RBC Technology. Proc., pp.295-324. Pa. U.S.A., Feb.1980.
11. Iwai S., Kusumoto M., and Ishiguro M. et.al., "Fixed-Film Biological Process", Chapter 4, RBC.process. Sangyo-yosui-Chosakai. Co., Nov.1980. (in Japanese)
12. Okuno N.,"Upgrading of the Existing Activated Sludge Process by the RBC", Proc., 18th Annual Conference of JSWA. May.1981 pp.198-200. (in Japanese)
13. Kurita Kogyo,"Pilot Scale Studies on the Combination of a new developed RBC process and the existing Activated Sludge process", SK NEWS. No.36, March.1981. (in Japanese)
14. Meiden-sha,"RBC Process with Auto Speed Regulator", Technical Data. No.GP-2028. Sept.1980. (in Japanese)
15. Meidensha,"Meiden Wastewater Recycling System for Building", Catalog.No.BB52-2038. Sept.1981. "Bio-rotacon: Rotating Disc Type Biological Contactor for Water Treatment", MP-9118. March.1980.
16. Mori T.,"The latest dates Status of RBC in U.S.A.", ECE. Vol.4, No.12, Dec.1975, pp.44-47. (in Japanese)
17. Ishiguro M.,"Current Status of RBC in U.S.A.", ECE., Vol.7, No.4, April.1978, pp.40-46, ."Current Status of RBC in Europe", ECE., Vol.7, No.11, Sept.1978, pp.9-23, "Current Status of RBC in North America", Proc., The 2nd National Simposium on RBC Technology, Oct.1980, pp.13-22 . (in Japanese)
18. Water Supply Department of Nakama-City,"Experimental Report on Pre-Treatment by the RBC for Polluted Water Sources to use of Water Supply", June.1978. The Task Committee Report on Biological Treatment Process for Polluted Water Sources to use Water Supply. Jan.1980. (in Japanese)
19. Shinan Tochi Kairyouku and Nihon Meintenace Engineering Co., "Experimental Study Report on the Purification Effect of Rice Field Irrigation Water by RBC Pilot Plant" Dec.1978. (in Japanese)
20. Sasaki A.,"Treatment of Agricultural Water containing low level of Contaminated by RBC Process", ECE., Vol.8, No.10, Oct.1979, pp.22-29. Proc., The First National Symposium on RBC Technology, ECE Association, Nov.1979 pp.55-58. (in Japanese)
21. Proc.,"1st. National Symposium on RBC Technology" Environmental Conservation Engineering Association. Nov.

1979. (in Japanese)
22. Proc. ,"1st Annual Conference on Fixed-Film Biological Processes", FFBP. Research Association. Oct.1979. (in Japanese)
23. Nakamura K., Matsumoto J., and Noike T. ,"With a Iron Bacteria RBC Treatment of Strength Accid Mine Drainage" Proc., The 35th Annual Conference, JSCE. Sept.1980, pp. 670-671. (in Japanese)
24. Ito K. and Matsuo T. ,"The Effect of Organic Loading on Nitrification in RBC Wastewater Treatment Processes" Proc., First National Symposium on RBC Technology, Pa. U.S.A., Feb.1980, pp.1165-1176.
25. Kubota H., and Takahashi M. ,"Effect of Daily-Fluctuation on Rotating Biological Wastewater Treatment Contactor", Japan Journal of Water Pollution Research, Vol.1, No.3 Dec.1978, pp.175-182. (in Japanese)
26. Kubota H. ,"Operation and Design Proceedures of RBC", ECE., Vol.9, No.6, June.1980, pp.50-56. (in Japanese)
27. Ishiguro M., Watanabe Y., and Masuda S. ,"Advanced Wastewater Treatment by RBC", Proc., 11th Symposium of Sanitary Engineering, JSCE., Jan.1975, pp.109-114. (in Japanese)
28. Watanabe Y., Ishiguro M., and Nishidome K. ,"Kinetic Analysis of Denitrification by RBC Unit", Proc. of JSCE No.276, Aug.1978, pp.35-43. (in Japanese) and Transactions of JSCE., Vol.10, Aug.1978, pp.166-169.
29. Watanabe Y., Ishiguro M., and Nishidome K. ,"A Mechanism of Substrate Removal in RBC Reactor (I), (II)", Journal of JSWA., Vol.15, No.172, Sept.1978, pp.24-34. Vol.17, No.195, Aug.1980, pp.14-23. (in Japanese)
30. Masuda S., Ishiguro M., and Watanabe Y. ,"Nitrogen Removal in RBC (1). Simultaneous Denitrification with Nitrification in Biofilm", Journal of JSWA., Vol.16, No.187, Dec.1979, pp.24-32. (in Japanese)
31. Sudo R., Okada M., and Mori C.,"An Aspect of Micro-Organism in the Bio-film for RBC", Journal of Japan Society of Fermentation Engineering. Vol.56, Feb.1978, p.580. (in Japanese)
32. Sudo R., Okada M., and Mori C. ,"The Microorganisms Control in a RBC units", Water and Wastewater, Vol.19 No.7, July.1977, pp.61-70. (in Japanese)
33. Takahata S., and Fujishima M. ,"The Rational Design Proceedure of RBC", Proc., The First National Symposium on RBC Technology, Nov.1979, pp.27-34. (in Japanese)
34. Kato Y. ,"Energy Saving in Wastewater Treatment Plants

and RBC", ECE.,Vol.9, No.8, Dec.1980, pp.69-77. (in Japanese)
35. Ishiguro M., Watanabe Y., Masuda S., Yamaguchi K., and Uchida H.,"Advanced Wastewater Treatment by RBC Unit (I-IV)", Journal of JSWA., Vol.12, No.129, Feb.1975, pp. 46-54, Vol.14, No.152, Jan.1977, pp.32-41, Vol.14, No. 161, Oct.1977, pp.53-59 Vol.16, No.185, Oct.1979, pp. 40-48. (in Japanese)
36. Watanabe Y., and Ishiguro M.,"Denitrification Kinetics in a Submerged Biological Disk Unit", Progress Water Technology, Vol.10, Nos.5/6, pp.187-195. IAWPR.,9th International Conference, Stockholm, Sweden, June.1978.
37. Watanabe Y., Ishiguro M., and Nishidome K.,"Nitrification Kinetics in a Rotating Biological Disk Reactor", Progress Water Technology, Vol.12, pp.233-251. IAWPR., 10th International Conference, Toronto, Canada, June. 1980.
38. Watanabe Y., Bravo H.E., and Nishidome K.,"Simulation of Nitrification and its Dynamics in a Rotating Biological Contactor", IAWPR., 11th International Conference, Kapetown, South Africa, March.1982.
39. Kojima S., and Ishiguro M. et.al., "Newest Technique of Wastewater Treatment by Biochemical Process" Management Co., July.1975. (in Japanese)
40. Iwai S.,"Guide book of Night Soil Treatment", Kankyo-gijutsu Co., May.1978. (in Japanese)
41. Research Groupe of RBC,"Wastewater Treatment by RBC Process", Sankaido Co., Sep.1978. (in Japanese)
42. Ishiguro M. et.al.,"Compillation book of Wastewater Treatment Technique by RBC", IPC Co., March.1979. (in Japanese)
43. Iwai S.,"Guide book of Sewage, Industrial Wastewater and Sludge Treatment", Kankyo Gijutsu Co., Aug.1979. (in Japanese)
44. Iwai S., Kato Y. et.al.,"Guide Book of Domestic Wastewater Treatment", Kankyo Gijutsu Co., Aug.1981. (in Japanese)
45. Special Edition,"Sewerage in Japan",Journal of WPCF., Vol.52, No.5, May.1980, pp.841-1054.
46. Nakamoto I.,"Current Status of Sewage Treatment Technology and RBC Process in Japan", The Third National Symposium of RBC Technology, Oct.1981, Proc. pp.72-78. and ECE., Vol.10, No.12, Dec.1981, pp.47-51. (in Japanese)
47. Supervisions of Ministry of Construction, Ministry of

Health & Welfare and Environmental Protection Agency
"Design Criteria of Night Soil Purification Tank and its Explanation", Japan Architecture Center, Aug.1980. (in Japanese)

RBC UNIT: BEST IN SEWAGE TREATMENT FOR SAUDI ARABIA

Sharaf Eldin I. Bannaga. Directorate for Housing and Military Cities, Saudi Arabian National Guard, Riyadh, Saudi Arabia.

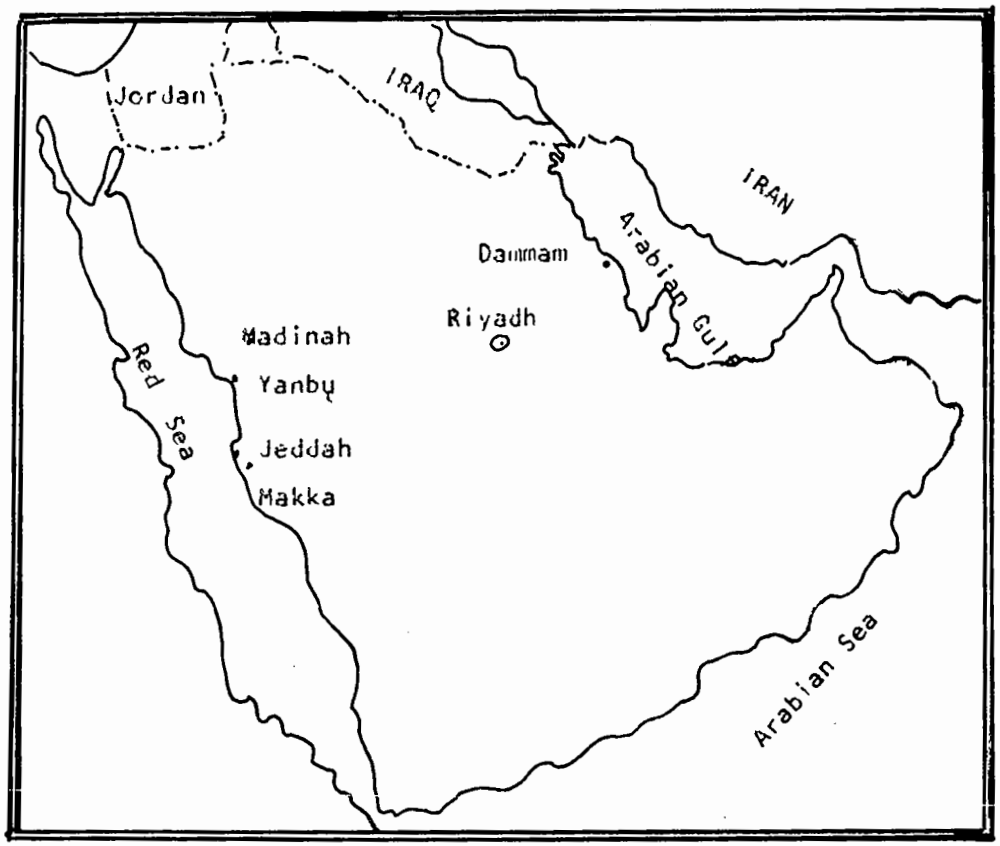
INTRODUCTION: THE NEED FOR ADEQUATE WASTEWATER TREATMENT FOR SAUDI ARABIA.

The Kingdom of Saudi Arabia which boundaries are the Red sea from the West, the Arabian Gulf from the East, Latitude 17 from the South and Latitude 30 from the North, covers an area of more than two million Sq. km and is inhabited by seven million people approximately. The region occupies a leading place in the Islamic World and sustains its heritage and culture.

Being a major exporter of petroleum, the Kingdom has been spending vast sums of money on ambitious development programmes, aimed at every sector of the economy. This very fact provides a link with the question of water supply and waste water purification, for just as all vital processes depend functionally on water as the medium, so do almost all Industrial production processes.

Due to the scarcity of water, which imposes a major problem for the Kingdom of Saudi Arabia, waste water treatment should not be aimed only at disposal for public health considerations,but also at recirculation of treated effluent for special use. Use of treated effluent is at the moment

KINGDOM OF SAUDI ARABIA



limited to Irrigation and landscaping, but further developments could include use in industries and drinking water supplies. It is encouraging to note the Fatwa(legal opinion) announced by the Board of Scientific(Religious)Research, Ifta (delivery of legal opinion), Dawa (invitation to Islam) and Guidance regarding the possible use of adequate, safe treated effluent for religious purposes.

It is therefore imperative that great attention be given to development of an adequate waste water treatment processes that will optimise in the Kingdom capability,manpower and materials. This need can best be demonstrated by some schemes executed in similar countries which are not in-country compatible.

To pick a process that will discharge total ability and satisfy the Kingdom requirements for waste water treatment the writer has to recommend the RBC(Rotating Biological Contactors)process.

The purpose of this report is to present a short account of literature about the RBC process which may be beneficial to engineers, consultants and research workers working in Saudi Arabia and elsewhere. However, the writer wishes to state that the opinions expressed in this report are entirely his own and do not necessary reflect the views of the governmental organization by whom he is employed.

RBC UNIT: PRACTICAL APPLICATION AND TECHNICAL PARTICULARS

The Rotating Biological Contactors (RBC) unit is a secondary biological treatment unit for waste water. The system consists of a number of large diameter plastic or expanded metal discs mounted on a horizontal shaft and placed in a reaction vessel which is often of a semi-circular cross-section. Numerous terms are used throughout the wastewater treatment literature to designate RBC's. Among the trade terms in current use are the Bio-Disc, Bio-Surf, Aero-Surf, Surfact, Bio-Sperial, Rotating Disc, etc.

The RBC process was developed initially in Europe in the 1950's. Further development of the process began in the United States in 1965 and has continued to the present time. However, active commercial use of RBC plants had not started until only in the early 70's in the U.S.

In Saudi Arabia practical application of the RBC process started a few years ago in the form of package plants serving small communities, such as university campuses, Hospitals, army bases etc. Initial emphasis was directed to

secondary treatment of municipal wastewater and may continue years ahead.

There are now more than 50 RBC plants operating in Saudi Arabia treating of over 20 MGD of mainly domestic wastewater These plants range in size from one to twenty four RBC assemblies and treat wastewater flows up to 4.8 MGD plant, King Abdul Aziz International Airport at Jeddah, is the largest Aero-Surf facility in Saudi Arabia.The RBC process is now gaining wide acceptance in the Kingdom for a large number of plants are under construction. These include the 7.5 MGD Yanbu RBC plant near the Red Sea and those awaiting construction for use of the Saudi Arabian National Guard Housing Project-Phase one at five different localities and capacity of which exceeds 8 MGD.

It should be emphasised that the growth of RBC process with regard to product development and commercial utilization seems promising and could be rapid when its applicability is recognised by the local authorities.

The RBC unit consists of;

a) Large plastic discs mounted on a horizontal metallic shaft (refer to figure 1). The discs are so mounted that slightly less than half of their surface area is immersed in waste water.
b) The discs and shaft assembly is placed in a tank which has a rectangular surface area. The tank is usually constructed of reinforced concrete.
c) A driving system is incorporated with the disc and shaft assembly. The mechanical driving system incorporates an electric motor that rotates the disc and shaft assembly.In Autotrol Aero-Surf units the disc and shaft assembly is rotated by buoyant force exerted by captured air(refer to figure 1). Aero-Surf assembly consists of corrugated media with plastic cups attached around the outer perimeter and over the entire length of the contactor. The media assembly is installed in a tank in the same manner as a conventional unit with the addition of an air header at a low pressure into the attached cups. The captured air exerts a buoyant force, which in turn exerts a torque on the shaft sufficient for rotation.
d) The tank is divided by a number of baffles for flow direction as well as creation of stages to the

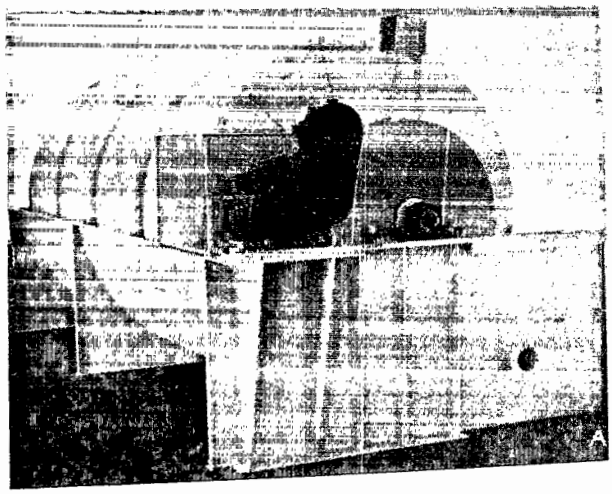


Figure 1

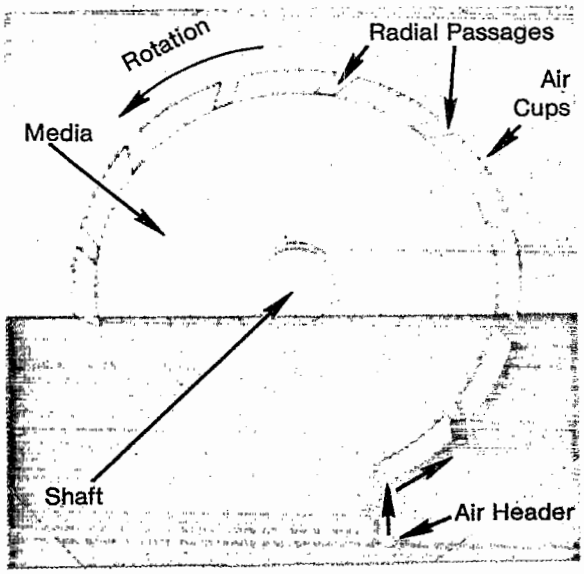


Figure 2

process.

Statically the disc and shaft assembly support the biological growth. The micro-organisms which are present naturally in wastewater adhere to the disc surface and multiply quickly within a few days of start-up and they cover the entire surface area of the discs.

By rotating the disc and shaft assembly two functions are fulfilled:

a) Increasing the dissolved oxygen content of the mixed liquor.
b) Provide contact between the biological growth and the waste water. The rotation alternately submerges the attached biological growth and then exposes them to air. The mixing and agitation enables food and oxygen to penetrate further into the biomass.

The process is continuous as the biological slime is alternately exposed to waste water and then to air. This provides a means of exposing the biological growth to the organic polluting load and of aerating the waste water. Replacing the mechanical drive system with an air drive system serves a number of purposes:

a) Mixed liquid dissolved oxygen concentration is increased through supplemental aeration.
b) Thinner biomass is achieved as a result of increased shearing action as air bubbles rise through the radial passages and corrugations in the media.
c) Power consumption is optimized through variable speed control and reduction of biomass.

RBC PROCESS: COMPARISON OF OPERATIONAL CHARACTERISTICS WITH CONVENTIONAL PROCESSES

The RBC process may be defined as the biological decomposition of organic waste materials in aerobic condition and without offensive odours as opposed to the anaerobic process of putrefaction with which small nuisances are inveriably associated. In this respect the RBC process is comparable to the conventional aerobic processes of the percolating filter and the activated sludge.

Comparison of the RBC Process with the Percolating Filter Process:

Both the RBC process and the percolating filter process are fixed film biological reactors. The growth in both units is supported by fixed solid surfaces, the discs in the RBC unit and the media in the percolating filter. The difference between the two biological processes is that the microbial mass in the RBC system is passed through the waste water, while in the percolating filter the waste water is passed over the microbial mass.

A factor contributing to the advantage of the RBC process is that the rotating discs provide an intimate contact between the biological slime and the waste water. The rotating discs also increase the degree of mixing, agitation and turbulence in the reaction vessel and in doing so, the organic pollutants in the waste water will stand better chances of diffusing into the biological film. Efficiency of the percolating filter process is usually impeded by unevenness of the settled sewage over the whole surface of the filter and of bad circulation of air through the bed which must reach the surface of each piece of the filter material to keep the right kind of bacteria and other organisms alive and active.

The clogging that occurs in the percolating filter system is prevented with the RBC unit by the sloughing action of the excess biomass from the discs caused by the shearing forces developed as the discs rotate. Percolating filters are susceptible to clogging by grit settlements, moss,and weeds.

The dissolved oxygen content of the waste water is increased by the rotating discs in the RBC units and the supplemental air used for driving the system of the Aero-Surf unit. This may prevent the development of anaerobic conditions and hence avoiding foul smells and objectionable sights both on and off the sewerage works.

Ronald Antonie[6] and associates, reported that there were no nuisances, no objectionable odours and no files at the village of Millwankee, Wis., USA and so did Simpson[20]. This is becuase the development of flies, which are often associated with the percolating filter operation, is prevented in the RBC unit operation by the continuous wetting of their biological growth.

A substantial amount of research work has been carried out by a number of research workers to specify the BOD

Loading that would be acceptable to the RBC process. The Water Pollution Research Laboratory[23] at Stevenage, UK, suggested a BOD Loading of 5-6 g/m^2 of disc median, Ellis and Bannaga[15] reported 20 g/m^2 while Autotrol Corporation used 12 g/m^2. For comparison the appropriate BOD Loading on a low-rate single pass percolating filter containing a medium of 50mm nominal size would only be 1-2 g/m^2 according to the Water Pollution Research Laboratory. This Indicates that the area required by a percolating filter to purify the same amount of settled sewage is much greater. Borchardt[11] reported that the actual area occupied by the RBC unit was about 1/10th of that required by a percolating filter.

It is unnecessary to recycle the effluent to achieve maximum wetting, dilution and flushing action in the RBC process which is required for the percolating filter. The report produced by the British Ministry of Housing and Local Government[12] recommended the use of recirculation for strong waste that makes the sewage more difficult to treat using percolating filters.

Both the RBC and the percolating filter systems are simple to maintain and have a relatively low cost. However the percolating filter requires more labourfor the sparge holes of its distribution arms and the arms themselves need to be regularly cleaned and brushed out and its dosing chamber and air pipes need to be maintained properly. The RBC unit requires minimal attention from operators but its belt and chains require checking for alignment.

Comparison of the RBC Process with the Activated Sludge Process.

The RBC process is somewhat similar to the activated sludge process in that it has a suspended culture of bio-mass in its mixed liquor and both processes possess aeration devices. However, the part of the bio-mass that is in suspension in the mixed liquor is too small to compare with the total amount of the biological growth supported by the surfaced of the discs and would therefore contribute only marginally to the treatment.

The RBC process retains a large fixed biological film and a great micro-organism population and because of this the RBC process is less upset by the variation in hydraulic loading than the activated sludge process. Activated sludge process is easily upset by industrial wastes and is incapable of handling shock loads.

The RBC unit is more efficient per unit volume than activated sludge unit. Ainsworth[1] reported that a settled sewage BOD Loading of about (0.48-1.28)kg/m^3 of tank capacity is suitable for fairly good purification by an activated sludge process. For comparison the appropriate BOD loading on a RBC unit used by Ellis and Bannaga was 3 kg/m^3 of tank capacity. This indicates that the treatment capabilities of the RBC process are much greater than those of the activated sludge process.

Unlike the activated sludge process oxidation of ammonia can be attained in the RBC process within normal retention periods.

The sludge solids from the RBC process have favourable concentration characteristics thus eleminating the need for special thickening. De-watering of sludge generated by a RBC unit through vacuum filteration was satisfactorily accomplished according to Ellis and Bannaga. Sludge generated by an activated sludge unit was not amenable to de-watering by vacuum filters according to Quirk[19].

The only disadvantage to RBC process is the need for covering the unit to protect the discs from wind, sand storms and rains.

The RBC unit requires little maintenance and minimal operator's attention when compared with the activated sludge unit for the RBC unit is mechanically simple. The activated sludge unit requires careful supervision. The British Ministry of Housing and Local Government recommends that specialist advice from the manufacturer should be obtained because the great complexity of plant piping arrangement and multiplicity of aeration devices etc and because the effectiveness of the plant is dependent upon the human element.

The power requirements for the RBC system are considerably less than an activated sludge system, because power is only required to rotate the discs.

RBC: ECONOMIC FEASIBILITY AND SUITABILITY FOR SAUDI ARABIA.

In order to examine the economic feasibility of the RBC unit, the unit has to be matched with the conventional ones in regard to capital expenditure and running costs. It is very difficult to compare the capital costs of waste water treatment units in Saudi Arabia since these units are mostly parts of large projects usually awarded on

lumpsum basis. The capital cost required for supplying,constructing and installing a RBC unit may be comparative to that of a percolating filter of same capacity for the cost of land is becoming very expensive in urban areas of Saudi Arabia. The RBC is now recognized as a cost-effective and cost-competitive since the annual operation and maintenance costs play an important factor in determining the selective. Lundberg and Plerce[18] present a summary of the results of cost-effectiveness analyses which compared the air-drive and mechanical RBC processes with air and pure oxygen activated sludge processes over a range of design flow capacities. The results of their studies indicated that RBC process,throughout the range of design flow capacities they used in analysis were less costly than activated sludge processes in supply and construction as well as operation and maintenance. The comparison of an extended aeration plant, which operates on activated sludge principles, and RBC plant for the Makkah, 14 Saudi Arabia, municipality showed that the extended aeration plant was about 70% more expensive to operate and maintain than Aero-Surf.

The particular problems of waste water treatment plants for Saudi Arabia are related to such factors as lack of skilled supervision, high operating temperatures, high rate of expansion due to intensive development and urbanization, recognition of scattered small communities such as the presence of military cantonments, isolated camps, special settlements etc, scarcity of water, dry weather, supply of local materials, availability of funds etc. All these factors may be adequately dealt with by the application of RBC system for the following considerations.

a. The system claims the benefit of reliability without frequent supervision.
b. The system does not depend substantially on oxygen dissolved in water which saturation concentration decreases as temperature rises.
c. The system is susceptible to upgrading or extension.
d. The system is well established for small communities application.
e. Due to scarcity of water, recirculation of waste water may be required for supplementing drinking water supplies. The system is capable of removing objectionable ammonia and producing nitrate required for potable water.
f. The system is capable of producing sludge of adequ-

ate quality liable for disposal on drying beds which effectively operate in dry weather conditions.

g. The discs which cost makes a large proportion of the system equipment costs can be manufactured locally because they are of plastic material which is a petrochemical product. Firms such as SABIC(Saudi Arabian Basic Industries Co), SAPPCO (Saudi Arabian Plastic Products Co) etc are well established for manufacture of such product.
h. Funds for installation of a technically sound system are easily allocated and therefore there is no necessity for application of systems operating on intermediate technology principles or committing nuisance and obnoxious to the community or restricting the freedom of expansion and advancement.

DESIGN CRITERIOR

One would ask, why the RBC process is not widely used in Saudi Arabia if it is adequately acceptable for waste water treatment and particularly satisfies the Kingdom requirements.

Being comparatively young in the market, in contrast to the conventional treatment processes, use of RBC process in Saudi Arabia is hampered partly by lack of adequate literature needed for engineers and consultants who handle waste water treatment in the Kingdom and partly by manufacturers who make little effort to pass on knowledge and convey correct information. Absence of such valuable information may subject the RBC process to reservation within the engineering community. Since waste water treatment is a supporting facility in most large scale projects. which are usually awarded on lump-sum contract basis to general contractors, its emphasis is not greatly aknowledged.

The ambitious development programmes launched in Saudi Arabia are so great that justified employment of multi-nationals who possess different technical background and approach and that coordination between different organizations could hardly be secured.Some programmes have been rushed and their periods squeezed for time saving. In absence of such a governmental body whose main task would be to furnish engineering departments of all organizations with sufficient and correct data and monitor their performance accordingly, the performance of such departments will depend totally on the type,quality and talent of personnel employed. In these circumstances if personnel employed are European for example they will

obviously choose European products since they are familiar with them, the Americal will choose American products etc.

At present designs of waste water collection and disposal schemes are generally based either on the existing facilities with annual percentage growth factor, or on current design criterior in industrial countries.Neither of these bases is satisfactory. Since literature and studies leading to identification of such design criterior are limited, it is insufficient for the engineer practising in Saudi Arabia just to exercise his technical competence and skill. His task has to be extended, within his own initiatives, to include gathering of information to arrive at reasonable design criterior applicable to Saudi Arabia. His success depends on the ability to take a positive interest in this direction.

CONCLUSIONS

The RBC units have become widely accepted in recent years, notably in the United States, which is evidenced by the great number of plants operating or under construction and by the influx of several new RBC manufacturers competing into the market. The unit suits best the requirements of Saudi Arabia and its manufacture can be carried out locally.

Engineers operating in Saudi Arabia should be encouraged to keep their designs as simple as possible and to avoid complicated features and sophisticated equipment for the completed plants cannot operate satisfactorily without skilled, talented operators particularly if they depend substantially on mechanical and electrical equipments.

There is a need to establish an organization whose prime tasks would be to identify the objectives of waste water treatment, review waste water treatment processes in application and recommend their use for varied purposes in Saudi Arabia, Lay design criterior, conduct research leading to development of suitable processes as well as their local manufacture etc.

REFERENCES

1. Ainsworth, G. "The Activated Sludge Process" Water Pollution Control Engineering, 6, pp 60-74 (H.M.S.O, 1970)
2. A.C.M., U.K. "Bio-Disc Process for small Community Sewage Purification". Patents granted or pending throughout the world.
3. Antonie, R.L. and Hynek, R.J. "Operating Experience with Bio-Surf Process Treatment of Food-Processing Wastes". Paper presented at the 28th Annual Purdue Industrial Waste Conference, West Lafayette, Indiana May 1973.
4. Antonie, R.L. and Welch, F.M. " Preliminary Results of a Novel Biological Process for Dairy Wastes". Presented at the 24th Purdue Industrial Waste Conference, May 6th-8th, 1969.
5. Antonie, R.L."Application of the Bio-Disc Process to Treatment of Domestic Waste Water" Federal Water Quality Administration of the U.S. Department of Interior, Contract No.14.12.810.
6. Antonie R.L. "Response of the Bio-Disc Process to Fluctuating Waste Water Flows". Presented at the 25th Purdue Industrial Waste Conference, May 5th-7th, 1970.
7. Antonie, R.L."Rotating Disc Dual Waste Water Role". Wat. Wasts Engng. 1971, 8NO. 1, pp 37-38.
8. Autotrol Corporation, U.S.A."Industrial Waste Water Treatment". Bio-Disc Information Bulletin.
9. Autotrol Corporation, "Waste Water Treatment Systems Design Manual". Bio-Systems Division, Milwaukee, Wisconsin, U.S.A., 1978.
10. Autotrol Corporation, U.S.A. "The Aero-Surf Process" Information Bulletin.
11. Borchardt, J.A. "Biological Wastetreatment using Rotating Disc". Biotechnol. Bio-Engineering Symp. 1972, No. 2, pp 131-140.
12. The British Ministry of Housing and Local Government "Operation and Management of Small Sewage Works". H.M.S.O., 1978.
13. C.J.B. Developments Ltd. "The C.J.B. Launches Stenglin Rotating Disc Plant". Surveyor Local Government Technol. 1971, 138 No. 4138, 44.
14. COM Consultants "Comparison of Extended Aeration Plant and RBC Plant for Makkah, Saudi Arabia, November 10, 1980.

15. Ellis, K.V.and Bannaga, S.E.I."A Study of Rotating -Disc Treatment Unit Operating at Different Temperatures". Journal of the Institution of Water Pollution Control, London, January 1976.
16. Hartmann, H. " A Dipping Percolating Filter Plant" G.P. 1,275, 967, Korresp.Abwasss, 1969, No.5, 102.
17. Hartmann, H. " The Dipping Contact Filter". Ost. Wasserw, 1965, 17, 264-269.
18. Lundberg, L.A. and Pierce, J.L. "Comparative Cost-Effectiveness Analysis of RBC and Activated Sludge Process for Carbon Oxidation". Schneider Consulting Engineers, Bridgville, Pennsylvania, U.S.A.
19. Quirk, T.P. and Hellman, J. "Activated Sludge and Trickling Filter Treatment of Whey Effluents". Wat. Poll. Cont. Fed. 1972, 44, 2277
20. Simpson, J.R. "Waste Water Treatment for Small Communities". Process Biochem., 1972, Vol. 7 18-21.
21. Simpson, J.R. "Technical Basis for Assessing the Strength Charges for Treatment and Treatability of Trade Wastes". Wa. Poll, Control, 1967, Vol 66, pp 165-181.
22. Tropey, W.H., Hewkelekian, H., Kaplovsky, A. and Epstein, L. "Effect of Exposing Slimes on Rotating Discs, to Atmosphere Enriched with Oxygen". Presented at the 6th International Water Pollution Research, June 18th-23rd, 1972.
23. Water Pollution Research Laboratory, Stevenage, U.K. " Rotary Biological Contactors". J. Inst. of Pub. Health Eng., May, 1973(3)pp 116-130.

THE FUTURE OF BIOLOGICAL FIXED-FILM PROCESSES AND THEIR APPLICATION TO ENVIRONMENTAL PROBLEMS

Stanley L. Klemetson. Department of Civil Engineering, Brigham Young University, Provo, Utah.

Gary L. Rogers. Department of Civil Engineering, Brigham Young University, Provo, Utah.

INTRODUCTION

The needs and design of municipal and industrial wastewater treatment facilities will be much different in the late 1980's than in the past. Cutbacks in federal funding and strained financial condition of local taxpayers will require that less expensive and more reliable wastewater treatment systems be built. While it was once desirable to build significant excess capacity in new plants to improve the economies of scale at lower current costs, the goal now is to build smaller plants in hopes of technological advances in the future. In many cases older plants are being upgraded for both treatment efficiency and flow capacity. Operational costs are being carefully reviewed by both municipalities and industries. The systems with the lowest energy, maintenance, and labor costs will be chosen when possible.

To appreciate the changes that will occur it is necessary to review the past. In both the United States and Britain the development of biological treatment progressed from sewage farms and discharge into waterways, through intermittent sand filters and contact beds, to trickling

filters, activated sludge basins, rotating biological contactors, and land applications. The term Biological Fixed-Film Process is really a new name for an old process. As will soon be shown we have gone full circle and are returning the semi-passive fixed-film systems.

EARLY DEVELOPMENTS

The disposal of raw and partially treated sewage onto the land was a natural consequence of good farming practices. The sewage placed or poured on the fields from the urban drainage ditches could irrigate the land and provide needed nutrients to the soils and the crops. Since disease control was not a great concern in the nineteenth century, it was not until land became scarce and the potential for profit from sewage-irrigated lands was diminished that that sewage farming was abandoned.

In areas where sandy soils existed, the practice of land application continued in the form of intermittent sand filtration. Dosages were continually intensified, and in many cases, wastewaters were pretreated by use of settling tanks or other biological treatment units. This practice continued until more advanced treatment units were developed.

In areas where tight clay soils existed, it was necessary to build contact beds, which were relatively shallow tanks containing many layers of slate supported on a layer of bricks or filled with broken stone or slag. The original beds used fill-and-draw and resting cycles. The beds provided an excellent site for large populations of microorganisms and removed dissolved as well as suspended solids from the wastewaters. The efficiency of the beds were soon increased by the addition of sprays which permitted application of the wastewaters to the beds on a more continuous basis. The discharged wastewater was also saturated with oxygen. This modification no longer required that the beds be flooded, but rather permitted oxygen to pass through the bed continuously to keep it aerobic. These improved beds were first called bacteria beds, but this was later changed to trickling filters.

Much later the activated sludge process was developed, and promoted as the ultimate in biological treatment. Even this method used a biological film provided by the microbial floc, however, it is not considered a fixed-film process.

With the advent of PL 92-500, a variety of new processes were developed to advance the wastewater treatment technology available. Innovative new processes were developed, not all of which worked well. The Environmental Protection Agency provided funds to encourage the application of these processes in new plants. Even industries were willing to try some of these processes that would reduce their total costs.

Among the developments were the high rate aerated contact bed and the high rate anaerobic system . Both of these systems are improvements of original contact bed. Both are suitable for municipal wastes (1), but the latter will see its greatest use in treating high strength industrial wastes. Specialized bacteria are also being developed to degrade non-conventional organic industrial waste compounds in the fixed film system (2).

While each of the processes have gone through a variety of revisions and updatings, about the last major different type of treatment unit is the rotating biological contactor. In this system, the microbial fixed-film alternatively is rotated into the wastewater and into the air. While this is based on the same principles as the trickling filter, it should provide a higher level of treatment in a smaller area.

PROCESS APPLICATIONS

The fixed-film treatment systems have been applied successfully to a variety of applications. Its role in wastewater treatment has been long and successful. Industrial wastewater pretreatment is more recent. Anaerobic have frequently been used for strong organic wastewaters. The use of anaerobic fixed-film filters is much more recent.

With the tightening of effluent standards, the concern for nitrification and denitrification became extremely important. Again, fixed-film systems proved very adequate.

Water conservation efforts have required the evaluation of many treatment methods. In the power industry, fixed-film systems have been used to remove organic contaminates before reuse within the plant.

Another use that will receive increased attention is aquaculture. The wastewaters used produced by fish and prawn farming must be treated before discharge. Also the internal recycle in the ponds requires that harmful biproducts be removed from the water on a continuous basis to maintain the aquatic population.

CURRENT APPLICATIONS

The future of biological fixed-film system will be discussed in the following sections in the context of their advantages and disadvantages.

Introductory Comments

The activated sludge process has, during the past fifteen years, received significant development to meet the needs of the wastewater treatment industry. If it were not for its high power requirements and the current high costs of buying this power, it is likely that this process would continue to be highly favored. However, the need for economies of operation for treatment plants requires that alternative treatment methods be considered in the design of new treatment plants and the up grading of old plants. The U.S. Environmental Protection Agency has also required this plan.

Land Applications

Land applications of wastewaters, while not considered by most of the industry to be fixed-film process, is a return to the concept of the sewage farm. However, the concern about disease is very important now. While raw sewage is no longer applied directly to the land, treated effluents and partially treated sludges are being applied. The three methods of application: Spray irrigation, overland treatment, and rapid infiltration, are modifications of the fixed-film process. In this case the biological growth occurs on the plant structure or within the soil. This system has limited usage in some regions of the country because of land costs, cold weather operational requirements, and the loss of water discharges to subsequent users.

Coupled Trickling Filter/Activated Sludge Systems

Efforts to upgrade existing wastewater treatment plants have led the development of coupled trickling filter/activated sludge treatment systems. These systems have the advantage of reducing future operating costs while meeting required effluent limitations. In some cases the trickling filter is only for roughing to reduce the load on the activated sludge.

This system will continue to be built for upgrading of existing plants, the multiplication of equipment for this dual system does not recommend it for new small and medium size wastewater treatment plants. In large treatment systems, it is quite possible that the operational advantages and treatment efficiencies of activated sludge systems will be combined with the economies of trickling filters (or other fixed film system) to provide the least cost alternative treatment system.

Trickling Filters

Trickling filters have been subjected to a variety of modifications to improve their operation. They have relatively low operating costs but suffer from inadequate removal efficiencies. Probably the most significant recent modification has been the introduction of plastic media. While rock media was significantly affected by changes in flow, the plastic media only requires a minimum quantity of wastewater for wetting and nutrient source. Beyond that flowrate the variation in flow does not significantly affect treatment efficiency. Trickling filters have, in the past, been considered inadequate to meet effluent standards without additional treatment. Therefore some additional treatment has been required.

Trickling filters will increase in their importance in the design of new wastewater treatment plants. However, in some applications, other fixed film biological treatment system will be more applicable. Among the limiting factors will be loading rate and area requirements.

Biological Towers

A modification of the trickling filter is the biological tower. Either plastic media or redwood can be used in the system. Either natural or forced aeration can be used, depending upon the design of the unit. Depending upon the aeration requirements, the operating costs are low. Additionally, the manpower requirements are low. Very high loadings can be applied to the filter and nitrification can be achieved in some of the filters. It is therefore possible to achieve quite adequate treatment efficiencies.

Biological towers will enjoy a strong role in the future of wastewater treatment systems. Industrial applications are expected to be significant. A number modifications will be developed in the next few years that will make the system more reliable for specific applications.

Biological Aerated Filter

Variations of the biological contact basin have been developed and will continue to receive development. These systems have low area requirements and low capital costs. About 10 years ago a fluidized activated carbon system was developed by Weber (3). There were other modifications, including the addition of air and pure oxygen. Each of these systems had a variety of advantages and disadvantages. A more recent development was the Biological Aerated Filter which uses a fixed bed of granular materials (2). The basic difference in the systems are structural and operational modifications.

The current systems will continue to experience popularity in the future, however, wide-spread full-scale applications will be slow in coming. Special purpose and industrial applications are and will continue to be the most likely use.

Anaerobic Filters

Anaerobic systems have often been used for strong organic wastewaters. The high rate anaerobic filter, a modification of the old contact bed system, effectively treats high strength organic wastes. It has a low operating cost, has high removal efficiencies, but does require some post-treatment system prior to discharge of effluent. Among the wastes that have been treated are food processing, pharmaceutical, sugar, potato, and beet sugar.

These systems will experience an increasing demand from industry and little demand from municipal waste treatment systems.

Rotating Biological Contactors

The development of the rotating biological contactors has opened up a new process of treating wastewaters in small to medium size systems with a minimum of equipment or manpower. Power costs are low for each shaft, and both rotation and aeration can be achieved by using the air drives. The systems are quite suitable for upgrading existing plants by adding the units to existing aeration tanks.

The systems have moderate land requirements, but have high capital costs. While the concept is good, not all of the manufacturers have produced good equipment. There have been a higher than expected number of equipment failures, including shaft failures and media failures. In addition, the published design curves are unrealisticly low and promote under design of treatment systems. Once these difficulties have been cleared up the systems have a good potential for the future.

Summary Comments

All of the systems can be compared on the basis of loading rates and capital costs. Starting with rock trickling filters as having the lowest loading rate, improvements can be obtained by using plastic media. At the top of the loading scale, the anaerobic filters can receive

the highest loading rates. The other units fall in between these limits. The required areas are inversely proportional to the loading rates.

The biological anaerobic and aerated filters have the lowest costs with rotating biological contactors being the highest. The other units fall in between these limits.

The overall comparisons are more clearcut than they should be. Each of the fixed film units have an optimum application and optimum size of operation. Each application will have to be analyzed for specific needs and locations.

FUTURE DEVELOPMENTS AND NEEDS

Some of the fixed-film biological treatment systems have been unable to meet effluent standards. The current push to relax those standards to about 50 mg/l BOD and no limit on suspended solids on selected waterbodies has made the return to trickling filters for the complete secondary treatment, much more realistic.

As the cost of energy increases, more treatment plants will be equipped with energy conserving equipment. This requirement will mean that more fixed-film systems will be constructed. The trickling filter, with its low energy requirements will experience continued improvements in design and media to meet effluent requirements. Combined Activated Sludge/Trickling Filter systems will be built to reduce the costs of operation. Improvements in media design to improve efficiency and to reduce cost will continue to be made. Alternative media will have to be developed to sever the ties to petroleum products.

Rotating Biological Contactors, which still hold a future promise of success, have several problems to overcome. The design curves need to made realistic so more success plants can be designed and built. The great, uneven, weight of the rotating biomass will require that design changes be made to prevent failures of the shafts. The media does not have the lifetime necessary for economical operation at all plants. These are challenges that need to be met since the system can provide a low cost operation for both power and manpower.

Biological towers, basicly a tall trickling filter, can provide an economical operating unit for selected applications. The biological aerated filter will continue to

be developed for specific applications. It is unlikely that it will be used as the sole biological treatment process for many plants.

SUMMARY

Biological fixed-film processes have been around for a long time. While many of them have been placed on the shelf for many years, their usefulness is being re-established, and they are being used again as a viable method of wastewater treatment.

REFERENCES

1. Switzenbaum, M.S. and Jewell, W.J., "The Anaerobic Attached Film Expanded Bed Reactor for the Treatment of Dilute Organic Wastes." TID-29398, National Technical Information Service, Department of Commerce, Springfield, Virginia, August 1978.
2. Stensel, H.D., "Biological Treatment Systems for the 1980's." Proceedings Utah Water Pollution Control Association Annual Meeting, April 16-17, 1982, Salt Lake City, Utah, Ed. Stanley L. Klemetson, Brigham Young University, Provo, Utah, 1982, pp. 22-26.
3. Weber, W. J., Jr., Physicochemical Process for Water Quality Control. John Wiley & Sons, Inc., p 164, 1972.

PART III: BIOFILM AND BIOKINETICS

PROCESSES INVOLVED IN EARLY BIOFILM FORMATION

James D. Bryers. Engineering Science, Swiss Federal Institute for Water Resources and Water Pollution Control (EAWAG), Dübendorf CH-8600 Switzerland.

INTRODUCTION

Fundamental and applied research in fixed-film biological processes has steadily progressed in the past ten years. Atkinson and Fowler (1) review the significance of microbial films in the fermentation industry while Cooper and Atkinson (2) and Smith et.al., (3) provide state-of-the-art symposia on fixed-film bioreactors in wastewater treatment.

A large portion of this research has focused on the mathematical description of substrate depletion within a biofilm - i.e., "biofilm kinetics". Typically, such kinetic models describe one dimensional mass transfer of substrate with simultaneous biological reaction; the resulting differential equation is

$$D\, d^2S/dx^2 = -r_i \qquad (1)$$

where S = substrate concentration in biofilm (ML^{-3}), D = effective substrate diffusivity (L^2t^{-1}), r_i = intrinsic substrate depletion rate ($ML^{-3}t^{-1}$), x = direction of substrate

flux (L.). Solutions to Equation 1 depend upon (a) prevailing boundary conditions and (b) the dependency of r_i on substrate concentration. Harremöes (4) and Riemer (5) provide excellent reviews of the extensive literature on various solutions to Equation 1. Those intrinsic kinetic forms assumed for r_i most relevant to sanitary engineers are the following:

First order	(ref. 6) :	$r_i = k\ S$	(2)
Zero order	(ref. 7) :	$r_i = k_o$	(3)
Half order	(ref. 8) :	$r_i = k_{1/2} S^{1/2}$	(4)
Saturation	(ref. 1, 8, 9, 10, 11, 12) :	$r_i = k\ S/K_s + S$	(5)

Unfortunately, these models only deal with substrate removal kinetics and ignore biofilm development. Equation 1 tacitly requires that r_i be either zero order in biofilm concentration or, if first order, that biofilm mass remain constant. Otherwise, an additional equation describing biofilm accumulation is required. Past works have either simply ignored biofilm production (8) or assumed zero biofilm accumulation by equating biofilm production to endogeneous decay processes (11, 12). In most cases, processes governing biofilm formation and, thus eventual fixed-film reactor performance are neglected; consequently, important information about reactor design, start-up procedures, and control of biofilm thickness remains unknown.

Contributing Processes

Biofilm net accumulation within a turbulent flow field proceeds as shown in Figure 1. Five stages are evident: (1) induction or lag, (2) exponential accumulation, (3) decreasing rate, (4) plateau, and (5) sloughing. Processes involved in this net accumulation can include (Figure 2):

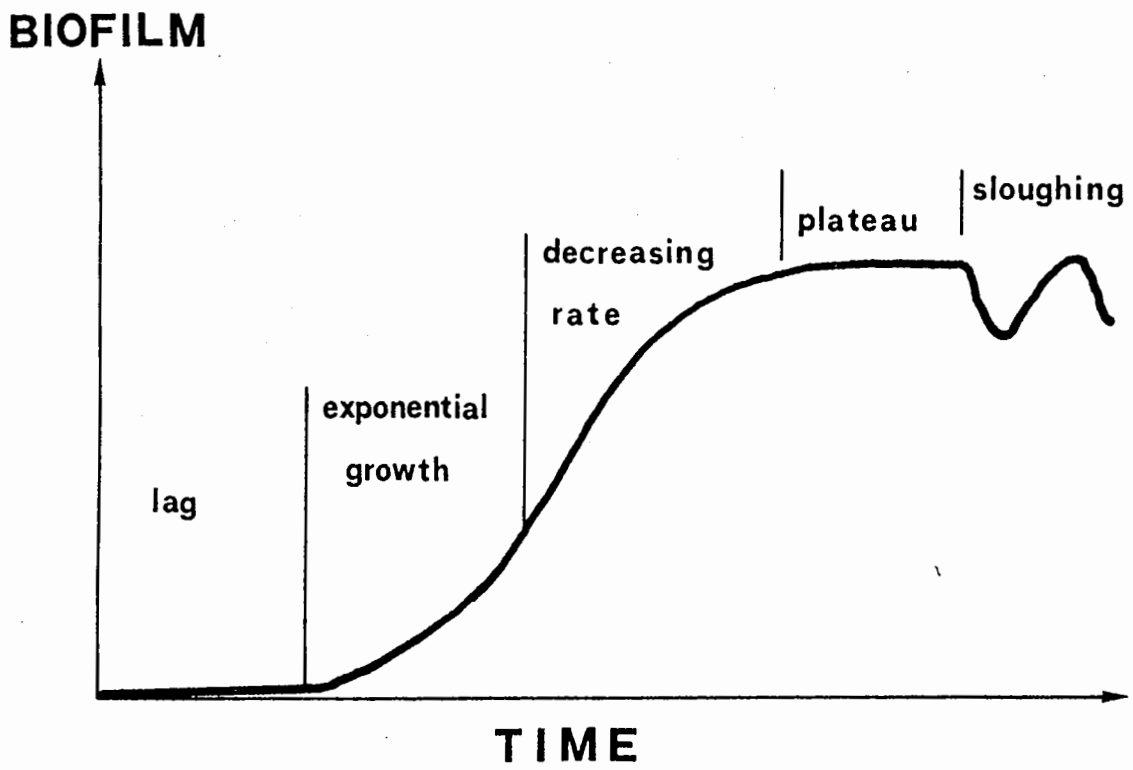


FIGURE 1. FIVE STAGES OF BIOFILM DEVELOPMENT

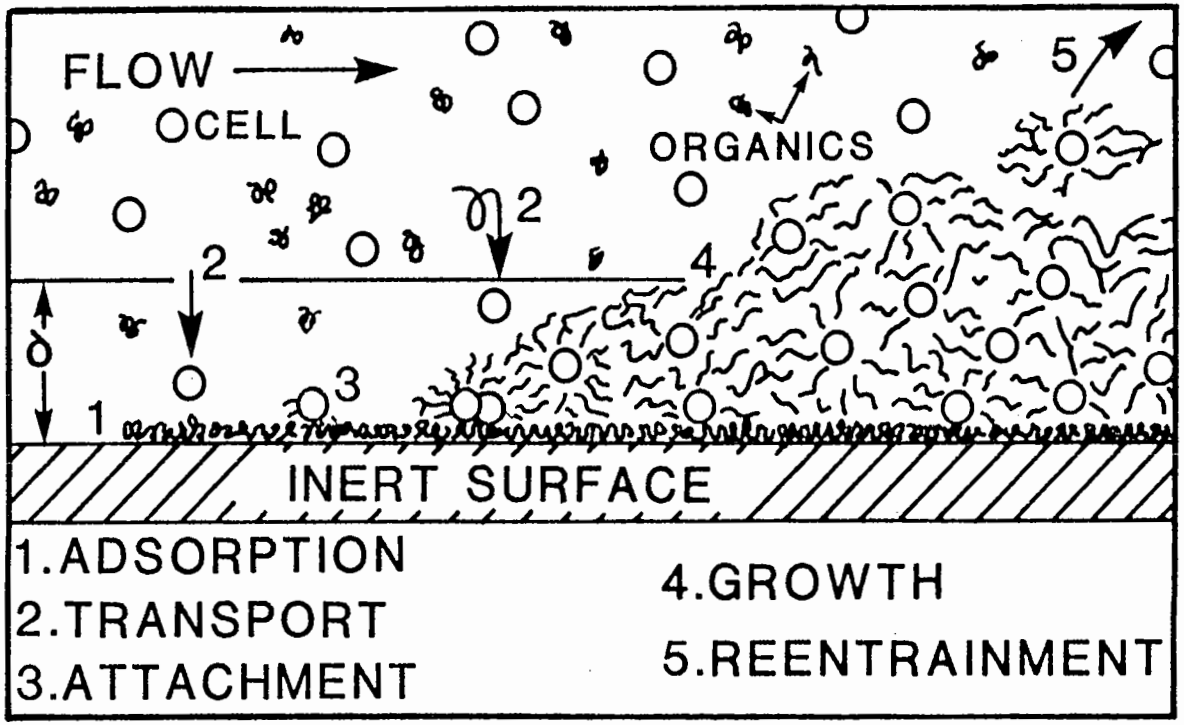


FIGURE 2. PROCESSES INVOLVED IN BIOFILM ACCUMULATION.

1. adsorption of organic molecules to the wetted surface.

2. deposition of bacterial cells to the organic-treated surface. Deposition rate can be considered the sum of bacterial cell transport and cellular attachment rates.

3. cellular growth, reproduction, and extracellular polymer formation.

4. detachment of biofilm and entrainment of debris into the fluid.

Trulear and Characklis (13), Bryers (14), and Characklis (15) provide extensive reviews on these processes and their involvement in *fouling* biofilm development. This paper will present methodology used to quantify the physical transport and microbiological processes involved in early biofilm formation.

EXPERIMENTAL PROTOCOL

Individual processes, and thus, net biofilm developement are considered in this study to be functions of the following:

1. prevailing hydrodynamic conditions - i.e., linear velocity, shear stress at the wetted surface, or Reynolds number, Re.

2. concentration of bacteria suspended in the ambient fluid, X.

3. metabolic activity of suspended bacteria as indicated by their growth rate, μ.

4. biofilm concentration as COD mass per area, B.

Consequently, the laboratory reactor system shown in Figure 3 was employed for it allowed continuous surveillance of biofilm development under defined conditions of Re, μ, and X. The system is operated as two completely stirred tank

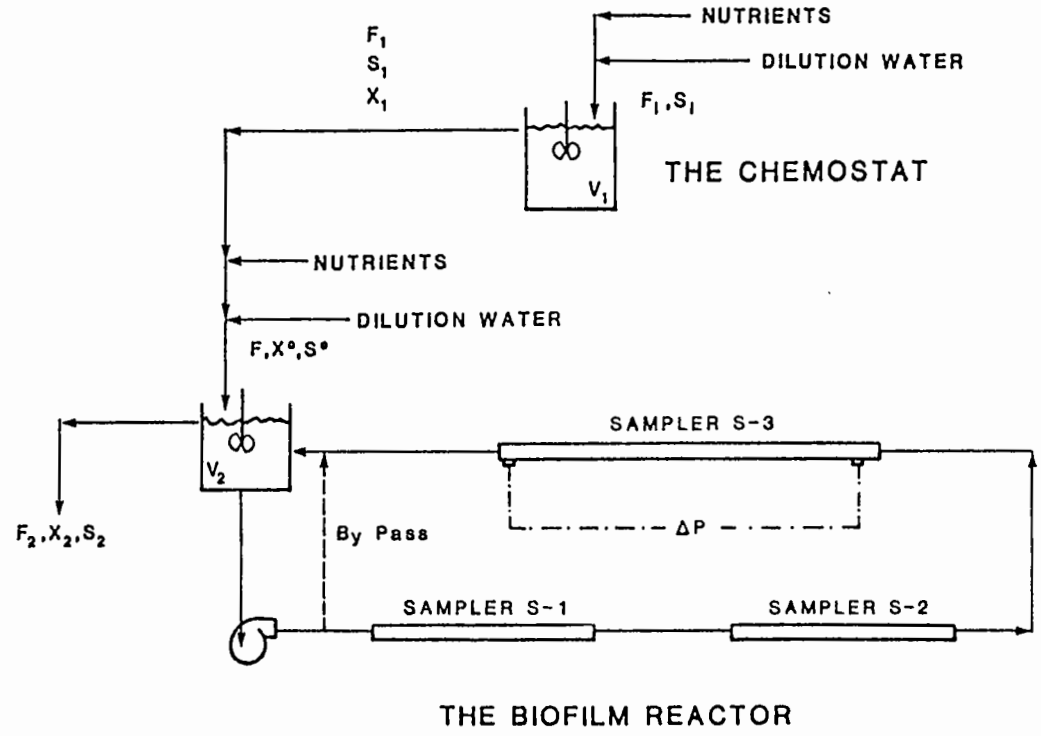


FIGURE 3. REACTOR SYSTEM DIAGRAM. CSTR 1 OPERATED AS A CHEMOSTAT WHILE CSTR 2 WAS THE BIOFILM REACTOR. OPERATING CHARACTERISTICS GIVEN IN TABLE I.

Table I.
Pertinent Characteristics of CSTR 1 and CSTR 2.

	CSTR 1: The Chemostat	CSTR 2: The Biofilm Reactor
System Specifics		
Reaction Volume (cm^3)	3000	4750
Total Wetted Surface Area (cm^2)	1070	5934
Surface Area: Volume (cm^{-1})	0.36	1.25
Dilution Rate (h^{-1})	0.33	4.0
Mean Residence Time (h)	3.0	0.25
Dilution Water Flow (cm^3h^{-1})	998	13000
Effluent Flow Rate (cm^3h^{-1})	1000	19000
Growth Specifics		
Inlet Substrate		
TSB: Glucose (wt: wt)	9:1	1:1
Combined Concentration		
(mg l^{-1})	1000	20.0
(mgCOD l^{-1})	850	28.0
Microbial Feed	Initial inoculation with heterogeneous population	CSTR 1 effluent
Temperature (°C)	31	31
pH	8.1	7.3
Recycle Loop Specifics (CSTR 2 only)		
Recycle Loop Tube length (cm)		1219.0
Inside Tube Diameter (cm)		1.27
Recycle Reynolds Number	13000	26000
Recycle Flow Rate ($cm^3 - s^{-1}$)	104	208
Recycle Velocity ($cm - s^{-1}$)	82	164
Test Sections	S1	S2,3
Length (cm)	91.4	104
Inside Diameter (cm)	1.27	1.27
Sample Tubes (S2,3)		
Number		40
Length/Tube (cm)		5.2
Inner Surface Area/Tube (cm^2)		20.7
Total Sampling Surface Area (cm^2)		830.0
% Sampling Area of Total System Area		14.0

reactors (CSTR), in series, such that operation of the first reactor is independent of the second. The first reactor (CSTR 1) is a conventional chemostat. The second reactor (CSTR 2) is a tubular reactor with internal recycle flow. Recycle flow rate in the CSTR 2 tube is far greater than the volumetric flow rate of influent to CSTR 2. The tubular geometry of CSTR 2 is choosen to simulate biofilm development under known hydrodynamic conditions.

CSTR 1 was operated at a residence time of 3 h and serves only as the source of suspended biomass for CSTR 2. CSTR 2 is operated at a residence time of 0.25 h; consequently, the majority of biological activity in CSTR 2 is due to biofilm development.

CSTR 2 contains two sampling sections which allows for periodic determination of biofilm accumulation as COD mass per area. A third section of the tubular reactor serves to monitor the increase in frictional resistance due to biofilm development (15). For this study, the early biofilm formation period is defined as that amount of biofilm accumulated prior to any increase in frictional resistance.

Table I summarizes pertinent operating characteristics of both reactors. Details of the reactor system, start-up procedures, sampling and analytical methods are provided elsewhere (14, 16).

Experimentation is divided into two parts:

Series I. development of an empirical rate expression describing net biofilm development as a function of Re, x, and μ.

Series II. Estimation of the relative magnitudes of individual processes contributing to net biofilm development and the effect of Re and X on those magnitudes.

RESULTS

Series I

Conditions for Series I experiments are given in Table II.

Table II. Series I Experimental Conditions.

Experiment Number	Group	X (mg-TSS l^{-1})	μ (h^{-1})	Re
1	Biomass	4.4	0.28	17,200
2		12.0	0.28	17,200
3		2.8	0.28	17,200
4		13.0	0.28	17,200
5		23.0	0.28	17,200
6		4.0	0.28	17,200
7		10.1	0.28	17,200
8		2.5	0.28	17,200
9	Reynolds	12.0	0.28	17,200
10	number	12.0	0.28	10,600
11		12.0	0.28	19,300
12		12.0	0.28	23,900
13		12.0	0.28	28,800
14	Suspended	18.0	1.0	17,200
15	growth rate	18.0	0.28	17,200
16		18.0	0.16	17,200
17		18.0	0.13	17,200
18		18.0	0.13	17,200

Biofilm development, as COD mass per area, is shown in Figures 4a-c as a function of either Re, X, or μ. Results in Figures 4a-c suggest a first order rate expression of the form:

$$dB/dt = k_N B \qquad (6)$$

The numerical value of the rate constant for biofilm net accumulation, k_N, is determined from statistical regression of B vs time according to the integrated form of Equation 6. These values of k_N are illustrated for each experimental group in Figures 5a-c.

The net accumulation rate constant, k_N is actually a function of the three parameters considered:

$$k_N = k_i X^a Re^b \mu^c \qquad (7)$$

where k_N = biofilm net accumulation rate constant (t^{-1}), k_i = intrinsic biofilm accumulation rate constant, and $(a, b, c)^i$ = empirical constants. Linear regression of data in Figures 5a-c provides the following estimates of the empirical con-

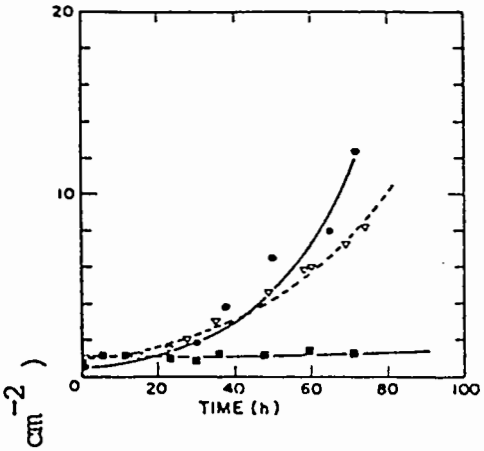


FIGURE 4A. BIOFILM ACCUMULATION AS A FUNCTION OF SUSPENDED BIOMASS. X = 23.0 mg/l (●), 12.0 mg/l (▽), and 2.4 mg/l (■). Re = 17,200 and μ = 0.28 h^{-1}.

(CURVES BASED ON EQ. 8)

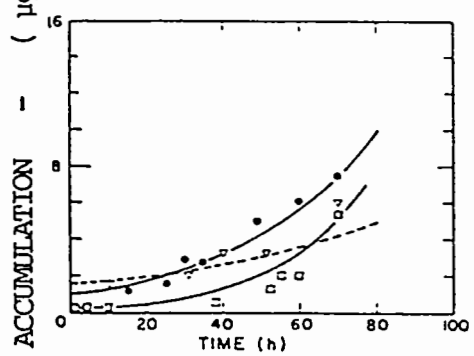


FIGURE 4B. BIOFILM ACCUMULATION AS A FUNCTION OF REYNOLDS NUMBER. Re = 10,000 (□), 17,200 (●), and 23,900 (▽). X = 12.0 mg/l and μ = 0.28 h^{-1}.

(CURVES BASED ON EQ. 8)

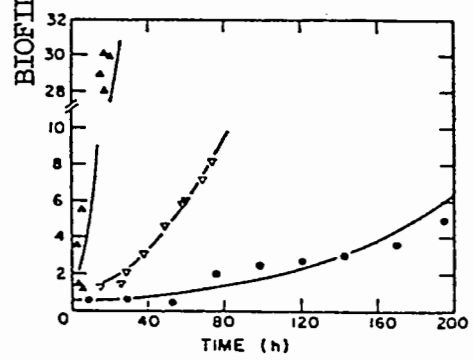


FIGURE 4C. BIOFILM ACCUMULATION AS A FUNCTION OF SUSPENDED BIOMASS GROWTH RATE. μ = 1.0 h^{-1} (▲), 0.28 h^{-1} (▽),and 0.13 h^{-1} (●). X = 18.0 mg/l and Re = 17,200.

(CURVES BASED ON EQ. 8)

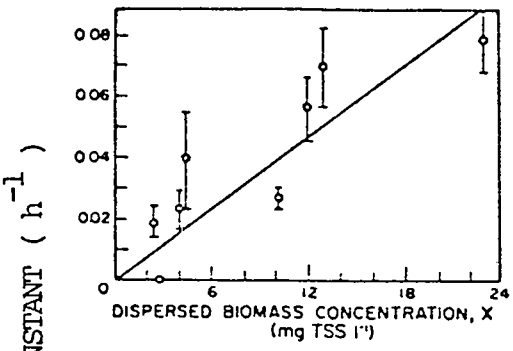


FIGURE 5A. ACCUMULATION RATE CONSTANT, k_n, AS A FUNCTION OF SUSPENDED BIOMASS CONCENTRATION.

Re= 17,200 and $\mu = 0.28\ h^{-1}$

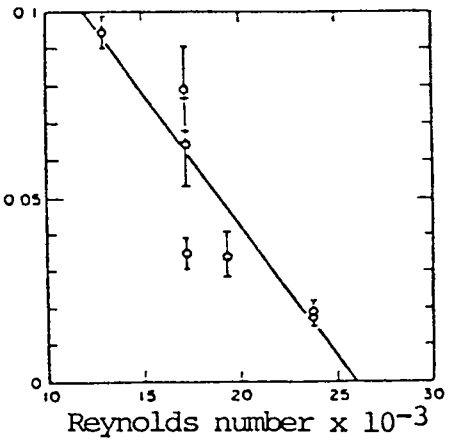


FIGURE 5B. ACCUMULATION RATE CONSTANT, k_n, AS A FUNCTION OF REYNOLDS NUMBER.

X=.18 mg/l and $\mu = 0.28\ h^{-1}$

k_n, BIOFILM SPECIFIC ACCUMULATION RATE CONSTANT (h^{-1})

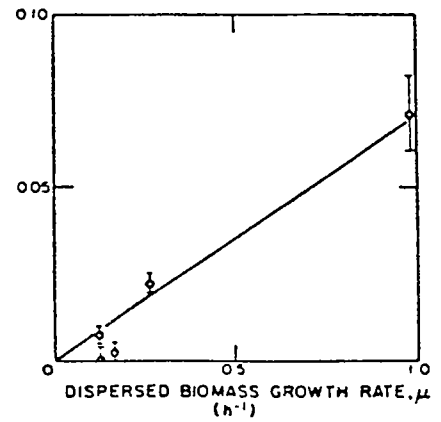


FIGURE 5C. ACCUMULATION RATE CONSTANT, k_n, AS A FUNCTION OF SUSPENDED BIOMASS GROWTH RATE.

X= 18 mg/l and Re= 17,200

stants: a = 1.0, b = -1.0, and c = 1.0. Once the empirical constants a, b, c are known, the intrinsic rate constant, k_i, can be calculated from any set of known experimental conditions (16). The resultant integrated form of Equation 6 can be written as follows:

$$B(t) = B_o \exp [(k_i x\mu/Re) \cdot t] \qquad (8)$$

where $k_i = 125.0 \pm 25$ mg $TSS^{-1}l^{+1}$ and B_o = biofilm COD per area at time zero (ML^{-2}).(Range of B_o observed was 0.5-1.0 µg COD cm^{-2}).

Series II

Details of CSTR 1 operation are given elsewhere (14) and are only summarized here in Table III.

Table III. Operational Results of CSTR 1, Series II Experiments.

Duration of CSTR 1 continuous operation	= 44 days
Dilution rate	= 0.33 h^{-1}
Inlet soluble COD concentration	= 640-850 mg COD l^{-1}
Effluent total COD concentration	= 390-400 mg COD l^{-1}
Effluent soluble COD concentration prior to dilution water to CSTR 2	= 40-50 mg COD l^{-1}
Dilution rate at culture "wash-out"	= 2.2 h^{-1}
Biomass yield (g biomass/g-COD)	= 0.42-0.56

Microorganisms present : Klebsiella oxytora, Klebsiella pneumoniae, Enterobacter cloace.

Operating conditions for CSTR 2 are given in Table IV. Inlet flow to CSTR 2 consists of dilution water, fresh sterile substrate, and CSTR 1 effluent. Primary substrate fed to CSTR 2 in all experiments is 10 mg l^{-1} trypticase soy broth and 10 mg l^{-1} glucose, after dilution.

Table IV. Operating Conditions of CSTR 2. The Biofilm Reactor.

	EXPERIMENT 1	2	3	4
CSTR 1 Effluent Suspended Biomass Concentration (mgCOD l^{-1})	370.	359.	368.	380.
CSTR 1 Effluent Delivered to CSTR 2 (cm^3 h^{-1})	1000.	1000.	1000.	200.
Measured Fresh[1] Inlet Substrate Concentration (mgCOD l^{-1})	22.7	22.9	33.0	23.8
Measured Inlet Suspended Biomass Concentration (mgTSS l^{-1})	19.5	18.9	19.4	4.0
(mgCOD l^{-1})	22.2	21.6	22.1	4.6
Reynolds Number	13000.	13000.	26000.	13000.
Mean Residence Time (h)	0.25	0.25	0.25	0.25

[1] Fresh substrate delivered to CSTR 2 consisted of 10 mg l^{-1} TSB and 10 mg l^{-1} glucose. Reported concentration is after dilution with CSTR 1 effluent and fresh dilution water.

Material Balances

Presentations of Series II experiments is facilitated by material balances for substrate and suspended biomass as well as a constitutive equation for biofilm accumulation:

Substrate: $$V\,dS/dt = F(S_i - S) - \mu VX/Y - R_g A/Y \quad (9)$$

Suspended Biomass: $$V\,dX/dt = F(X_i - X) + \mu XV + R_r A - R_d A \quad (10)$$

Biofilm: $$[\,dB/dt = R_g + R_d - R_r]\,A \quad (11)$$

where S = substrate concentration measured as COD (ML^{-3}), X = suspended biomass concentration measured as COD (ML^{-3}), B = attached biomass measured as COD (ML^{-2}), t = time (t), S_i = inlet substrate concentration measured as COD (ML^{-3}), V = reactor volume (L^3), A = reactor surface area (L^2), F = volumetric flow rate (L^3t^{-1}), μ = specific growth rate of suspended biomass (t^{-1}), Y = biomass yield measured as COD (MM^{-1}), R_g = net biofilm production rate due to metabolic processes measured as COD ($ML^{-2}\ t^{-1}$), R_d = deposition rate of suspended biomass measured as COD ($ML^{-2}t^{-1}$), R_r = detachment rate of biofilm measured as COD ($ML^{-2}t^{-1}$).

These material balances can be simplified with the following assumptions:

1. Rates of accumulation of S and X (i.e., dS/dt and dX/dt) are negligible and the system can be considered at steady state.

2. Although an increase in suspended cell numbers is unlikely at residence times of 0.25 h, increases in suspended biomass concentration may be significant. Consequently, suspended biomass growth in CSTR 2 is not ignored.

3. Substrate depletion rate by suspended biomass is also considered significant in CSTR 2 (see Assumption 2).

4. Net biofilm production rate is assumed the sum of biofilm production processes (i.e., growth of organisms and product formation) and maintenance energy requirements, i.e.,

$$R_g = (\mu_p - k_e) B \qquad (12)$$

where μ_p = specific biofilm production rate (t^{-1}), k_e = decay rate (t^{-1}).

Equation 12 tacitly assumes specific biofilm production and decay rates are first order in biofilm accumulation. Consequently, Equations 9,10, 11, and 12 reduce to the following:

$$F(S_i-S) = (\mu_p BA + \mu XV)/Y \qquad (13)$$

$$F(X_i-X) + (\mu XV) = R_d A - R_r A \qquad (14)$$

$$dB/dt = (\mu_p - k_e) B + R_d - R_r \qquad (15)$$

Determining Individual Process Rates

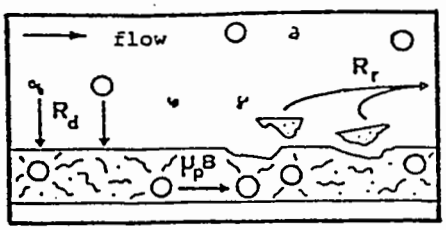
Equation 15 describes biofilm accumulation throughout an experiment as the sum of four processes: biofilm production ($\mu_p B$), biofilm mainenance decay ($k_e B$), suspended biomass deposition (R_d), and biofilm removal (R_r). However, analytical methods provide for direct measurement of only biofilm accumulation - e.g., B and dB/dt, and the decay rate, k_e.

Consequently, changes in CSTR 2 experimental conditions are made periodically to simplify Equation 15. These perturbations consist of depriving CSTR 2 of inlet substrate and/or inlet suspended biomass during four two-hour periods in each experiment. Figure 6 details these perturbations and their intended purpose. This technique allows estimation of the following:

1. Suspended biomass deposition rate (R_d) on the "clean" surface at time equal zero. In further calculations, R_d is assumed constant and independent of biofilm accumulation.

2. Equations 17 and 18 (Figure 6) can be used to calculate the biofilm removal rate, R_r, at each perturbation period knowing values of k_e, R_d and the slope of the biofilm accumulation curve (i.e., dB/dt).

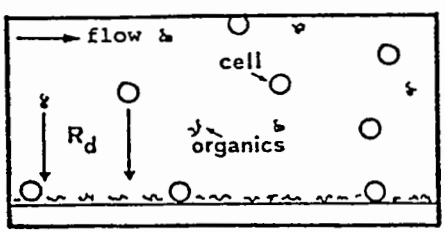
FIGURE 6. DEFINITION AND SIGNIFICANCE OF PERTURBATIONS TO CSTR 2.



NORMAL OPERATION

FRESH SUBSTRATE, SUSPENDED BIOMASS, AND DILUTION WATER SUPPLIED TO CSTR 2. BIOFILM NET ACCUMULATION DESCRIBED BY:

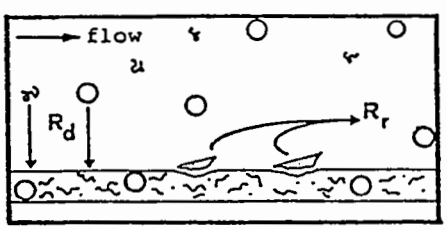
$$dB/dt = R_d + (\mu_p - k_e)B - R_r \quad (15)$$



PERIOD 1

ELAPSED TIME = 0-2 hours. NO SUBSTRATE TO CSTR 2, ONLY DILUTION WATER AND SUSPENDED BIOMASS. EQUATION (15) REDUCES TO:

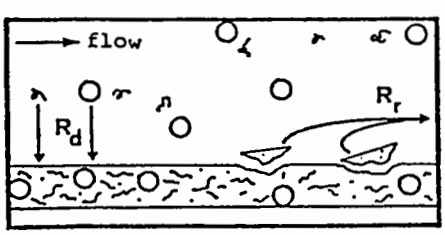
$$dB/dt = R_d \quad (16)$$



PERIOD 2

ELAPSED TIME = 18-20 hours. NO SUBSTRATE TO CSTR 2, ONLY DILUTION WATER AND SUSPENDED BIOMASS. EQUATION (15) REDUCES TO:

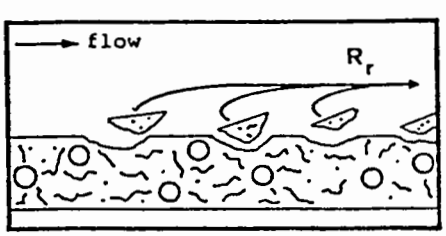
$$dB/dt = R_d - k_e B - R_r \quad (17)$$



PERIOD 3

ELAPSED TIME = 40-42 hours. SAME CONDITIONS AS PERIOD 2 EXCEPT BIOFILM ACCUMULATION IS GREATER.

$$dB/dt = R_d - k_e B - R_r \quad (17)$$



PERIOD 4

ELAPSED TIME = 50-52 hours. NEITHER SUBSTRATE NOR SUSPENDED BIOMASS TO CSTR 2, ONLY DILUTION WATER. EQUATION (15) BECOMES:

$$dB/dt = -R_r - k_e B \quad (18)$$

Figure 7 indicates biofilm COD accumulation including perturbed and non-perturbed intervals for a typical Series II experiment. Biofilm accumulation, dB/dt, during the perturbation periods only are presented, for all experiments, in Table V. Values of the biofilm decay rate, k_e, determined via respirometer measurements (14), are also given in Table V.

Calculations of R_r in Periods 2-4, from data in Table V and Equations 17 and 18, are summarized in Table VI. Figure 8 illustrates resultant R_r values versus the average biofilm COD present during the perturbation. Data from Trulear and Characklis (13), obtained from an annular rotating reactor, are also included and indicate a similar magnitude of biofilm removal rates.

Specific biofilm production rates, μ_p, throughout the unperturbed portion of each experiment, can be determined using Equation 15 and the values R_d, k_e, and R_r above. Resultant μ_p values as a function of biofilm COD are shown in Figure 9.

DISCUSSION

Deposition

Rate of deposition, R_d, was considered constant throughout any experiment at the value of dB/dt determined during Period 1 (ref., eq. 16, Figure 6). This assumption provides an estimate of deposition rate at "clean" surface conditions and most likely underestimates the enhanced effect a fouled surface would have on particle deposition at later stages of biofilm development.

Adsorption of organic molecules (e.g., polysaccharides and/or glycoproteins) can contribute to the total amount deposited (and rate of deposition) as detected by COD analysis. However, this adsorption occurs within minutes of exposure (17) and the maximum amount of adherent material due to organic adsorption in this system is estimated ≤ 0.01 μgCOD cm^{-1}. Consequently, rates of organic adsorption are assumed instantaneous and independent of Reynolds number (Re) and suspended biomass concentration (X).

Mass flux of particles, suspended in a turbulent flow field, across a boundary layer is directly proportional to the bulk fluid concentration of particles (18, 19, 20).

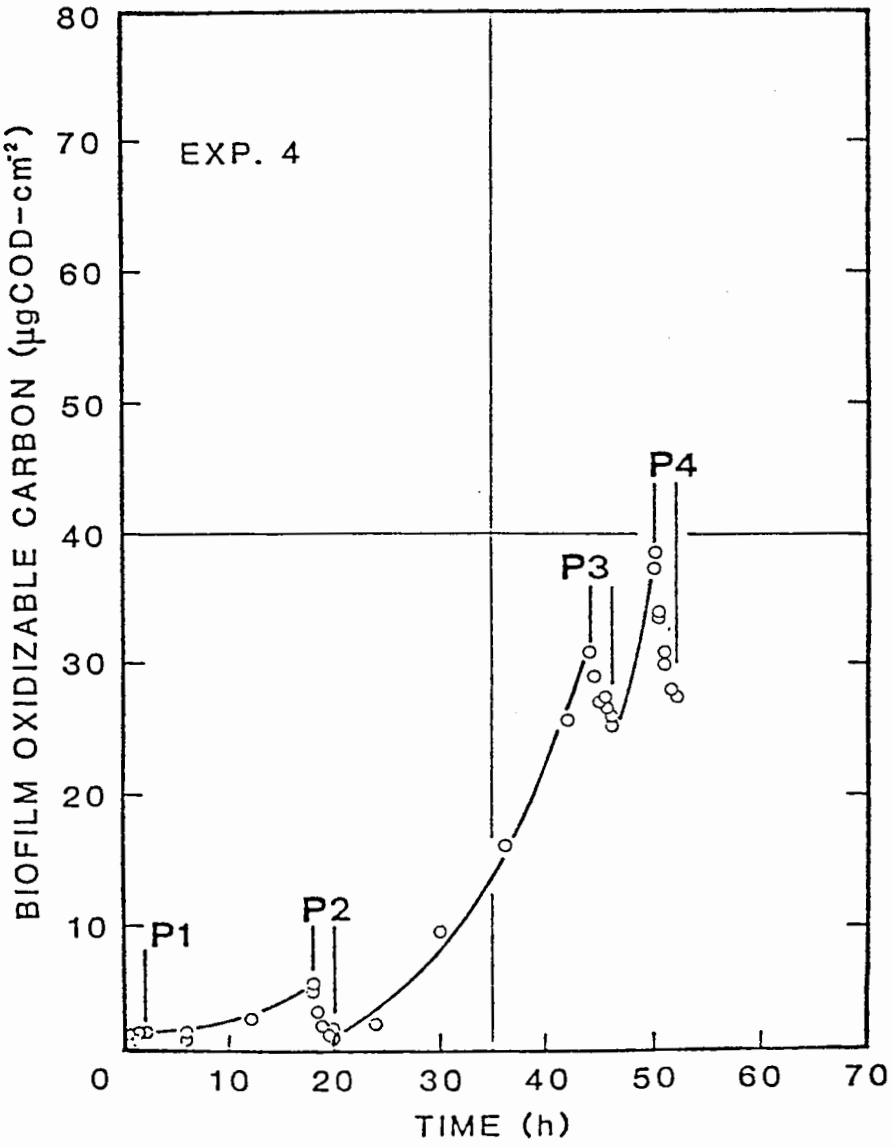


FIGURE 7. BIOFILM NET COD ACCUMULATION DURING EXPERI-4 INDICATING BOTH NORMAL GROWTH AND FOUR PERTURBATION PERIODS.

Table V. Biofilm Accumulation during CSTR 2 Perturbations.

	$B^{(1)}$ ($\mu gCOD\ cm^{-2}$)	$dB/dt^{(2)}$ ($\mu gCOD\ cm^{-2} h^{-1}$)	$k_e B$ ($\mu gCOD\ cm^{-2} h^{-1}$)
Experiment One			
Period 1	1.2	1.2	ND
2	ND	ND	ND
3	63.5	8.0	0.38
4	48.3	-5.3	ND
Experiment Two			
Period 1	1.1	1.1	NKD
2	7.3	-2.4	ND
3	38.8	-2.9	0.23
4	57.5	-2.5	ND
Experiment Three			
Period 1	0.8	0.8	ND
2	2.5	-0.4	ND
3	38.5	0	0.22
4	61.0	-10.8	ND
Experiment Four			
Period 1	0.3	0.3	ND
2	2.8	-2.2	ND
3	28.0	-3.0	0.17
4	33.0	-6.0	ND

(1) average biofilm during perturbation only. (2) accumulation during perturbation period

Table VI. Summary of Biofilm Removal Rate Calculations.

	B_V (μgCOD cm^{-2})	dB/dt	(μgCOD cm^{-2}h^{-1}) R_d	$k_e B$	R_r
			(a)	(b)	(c)
Experiment 1					
Period 1	1.8	+1.2	1.2	(d)	0
2	ND	ND	1.2	ND	ND
3	63.5	+8.0	1.2	0.4	-7.2
4	48.3	-5.3	(e)	0.3	5.0
Experiment 2					
Period 1	1.1	+1.1	1.1	(d)	0
2	7.3	-2.4	1.1	0.04	3.5
3	38.8	-2.9	1.1	0.2	3.8
4	57.5	-1.4	(e)	0.3	2.2
Experiment 3					
Period 1	0.8	0.8	0.8	(d)	0
2	2.5	-0.4	0.8	0.2	1.2
3	38.5	0	0.8	0.2	0.6
4	61.0	-10.8	(e)	0.4	10.4
Experiment 4					
Period 1	0.3	0.3	0.3	(d)	0
2	2.8	-2.2	0.3	0.02	2.5
3	28.0	-3.0	0.3	0.2	3.1
4	33.0	-6.0	(e)	0.2	5.8

ND = not determined; (a) = deposition rate assumed constant at value of dB/dt determined in Period 1; (b) = biofilm specific decay constant k_e= 0.006 h^{-1}for all experiments; (c) calculated from Equations 17 or 18(see Figure 6); (d) assumed zero during Period 1; (e) assumed zero during Period 4.

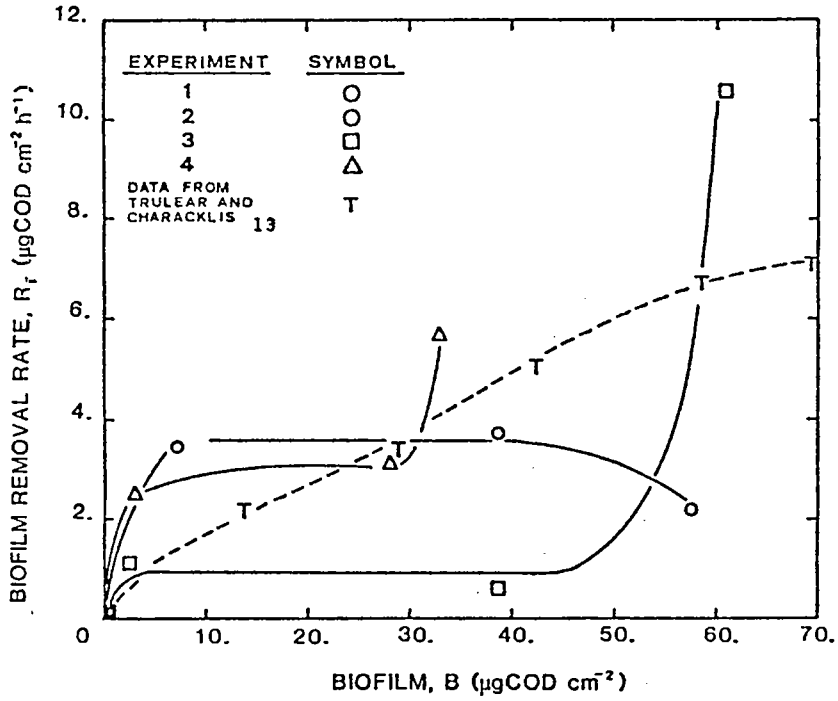


FIGURE 8. BIOFILM REMOVAL RATE, R_r, AS A FUNCTION OF BIOFILM COD.

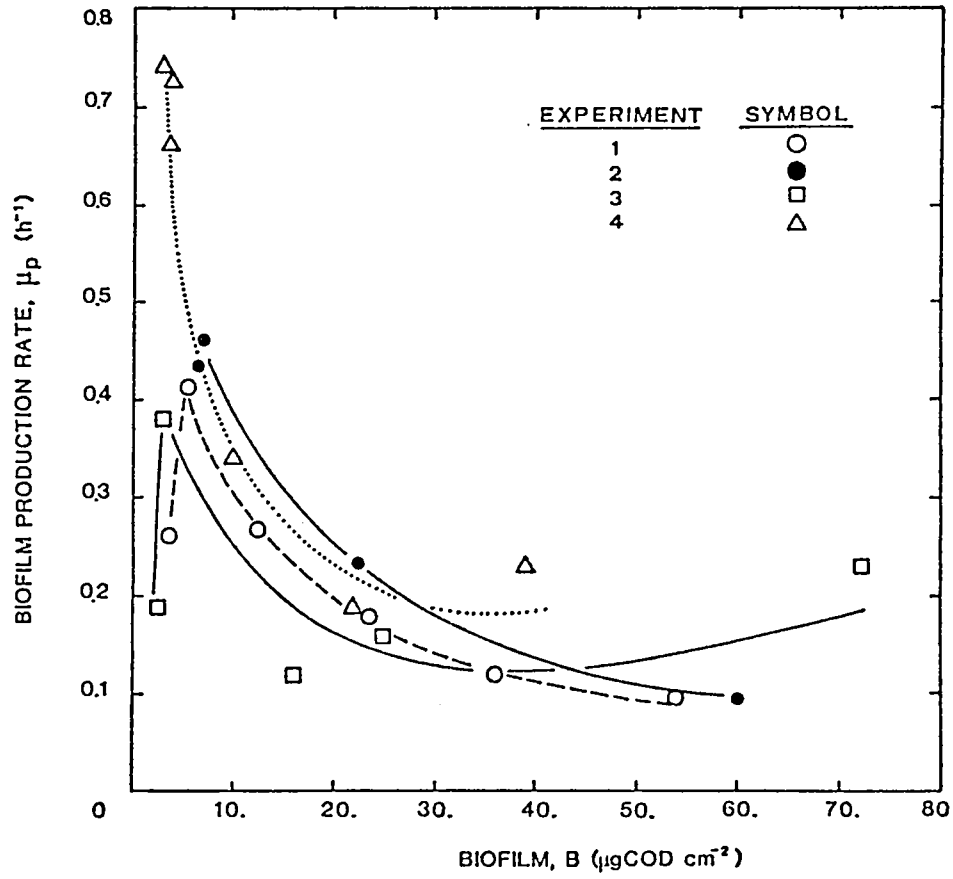


FIGURE 9. BIOFILM SPECIFIC PRODUCTION RATE, μ_p, AS A FUNCTION OF BIOFILM COD.

A reduction in suspended biomass concentration from 19.5 to 4.0 mgTSS-l^{-1}(Experiment 1 and 4, respectively) does result in a proportional decrease of 1.2 to 0.3 μgCOD $cm^{-2}h^{-1}$in deposition rate, R_d(ref.Table VI, Period 1 data).

The effect of changing fluid flow regime on particle transport is a complicated function of fluid velocity, fluid properties, particle size and particle physical properties. Increasing fluid velocity can have the following two effects:

1. Increased turbulence may increase or decrease the mass transfer coefficient depending on characteristics of the suspended particle (19).
2. Increased turbulence may decrease the boundary layer thickness and, thus, increase transport to the surface.

Suspended biomass generally has specific gravity less than 1.1 and suspended biomass aggregates in these experiments measured 3.0 - 5.0 μm in equivalent diameter. Consequently, the suspended biomass particles are assumed uniformily distributed and concentration gradients did not exist in the bulk fluid. Table VI, Period 1 data shows that deposition rate R_d decreases only slightly with a doubling in recycle Reynolds number. In experiments at Re = 13000 (0.8 m/sec), deposition rate was 1.1 - 1.2 μgCOD $cm^{-2}h^{-1}$ while at Re = 26000 (1.6 m/sec), R_d was 0.8 μgCOD $cm^{-2}h^{-1}$, yet Beal (19) predicts an increase in tranport rate. This discrepancy arises since deposition rate is the sum of both the particle transport rate and microbial cell adhesion rate. Therefore, while the particle transport rate may be increasing with Re, the deposition rate (the measured parameter) may not; suggesting that cell "sticking efficiency" is changing with Reynolds number.

Biofilm production

Figure 9 gives values of μp throughout each experiment as determined from Equation 15. In all cases, μp asymptotically decreases with increasing biofilm COD to the same value, 0.1-0.2 h^{-1}. Wide variations in μ_p initially may result from errors in biofilm measurements at very low levels or changes

in cellular metabolism upon attachment. Decreases in μp with increasing biofilm COD could result from changes in cell metabolism or increasing internal resistance to substrate mass transfer within biofilm.

Instantaneous yield values in all experiments were calculated from the substrate material balance, Equation 15 or upon rearrangement:

$$Y = (\mu_p BA + \mu XV)/F(S_i - S) \qquad (19)$$

which tacitly assumes that yield coefficients for attached and suspended growth are the same. A summary of yield calculations is given in Table VII and suggests an average yield of approximately 0.5 mg COD biomass per mg COD removed. This value compares favorably with those obtained by Stathopoulos (21) for similar experimental systems.

Biofilm Decay

The specific biofilm decay rate is 0.006 h^{-1} and corresponds to values reported by Lawrence and McCary (22) for suspended biomass (0.0019-0.22 h^{-1}) and Stathopoulos (21) for biofilm experiments at temperatures ranging from 15-60°C (0.04-0.22 h^{-1}).

Biofilm Removal

Biofilm removal rates, due to existing shear stresses, are shown in Figure 8 as a function of biofilm COD. Over the range of biofilm COD observed, biofilm removal rates, R_r, were less than 5 μgCOD $cm^{-2}h^{-1}$ and appeared independent of both Reynolds number and suspended biomass concentration. This is true except for the R_r value determined in Period 4 of Experiment 3; at Re = 26000 the removal rate suddenly increases from 1.0 μgCOD $cm^{-2}h^{-1}$ at a biofilm COD = 45 μgCOD cm^{-2} to 10.8 μgCOD cm^{-2} h^{-1} at a biofilm COD = 61 μgCOD cm^{-1}.

This increase in removal rate can be explained by considering the changes in hydrodynamic conditions that occur between these two levels of biofilm COD. Biofilm levels of 45 and 61 μgCOD cm^{-2} correspond approximately to biofilm

Table VII. Summary of Calculations for Yield Coefficients.

time (h)	$F(S_i-S)$ (mgCOD h^{-1})	B (mgCOD cm^{-2})	μ_p (1) (h^{-1})	$\mu_p BA$ (mgCOD h^{-1})	X (mgCOD l^{-1})	μXV (2) (mgCOD h^{-1})	Y (3) (mgCOD/mgCOD)
Experiment 1							
0	52.7	.001	.70	0.5	18.7	29.3	.57
10	79.6	.005	.40	11.8	24.0	37.6	.62
34	150.6	.019	.20	22.5	28.9	45.3	.45
50	149.6	.035	.13	27.0	26.5	41.5	.46
Experiment 2							
0	0	.001	.10	0.5	18.9	29.6	--
2	96.9	.003	.25	4.4	25.5	39.9	.46
6	101.1	.005	.47	13.9	24.	37.6	.51
20	103.4	.007	.45	18.7	22.5	35.3	.52
46	208.4	.030	.18	32.0	25.0	39.2	.34
Experiment 3							
6	81.6	.005	.35	10.4	22.7	34.6	.55
18	81.6	.005	.35	10.4	22.7	34.6	.55
24	84.0	.006	.32	11.6	22.1	34.6	.55
30	96.2	.011	.25	16.3	22.0	34.5	.53
50	106.4	.055	.16	52.2	22.5	35.3	.42
Experiment 4							
6	14.3	.001	.75	4.5	3.0	4.7	.64
20	31.4	.002	.70	8.3	4.2	6.6	.47
44	110.5	.020	.24	28.5	12.0	18.8	.43
50	106.8	.035	.24	49.8	12.0	18.8	.64
51	89.9	.033	.24	46.9	7.0	11.0	.64

(1) biofilm production rate constant taken from Figure 9 at specific biofilm COD.
(2) Suspended biomass growth rate, μ, assumed value = 0.33 h^{-1}. (3) From Eqn. 19.

thickness of 39.5 and 53.0 μm, respectively (where 1 mg biofilm = 1.74 mg biofilm COD and biofilm density = 10.0 mg biofilm cm^{-3}) (14). A viscous sublayer thickness, δ, of 44 μm can be calculated as follows (15):

$$\delta = 25\ d\ (Re)^{-.875} \qquad (20)$$

with d = pipe diameter (1.27×10^4 μm) and Re = 26000. This calculation indicates the biofilm thickness, just prior to Period 4 of Experiment 3, exceeded the viscous sublayer, therefore, increasing the system friction factor and, consequently, the shear stresses at the biofilm-fluid interface. This increase in shear stress could result in the dramatic increase in biofilm removal rate. Viscous sublayer thickness at Re = 13000 is 80 μm and biofilm thicknesses in the three expiriments at Re = 13000 never exceeded 55 μm; consequently, no radical increase in biofilm removal rate is expected and none are observed.

During Period 4 in all cases, suspended biomass from CSTR 1 is not supplied to CSTR 2. Therfore, after this two hour period (eight CSTR 2 reactor residence times) any suspended biomass leaving the system must originate as biofilm. Consequently, the rate of any suspended biomass leaving CSTR 2 after this perturbation is considered equal to the rate of biofilm removal - i.e.,

$$R_r = FX/A \qquad (21)$$

Table VIII indicates the biofilm removal rates determined from biofilm COD (Equation 18) are somewhat less, but of the same order, as values determined from the suspended biomass material balance, Equation 21.

Table VIII. Comparison of Biofilm Removal Rate Estimates

Exp.No.	Re	CSTR 2 Effluent Biomass (a) (μgCOD l^{-1})	R_r (Eq.21)[b] (μgCOD cm^{-2} h^{-1})	R_r(Eq.18) (μgCOD cm^{-2} h^{-1})
3	26000	5358	16.3	10.8
4	13000	1938	5.9	5.8

(a) determined after eight residence times from start of Period 4.

(b) evaluated using F = 18 l-hr^{-1} and A = 5934 cm^2

SUMMARY

Series I Experiments.

Results in Series I experiments provide the following information:

1. the rate of biofilm COD accumulation during its early formation stages was described mathematically using a first-order rate expression. The resultant first-order rate constant was a linear function of suspended biomass concentration and growth rate, and Reynolds number

Series II Experiments.

Results of Series II work, for the thin aerobic biofilms investigated, show the following:

1. Although particle deposition contributes significantly to the initiation of biofilm development, its relative role in biofilm net accumulation decreases with time.

2. Biofilm decay due to mainentance requirement is insignificant for the thin biofilms considered.

3. Biofilm production and shear removal processes contribute significantly to early biofilm accumulation. Shear removal rates drastically increase as the film exceeds the viscous sublayer.

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REFERENCES

1. Atkinson, B. and Fowler, H.W. "The Significance of Microbial Films in Fermenters" Chap 6. *Advances in Biochemical Engineering, Vol 3.* Ghose, T.K. *et.al.* (Eds). Springer-Verlag. NY., 1974, pp 221-277.
2. Cooper, P.F. and Atkinson, B. (Eds). Biological Fluidised Bed Treatment of Water and Wastewater Ellis Horwood Limited, Chichester, UK. 1981.
3. Smith, E.D. *et.al.*, (Ed). Proceedings First National Symposium on Rotating Biological Contactor Technology. Champion, Pennsylvania, February 4-6, 1980.
4. Harremöes, P. "Biofilm Kinetics". Chap. 4. Water Pollution Microbiology, Vol. 2., Wiley Interscience (Ralph Mitchell, Ed.), 1976.
5. Riemer, M.W. "Kinetics of Denitrification in Submerged Filters. Part I" Ph.D. Dissertation. Technical University of Denmark, 1978.
6. Levenspiel, O. Chemical Reaction Engineering, Wiley and Sons, New York, 2nd Ed. 1972.
7. La Motta, E. "Evaluation of Diffusional Resistances in Substrate Utilization by Biological Films" Ph.D. Dissertation, Univ.of North Carolina, Chapel Hill, 1974.

8. Harremöes, P. "Half order Reactions in Biofilm and Filter Kinetics" *Vatten 2* 1977, p 22.
9. Williamson, K.J. and McCarty, P.L. "A Model of Substrate Utilization by Bacterial Films", *J.W.P.C.F. 48 (1),* 1976, pp. 9-24.
10. Willimason, K.J. and McCarty, P.L. "Verification Studies of the Biofilm Model for Bacterial Substrate Utilization", *J.W.P.C.F. 48 (2)* 1976, pp 281-296.
11. Rittmann, B.E. and McCarty, P.L. "Evaluation of Steady-State-Biofilm Kinetics", *Biotechn. Bioengr. 22.* 1980, pp 2359-2373.
13. Trulear, M.G. and Characklis, W.G. "Dynamics of Biofilm Processes" in *34th Annual Purdue Industrial Waste Conference, West Lafayette, Indiana, May 8-10, 1979 (Ann Arbor Publishers, Ann Arbor, MI, 1979).*
14. Bryers, J.D. "Dynamics of Early Biofilm Formation", Ph.D. Dissertation, Rice University, Houston, Tx. 1980.
15. Characklis, W.G. "Bioengineering Report: Fouling Biofilm Developement - A Process Analysis" *Biotechn. Bioengr. 23* 1981, pp. 1923-1960.
16. Bryers, J.D. and Characklis, W.G. "Early Fouling Biofilm Formation in a Turbulent Flow System: Overall Kinetics." *Wat.Res.., 15* 1981, pp. 483-491.
17. Baier, R.E. "Influence of the Initial Surface Conditions of Materials on Broadhesion" *Proc. 3rd Initial Congr. Marine Corrosion and Fouling.* Nat'l Bureau of Standards, Gaithersburg, Maryland, Oct. 2-6, 1972.
18. Freidlander, S.K. and Johnstone, H.F. "Deposition of suspended particles from turbulent gas streams". *Ind. & Engr. Chem. 49 (7),* 1970, pp. 1-11.
19. Beal, S.K. "Deposition of particles in turbulent flow on channel and pipe walls". *Nucl.Sci. and Eng. 40* 1970 pp. 1-11.
20. Browne, L.W.B. "Deposition of particles on rough surfaces during turbulent gas flow in a pipe". *Atm.Envir. 8,* 1974, pp. 801-815.
21. Stathopoulous, N. "Influence of Temperature on Biofilm Processes", M.S. Thesis, Rice University, Houston, Tx 1981.
22. Lawrence, A.W. and McCarty, P.L. "A unified basis for biological treatment design and operation". *J.Sanit Eng. Div.,* ASCE, 96, 1970, SA3.

THE MICROBIOLOGY OF ROTATING BIOLOGICAL CONTACTOR FILMS

Nancy E. Kinner. Department of Civil Engineering, University of New Hampshire, Durham, New Hampshire.

David L. Balkwill. Department of Microbiology, University of New Hampshire, Durham, New Hampshire.

Paul L. Bishop. Department of Civil Engineering, University of New Hampshire, Durham, New Hampshire.

INTRODUCTION

The treatment of municipal and industrial wastewater generated by modern society is rapidly becoming an intractable problem. The continuing demand for a pollutant-free environment (1) is exceeding the ability of traditional waste treatment processes to produce high quality effluents at reasonable costs. A recent GAO study reported that, of the 242 wastewater treatment facilities examined, 87 percent were in violation of their NPDES permits at least one month of the year, with 56 percent being in violation more than half of the year (2). Many of the violations resulted from the municipality's inability to afford the high operation and maintenance costs (3). Consequently, economical and innovative wastewater treatment techniques are needed immediately to meet the legal and public demand for water pollution control. The rotating biological contactor (RBC), a relatively new technique for aerobic biological wastewater treatment, offers a cost effective solution to this demand with the advantages of low energy and maintenance requirements, high organic removal efficiencies at short retention

times, modular flexibility in design, and adaptability to a wide range of wastewater types and flows.

Until recently, most of the research on RBCs has been conducted using traditional engineering methods in an effort to determine their overall organic removal efficiency and design parameters during the treatment of various kinds of wastes (4,5,6,7,8,9,10,11). Design equations, based on hydraulic [m^3 applied/$m^2 \cdot d$] (12,13) and organic [gms organic matter as BOD or COD applied/$m^2 \cdot d$] (14,15) loading rates, have employed general empirical relationships and large, conservative safety factors. With the increasing demand for cost effective designs, optimization of RBCs has become important. To optimize organic removal, one must understand the interactions between RBC microorganisms and their physico-chemical environment. This results from the fact that the RBC process is a product of the microbial ecosystem which operates within its confines. As a first step towards understanding the microbial interactions which occur in the RBC it is necessary to have a general knowledge of 1) the microorganisms present, 2) their relative abundance, and 3) their ultrastructural characteristics which may be indicative of their physiological state.

The bacteria inhabiting RBCs during secondary wastewater treatment have not been thoroughly examined. Most published research which contains a description of the biofilm constituents provides it as supplementary information. No examination of the microflora of the suspended flocs has been conducted, though Kincannon and Groves (16) assert that they can play a major role in organic removal. Information on the biofilm has been collected by observing its gross morphology and by examining wet mount slides.

The RBC biofilm is usually characterized as shaggy and filamentous (5,17). The effects of compartmentalization, however, are apparent; the biofilm's color and density varies along the length of the unit. When treating municipal or artificial sewage the first compartment usually contains a thick, white to gray growth (7,12,18) which grades to a dark, brown-black and thinner biofilm in the final compartments (7,18,19). The sparse growth is attributed to protozoan predation and low organic concentrations. These characteristics may differ when industrial wastes are being treated (6).

The first attempt at a complete categorization of the biofilm constituents was made by Antonie and Welch (20) as

part of a study of RBCs during dairy waste treatment. Several microorganisms were identified (Table I). They concluded that the most important species were the filaments *Geotrichium candidum* and *Bacillus cereus*, and the nonfilamentous bacteria *Zoogloea filipendula, Pseudomonas denitrificans, Aerobacter aerogenes,* and *Escherichia coli*. Unfortunately, the authors did not discuss the techniques used to isolate and identify the bacteria to the species level nor did they mention their relative numbers and distribution within the RBC.

Table I

Organisms Identified in an RBC Biomass (20)

Predominant Organisms	Non-Predominant Organisms
Zoogloea filipendula	*Pseudomonas fluorescens*
Pseudomonas denitrificans	*Pseudomonas aeruginosa*
Aerobacter aerogenes	*Neisseria catarrhalis*
Escherichia coli	*Geotrichium candidum*
Escherichia freundii Type I	*Torula spp.*
Escherichia spp.	*Rhodotorula spp.*
Bacillus cereus var. mycoides	
Bacillus cereus	
Micrococcus conglomeratus	
Micrococcus luteus	

Several authors have described the indigenous biofilm populations inhabiting properly loaded RBCs treating municipal or artificial sewage. They have identified the bacteria present by examining wet mount slides of the biofilm. In the first compartments, the most commonly observed filamentous bacterium is *Sphaerotilus* (21,22,23,24,25,26). *Beggiatoa* (22,23,27), *Fusarium* (26), *Nocardia* (25), *Cladothrix* (23), and *Oscillatoria* (26) are found less frequently. Nonfilamentous forms observed in the first compartments are *Zoogloea* and zoogloeal masses (21,23); unicellular algae (26); and unicellular rods, spirilla, and spirochaetes (27). The final compartments contain most of the same forms as well as *Streptomyces* (27) and *Athrobotrys* (22). Protozoan populations have been characterized microscopically, but will not be discussed in this paper.

In this research traditional light microscopy, interfence optics, and transmission electron microscopy were used

to examine the biofilm constituents of the first compartments of RBCs treating domestic wastewater. Two different types of RBC pilot plants were studied. The smaller unit consisted of a single compartment which had 18 cm diameter, polyurethane coated Masonite^{R} disks. The larger unit had four equally sized compartments each of which contained a section of 0.5 m diameter corrugated plastic disk media. Both RBCs were loaded at 0.04 $m^3/m^2 \cdot d$; typically hydraulic loading rates for RBCs vary from 0.04 to 0.08 $m^3/m^2 \cdot d$ (17). Biofilm from the first compartment of each unit was examined after steady state operation was achieved. Some staining was done for light and transmission electron microscopy. Filaments were isolated on special microbiological media. Particular attention was directed to determining 1) the identity of the predominant filaments, 2) the morphological characteristics of single-celled bacteria present, and 3) the ultra-structural characteristics of the bacteria as a possible indicator of their physiological and ecological conditions.

MATERIALS AND METHODS

RBC Pilot Plant Descriptions

One laboratory-scale RBC unit was operated under a fume hood in an environmental engineering laboratory. It had one compartment constructed from an acrylic half cylinder 30 cm long and 20 cm in diameter. A horizontal stainless steel shaft supported 16 disks, each with an 18 cm diameter, for a total wetted surface area of 0.78 m^2. The equally spaced disks were made of Masonite^{R} sealed with polyurethane. Effluent flowed from the RBC to an ajoining basin through each of four 1.3 cm diameter ports located at the base of the end wall and over a notched weir located at the top of the end wall. A peripheral disk velocity of 0.31 m/s was maintained by a mechanical drive. The unit was exposed to a low level fluorescent light of less than 100 lm/m^2 for a maximum of 12 hours per day. Ambient air and wastewater temperatures were 20°C. All disks were approximately 40 percent submerged in wastewater at any given time.

A second RBC pilot plant was housed in a laboratory trailer located at the Durham, New Hampshire wastewater treatment plant. It was a 0.5 m diameter, 4 compartment Bio-

Surf[1] unit with corrugated polyethylene disk media. Influent was delivered to a wet well and was distributed to the first compartment by four rotating scoops. Wastewater flowed from one compartment to the next through each of two 2.6 cm diameter ports located in the baffle walls. Effluent passed out of the fourth compartment via an overflow pipe. The peripheral velocity was 0.31 m/s (mechanical drive) and the submergence level was 40 percent. The unit was exposed to no longer than 10 hours of natural light per day. Ambient air temperature was maintained at 20°C; wastewater temperature was no less than 17°C.

Raw sewage, the influent for the 18 cm RBC, was obtained in 20 l carboys from the Durham sewage pumping station. The carboys were stored at 4°C until used (a maximum holding time of 3 days). During this period solids settling occurred. The settled sewage was transported from the carboys to the small RBC via a peristaltic pump set to deliver 67 l/d. This flow was sufficient to operate the unit at an hydraulic loading rate of 0.04 $m^3/m^2 \cdot d$ and an organic loading rate averaging 3.2g $TOC/m^2 \cdot d$.

The 0.5 m diameter RBC received 0.95 m^3 of fresh primary effluent from the Durham treatment plant per day. This was pumped continuously to the wet well of the RBC achieving an overall hydraulic loading rate of 0.04 $m^3/m^2 \cdot d$ and an organic loading rate averaging 3.2g $TOC/m^2 \cdot d$. The hydraulic loading to the first compartment was 0.16 $m^3/m^2 \cdot d$ and the organic loading rate averaged 12.8g $TOC/m^2 \cdot d$.

After a three week start-up period both RBCs had achieved steady state operation as determined by obtaining similar effluent total organic carbon (TOC) concentrations on three consecutive days. TOC measurements were performed on the RBC influent and effluent samples after filtration through Whatman #40 paper, according to the ampule method outlined for the Oceanography International[2] Model 526. The RBC influent wastewater: settled raw sewage and the fresh primary effluent, had average TOC's of 80 mg C/l. The effluent concentration from the 18 cm diameter and 0.5 m diameter RBCs were 17.5

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[1] Autotrol Corporation; Milwaukee, Wisconsin.

[2] Oceanography International Corporation; College Station, Texas.

and 23.0 mg C/l, respectively. Microbial samples of the 18 cm diameter RBC biofilm for both light and electron microscopy were randomly scraped from the surface of the first disk after steady state was achieved. Biofilm from the 0.5 m diameter unit was randomly scraped from the front, middle and end surfaces of the disk media in the first compartment.

Light Microscopy

The biofilm removed from the RBC disks was too dense to examine directly. To prepare samples for light microscopy the biofilm was rinsed in several petri dishes containing Nannopure^{R} water and then repeatedly drawn up into a Pasteur capillary pipette to separate the densely tangled mass. Several wet mount slides of each washed sample were examined under a Nikon Biophot Research Microscope equipped with Nomarski differential interference bright field optics. A photographic record of observations was made using Panatomic X (ASA 32) film.

Pieces of each sample of rinsed biofilm were then run through another series of four rinses in Nannopure^{R} water. These were further separated by the Pasteur pipette technique and by micromanipulation. Most of the constituents were removed from the sample by these procedures except for the filaments and zoogloeal masses. Staining methods were employed to determine the presence of poly-β-hydroxybutyrate (PHB) and ferric iron (Fe^{+3}) in these samples.

Burdon's method, as outlined in the Manual of Microbiological Methods (?8), was used to determine if the microorganisms contained PHB. After staining with 0.3% alcoholic Sudan Black B and counterstaining with 0.5% aqueous Safranin, PHB appeared blue-black while the rest of the cell was pink.

Ferric iron on the filaments was reacted with 0.1% aqueous potassium ferrocyanide, under acidic conditions, to produce the Prussian blue reaction (29). Special care was taken to insure that soluble ferric iron in the biofilm was removed by washing these samples in extra Nannopure^{R} water rinses.

Color photographs were taken of the samples after staining procedures were performed. An Olympus BHA microscope was used with Kodachrome ASA/25 and ASA/64 film.

Isolation Experiments

The isolation techniques developed by Dondero, Phillips, and Heukelekian (30) for *Sphaerotilus* were followed. Biofilm washed in four rinses of Nannopure^{R} water and teased apart using the Pasteur pipette technique was placed in a blender, containing 50 ml of Nannopure^{R} water, for 30 seconds. The homogenate was streaked on petri dishes of CGY and CG agar media and incubated for 48 hours at 28°C. After incubation the plates were observed under a dissecting microscope. Tangled, curled, filamentous growth suspected of being *Sphaerotilus*, was reisolated on fresh plates of the media and incubated for 48 hours at 28°C. The filamentous growth formed after reisolation was observed using the Olympus microscope and the PHB test was performed according to the procedures described above. Color photographs were taken of these samples.

Phototatic Experiments

To test the response of the biofilm constituents to light, a series of phototactic experiments were conducted. These procedures were recommended by Dr. Jane Gibson of Cornell University (31). Extract agar plates were prepared by adding 2 gm of agar to 1 liter of filtered (Whatman #40 paper) mixed liquor from the first compartments of the RBCs. Plates were poured after the medium was sterilized for 20 minutes at 15 psi.

Six plates of the RBC extract media were streaked with rinsed biofilm samples. Three plates were incubated in the dark; three plates were incubated in continuous light which was provided by two fluorescent lights. All incubations were at 25°C for one week.

Biofilm samples removed directly from the RBC were teased apart and placed on one side of each of nine plates containing RBC extract media. Six of the plates were then covered with aluminum foil. Three of these plates had a 1 mm diameter hole placed in the foil on the side opposite the sample. The pinhole provided a fixed light source to which photosynthetic organisms would migrate. All of the plates were placed in a 25°C incubator, which was continuously illuminated by two fluorescent lights, for one week.

Electron Microscopy

All biofilm specimens were prepared for electron microscopy by the thin sectioning technique. Two fixation procedures were used to prepare each sample for thin sectioning. For the Kellenberger fixation, pieces of biofilm material were suspended in Kellenberger buffer (32) and sufficient 1% OsO_4 (in Kellenberger buffer) was added to bring the final concentration of OsO_4 to 0.1%. The samples were prefixed in this suspension for 30 minutes at room temperature, after which they were concentrated and washed by centrifugation in Kellenberger buffer. The resulting pellet was resuspended in 2-3 drops tryptone-salt solution (1% tryptone, 0.5% NaCl) and mixed with approximately 0.5 ml molten 2% Difco Noble Agar[1] at 50°C. The agar-specimen mixture was then transferred to a glass slide, allowed to solidify, and cut into small blocks (less than 1mm on a side). These blocks were postfixed 12-18 h at room temperature in 1% OsO_4 (in Kellenberger buffer) and prestained 2 h at room temperature in 0.5% uranyl acetate (in Kellenberger buffer). For the glutaraldehyde-osmium tetroxide fixation, pieces of biofilm material were suspended in 0.1 M sodium cacodylate buffer (pH 7.5) and sufficient 12.5% glutaraldehyde (in 0.1 M sodium cacodylate buffer) was added to bring the final concentration of glutaraldehyde to 3%. Following prefixation in this suspension for 2 h at room temperature, the specimens were concentrated and washed twice by centrifugation in sodium cacodylate buffer. The final pellet was resuspended in tryptone-salt solution and embedded in agar as described above. The resulting blocks of agar were then postfixed 12-18 h at room temperature in 1% OsO_4 (in 0.1 M sodium cacodylate buffer).

Samples from both fixations were dehydrated through a graded ethanol series and then embedded in Spurr's low-viscosity epoxy resin (33). Thin sections were cut on an LKB Ultratome III ultramicrotome, using glass knives or a Diatome diamond knife. The sections were retrieved on uncoated, 400-mesh copper specimen grids, after which they were poststained 15 minutes with 0.5% uranyl acetate (in 50% methanol) and 2 minutes with 0.4% lead citrate (34).

Thin sections were viewed with a JEOL JEM-100S transmission electron microscope at an accelerating potential of 80 kV. The specimens were examined and photographed extensively in

[1]Difco Laboratories; Detroit, Michigan.

order to ensure that a representative sampling of microbial cells was obtained. Comparisons were also made with light microscopical observations (above) for this purpose. Both fixation procedures used for thin sectioning gave equivalent results. Micrographs of samples prepared with the glutaraldehyde-osmium tetroxide fixation were chosen for purposes of illustration in this study.

RESULTS

Gross Morphology

The biofilm on the first disk in the 18 cm diameter RBC and in the first compartment of the 0.5 m diameter RBC was gray-brown and filamentous with a subsurface black layer. Growth was fairly uniform; maximum film thickness was 1 mm. Sloughing occurred randomly and recolonization was immediate. On the terminal disks in the 18 cm RBC and in the last compartments of the 0.5 m RBC the biofilm was thinner and dark brown appearing mottled because recolonization occurred more slowly. These RBC biofilm characteristics were similar to those observed previously during domestic wastewater treatment (7,12,18,19).

Light Microscopy

Biofilm from both of the RBC pilot plants was extremely dense forming an interwoven mat. It was composed of two principal constituents: filaments and single-celled bacteria grouped together in amorphous clumps. The biofilm constituents were similar to ones previously observed in RBC biofilms in this laboratory (35). The filaments consisted of a series of sausage-shaped cells, approximately 1-2 μm in diameter and 2-5 μm long, which were tightly encased in an outer sheath. The sheath was most visible at the ends of the filaments where the cell chain terminated leaving only the empty casing. No flagellated cells were observed exiting from the broken ends of the filaments. Concurrently, no holdfasts were seen, though these may have been lost when the sample was scraped from the RBC. The sheaths were very flexible and were often bent to severe angles without rupturing. It was impossible to determine the overall length of the filaments because they were too intertwined with one another. False branching was rarely observed. The filaments did not move or oscillate during examination.

Most of the cells within the filaments contained blue-black inclusion bodies after staining with Sudan Black B indicating that PHB was stored. A small number of filaments did not contain PHB or contained it only in a localized region. Cells with PHB usually contained at least three of the blue-black inclusion bodies; in some filaments the PHB storage appeared to involve as much as 3/4 of the cell. The zoogloeal masses also contained these inclusion bodies.

The sheaths of the filaments stained a dark Prussian blue after exposure to potassium ferrocyanide under acidic conditions. As great care was exercised to insure that no soluble ferric iron was in solution, it appears that the iron precipitated out onto the sheaths of the filaments during wastewater treatment.

Isolation Experiments

Tangled and curled filamentous growth appeared on all of the initial isolation plates of CGY and CG media. After reisolation the filaments were examined using the light microscope. They were similar to those observed in the biofilm samples: sausage-shaped cells within a sheath exhibiting the same morphological characteristics. Most of the individual cells contained PHB.

Phototactic Experiments

There was no visible difference in the amount of growth on the RBC extract media plates after the one week incubation. Both the plates exposed to continuous light and complete darkness contained a substantial number of bacterial colonies. No microorganisms moved toward the light on those plates with the pinhole in the aluminum foil covering, except for nematodes which burrowed throughout the media.

Ultrastructure of the Non-filamentous Population

Transmission electron microscopy of thin-sectioned samples confirmed the presence of the non-filamentous bacterial cells seen by light microscopy. From low-magnification micrographs, it was evident that both a large number and a wide variety of these organisms were present (Fig. 1). Cell diameters varied considerably, ranging from 0.25 to 1.5 μm.

The ultrastructural characteristics of several representative types of the non-filamentous bacteria are shown in Fig. 2. A number of these organisms regularly contained one

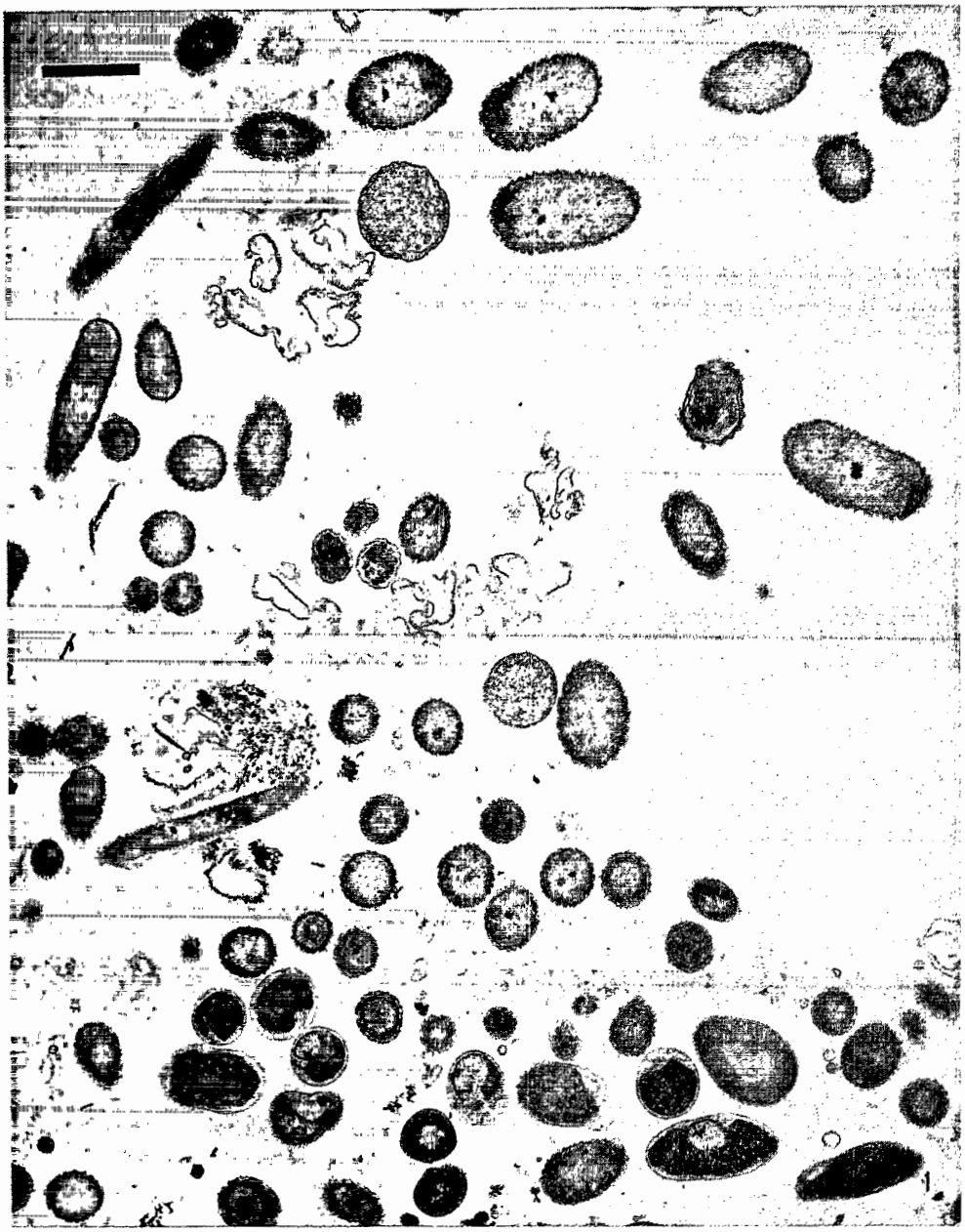

Fig. 1. Low-magnification electron micrograph of the non-filamentous bacterial forms present in the RBC biofilm. Bar = 1.0 μm. Note variety of morphological types and the tendency for similar types to be grouped in a relatively confined area.

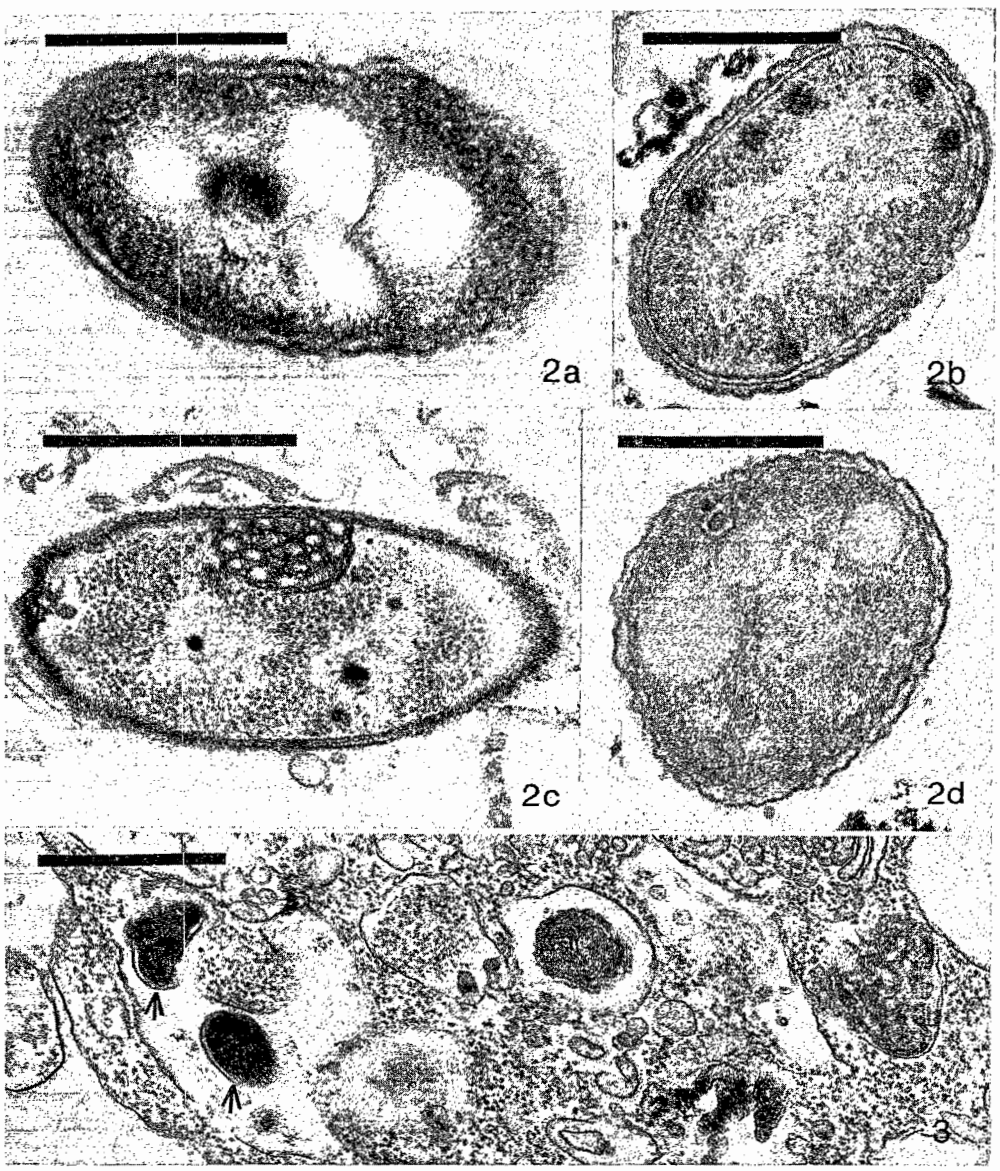

Fig. 2. Electron micrographs of representative non-filamentous forms, showing typical ultrastructural features. Bars = 0.5 µm. a. Cell with PHB inclusion bodies. b. Cell with electron-dense inclusions, possibly polyglucoside granules. c. Cell with prominent mesosome. d. Cell with unidentified inclusions of medium electron density.

Fig. 3. Electron micrograph of amoeboid cell containing two bacterial cells (arrows) within a vacuole. Bar = 1.0 µm.

or more inclusion bodies, and most of them possessed mesosome-like structures. In several instances, cells were seen that appeared to be infected with bacteriophage.

The non-filamentous bacteria often appeared as groups of cells possessing identical morphological and ultrastructural characteristics. These groups, which included 3 to 25 cells, apparently represented microcolonies or colonies. One or more of the cells in such groups were sometimes seen to be undergoing cell division.

The only eukaryotes detected with any regularity in the electron microscopical investigations were amoeboid organisms because the larger protozoa and metazoa were lost in the fixation process (Fig. 3). Interestingly, these organisms always appeared to have several vacuole-like structures which contained one or more intact bacterial cells.

Ultrastructure of the Predominant Filaments

Transmission electron microscopy also confirmed that the predominant filaments seen by light microscopy (above) consisted of independent bacterial cells surrounded by a common sheath (Fig. 4). The filaments ranged from 1.35 to 1.55 μm in diameter, including the sheath. Individual cells within the sheaths ranged from 1.0 to 1.2 μm in diameter and from 1.9 to 4.5 μm in length.

All filaments possessed a relatively dense layer of sheath material that was situated quite close to the surface of the underlying cell walls. Many filaments possessed an additional layer of sheath material which was external to the dense layer.

The cells within the filaments were independent of one another. The cells possessed a typical Gram-negative cell envelope (36). The cells in some, but not all, filaments contained as many as 15 electron-transparent inclusions surrounded by a single electron-dense bounding layer (Fig. 5). They corresponded in size and location to the Sudan Black B-staining granules seen by light microscopy, and in their ultrastructural characteristics to PHB bodies (37,38). In one instance, a filamentous organism appeared to be infected with bacteriophage.

Most cells in the filaments contained prominant mesosome-like structures. These were peripherally located and sometimes appeared to be associated with the polar walls.

Fig. 4. Electron micrograph of predominant *Sphaerotilus*-like filaments, showing arrangement of cells within the common sheath (S). Note the prominent mesosome-like structures (M). Bar = 1.0 µm.

Fig. 5. Electron micrograph of cell within a *Sphaerotilus*-like filament containing many PHB inclusion bodies (P). Note sheath (S) structure, mesosome-like structure (M), and typical Gram-negative cell wall (W). Bar = 0.5 µm.

DISCUSSION

This study has served to document for the first time the morphological and ultrastructural characteristics of microorganisms living within the biofilm formed on a rotating biological contactor. It makes a significant contribution to the RBC literature because the information available on the morphology, physiology, and ecology of RBC biofilm microorganisms is extremely limited. Ultimately a greater understanding of the biofilm's function can help the engineer optimize RBC design and performance. In addition, this study provides information on microorganisms operating in their natural environment; making the conclusions drawn from this information available for practical application.

The predominant organism in the biofilms examined here was a filamentous bacterium consisting of rod-shaped cells enclosed by a common sheath. The data presented suggest strongly that this bacterium is a *Sphaerotilus* species according to the taxonomic structure of the *Sphaerotilus-Leptothrix* group recently established by van Veen et al (39). It is important to note, however, that the species-level taxonomy of this group has been controversial (39,40,41). The principal methods of identification for *Sphaerotilus* are based on microscopic examination, plating and isolation, and on discounting the possibility of the specimen being another type of filamentous form (40). The light microscopical morphology of the organism was identical to that described for *Sphaerotilus* by several authors (39,40,42,43,44). The RBC filaments had: 1) smooth, thin, colorless, Fe^{+3} encrusted sheaths which tightly encased the cells and were often partially evacuated on one end, and 2) Gram-negative cells within the size range 1.-3.5 μm wide and 2.5-16 μm long arranged in a single row. Other filamentous forms were ruled out because the RBC species lacked endospores and crosswalls, were nonmotile, and did not demonstrate a positive phototactic response. The filaments isolated on both the CGY and CG media were similar to those described by Dondero et. al. (30) further indicating that they were *Sphaerotilus*. The ultrastructural characteristics of the filaments were also similar to those of *Sphaerotilus* species noted in other studies (39, 43,44,45), especially the strain examined by Petitprez et. al. (46). The principal similarities included sheath morphology, wall structures, presence of PHB granules, and presence of prominent mesosomes.

Most microscopic studies of the biofilm have not involved a thorough examination of the filaments present. As a result they have been identified as various types of bacteria, algae, and fungi. This study has confirmed that the predominant filament growing in healthy RBC biofilms during domestic wastewater treatment is the bacterium, *Sphaerotilus*. Initially the iron encrustation on its sheath led investigators to believe *Sphaerotilus* was an autotroph (47,48); however, it is now considered an aerobic heterotroph (39,49,50). The role of iron deposition in *Sphaerotilus* remains unknown, though the mechanism may be associated with a moiety of the organisms sheath which catalyzes the reaction (51,52). *Sphaerotilus*-based films growing in laboratory and natural environments can remove 0.5-7.4 g organic $C/m^2 \cdot d$ (53,54); suggesting that this bacterium may contribute significantly to the organic uptake capacity of the RBC biofilm. Though *Sphaerotilus* requires oxygen as a terminal electron acceptor it can function in microaerophillic conditions (39,55). This is particularly significant because the filaments may continue to assimilate organic matter in spite of the rapid decrease in oxygen concentration with depth in the RBC biofilm. *Sphaerotilus* can exist as a filament or free-swimming flagellated cell. This flexible morphology is also uniquely suited to the RBC process. Swarmers can rapidly recolonize disk surfaces after sloughing and filaments can attach to the disks and/or serve as a stabilizing force within the biofilm in a manner similar to that of reinforcing rods in concrete. The maximum growth of *Sphaerotilus* occurs when the fluid velocity is between 0.18 and 0.45 m/s (56) which coincides with the peripheral velocities used to optimize effluent quality in RBCs (17,18).

The fixation procedure used here to prepare the biofilm for electron microscopy did not specifically preserve eukaryotic cells, especially large protozoa and metazoa. Amoebae, however, were regularly observed indicating that they may play a significant role in the trophic structure of the biofilm. Most previous studies have determined that ciliates are the major protozoa present in wastewater treatment systems (57,58,59). A few researchers (60,61,62) found that amoebae were often overlooked or identified as detritus. Sydenham (61) concluded that they may be ecologically as important as ciliates in improving the efficiency of wastewater treatment systems. The amoebae in this study contained single-celled bacteria in individual vacuoles. Ciliates in activated sludge systems have been observed to prey upon single-celled

bacteria; predominantly on enteric species from the raw sewage (63,64,65). The ciliate's predatory activity results in lower organic and suspended solids concentrations in the effluent. The amoebae may play a similar role in the RBC biofilm. It is likely that they live on or near the biofilm's surface where oxygen and influent bacteria are more abundant.

A cell's ultrastructural characteristics can indicate the microorganism's physiological condition. The data presented in this study confirm the presence of a metabolically active population in the biofilm. It was apparent that the population was quite active from: the numbers of cells seen, the variation in cell size, the presence of microcolonies, and the presence of dividing cells within these microcolonies. The presence of mesosomes in both filaments and non-filamentous cells may also be evidence for active metabolism and growth. Although mesosomes are currently somewhat controversial in terms of their true ultrastructure and their functions (if any) in the bacterial cell, they are often seen in dividing cells or metabolically active cells (66).

Both the *Sphaerotilus* filaments and many of the non-filamentous bacteria contained PHB granules. PHB is stored by bacterial cells when carbon concentrations available in the environment are not limiting (44,67). The large number of PHB granules found in the RBC bacteria indicates that excess carbon was present and had been metabolized. Organic carbon assimilated by bacteria may be used in 1) respiration, 2) cell growth and division, 3) PHB production, or 4) extracellular matrix and sheath production. The storage of PHB by biofilm bacteria may serve as an important intracellular sink for organic carbon in RBCs. PHB can account for 11-22.5% of the dry weight of *Sphaerotilus* (68) and 12.0-50.5% of the dry weight of *Zoogloea* (69). The variation in the percent of cell volume involved in PHB storage may be a function of the amount of organic matter available to the cells. Therefore, as the organic loading in the RBC increases the bacteria may store more carbon as PHB until some critical amount of the cell's volume is occupied by this substance. PHB storage, however, cannot be considered exclusively of the other cellular metabolic processes because it acts concommitantly with them in determining the fate of assimilated carbon in the biofilm. PHB also serves as a carbon and energy source for the cells during low nutrient concentrations (70,71) and in this capacity it may mitigate against the effect of fluctuating hydraulic and organic loadings in the RBC.

In *Sphaerotilus* the thickness of the sheath and the formation of an additional layer of sheath-like material has been observed in cells exposed to high organic loadings (44, 72). These external cell structures may function in a manner similar to PHB and/or may additionally function like the extracellular polysaccharide matrices described for zoogloeal bacteria. In aerobic waste treatment systems zoogloeal matrices are important as 1) a storehouse of carbon and energy, 2) an effective adsorbent of metals and organic compounds, 3) an adhesive mechanism, and 4) a buffer during high carbon and nitrogen growth conditions (73).

Some understanding of the ecological conditions in the biofilm may also be drawn from examining the biofilm microorganisms. Both light and transmission electron microscopy revealed the presence of many different types of bacterial cells. Eukaryotic organisms were seen as well. This work supports the contention of other researchers who found various types of bacteria present in wastewater treatment systems (74,75,76,77). The greater the diversity of the biofilm community, the greater its stability, which increases its ability to efficiently degrade wastes and withstand fluctuations in the environment. The presence of different types of bacteria, protozoa, and metozoa indicate that a complex trophic structure may be operating in the biofilm which helps it to continue functioning in spite of external perturbations. The appearance of groups of cells, either as filaments or microcolonies, suggests that these forms are favored over single cells. The presence of phage within some of the bacteria may be indicative of deteriorating conditions in the biofilm. Whatever the cause, bacteriophage may act as natural enemies of biofilm bacteria by reducing their ability to assimilate organic matter from the wastewater.

While this study has shown that the RBC biofilm contains a large and diverse population of microorganisms which form a metabolically active ecosystem it leaves many questions about the microbial ecology of the film unanswered. Its greatest significance may be that it prompts more research aimed at optimizing RBC design and evaluation by increasing the engineer's understanding of the biofilm's mode of operation. Initially additional studies must be performed on the biofilm in the other RBC compartments and as a function of radial distance from the center of the disks. Similarly, the profile of microorganisms must be examined as a function of time and depth within the biofilm. The role of PHB in the physiology of the cells and as a function of organic loading

must be understood to determine the limits of its ability as an intracellular carbon sink in the RBC. The presence or absence of extracellular polysaccharide matrices in the biofilm should be determined because the role of these structures in the metabolism of organic carbon is suspected and deserves further examination. Finally the role of the protozoa and the overall predator-prey relations of the RBC biofilm must be determined to give a clearer picture of its tropic structure. Research of this kind is continuing in our laboratories in an effort to answer some of these questions and to ultimately optimize RBC design and evaluation through an increased understanding of microbial interactions and processes.

REFERENCES

1. _____, 1980, "Environmental Support Still Strong," EPA J. 6(10): 16-18.

2. _____, 1980, "Costly Wastewater Treatment Plants Fail to Perform as Expected," Govt. Acct. Office CED-81-9.

3. Hill, W.R., Regan, T.M., and Zickefoose, C.S., 1979, "Operation and Maintenance of Water Pollution Control Facilities: a WPCF White Paper," JWPCF 51: 899-906.

4. Birks, C.W., and Hynek, R.J., 1971, "Treatment of Cheese Processing Wastes by the Bio-Disc Process," Proc. Purdue Ind. Waste Conf. 26: 89-105.

5. Corneille, R.W., and O'Shaughnessy, J.C., 1975, "Treatment of Apple Wastes Using Rotating Biological Contactors," Proc. Purdue Ind. Waste Conf. 30: 675-688.

6. Hudson, J.W., Smith, J.P., and Pohland, F.G., 1976, "Rotating Biological Contactor Treatment of Shellfish Processing Wastewaters," Proc. Purdue Ind. Waste Conf. 31: 193-205.

7. Griffith, G.T., Young, R.H.F., and Chun, M.J., 1978, "Rotating Disc Sewage Treatment Systems for Suburban Developments and High-Density Resorts of Hawaii," Technical Rept. 116, Water Resource Center, University of Hawaii, Honolulu.

8. Mikula, W.J., et. al., 1980, "A Kinetic Model for Treatment of Cheese Processing Wastewater with a Rotating Biological Contactor," First Nat. Symp. on Rotating Biological Technology 1: 491-514.

9. Huang, J.C., 1980, "Operational Experience of Oxygen Enriched Rotating Biological Contactors." First Nat. Symp. on Rotating Biological Contactor Technology 1: 637-659.

10. Tanacredi, J.T., 1980, "Removal of Waste Petroleum Derived Polynuclear Aromatic Hydrocarbons by Rotating Biological Discs," First Nat. Symp. on Rotating Biological Contactor Technology 1:725-732.

11. Pano, A., Reynolds, J.H. and Middlebrooks, E.J., 1980, "The Kinetics of a Rotating Biological Contactor Treating Domestic Sewage," First Nat. Symp. on Rotating Biological Contactor Technology 1: 449-467.

12. Antonie, R.L., Kluge, D.L., and Mielke, J.H., 1974, "Evaluation of a Rotating Disk Wastewater Treatment Plant," JWPCF 46: 498-511.

13. Barnes, D., and Wilson, F., 1976, The Design and Operation of Small Sewage Works, E. and F.N. Spon, Ltd., London.

14. Stover, E.L., and Kincannon, D.F., 1976, "Evaluating Rotating Biological Contactor Performance," Wat. Sew. Works 123: 88-91.

15. Poon, C.P.C., Chao, Y.L., and Mikucki, W.J., 1979, "Factors Controlling Rotating Biological Contactor Performance," JWPCF 51: 601-611.

16. Kincannon, D.F., and Groves, S., 1980, "Role of Suspended Solids in the Kinetics of RBC Systems," First Nat. Symp. on Rotating Biological Contactor Technology 1: 433-488.

17. Antonie, R.L., 1976, Fixed Biological Surfaces Wastewater Treatment - The Rotating Biological Contactor, CRC Press Inc., Cleveland.

18. Friedman, A.A., Robbins, L.E., and Woods, R.C., 1979, "Effects of Disk Rotational Speed on Biological Contactor Efficiency," JWPCF 51: 2678-2691.

19. Ellis, K.V., and Banaga, S.E.I., 1976, "A Study of Rotating Disc Treatment Units Operating at Different Temperatures," Wat. Pollut. Control 75: 73-91.

20. Antonie, R.L., and Welch, F.M., 1969, "Preliminary Results of a Novel Biological Process for Treating Dairy Wastes," Proc. Purdue Ind. Waste Conf. 24: 115-125.

21. Torpey, W.N., et. al., 1971, "Rotating Disks with Biological Growths Prepare Wastewater for Disposal or Reuse," JWPCF 43: 2181-2188.

22. Pretorius, W.A., 1971, "Some Operating Characteristics of a Bacterial Disc Unit," Wat. Res. 5: 1141-1146.

23. Pescod, M.B., and Nair, J.V., 1972, "Biological Disc Filtration for Tropical Waste Treatment. Experimental Studies," Wat. Res. 6: 1509-1523.

24. Sudo, R., Okada, M., and Mori, T., 1977, "Rotating Biological Contactor -- Microbe Control in RBC, "J. Wat. Waste 19: 1.

25. Hoag, G.E., Widmer, W.J., and Hovey, W.H., 1980, "Microfauna and RBC Performance: Laboratory and Full-Scale Systems. First Nat. Symp. on Rotating Biological Contactor Technology 1: 167-187.

26. Khan, A.N., and Raman, V., 1980, "Rotating Biological Contactor for the Treatment of Wastewater in India," First Nat. Symp. on Rotating Biological Contactor Technology 1: 235-258.

27. Clark, J.H., Moseng, E.M., and Asano, T., 1978, "Performance of a Rotating Biological Contactor Under Varying Wastewater Flow," JWPCF, 50: 896-911.

28. Pelczar, M.J., Manual of Microbiological Methods, McGraw-Hill, New York, 1957.

29. Sherman, V.B.D., A Guide to the Identification of the Genera of Bacteria, Williams and Wilkins, Baltimore, 1967.

30. Dondero, N.C., Phillips, R.A., and Heukelekian, 1961, "Isolation and Preservation of Cultures of *Sphaerotilus*," Appl. Microbiol. 9: 219-227.

31. Gibson, Jane, 1980, Personal Communication.

32. Kellenberger, E., Ryter, A., and Sechaud, J., 1958, "Electron Microscope Study of DNA-Containing Plasms. II. Vegetative and Mature Phage DNA as Compared with Normal Bacterial Nucleoids in Different Physiological Studies", J. Biophys. Biochem. Cytol. 4: 671-687.

33. Spurr, A.R., 1969, "A Low-Viscosity Epoxy Resin Embedding Medium for Electron Microscopy," J. Ultrastruct. Res. 26: 31-43.

34. Reynolds, E.S., 1963, "The Use of Lead Citrate at High pH as an Electron-Opaque Stain in Electron Microscopy," J. Cell. Biol. 17: 208-212.

35. Kinner, N.E., and Bishop, P.L., 1981, "Rotating Biological Contactor Microbiology During Saline and Non-Saline Wastewater Treatment," Submitted for Publication in JWPCF.

36. Costerton, J.W., Ingram, J.M., and Cheng, K.J., 1974, "Structure and Function of the Cell Envelope of Gram-Negative Bacteria," Bacteriol. Rev. 38: 87-110.

37. Dunlop, W.F., and Robards, A.W., 1973, "Ultrastructural Study of Poly-β-Hydroxybutyrate Granules from *Bacillus cereus*, J. Bacteriol. 114: 1271-1280.

38. Snively, J.M., 1974, "Inclusion Bodies of Prokaryotes," Annu. Rev. Microbiol. 28: 167-187.

39. van Veen, W.L., Mulder, E.G., and Deinema, M.H., 1978, "The *Sphaerotilus-Lepthothrix* Group of Bacteria," Microbiol. Rev. 42: 329-356.

40. Dondero, N.C., 1975, "The *Sphaerotilus-Leptothrix* Group," Annu. Rev. Microbiol., 29: 407-428.

41. Pringsheim, E., 1949, "Iron Bacteria," Biol. Rev. Cambridge Phil. Soc. 24: 200-245.

42. Mulder, E.G., and van Veen, W.L., 1963, "Investigations on the *Sphaerotilus-Leptothrix* Group." Ant. van Leeuwenhoek 29: 121-153.

43. Stokes, J.L., 1954, "Studies on the Filamentous Sheathed Iron Bacterium *Sphaerotilus natans*," J. Bacteriol. 67: 278-291.

44. Hoeniger, J.F.M., and Tauschel, H.D., 1973, "The Fine Structure of *Sphaerotilus natans*," Can. J. Microbiol. 19: 309-313.

45. Doetsch, R.N., 1966, "Notes on Some Structural Features of *Sphaerotilus natans*," Arch. Mikrobiol. 54: 46-55.

46. Petitprez, M., et. al., 1969, "Quelques Aspects Structuraux de *Sphaerotilus natans*," Ann. Inst. Pasteur-Lille 20: 103-114.

47. Winogradsky, S., 1888, "Ueber Eisenbakterien," Botan. Ztg. 46: 262.

48. Winogradsky, S., 1922, "Eisenbakterien als Anorgoxydanten," Zbl. Bakteriol. 57: 1.

49. Petitprez, M., and LeClerc, H., 1969, "Les Bacteries du Groupe *Sphaerotilus-Leptothrix*," Ann. Inst. Pasteur-Lille 20: 115-140.

50. Dondero, N.C., 1961, "*Sphaerotilus*, Its Nature and Economic Significance," Adv. Appl. Microbiol. 3: 77-107.

51. Rogers, S.R., and Anderson, J.J., 1976, "Measurement of Growth and Iron Deposition in *Sphaerotilus discophorus*," J. Bacteriol. 126: 257-263.

52. Rogers, S.R., and Anderson, J.J., 1976, "Role of Iron Deposition in *Sphaerotilus discophorus*," J. Bacteriol. 126: 264-271.

53. Curtis, E.J.C., Delves-Broughton, J., and Harrington, D.W., 1971, "Sewage Fungus: Studies of *Sphaerotilus* Slimes Using Laboratory Recirculating Channels," Wat. Res. 5: 267-269.

54. Capblancq, J., and Cassan, M., 1979, "Etude du Periphyton d'une Riviere Polluee (L'Agout). I. Structure et Developpement des Communautes sur Substrats Artificiels," Annls. Limnol. 15: 193-210.

55. Dias, F.F., Dondero, N.C., and Finstein, M.S., 1968, "Attacked Growth of *Sphaerotilus* and Mixed Populations in a Continuous-Flow Apparatus," Appl. Microbiol. 16: 1191-1199.

56. Phaup, J.D., and Gannon, J., 1967, "Ecology of *Sphaerotilus* in an Experimental Outdoor Channel," Wat. Res. 1: 523-541.

57. Curds, C.R., and Cockburn, A., 1970, "Protozoa in Biological Sewage-Treatment Processes, I. A Survey of the Protozoan Fauna of British Percolating Filters and Activated Sludge Plants, "Wat. Res. 4: 225-236.

58. Pike, E.B., and Curds, C.R., 1971, "The Microbial Ecology of the Activated Sludge Process." In Microbial Aspects of Pollution, Ed. Sykes and Skinner, Academic Press, pp. 123-147.

59. Curds, C.R., 1975, "Protozoa," In Ecological Aspects of Used-Water Treatment, Ed. Curds, C.R., and Hawkes, H.A., Academic Press, London, Vol. I: 203-268.

60. Brown, T.J., 1965, "A Study of the Protozoa in a Diffused-Air Activated Sludge Plant," Wat. Pollut. Control 64: 375.

61. Sydenham, D.H.J., 1971, "A Re-Assessment of the Relative Importance of Ciliates, Rhizopods, and Rotatorians in the Ecology of Activated Sludge," Hydrobiol. 38: 553-563.

62. Schofield, T., 1971, "Some Biological Aspects of the Activated Sludge Plant at Leicester," Wat. Pollut. Control 70: 32.

63. Curds, C.R., Cockburn, A., and Vandyke, J.M., 1968, "An Experimental Study of the Role of the Ciliated Protozoa in the Activated Sludge Process," Wat. Pollut. Control 67: 312-324.

64. Curds, C.R., and Fey, G.J., 1969, "The Effect of Ciliated Protozoa in the Fate of *Escherichia coli* in the Activated Sludge Process," Wat. Res. 3: 853-867.

65. Curds, C.R., 1971, "Computer Simulations of Microbial Population Dynamics in the Activated Sludge Process," Wat. Res. 5: 1049-1066.

66. Greenawalt, J.W., and Whiteside, T.L., 1975, "Mesosomes: Membranous Bacterial Organelles," Bacteriol. Rev. 39: 405-463.

67. Fukui, T. et. al., 1976, "Enzymatic Synthesis of Poly-β-Hydroxybutyrate in *Zoogloea ramigera*," Arch. Microbiol. 110: 149-156.

68. Rouf, M.A., and Stokes, J.L., 1962, "Isolation and Identification of the Sudlanophilic Granules of *Sphaerotilus natans*," J. Bacteriol. 83: 343-347.

69. Crabtree, K., McCoy, E., Boyle, W.C., and Rohlich, G.A., 1965, "Isolation, Identification, and Metabolic Role of the Sudanophilic Granules of *Zoogloea ramigera*," Appl. Microbiol. 13: 218-226.

70. Stokes, J.L., and Parson, W.L., 1968, "Role of Poly-β-Hydroxybutyrate in Survival of *Sphaerotilus discophorus* during Starvation," Can. J. Microbiol. 14: 785-789.

71. Parsons, A.B., and Dugan, P.R., 1971, "Production of Extracellular Polysaccharide Matrix by *Zoogloea ramigera*," Appl. Microbiol. 21: 657661.

72. Phaup, J.D., 1968, "The Biology of *Sphaerotilus* Species," Wat. Res. 2: 597-614.

73. Joyce, G.H., and Dugan, P.R., 1970, "The Role of Floc-Forming Bacteria in BOD Removal from Wastewater," Dev. Ind. Microbiol. 11: 377-386.

74. James, A., 1964, "The Bacteriology of Trickling Filters," J. Appl. Bacteriol. 27: 197-207.

75. Taber, W.A., 1976, "Wastewater Microbiology," Annu. Rev. Microbiol. 30: 263-277.

76. Lighthart, B., and Loew, G.A., 1972, "Identification Key for Bacteria Clusters from an Activated Sludge Plant," JWPCF 44: 2078-2085.

77. Unz, R.F., and Dondero, N.C., 1970, "Nonzoogloeal Bacteria in Wastewater Zoogloeas," Wat. Res. 4: 575-579.

ROTATING BIOLOGICAL CONTACTORS - SECOND ORDER KINETICS

by

Edward J. Opatken
U.S. Environmental Protection Agency
Wastewater Research Division
Municipal Environmental Research Laboratory
Cincinnati, Ohio 45268

The RBC process is uniquely adaptable for kinetic studies on secondary treatment of wastewater. Secondary treatment, for this specific kinetic study, is defined as the removal or reduction of soluble substrate with time. The substrate is identified in the reaction phase as soluble chemical oxygen demand (sCOD) and/or soluble biochemical oxygen demand (sBOD). The reduction of insoluble oxygen demanding material is not applicable since:

1. It is the function of the reactor (RBC) to convert soluble organic matter into carbon dioxide and insoluble matter for later removal by the secondary clarifier.
2. The use of unfiltered oxygen demand would require the kinetic study to treat the RBC process as a heterogeneous reaction instead of a homogeneous reaction.

The applicability of the RBC process to study reaction kinetics is attributed to its process configuration, and operation mechanism. The RBC process usually consists of modular units (shafts) that are normally installed in series. Each RBC shaft contains either 100,000 sq ft (9,300 m^2) or 150,000 sq ft (14,000 m^2) of surface area. The volume of the trough to surface area of the discs (V/SA) ratio is fixed by the manufacturer at 0.12 gal/sq ft (4.9 L/m^2). These basic geometric standards enable the reaction time to be determined at each stage. Each RBC shaft rotates at approximately 1.6 rev/min or at a peripheral speed of 60 ft/min (18.3 m/min). The rotation of the RBC mixes the wastewater, and thus simulates a stirred tank reactor. The RBC is divided into independent stages by a baffle which enables the disappearance of soluble oxygen demand to be quantified at each stage for specific time intervals.

Second order kinetics.

The published data by A. A. Friedman[1] on the disappearance of sCOD was incorporated into the rate expression for a second order equation.

where $r = \frac{\Delta C}{\Delta t} = kC_n^2$

$r = \frac{\Delta C}{\Delta t}$ = rate of disappearance of sCOD, mg sCOD/L

k = reaction rate constant, L/mg·h

C_n^2 = the square of the concentration of sCOD in the nth stage, $(mg/L)^2$

$\Delta C = (C_{n-1}-C_n)$ = the difference in concentration of the influent into a stage from the concentration within that stage in mg sCOD/L

Δt = reaction time in h

A plot of $\Delta C/\Delta t$ vs C^2 is shown in Figure 1. The slope of the line is the reaction rate constant, k, which has a value of 0.0062 L/mg·h. The intercept on this curve should theoretically go through zero; however, there is a fraction of sCOD that can be assumed as refractory. This fraction will not undergo biochemical conversion and is represented by the "x" intercept. This fraction is 33 mg sCOD/L for the synthetic influent used by A. A. Friedman in his RBC pilot plant study.

Another published paper by R. J. Hynek(2) was used to obtain interstage data on the disappearance of sCOD. The hydraulic loading rate was used to calculate the residence time in each stage and again the data was incorporated into a second order rate expression. The data consisted of results from both a mechanical drive and an air drive RBC shaft.

Data for the first five runs on both air and mechanical drive systems were plotted showing the concentration of sCOD at specific time intervals based on the retention time within a stage. The curves for only three of these runs are shown in Figures 2 and 3 to improve the clarity of the plot. The plot indicated that the removal rate of sCOD decreased as the reaction time increased, which indicated second order rates of reaction. The data

were then used to plot $\Delta C/\Delta t$ vs C^2 to obtain the reaction rate constant from the slope of the plot. For the five runs the mean reaction rate constant was 0.015 L/mg·h for the mechanical drive system and the mean refractory concentration for these five runs was 25 mg sCOD/L. The air drive system had a mean reaction rate constant of 0.025 L/mg·h and the mean refractory concentration was 21 mg sCOD/L. These data are shown in Table 1.

Table 1. Reaction Rate Constants Derived from R. J. Hynek Data

Run Number	k (L/mg·h)	Correlation Coefficient	Refractory sCOD (mg/L)
VA-M	0.013	0.999	32
VA-A	0.015	0.999	30
VB-M	0.018	0.997	27
VB-A	0.024	0.995	26
V1A-M	0.019	0.998	27
V1A-A	0.022	0.996	24
V1B-M	0.013	0.999	14
V1B-A	0.056	0.987	0
V1C-M	0.014	0.997	26
V1C-A	0.010	0.999	25
mean value (M):	0.015	0.998	25
mean value (A):	0.025	0.995	21

M = Mechanical drive
A = Air drive

Field verification of second order reaction.

Three RBC facilities within a 80 km radius of the Andrew W. Breidenbach Environmental Research Center were sampled to obtain interstage data on the disappearance of sCOD. The three facilities were LeSourdsville, Ohio; Indian Creek (Cleves, Ohio); and Brookville, Indiana. Table 2 summarizes the characteristics of these facilities.

Table 2. Characteristics of RBC Facilities

	LeSourdsville	Indian Creek	Brookville
No. of trains	4	2	3
No. of shafts	20	6	3
No. of stages per train	5	3	4
Diameter of disc, ft (m)	12(3.7)	12(3.7)	12(3.7)
Stages per shaft	1	1	4
Total surface area, ft^2	2.6×10^6	4.8×10^5	3.0×10^5
(m^2)	(2.4×10^5)	(4.5×10^4)	(2.8×10^4)
Surface area, per stage ft^2	1×10^5* 1.5×10^5**	8×10^4	2.5×10^4
(m^2)	(9,300)* (14,000)**	(7,500)	(2,300)
Design flow, mgd	4.0	0.5	0.6
(m^3/d)	(15×10^3)	(1.9×10^3)	(2.3×10^3)
Design hydraulic load, gpd/sq ft	1.5	1.0	2.0
($m^3/m^2 \cdot d$)	(0.062)	(.042)	(0.081)

*Stages 1&2
**Stages 3,4,5

On each sampling date, the following were obtained:

1. Influent, effluent, and stage samples
2. Influent and effluent temperature
3. Plant flow rate during the sampling period

The samples were filtered and stabilized with acid before submittal to the MERL Waste Identification and Analysis Section for sCOD analysis.

The data were then incorporated into a second order reaction rate equation to determine the rate constant for these systems.

The interstage data on sCOD obtained for LeSourdsville were treated in the following two modes. The first mode consisted of plotting sCOD against time and a curve was drawn to represent an approximate fit.

The data from this curve were used to determine the reaction rate constant by determining the slope when $\Delta C/\Delta t$ was plotted against C^2. The C^2 intercept was used to predict the refractory portion of the sCOD. The results are shown in Table 3.

Table 3. Reaction Rate Constants for LeSourdsville

Run Number		k (L/mg·h)		Refractory sCOD, (mg/L)
L0815		0.016		37
L0825		0.028		41
L0903		0.032		6
L0909		0.018		21
L0925		0.023		36
L1002		0.015		28
L1008		0.022		43
LI1017		0.015		47
LII1017		0.026		26
LI1024		0.019		30
LII1024		0.026		36
LI1031		0.026		27
LII1031		0.021		16
LI1105		0.009		49
	mean value:	0.021	mean value:	32
	σ:	0.0062	σ:	12

A second approach was to combine the data from all the runs to obtain a mean value for the influent, the four intermediate stages, and the effluent. A plot of the disappearance of sCOD with time is shown in Figure 4. The reaction rate constant, k, was determined from the slope of the line shown in Figure 5, where $\Delta C/\Delta t$ was plotted against C^2. The reaction rate is 0.024 L/mg·h and the refractory portion is 40 mg/L sCOD. The k value of 0.024 L/mg·h is similar to the k value of 0.021 L/mg·h obtained by determining the mean of the 14 individual runs at LeSourdsville.

The k value at LeSourdsville also is similar to the k value obtained from the Hynek data at the South Shore plant, which is 0.015 for mechanical drive and 0.025 L/mg·h for air drive RBC.

There are wide variations in the analytical data from Indian Creek and these may be attributed to the low level of sCOD in the influent and the long residence time, which at times were over 6 hours. The maximum sCOD obtained at Indian Creek was 105 mg/L and only one sample out of 36 was above 100 mg/L. Another factor that impaired the analyses at Indian Creek was the physical layout which consisted of only three stages. This limited the number of sample points and reduced the probability of determining a curve for representing the disappearance of sCOD.

There were nine sampling dates at Indian Creek. Of the nine dates, four were discarded because a curve could not be drawn that would adequately represent the data to describe the disappearance of sCOD. Figure 6 is an example of a wide scatter analytical result that could not be used in the data reduction. Figure 7 is an example of the disappearance of sCOD with time that could be represented by drawing a curve to represent the selected data. For the five dates that could be described by drawing the best curve for the disappearance

of sCOD with time, the reaction rate constant was 0.018 L/mg·h and the mean sCOD refractory was 27 mg/L. Again, the data from these five runs showed a k value similar to the value at LeSourdsville and by Hynek, even though the level of influent sCOD was significantly below the sCOD levels at the other locations.

The result at Indian Creek behaves as though it were biochemical reaction rate limited and the kinetics obey a second order rate expression. The Indian Creek results show that the low level of sCOD in the influent does not alter its kinetic behavior, and obeys a second order rate expression, whose reaction rate constant is similar to the values obtained at LeSourdsville and with Hynek data.

Oxygen transfer limitation.

During Hynek's test, four runs were operated at a significantly higher hydraulic loading rate, ranging between 2.1 and 2.9 gpd/sq ft (86 to 120 $L/d \cdot m^2$). This, in effect, reduced the reaction time by approximately 50%. To accomplish the same sCOD reduction at the high hydraulic loading, as was obtained at the low hydraulic loading, would require doubling the oxygen transfer rate and an adequate level of biomass to handle the additional sCOD removal requirements resulting from the increase in the hydraulic loading rate.

The plot of sCOD with time is represented by Run VIII for the air drive system and is shown in Figure 8. The data show a linear relationship for the disappearance of sCOD with time. This relationship indicates zero order kinetics. A possible explanation is; as the hydraulic loading increased there was insufficient time to transfer the oxygen required for converting the sCOD; thus changing the system from a biochemical reaction limiting process into an oxygen transfer limiting process; and the kinetic rate changed from a second order expression into a zero order expression.

The Brookville, Indiana facility was sampled on 10 dates. The data reduction for the ten sampling dates resulted in an apparent oxygen limiting operation.

A plot of sCOD against time for the Brookville data showed that seven of the ten dates could best be represented by a zero order rate equation. The data were combined to obtain a mean value of sCOD at each of the four stages and the average retention time at each stage. These values are plotted in Figure 9 and show an excellent correlation for a zero order rate equation.

A comparison was made of the loading levels at LeSourdsville, Indian Creek, and Brookville. A significantly higher loading is evident at Brookville when compared with LeSourdsville or Indian Creek.

The pseudo oxygen mass transfers were calculated for the first stage of the RBC at LeSourdsville, Brookville, and Indian Creek. A sample calculation for the pseudo oxygen transfer at LeSourdsville follows.

The hydraulic loading at LeSourdsville averaged 0.82 gal/d·sq ft

$$0.82 \frac{\text{gal}}{\text{d·sq ft}} \times \frac{\text{d}}{24\text{h}} \times [2 \times 100{,}000 + 3 \times 150{,}000]\ \text{sq ft} \times \frac{3.8\ \text{L}}{\text{gal}} =$$

$$0.82 \frac{\text{gal}}{\text{d·sq ft}} \times \frac{\text{d}}{24\text{h}} \times 650{,}000\ \text{sq ft} \times 3.8 \frac{\text{L}}{\text{gal}} = 85{,}000\ \text{L/h}$$

The oxygen required to satisfy the disappearance of 54 mg/L of sCOD in the first stage is determined by:

$$85{,}000 \frac{\text{L}}{\text{h}} \times 54 \frac{\text{mg}}{\text{L}} \times \frac{1}{100{,}000\ \text{sq ft}} = \frac{37\ \text{mg}\ O_2}{\text{h·sq ft}} \quad \frac{(400\ \text{mg}\ O_2)}{\text{h·m}^2}$$

Table 4. Loading Levels at the RBC Facilities

	LeSourdsville	Brookville	Indian Creek
Hydraulic loading, gpd/sq ft ($L/d \cdot m^2$)	0.88 (36)	1.5 (62)	.5 (20)
Influent, mg sCOD/L	118	288	65
Retention time, h	3.5	2.0	6.0
Pseudo oxygen transfer, $mgO_2/h \cdot sq$ ft, (first stage) ($mgO_2/h \cdot m^2$)	37 (400)	48 (520)	9 (100)

It is evident from this comparison that Brookville has 170% greater hydraulic loading than LeSourdsville and the influent concentration in sCOD is 240% greater at Brookville resulting in an exceptionally high organic loading. The oxygen mass transfer is assumed to be at a maximum, and the limiting factor at Brookville appears to be oxygen transfer rate limited.

If it is assumed that Brookville were limited by a second order rate equation with a rate constant equal to the rate constant obtained at LeSourdsville, 0.021 L/mg/·h, then the disappearance of sCOD would follow the curve as shown in Figure 10. The concentration leaving the first stage, 0.5h reaction time, is 123 mg/L. The pseudo oxygen transfer rate to achieve this level of reduction is 158 mg/h·sq ft (1700 $mg/h \cdot m^2$). This is more than 4 times the pseudo oxygen transfer rate calculated for LeSourdsville and 3 times the actual rate calculated for Brookville. It is for these reasons that the oxygen transfer rate is believed controlling the reaction mechanism at Brookville.

The selection of the three facilities, LeSourdsville, Brookville and Indian Creek, was based on the proximity of these sites to MERL. Yet these three facilities provide a good mix for this evaluation because of the wide

variation in their loadings.

1. LeSourdsville operates at an organic loading that appears to be within 20% of the upper limit for oxygen mass transfer rates and behaves as though it were biochemical reaction rate limited.
2. Brookville operates at an organic loading that appears to be limited by the oxygen mass transfer rate.
3. Indian Creek operates at an organic loading considerably below LeSourdsville and appears to follow a biochemical reaction rate limting process, whose rate constant is similar in value to the rate constant obtained at LeSourdsville and from Hynek.

When the hydraulic load increased, as Hynek did in his evaluation, then the process appears to change from a kinetically limited system to an oxygen limited process.

These results present a new approach in the analyses of RBC performance. The applicability of a second order reaction rate expression to follow the disappearance of the soluble organic fraction was demonstrated with Friedman's pilot plant data, Hynek data, LeSourdsville, and Indian Creek.

The second order expression failed to follow the disappearance of sCOD with time at Brookville, and with Hynek results when the hydraulic loading was doubled. These two operations obeyed zero order kinetics and were assumed to be oxygen mass transfer limited.

The similar k values obtained from RBC's at different field sites treating municipal wastewaters indicates that the reaction rate constant, k, can be used to predict the performance for RBC's when they are employed for secondary treatment. For a series of stirred tank reactors or RBC stages, the concentration of soluble organics can be determined at any stage in the process by use of Levenspiel's equation[3] if the following parameters are known:

1. Reaction rate constant based on sCOD or sBOD, L/mg·h
2. Residence time, h
3. Influent organic concentration, mg/L

Levenspiel's equation for staged reactors that follow second order kinetics is mathematically derived from a mass balance, and is applicable for calculating the soluble organic concentration at any stage. The equation is:

$$C_n = \frac{-1 + \sqrt{1 + 4\,(kt)(C_{n-1})}}{2(kt)}$$

where C_n = concentration of soluble organics in n-stage, mg/L

k = second order reaction rate constant, L/mg·h

t = residence time, h

C_{n-1} = influent soluble organic concentration to stage n, mg/L

This equation can then be programmed into a computer and by inserting the number of stages, n, the initial concentration, C_{n-1}, the residence time within each stage, t, and the reaction rate constant, k; the concentration, C_n, in terms of soluble organics can be readily obtained at any stage in the process train.

To test the applicability of second order kinetics to predict the concentration of soluble organics in any stage of a RBC train, interstage data was obtained from Ianone[4] on the disappearance of sBOD at 9 plants using air drive RBC. The results obtained by Hynek[2] and analyzed earlier in this paper to obtain k values based on sCOD for both air and mechanical drive RBC also included interstage data on the disappearance of sBOD. Hynek's data with sBOD using air drive shafts were incorporated into the second order rate expression to obtain a reaction rate constant for sBOD of 0.083 L/mg·h. The k value was incorporated into Levenspiel's equation to predict the sBOD concentration at any stage for each of the 9 air drive RBC plants. These results are shown in Table 5, and are displayed with the actual results for comparative purposes.

Table 5. Comparison of the Predicted and Actual Disappearance of sBOD

Cleves

Shafts/Stage = 1-1-1

t(h) = 2.5, 2.5, 2.5

	Predicted	Actual
C_{in} =		40
C_i =	12	8
C_2 =	5	5
C_3 =	3	3

Enumclaw

Shafts/Stage = 3-1-1-1

t(h) = 1.4, .46, .46, .46

	Predicted	Actual
C_{in} =		168
C_1 =	34	14
C_2 =	20	9
C_3 =	13	7
C_4 =	10	6

Lancaster

Shafts/Stage = 1-1-1-1.5**

t(h) = 1.4, 1.4, 1.4, 2.2

	Predicted		Actual
C_{in} =			218
C_1 =	39	78*	78
C_2 =	15	22	22
C_3 =	8	10	14
C_4 =	4	5	8

Lower East Fork

Shafts/Stage = 1.5-1-1-1

t(h) = .97, .64, .64, .64

	Predicted	Actual
C_{in} =		20
C_1 =	11	11
C_2 =	8	6
C_3 =	6	5
C_4 =	5	5

*Assume overloaded first stage and determine concentrations of sBOD in succeeding stages.

**High density media

C_{in} = influent

Table 5. Comparison of the Predicted and Actual Disappearance of sBOD (cont'd)

Woodburn

Shafts/Stage = 4-2-1.5-1.5

t(h) = 1.69, .84, .63, .63

	Predicted	Actual
C_{in} =		226
C_1 =	37	28
C_2 =	17	12
C_3 =	11	7
C_4 =	8	7

Greenwood Springs

Shafts/Stage = 1-1-1-1.5

t(h) = .56, .56, .56, .84

	Predicted	Actual
C_{in} =		43
C_1 =	22	20
C_2 =	13	14
C_3 =	9	4
C_4 =	6	5

Dodgeville

Shafts/Stage = 2-1-1

t(h) = 2.6, 1.3, 1.3

	Predicted	Actual
C_{in} =		37
C_1 =	11	9
C_2 =	7	7
C_3 =	4	4

West Dundee

Shafts/Stage = 1-1-1.5

t(h) = .76, .76, 1.2

	Predicted	Actual
C_{in} =		101
C_1 =	33	33
C_2 =	16	15
C_3 =	9	8

Hartford

Shafts/Stage = 1-1-1-1

t(h) = .25, .25, .25

	Predicted	Actual
C_{in} =		17
C_1 =	13	13
C_2 =	11	12
C_3 =	9	9
C_4 =	8	8

There is good agreement between the predicted and actual sBOD at seven of the nine plants. There was a difference at Enumclaw and Lancaster. The calculation for Lancaster was modified by assuming an inadequate oxygen transfer rate in the first stage, due to the high organic loading, and then applying second order kinetics to the following stages. By using the actual value of 78 mg/L sBOD, that was obtained in the second stage, as the initial concentration, and then calculating the sBOD in the ensuing stages, good agreement was then obtained for Lancaster between the predicted and actual sBOD concentrations. There is no explanation that can be theorized at this time for the discrepancy at Enumclaw. An analysis similar to Lancaster is not valid because the actual concentration of sBOD in the first stage was considerably below the predicted value, and therefore oxygen transfer requirements were satisfied at Enumclaw.

These results provide added evidence that RBC obey second order kinetics and when the reaction rate constant is known, can be used to predict performance, design optimum train configurations, and can be used to reduce capital costs.

References

1. Friedman, A. A. "Kinetic Response of Rotating Biological Contactors," 31st Annual Purdue Industrial Waste Conference, 1976.

2. Hynek, R. J., and Chou, C.S. "Development and Performance of Air Driven Rotating Biological Contactors," 31st Annual Purdue Industrial Waste Conference, 1976.

3. Levenspiel, O. Chemical Reaction Engineering, John Wiley & Sons, 124-149, 1972.

4. Iannone, J., Personal communication (Roy F. Weston), December 7, 1981.

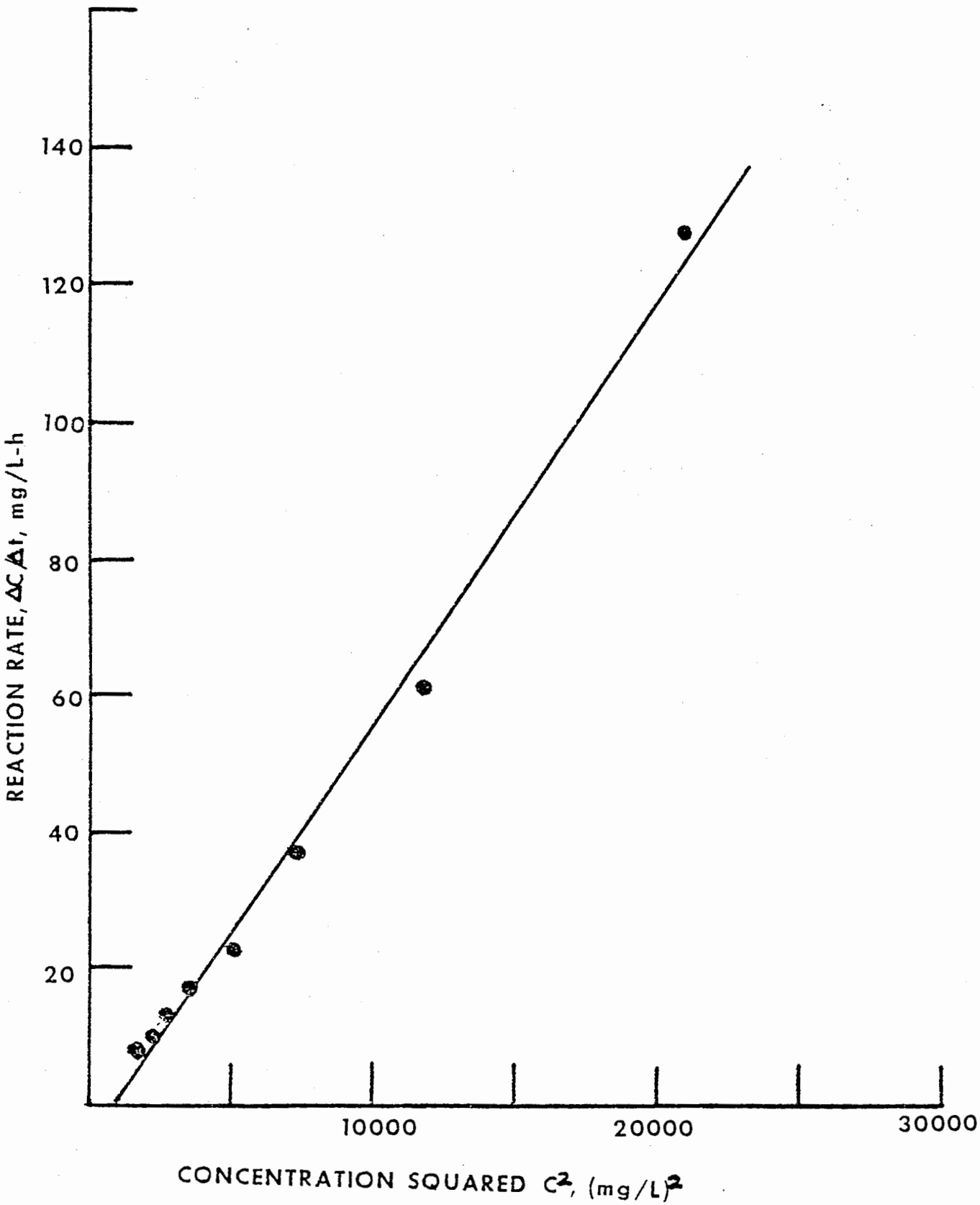

Figure 1.Reaction rate at various concentrations squared.

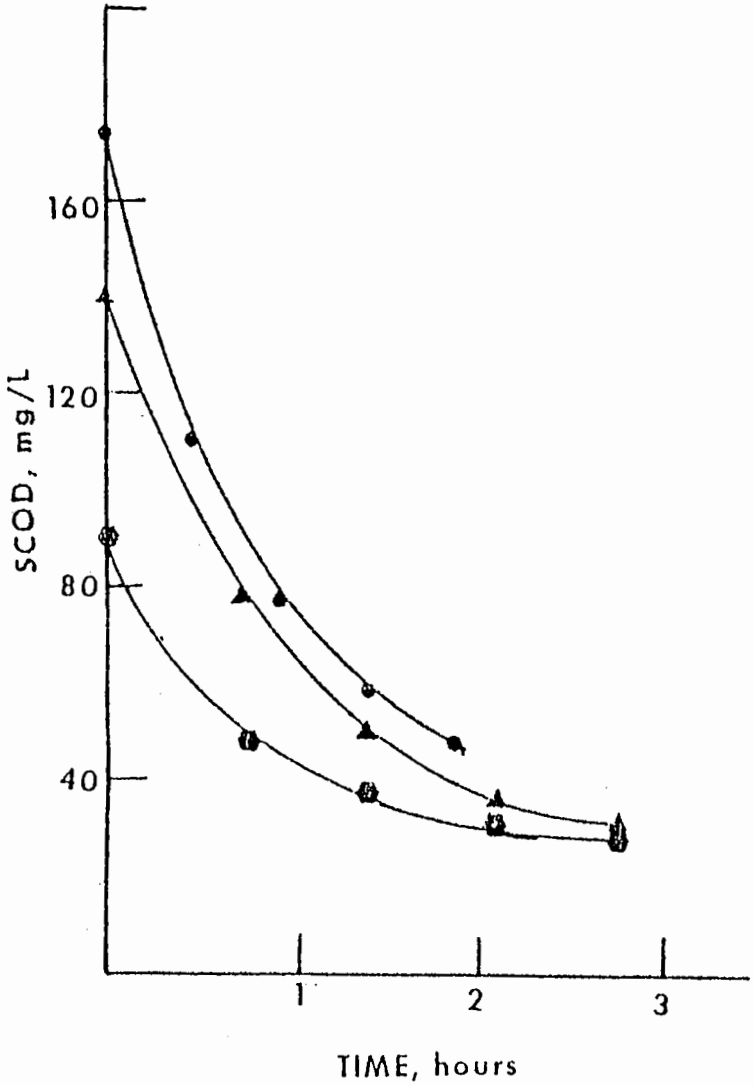

Figure 2. Air drive — Disappearance of sCOD with time.

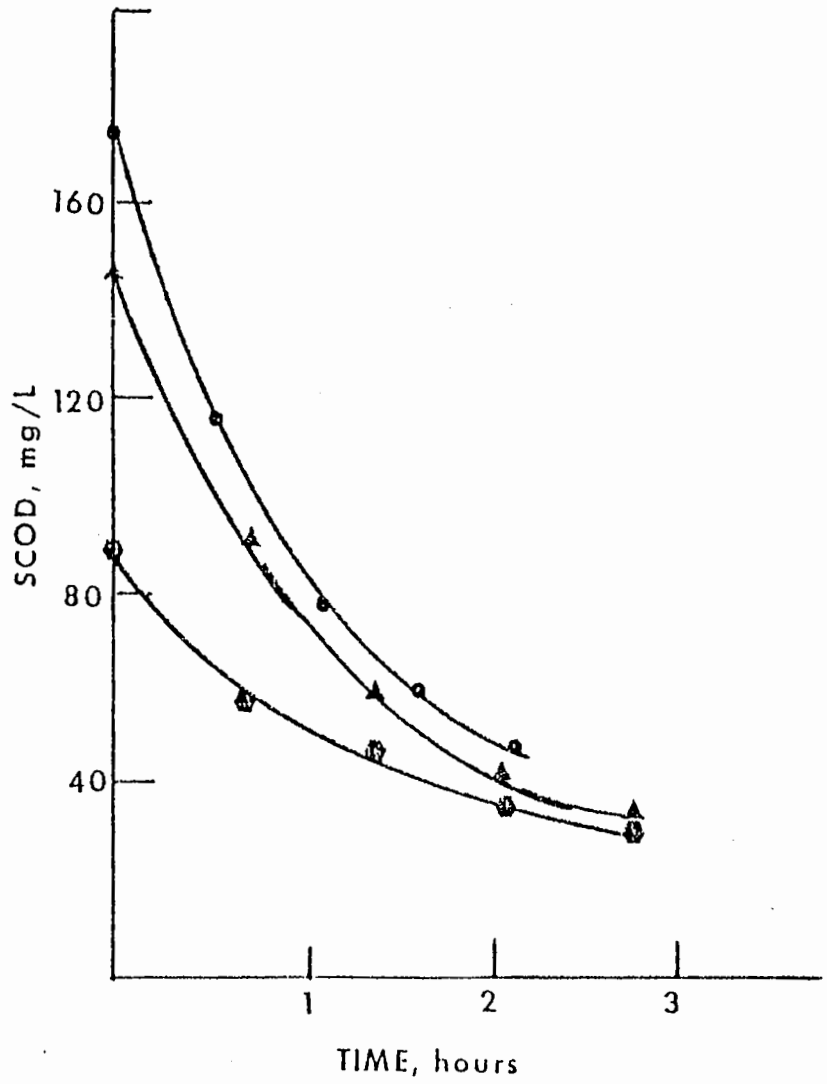

Figure 3. Mechanical drive — Disappearance of sCOD with time

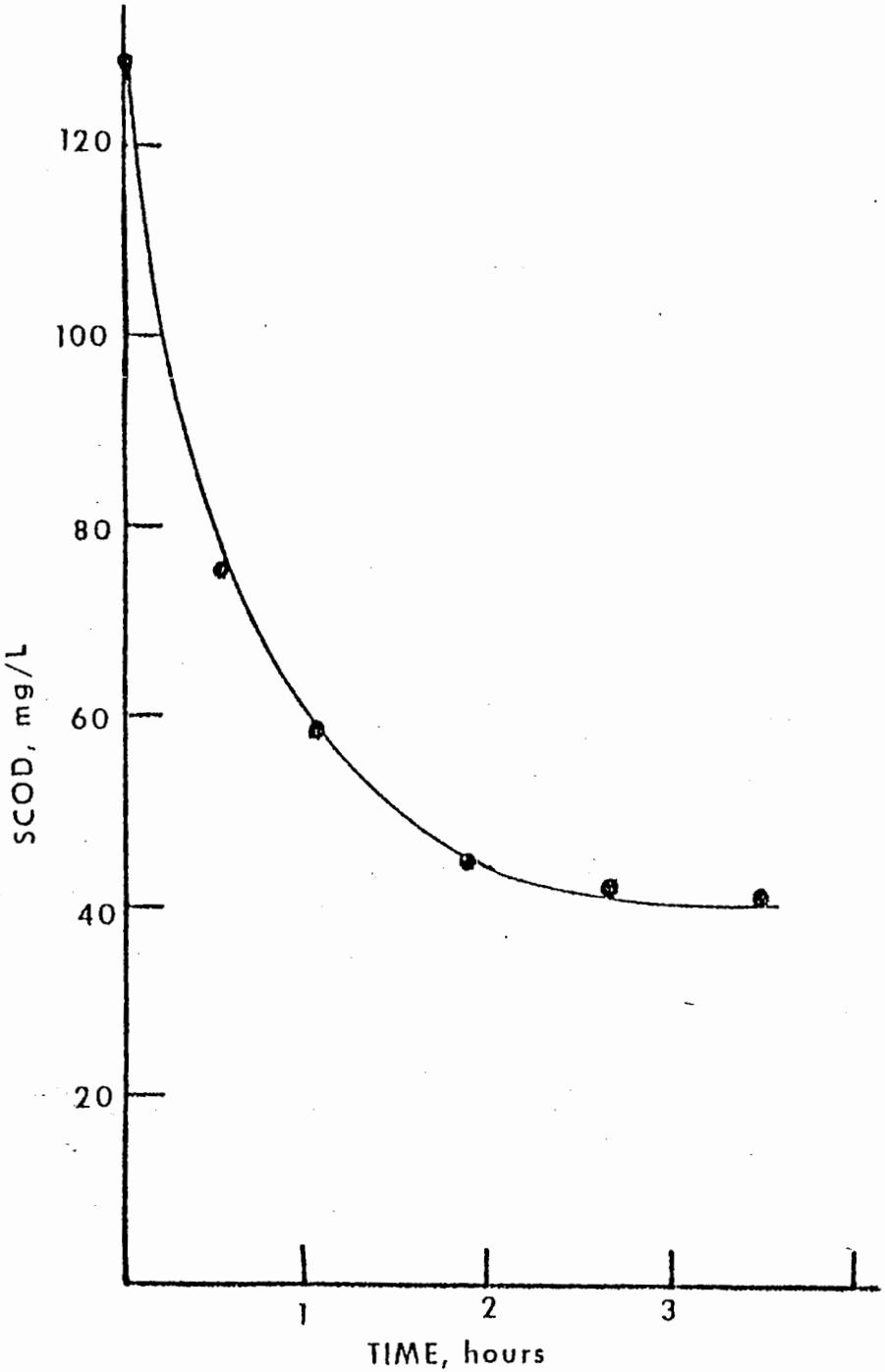

Figure 4. LeSourdsville—
Disappearance of sCOD with time.

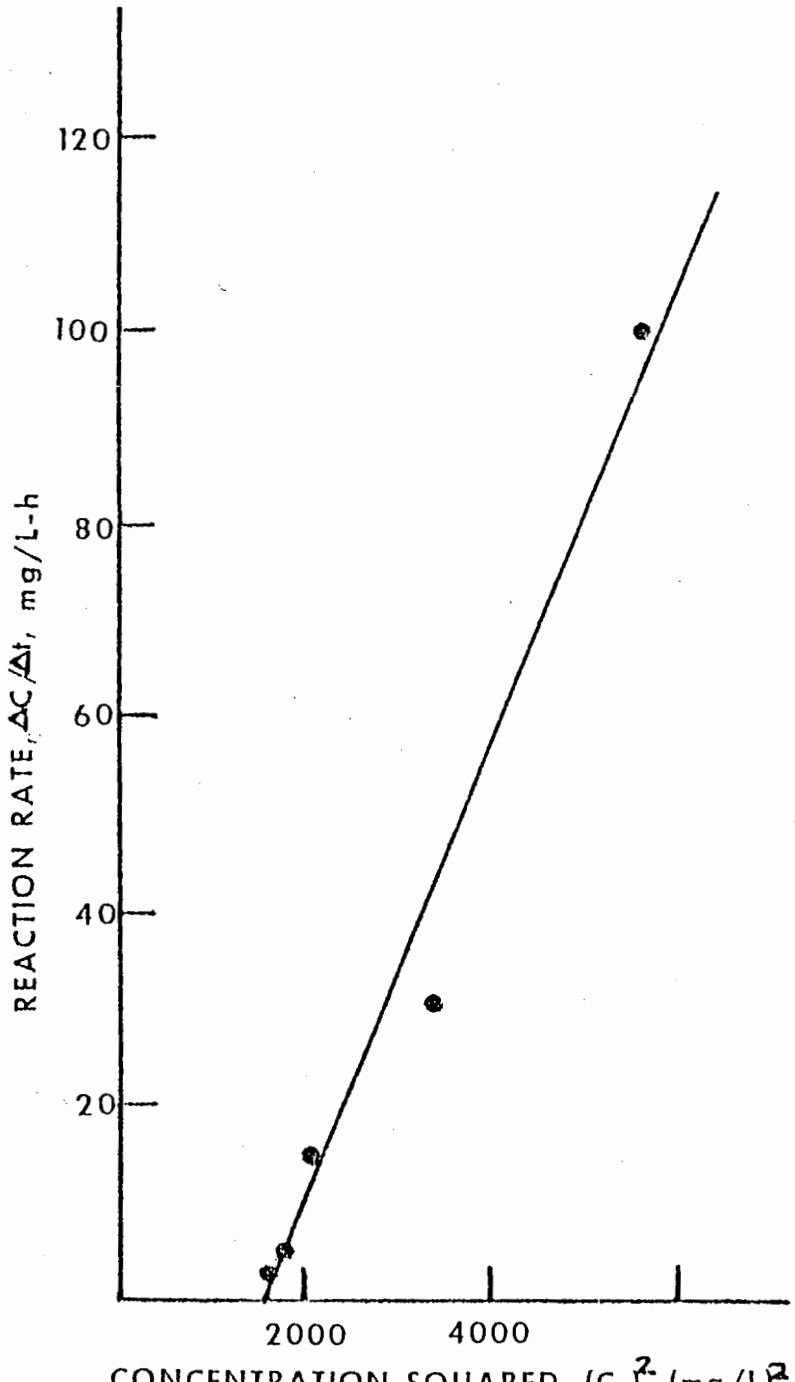

Figure 5. LeSourdsville.
Reaction rate at various concentrations squared.

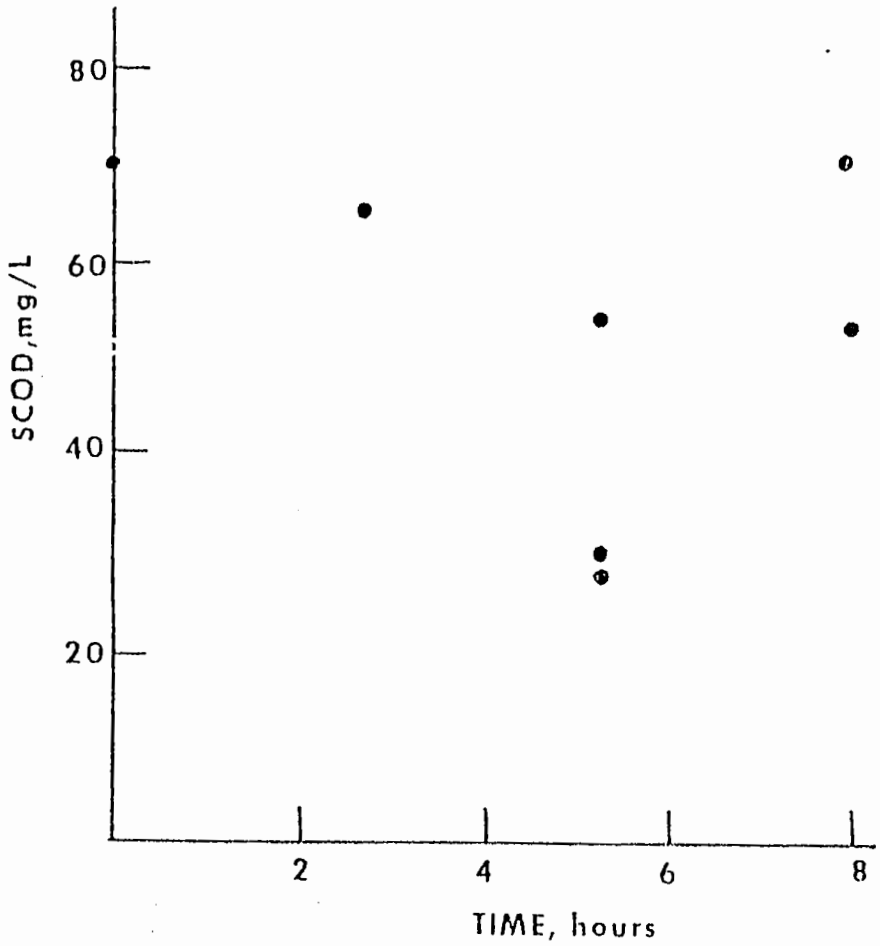

Figure 6. Scatter on disappearance of sCOD with time.

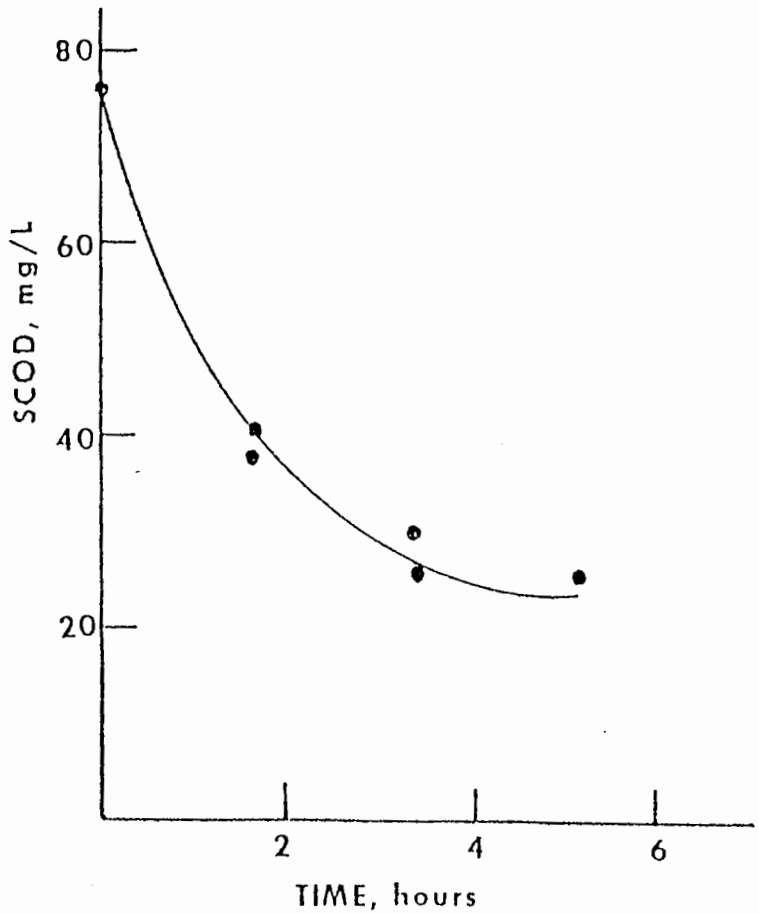

Figure 7. Indian Creek — Disappearance of sCOD with time.

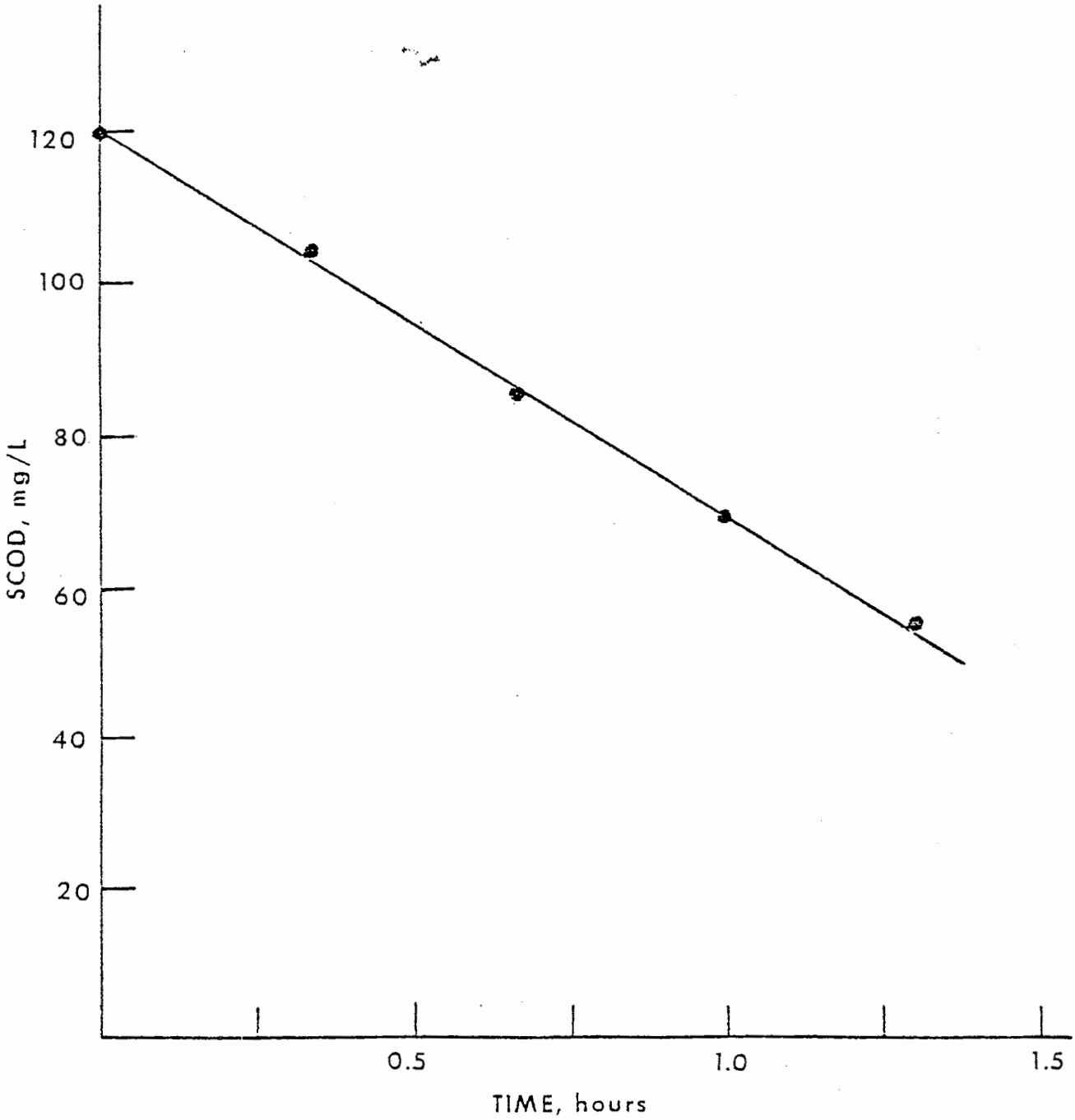

Figure 8. Hynek — Disappearance of sCOD with time.

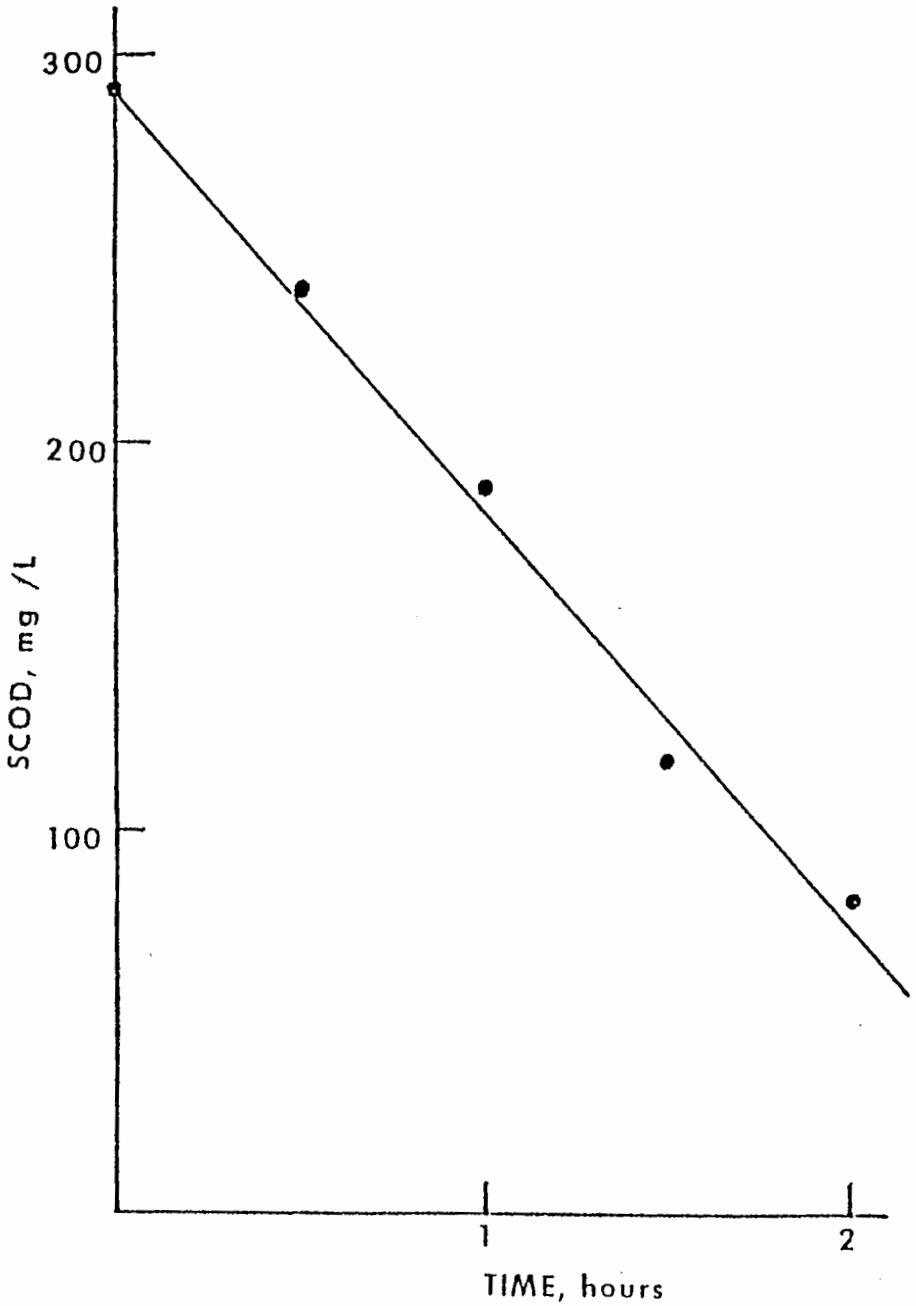

Figure 9 Brookville Disappearance of sCOD with time.

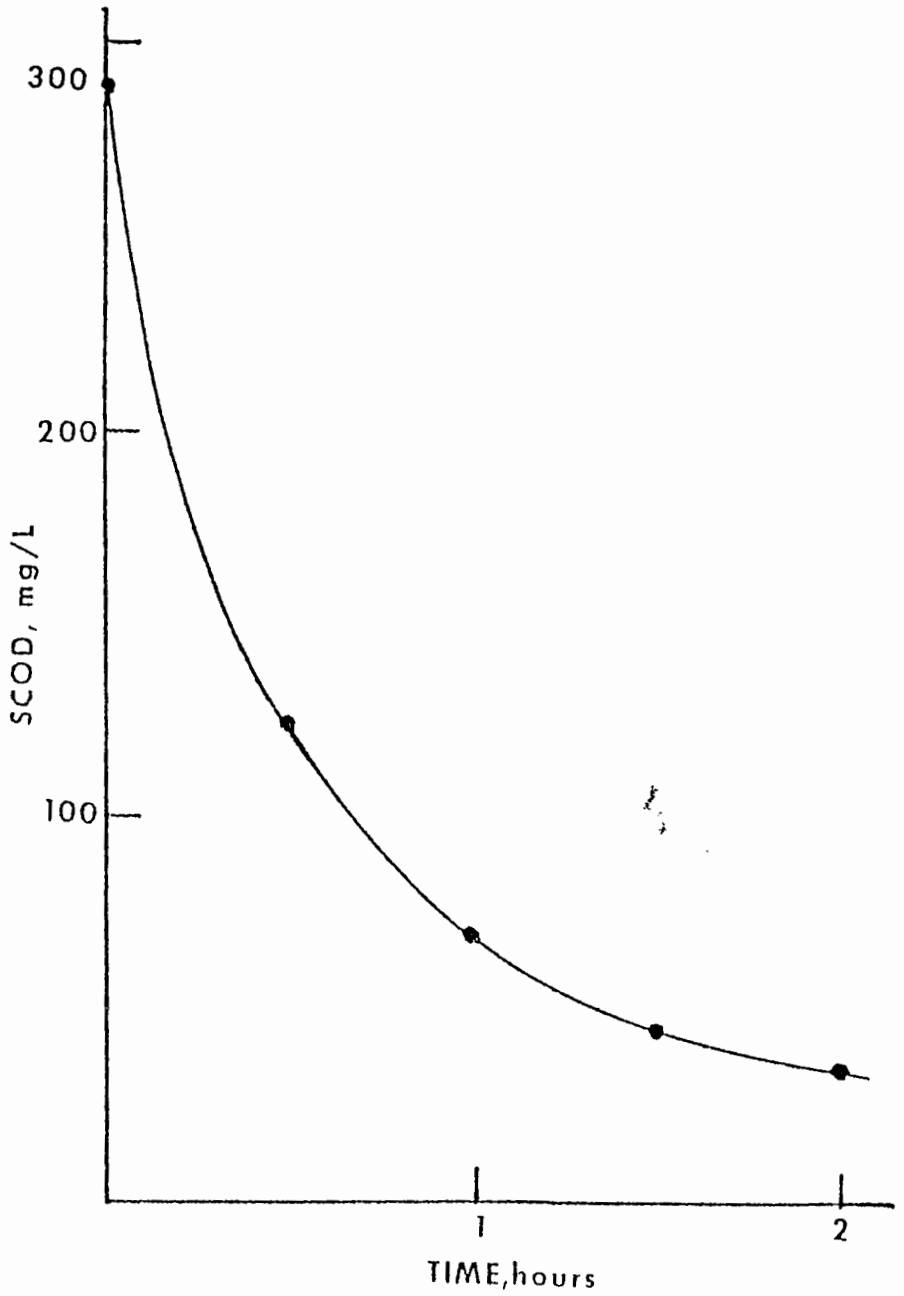

Figure 10. Disappearance of sCOD with time

ASSESSMENTS OF THE KINETIC PERFORMANCE OF A ROTATING BIOLOGICAL CONTACTOR SYSTEM

Ta-Shon Yu, Ph.D., P.E.
Office of Environmental Programs
State of Maryland

Randolph G. Denny
Office of Environmental Programs
State of Maryland

INTRODUCTION

The employment of a rotating biological contactor (RBC) for wastewater treatment was pioneered by Hans Hartmann and Franz Popel of Germany in 1955 on a scale of technological research basis. It had not been developed into the extent of commercial applications until early 1970's when the technological practice became economically competitive with the activated sludge process. From 1974 to 1980, escalation of energy costs and abundance of federal funds for construction in the United States prompted this wastewater treatment technique into its prospective market place within a short time frame during which the sales representatives of the biological contactor manufacturers were the only authorities in the structural design as well as the functional forecast. As a result, the owners and operation personnel associated with the biological contactors would either take in a pride of prudent decisions in selection of this specific treatment process, or be dismayed by the outcome of functional performance for the entire life span.

Whether the rotating biological contactor process can live up with the expectations of cost-effectiveness and

and functional reliability should not be assessed solely based on the numerous incidents of the structural failure or simply based on the reports of successful performance with short-time experience. The manufacturers have been pressed to improve the structural integrity for the reason of business survival. Only the time can tell if improvements have been made to a satisfactory manner that requires a functional life of at least 25 years to justify its cost-effective claim. The structural set-back could be viewed as a typical problem of any technological transition. However, it should be born in mind what damage can be done with the business once the reputation is ruined. Should the biological contactor industry strive to stay in business, it would be a viable wastewater treatment technique which deserves a fair consideration.

The first installation of the rotating biological contactors in the State of Maryland is at the St. Michael Wastewater Treatment Plant. The design capacity is 0.5 mgd to accommodate the projected needs for the year of 1990's. This is a tertiary plant which consists of primary sedimentation, biological treatment by rotating contactors, secondary sedimentation followed by filtration, chlorination, dechlorination and post-aeration. It is designed primarily to treat domestic wastewaters containing 240 mg/l of BOD_5 and suspended solids respectively to meet 20 mg/l of BOD_5 and 10 mg/l of suspended solids as monthly average effluent quality limitations set forth by the National Pollution Discharge Elimination System (NPDES) permit. The design criteria are shown in Table I. It should be noted that the design of primary and secondary clarifiers is not intended to be conservative, but to satisfy the performance reliability which requires at least two units for each sedimentation process.

The plant operation was initiated in late 1979. The current flows approximate 0.25 mgd with 210 mg/l of BOD_5 and 120 mg/l of suspended solids on a yearly basis. Because of the current low flow conditions, one primary clarifier and one secondary clarifier are in line with the remaining treatment processes. The biological contactors are driven by 5-hp gearmotors with a rotating speed of 1.6 rpm.

PERFORMANCE STUDY

Attempts were made to assess microbial behaviors of the rotating biological contactors on performance of

Table I - Criteria Used for Design of the
St. Michael Wastewater Treatment Plant

Average Daily Flow	
Initial (1979)	0.25 mgd
Design (1990)	0.50 mgd
Influent Characteristics	
BOD_5	240 mg/l
Suspended Solids	240 mg/l
Primary Clarifier (2 Units)	
Dimensions	30' dia. x 10' SWD
Surface Overflow Rate	350 gpd/sq. ft.
Detention Time	5 hrs.
Biological Contactor (3 Units)	
Operation Mode	in series
Shaft Dimensions	25' x 11'-6"
Surface Area - Each	100,000 sq. ft.
Nominal Volume - Each	10,500 gal.
Nominal Detention Time - Each	0.5 hr.
Secondary Clarifier (2 Units)	
Dimensions	30' dia. x 8' SWD
Surface Overflow Rate	350 gpd/sq. ft.
Detention Time	4 hrs.
Filtration	
Operation Mode	continuous backwash
Surface Area	180 sq. ft.
Filtration Rate	2 gpm/sq. ft.
Chlorination	
Detention Time	60 min.
Dechlorination/Post-aeration	
Detention Time	15 min.

carbonaceous removal and nitrification so as to acquire relevant information for optimal design. Evaluations were conducted under both normal and abnormal operating conditions by analyzing samples taken from each stage of the biological treatment. The area of relative microbial activity at each stage of the contactor was also exploited.

The magnitude of pollutants permissible for discharge, except for the toxic substances, is indicative of the assimilative capacity of the receiving water through the natural purification process to satisfy oxygen demands exerted by the carbonaceous and nitrogenous compounds. No matter these bio-degradable compounds are soluble or insoluble, the receiving water is obligated to replenish the total amount of oxygen required until the assimilative capacity is exhausted. The current approach in evaluation of the performance efficiency of the rotating biological contactor apparently tends to place its importance upon removal of soluble and readily oxidizable constituents. This study is intended to reiterate the significance of the fundamental principle of pollution abatement related to the capability of the biological contactor in removing insoluble bio-degradable organic substances.

The primary effluent was introduced into the biological contactor in a direction perpendicular to the shaft. The compartment of each stage was so confined that the mixed liquor in a practical sense represented a completely mixed system. Samples taken from the contactor compartments had been allowed to settle for 30 minutes before the supernatants were drained for laboratory analyses conducted by the Laboratories Administration of the Maryland State Department of Health and Mental Hygiene. The analytical results of the supernatants would provide the accessory information of relative settleability of the mixed liquor suspended solids in each stage of contactor in comparison with that of the secondary effluent.

The rotating biological contactor system installed at the St. Michael Wastewater Treatment Plant was manufactured by George A. Hormel & Co., EPCO - Hormel RBS Bio-Shaft, Model M3707, Serial No. 179. In two years operation, the structural failures were experienced. As a State regulatory agency in approving construction contract plans and specifications and in implementation of plant performance, such unwanted problems must be resolved. In order to live up with the expectation that the rotating biological contactor

is a viable and dependable technique in wastewater treatment, recommendations are made to control structural integrity in the process of the construction contract procurement.

RESULTS AND DISCUSSIONS

Under Normal Operating Conditions

The current domestic flows at the St. Michael Wastewater Treatment Plant average 0.25 mgd. Figure 1 represents the typical pattern of the rotating biological contactor performance in removal of carbonaceous compounds and achievement of nitrification under normal operating conditions. It is interesting to note that the first stage contactor is capable of performing two distinctly different metabolic functions simultaneously. The result indicates that the heterotrophic micro-organisms responsible for BOD removal and the autotrophic micro-organisms responsible for oxidation of ammonia nitrogen co-exist on the same environment favorable for their growth and propagation.

The bio-mass attached to the contactors is roughly equivalent to 10,000 mg/l of mixed liquor suspended solids in the first stage compartment, 7,500 mg/l in the second stage compartment, and 3,750 mg/l in the third stage compartment. The BOD_5 applied to the first stage contactor approximates 100 mg/l that is 200 pounds of BOD_5 at the flow of 0.25 mgd. The corresponding organic loading lies in the neighborhood of 2 lbs. BOD_5 / day / 1000 sq. ft. or 0.2 lb. of BOD_5 per pound of bio-mass. The organic loading of this magnitude is comparable to the operation mode of the extended aeration process.

In the presence of high concentrations of alkalinity (200 mg/l to 300 mg/l) and slightly alkaline pH conditions (7.5 to 8.0), a complete nitrification can be expected at temperatures above 10°C. As the nitrification takes place, it consumes approximately 8 mg/l of alkalinity for 1 mg/l of ammonia nitrogen oxidized. Other than the favorable environmental factors with respect to alkalinity, pH and temperatures, the successful nitrification may have been attributed to the low BOD loading which refrains the heterotrophic micro-organisms from rapid growth to the extent that permits Nitrosomonas and Nitrobacters to reproduce themselves.

As shown in Figure 1, the second and the third stages of contactors contribute little wastewater treatment under

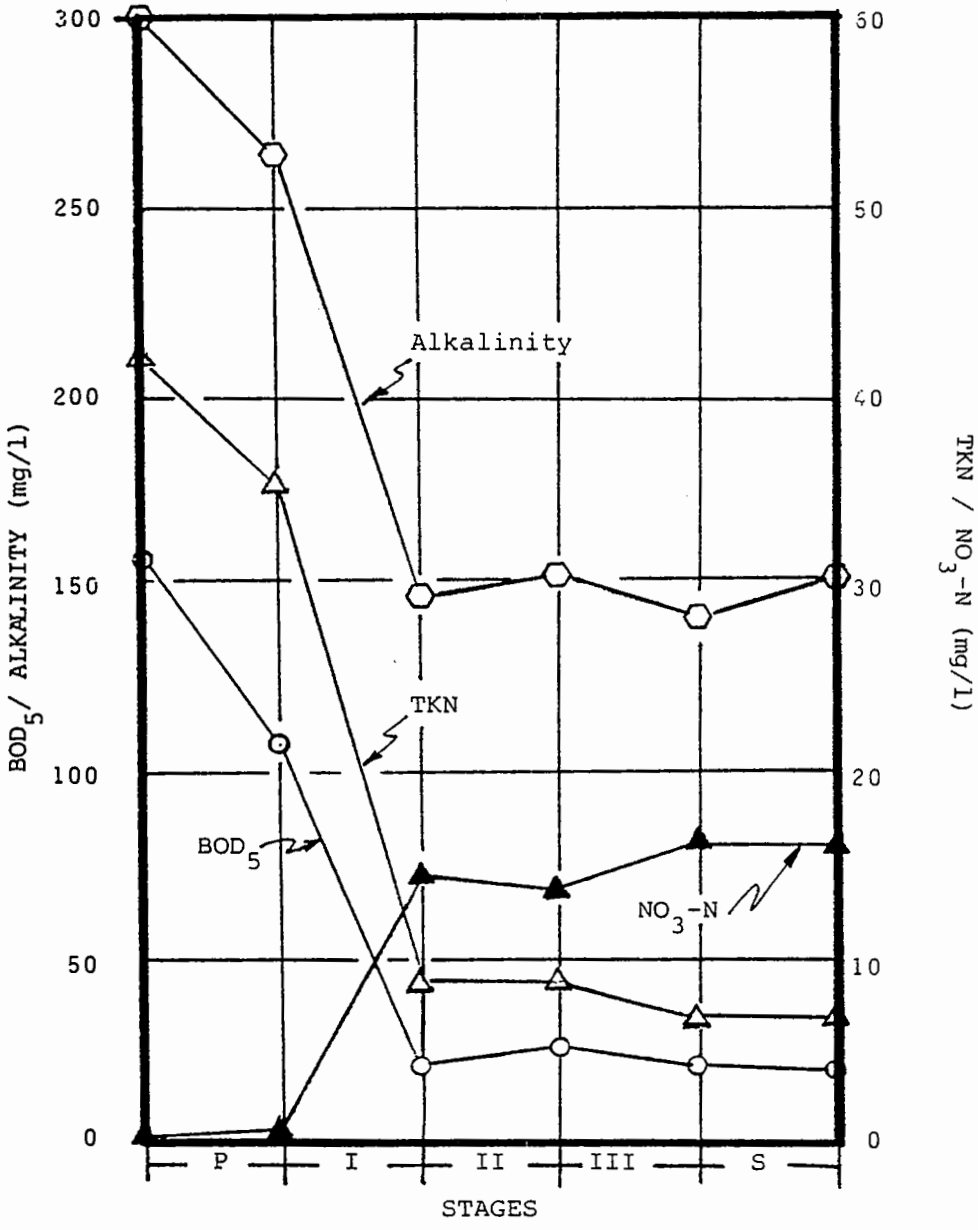

Figure 1 - Metabolic Responses on Oxidation of Carbonaceous and Nitrogenous Compounds Under Normal Operating Conditions

the normal operating conditions. It is conceivable that the first stage contactor alone will be able to treat the same characteristics of sewage at the design flow of 0.5 mgd under the organic loading condition of 4 lbs. BOD_5 /day/ 1000 sq. ft. The question arises as to whether this single-train system with three stages in series was over-designed or was intended to provide the necessary redundant capability for operations. It is of the opinion that if the structural reliability is sound, the second stage contactor should be incorporated into the single-train system with a reserved capacity to treat the unexpected peak or concentrated wastewaters; however, if the structural reliability becomes questionable, there is no room for criticisms against a single-train system equipped with three stages of contactors.

Under Abnormal Operating Conditions

The rotating biological contactor system at the St. Michael Wastewater Treatment Plant has experienced both mechanical problem and structural failure.

Approximately one year after the system was installed, the first stage contactor's shaft bearings had to be replaced. The suspected cause of the bearing failure was thought to be due to the drainage of the mixed liquor down on the shaft and into the bearings. The problem was corrected by putting a bead of silicone rubber around the shaft to divert the mixed liquor from entering the bearings.

The system had been operated in a satisfactory manner for two years until a severe structural failure developed in the early winter of 1981. The tie rods holding the individual polyethylene discs of the first stage contactor began to shear and dismember the disc assembly. This problem caused noise and shaft vibration and the unit was taken out of service as the result. Nevertheless, in an attempt to alleviate the possible development of a differential torque applied to the shaft caused by non-uniform microbial growth, it was managed to operate the first stage contactor for 10 minutes twice daily under the stressed crippling conditions.

During the down-time, the primary effluent continued passing through the first stage compartment. The principal responsibility of wastewater treatment depended to a great extent upon the second stage contactor. The metabolic responses to the abnormal operating conditions shortly after shut-down of the first stage contactor are shown in Figure 2

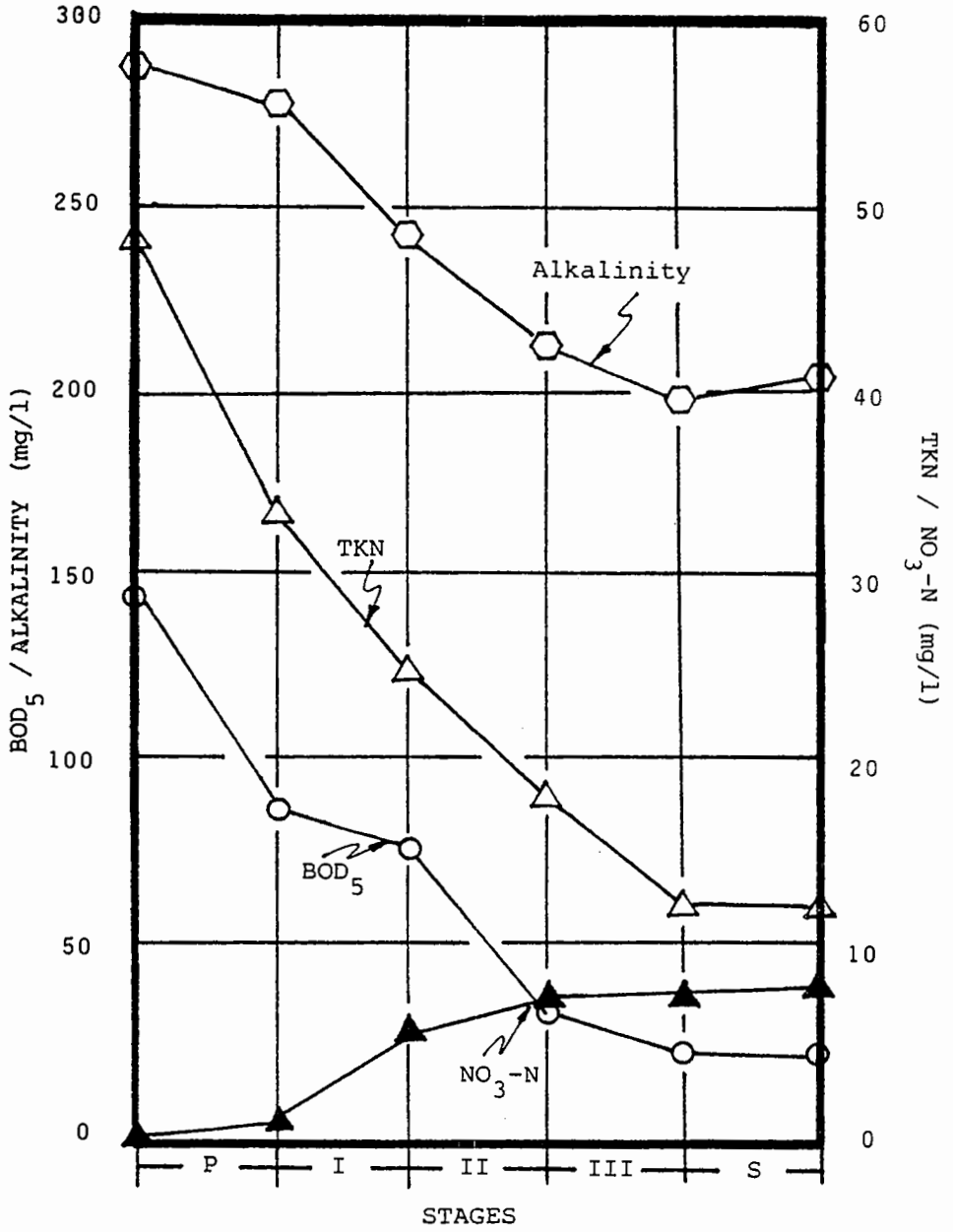

Figure 2 - Metabolic Responses on Oxidation of Carbonaceous and Nitrogenous Compounds Shortly After Shut-down of the First Stage Contactor

which indicates that the carbonaceous and nitrogenous removal rates were considerably low at the immediate juncture of the transition state. It is also noted in Figure 3 that the rates of BOD removal and nitrification achieved by the second stage contactor were lower than the corresponding rates accomplished by the first stage contactor under normal operating conditions. Notwithstanding the disruption of the first stage contactor operations, the over-all efficiency of the system performance in every respect remained exceptionally high. Such an accomplishment of a high degree treatment should be credited to the second stage contactor and the third stage contactor as well.

Since the shut-down of the first stage contactor, it was found that the bio-mass on both second stage and the third stage contactors was gradually developed. This natural phenomenon reflected higher organic loadings being applied to them.

In order to prevent an anaerobic environment from development and to prevent sedimentation from taking place in the first stage compartment, the operation personnel decided to remove the partition between the first stage and the second stage compartments two weeks after shut-down of the first stage contactor. This arrangement would permit fluxing the wastewaters in a common compartment in which oxygen was supplied and sedimentation was prevented as a result of the second stage contactor operations.

The metabolic responses to the operation improvement, as presented in Figure 4, illustrate that the metabolizable components of the carbonaceous and nitrogenous compounds were readily removed in the common compartment of the first stage and the second stage contactors. However, the low temperature at 9°C either curtailed the capability or diminished the population of the autotrophic micro-organisms to achieve nitrification. A slight reduction of ammonia nitrogen was reasoned on the grounds for supporting microbial growth in the processes of catabolism and bio-synthesis.

Several weeks later, the second stage contactor experienced the same problem. It was decided that the entire system should be taken out of service and repaired.

Modes of Substrate Removal

The primary effluent contains approximately 100 mg/l of BOD_5 in which 25 mg/l to 35 mg/l are soluble and 65 mg/l to

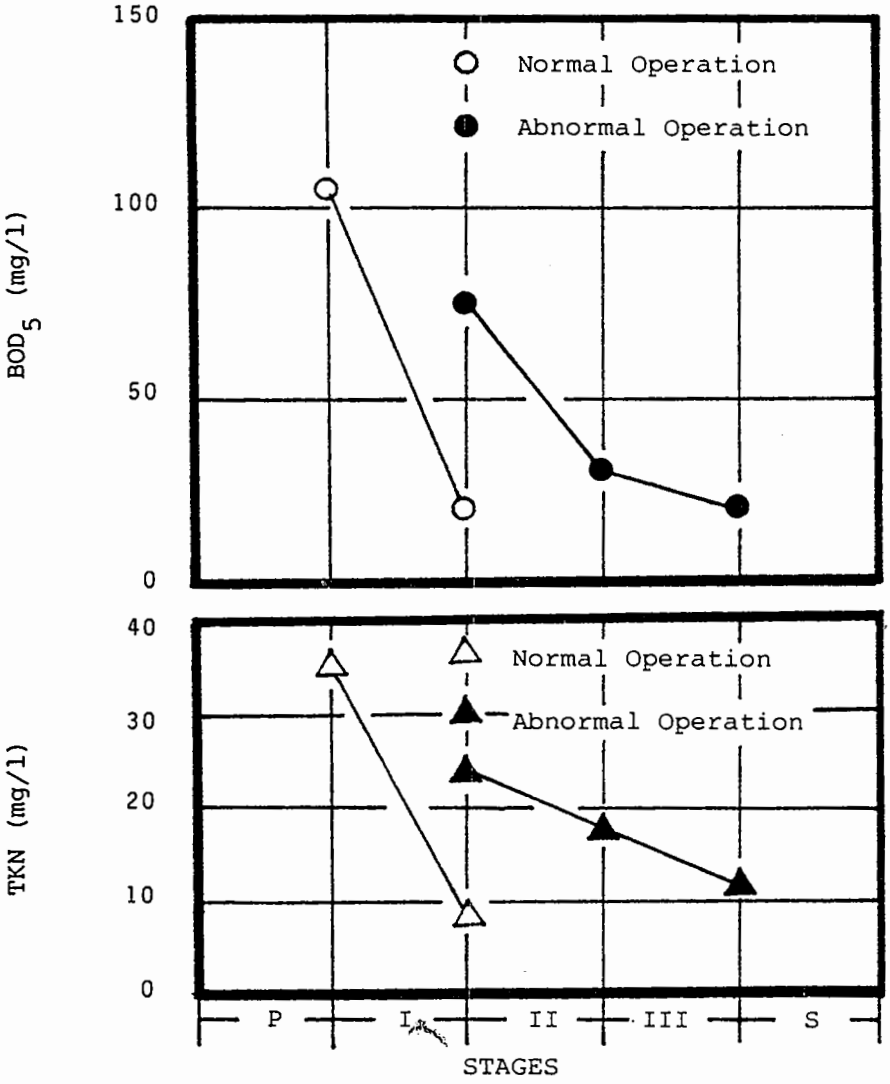

Figure 3 - Comparison of Metabolic Rates: First Stage Performance Under Normal Operating Conditions versus Second Stage Performance Under Abnormal Operating Conditions

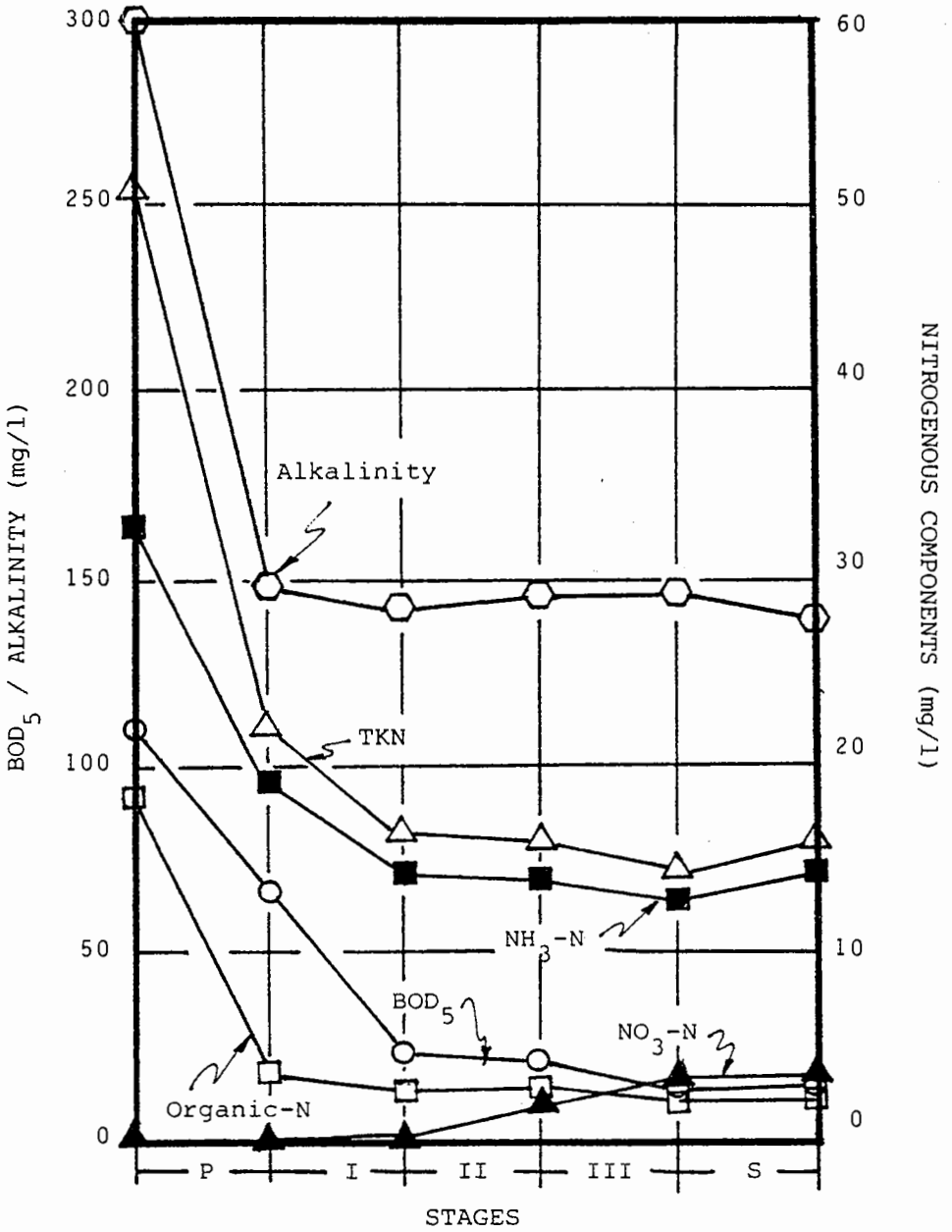

Figure 4 - Metabolic Characteristics After Removal of Partition Between First Stage and Second Stage Compartments

75 mg/l are insoluble. It also contains approximately 30 mg/l of Total Kjedahl Nitrogen (TKN) in which 20 mg/l to 25 mg/l are ammonia nitrogen and 5 mg/l to 10 mg/l are organic nitrogen. Since the heterotrophic and the autotrophic micro-organisms contained in the bio-mass are not distinguishable, the loadings cannot be meaningfully expressed on the mass ratio basis. The term expressed as "lb./day/1000 sq. ft." for the various loading conditions applied to the first stage contactor are given in Table II.

Table II - Various Loading Conditions Applied To The First Stage Contactor At 0.25 MGD

Constituents	Loadings ($lb./day/10^3$ sq. ft.)
Soluble BOD_5	0.5 to 0.7
Insoluble BOD_5	1.3 to 1.5
NH_3-N	0.4 to 0.5
Organic - N	0.1 to 0.2

Nitrosomonas and Nitrobacter are chemosynthetic nitrifiers, a kind of autotrophic micro-organisms. The biosynthesis is undertaken through utilization of energy supplied by oxidation of ammonia. On the contrary, the heterotrophic micro-organisms metabolize organic carbon as well as nitrogen and release nitrogen as ammonia which can be further oxidized by nitrifiers. The degree of nitrification of a heterogeneous microbial system is the measurement of the nitrifiers' capability to convert TKN into nitrate.

Under the normal operating conditions, as shown in Figure 5, the heterotrophic micro-organisms on the first stage contactor swiftly remove carbonaceous compounds of the constituents in forms of soluble BOD or insoluble BOD, while, organic nitrogen remained essentially untouched. At the same time, nitrifiers readily oxidized ammonia nitrogen. With respect to carbonaceous metabolism, the result indicates that the carbonaceous compounds required for the heterotrophic micro-organisms exceeded the amount of soluble BOD available as the low concentrations of soluble BOD failed to exert inhibitory effects on microbial utilization of insoluble BOD. Consequently, soluble BOD and insoluble BOD were removed concurrently. On the other hand, the pattern of nitrogen metabolism displayed a sequential mode. This phenomenon can be deduced as the result that the amount of

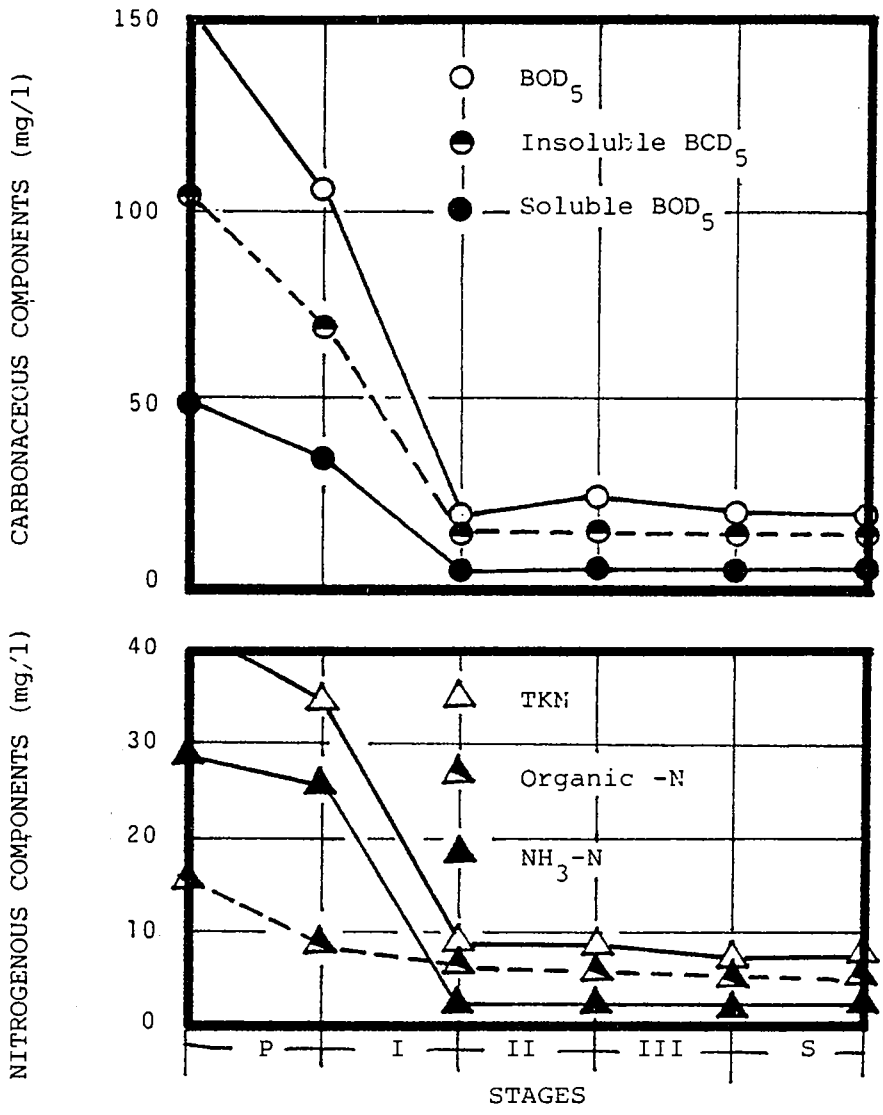

Figure 5 - Metabolic Responses on Oxidation of Carbonaceous and Nitrogenous Compounds Under Normal Operating Conditions

ammonia nitrogen in excess of what was required for metabolic needs inhibited the heterotrophic micro-organisms from further degradation of organic nitrogen.

Under the abnormal operating conditions, before the bio-mass on the second and the third stages of contactors was fully developed, the microbial activities decreased significantly. The slow metabolic rates provided an avenue to gain insight into the microbial behavior on the mode of substrate removal. As shown in Figure 6, it is evident that removal of insoluble BOD took place immediately after soluble BOD had been utilized. This mode of sequential substrate removal reflected the metabolic responses from the samples taken when the first stage contactor was not in service. The submerged heterotrophic micro-organisms utilized soluble BOD and by-passed insoluble BOD to the second stage contactor, where soluble BOD was not available and the heterotrophic micro-organisms must metabolize insoluble BOD for survival.

Figure 7 portrays a similar sequential mode of nitrogen metabolism as that illustrated in Figure 5. It is reasonable to conclude that only an inappreciable amount of organic nitrogen removal can be expected by the rotating biological contactor process when the wastewater contains an excessive amount of ammonia nitrogen. The inherent nature of a short detention time provided for biological treatment of wastewater also plays an important role in limiting microbial degradation of organic nitrogen. The combined effect of metabolic inhibition and short reaction time causes removal of organic nitrogen ineffective. Even if the environmental factors favor nitrification, achievement of nitrification in a large measure depends upon the amount of organic nitrogen contained in the wastewaters. In order to assure a greater degree of nitrification, organic nitrogen should be removed by the sedimentation process which proves to be the most effective and simplest means of treatment.

Evaluation of Kinetics

There are two unique features imbeded in the rotating biological contactor treatment process: (1) the predominating micro-organisms differ from one stage to another due to substrate gradient distribution, and (2) the mixed liquor in each stage of compartment displays a complete mix system due to a through agitation in a confined reactor. With these two inherent features coupled with a continuous flow pattern, assessment of kinetic performance within a specific stage of contactor becomes a matter of art of which beauty is in

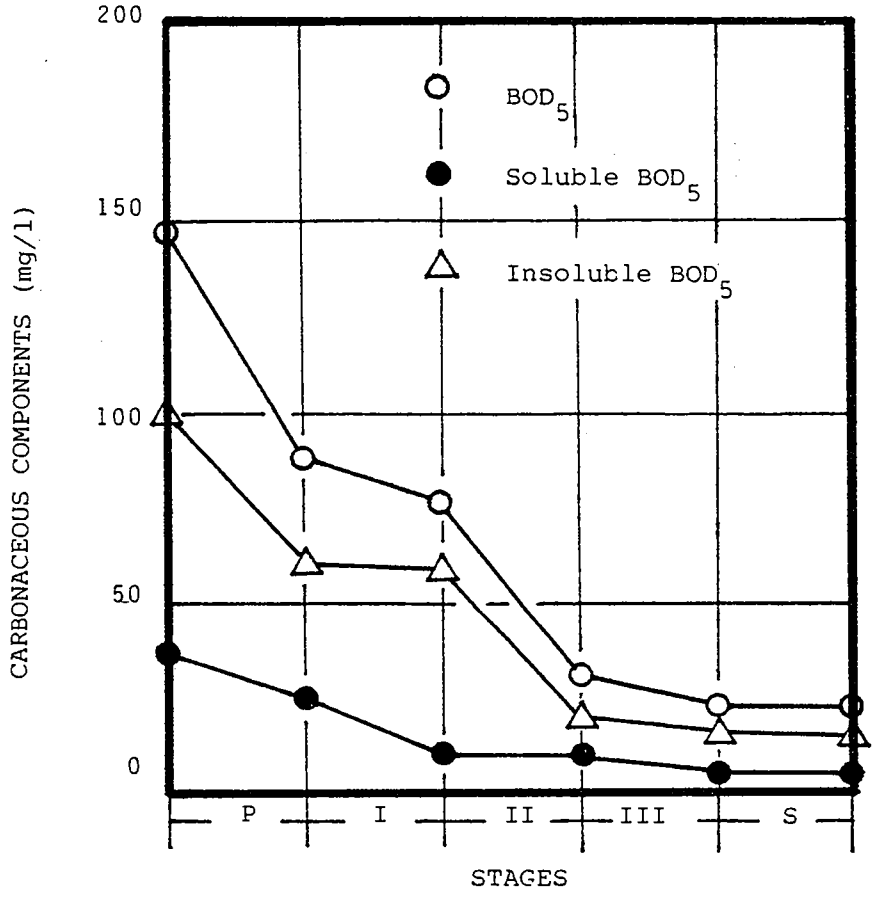

Figure 6 - Metabolic Responses on Oxidation of Carbonaceous Compounds by Heterotrophic Micro-organisms Under Abnormal Operating Conditions

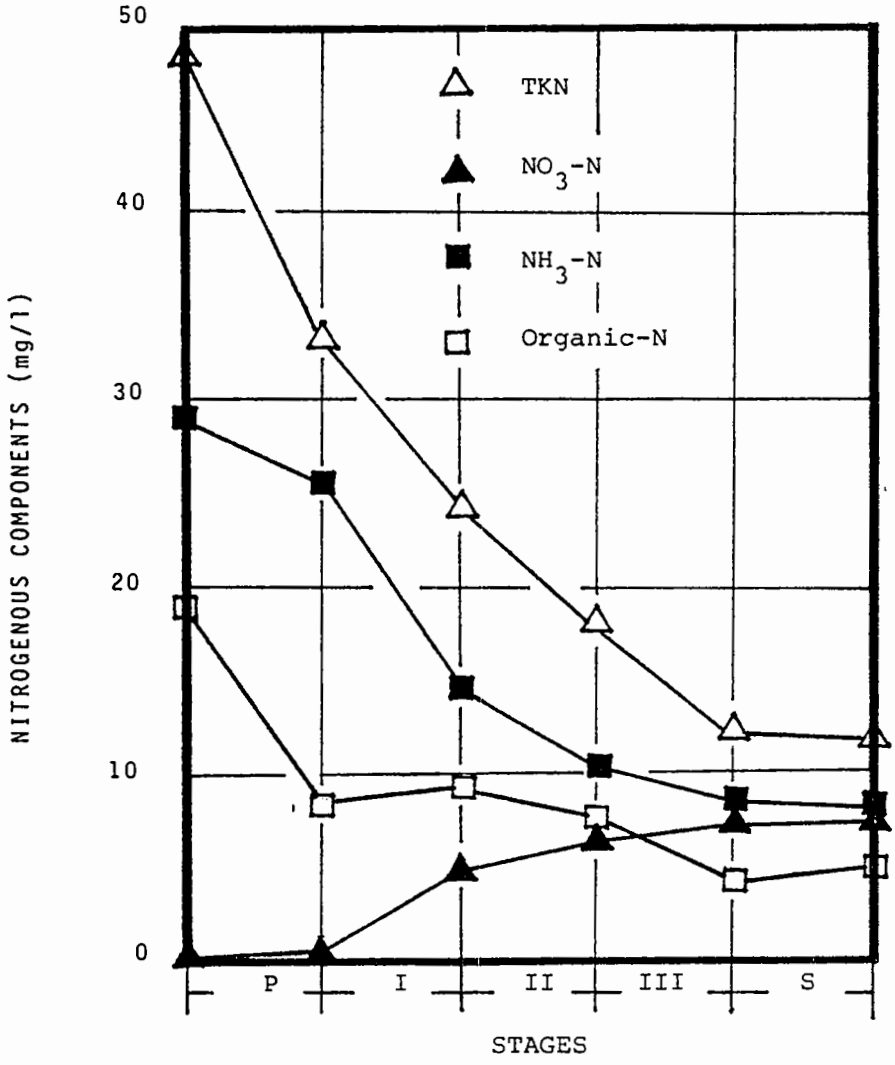

Figure 7- Metabolic Responses on Oxidation of Nitrogenous Compounds by Autotrophic Micro-organisms Under Abnormal Operating Conditions

the eyes of the beholder. The metabolic rates among stages portray responses of various predominating groups of microbial population to specific loading and environmental conditions. The environmental factors and the characteristics of the wastewaters which vary from time to time determine selection of certain predominating microbial species to grow on various stages of the contactors. Unless those influencing elements can be properly controlled, the kinetic order merely reflects the shape of a specific metabolic rate curve and the kinetic value simply stands for a numerical figure. No meaningful engineering application in the process design for wastewater treatment is expected.

The curves plotted in Figure 1 through Figure 7 are illustrations of the concentration changes in wastewater constituents from stage to stage. A line between two points where a slope exists, should not be construed as an implication of a gradual decrease or increase in the concentration of a specific constituent, because each compartment is a completely mixed reactor in which the concentration gradient does not exist. In order to convey this concept, all data points shown in Figure 1 and Figure 2 are respectively plotted in Figure 8 and Figure 9. The sampling points on the designated line number are explained below:

Line Number	Location of Samples Taken
L_1	Primary Influent
L_2	Primary Effluent
L_3	First Stage Compartment
L_4	Second Stage Compartment
L_5	Third Stage Compartment
L_6	Secondary Effluent

The primary and the secondary clarifiers are designed on the plug flow pattern. The changes in the concentration gradient are best represented by the lines connecting data points on L_1 and L_2 or L_5 and L_6. Nevertheless, the representative lines for the rotating biological contactors' performance should be drawn horizontally from points on L_3 to L_2, L_4 to L_3 and L_5 to L_4, and then vertically connecting points on L_2, L_3 and L_4 to where the horizontal lines intersect. In agreement with this concept, the metabolic responses should reflect zero order kinetics.

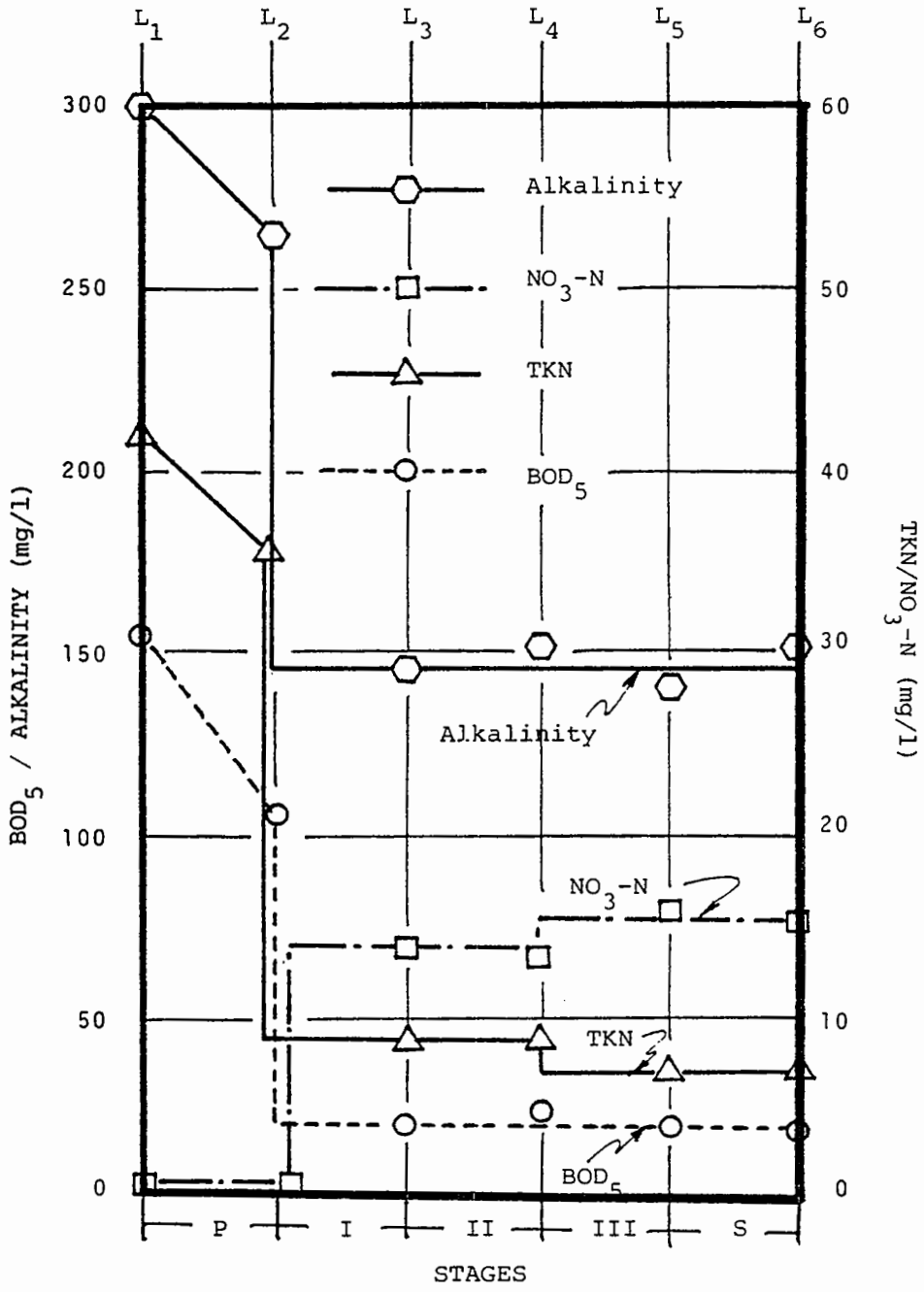

Figure 8 - Kinetic Performance at Various Stages Under Normal Operating Conditions

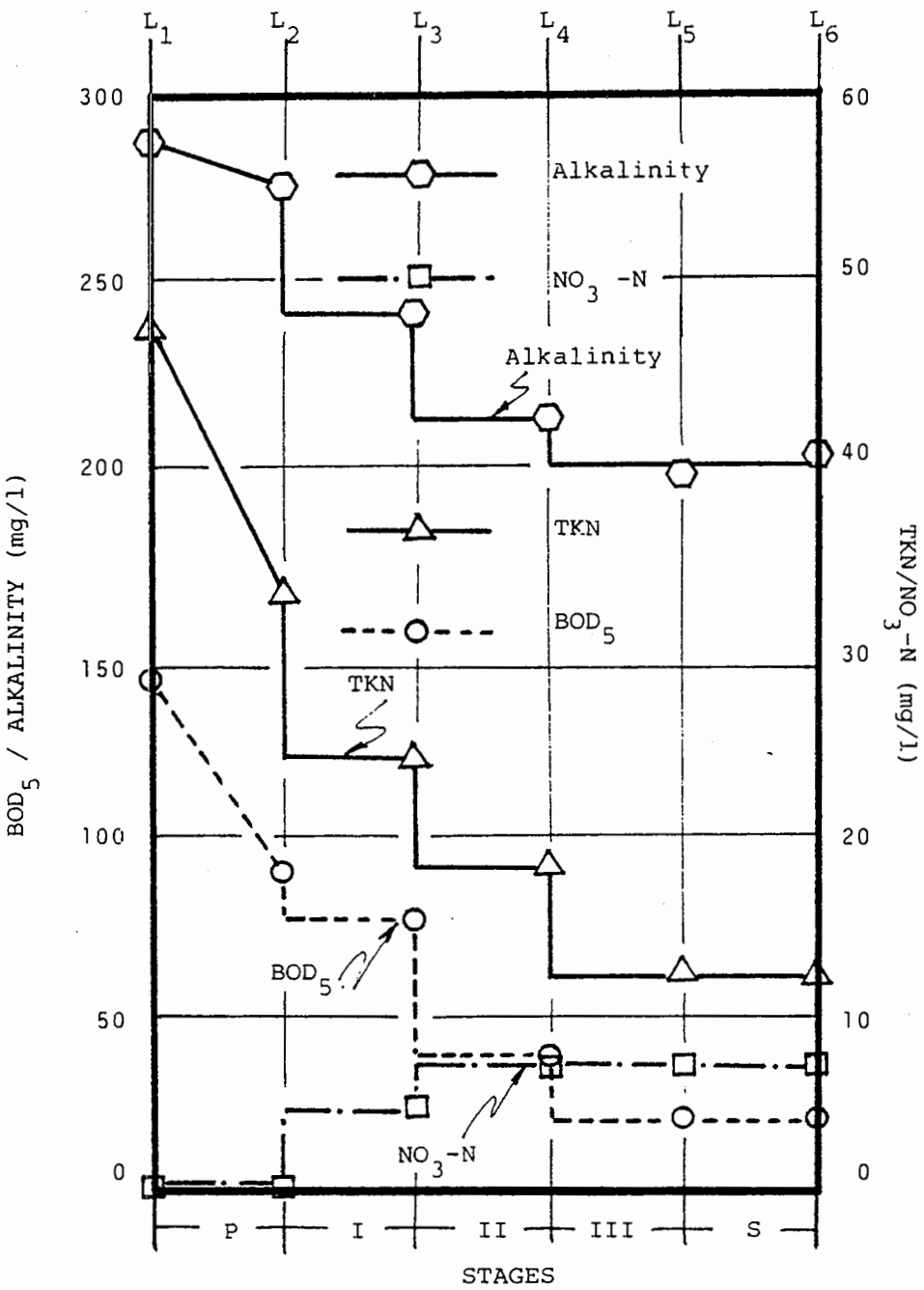

Figure 9 - Kinetic Performance at Various Stages Under Abnormal Operating Conditions

No matter how the rotating biological contactor process is designed, the engineers can control neither waste characteristics nor environmental factors. The design should therefore be based on the operation experience as well as the emperical equation to size the unit.

Wu and Smith (1) developed an emperical model based on full scale operations to predict the over-all system performance and assist engineers in the process design. The Wu's model as shown below describes the relationship between percent BOD removal and percent BOD remaining as a function of process variables including surface hydraulic loading, influent soluble BOD concentration, number of stages, and temperature.

Wu's Model

$$F = \frac{14.2 \times q^{0.5579}}{e^{0.32N} \times L_o^{0.6837} \times T^{0.2477}} \quad \text{-- Equation 1}$$

where,

F = fraction of influent soluble BOD remaining in the effluent, %

q = surface hydraulic loading, gpd/ft^2

N = number of stages

L_o = influent soluble BOD concentration, mg/l

T = temperature, ^oC

If the Wu's model represents a general characteristic of the system performance, the relationship among variables should be independent on the number of stages. Therefore, the model can be generalized as Equation 2.

$$F_n = \frac{14.2 \times q^{0.5579}}{e^{0.32} \times L_n^{0.6837} \times T^{0.2477}} \quad \text{-- Equation 2}$$

where, n = footnote referring to stage number

The number of stages can be determined by repeating calculations of Equation 2 until L_{n+1} meets the discharge quality limitations.

For example:

1st stage - Use L_1 to find F_1

$L_2 = L_1 \times F_1$

2nd Stage - Use L_2 to find F_2

$L_3 = L_2 \times F_2$

N stage - Use L_n to find F_n

$L_{n+1} = L_n \times F_n$ = discharge quality limitation

When Equation 1 is rearranged to solve N, Equation 3 is obtained.

$$N = 3.125 \times \log_e \frac{14.2 \times q^{0.5575}}{F \times L_o^{0.6837} \times T^{0.2477}} \text{ -- Equation 3}$$

A paradoxical relationship between N and L_o is found in Equation 3, i.e., the number of stages required decreases as the concentration of soluble BOD increases, when variables q, F, and T are constant. This relationship can be explained by the fact that the higher concentration of influent BOD stimulates higher microbial activities and the percent of BOD remaining can be easily maintained. As a result, it requires fewer contactors for treating wastewaters with higher concentrations of BOD than the number of contactors needed for treating wastewaters with lower concentrations of BOD in order to achieve the same degree of percent BOD reductions.

Clark, Moseng and Anaso (2) developed a complete - mix model and claimed that the principle of Monod's Equation should also apply to each stage of the contactor at the steady state.

Clark's Model

$$A_W = F(S_o - S_1)\left(\frac{K_s}{P} \times \frac{1}{S_1} + \frac{1}{P}\right) \text{-- Equation 4}$$

$$P = (\mu_{max} / Y_a)\, X_a \text{ ---------- Equation 5}$$

where, A_W = wetted area of bio-disc, m^2

F = wastewater flow rate, l/s

S_o = influent substrate concentration, mg/l

S_1 = effluent substrate concentration, mg/l

K_S = the Monod half-velocity coefficient, mg/l

P = area capacity constant, the amount of substrate removed per day per unit surface area of disc

μ_{max} = maximum specific growth rate for the attached bio-mass / day

Y_a = apparent yield of suspended oraganisms

X_a = concentration of suspended organisms, gm/m^2

When Equation 4 is rearranged to solve S_1, Equation 6 is obtained.

$$S_1 = \frac{[A_W P + FK_S - FS_o)^2 + 4F^2 S_o K_S P]^{\frac{1}{2}}}{2F} - \frac{(A_W P + FK_S - FS_o)}{2F} \text{ --- Equation 6}$$

Equation 6 can be generalized and expressed as Equation 7.

$$ {}_nS_1 = \frac{[(A_w P_n + F\ {}_nK_s - F\ {}_nS_o)^2 + 4F^2\ {}_nS_o\ {}_nK_s\ P_n]^{\frac{1}{2}}}{2F} $$

$$ - \frac{(A_w\ P_n + F\ {}_nK_s - F\ {}_nS_o)}{2F} \text{ -- Equation 7} $$

Where, n = footnote referring to stage number.

The number of stages can be determined by repeating calculation of Equation 7 until ${}_nS_1$ meets the discharge quality limitation.

Application of these models is restricted to the soluble BOD system. Such a restriction brings about a serious question as to their validity for design purposes, when insoluble BOD must be removed and the ratio of soluble BOD to insoluble BOD is not available. The complications are further extended to the system where sequential substrate removal occurs.

In application of the Clark's model, the fundamental problem lies in the fact that the rotating biological contactor process has never been operated under a steady state. Consequently, μ_{max}, K_s, X_a, and Y_a cannot be easily determined within a reasonable range of accuracy. In fact, the information relevant to μ_{max}, K_s, X_a and Y_a may not be available at the design stage.

The manufacturers (3) published various charts which correlate mass loading with hydraulic loading to predict effluent quality under specific influent wastewater characteristics and temperature conditions. These charts have been widely acceptable because of their simplicity in usage. The charts were developed on the assumptions that insoluble BOD and soluble BOD would be removed concurrently at the same rate, and the ratio of these two components was 1. These assumptions may not present a problem for domestic wastewater treatment design because of low substrate concentrations in both insoluble BOD and soluble BOD. However, it is hard to comprehend that the same charts can also be applicable to the

design of an industrial wastewater treatment process without a pilot plant study.

In accordance with the design procedures, the total surface area required is calculated by dividing the design hydraulic loading (gpd/ft^2) into the average design flow (gpd) The hydraulic loading is in turn figured from a chart showing hydraulic loading rate (gpd/ft^2) vs. effluent BOD concentration (mg/l). There are two linear relationships existing between hydraulic loading rate and effluent concentration: one above 15 mg/l of soluble BOD and the other below 15 mg/l of soluble BOD. The design manual did not explain why the number of 15 mg/l was so magic as to render the microbial population to behave differently in the process of metabolism. The existence of linear relationship claimed by the manufacturer is principally in contradition to the Wu's and the Clark's models.

If the required total surface area is proportional to the hydraulic loading rate, it implies that the microbial population will uniformly grow on the surface of the contactors and the metabolic rates will be identical among the stages. Of course, the manual for design purposes may not be intended to address the kinetic matter. Nevertheless, it may consequently overload the up-stream stages and underload the down-stream stages of the contactors. In order to achieve the most cost-effective design and the most efficient operation, the flow distributions into parallel trains must be carefully arranged. For example, the treatment capacity of a system consisting of 4 stages in 2 trains is not as great as a system consisting of 2 stages in 4 trains. The latter arrangement not only distributes a great magnitude of the organic loading into the four first stage contactors of which the operation reliability can be backed up by the four second stage contactors. In addition, it may also avoid the overloading condition imposed upon the four first stage contactors.

All models were developed under different theoritical assumptions. It is impossible to correlate and express them in an explicit mathematical language. However, the model makers confidently insist that the rotating biological contactor wastewater treatment plants can be easily and precisely designed and performed in accordance with the models. This conclusion may be statistically correct without guaranty, because there are numerous factors uncontrollable. The importance of the water pollution abatement program is what quality of the plant effluent discharges, not what model is based for

the plant design.

Historically, many models have been developed for the design of the activated sludge process. With the valuable information given, engineers still felt uncomfortable to use a specific model because they must consider all factual conditions and include built-in redundancy as required by the governmental guidelines or regulations. As a result, almost all designs on the activated sludge process followed the established criteria of the organic and hydraulic loadings. If the history repeats itself, engineers are bound to adopt the same kind of criteria published by the manufacturers for the future design of the rotating biological contactor process.

Recommendations for Structure Design

In the name of cost-effectiveness, the public has been led to believe that the rotating biological contactor technique would be a dependable wastewater treatment process. In fact, the application history has been too short to assess its success or to condemn its failure, especially many systems in operations have not reached the design loading conditions. At the early stage of the market promotion, few consulting engineers undertook stress range analysis of the rotating biological contactor structures. This caused general concern as well as disappointment of the technique dependability.

The reliable plant performance lies in the structural dependability of which the importance cannot be over-emphasized. Historically, the shaft has been the main issue of the problem. In order to meet the quality of the structural design, it is strongly recommended that the maximum stress range for the main central shaft, stub shafts and all weldments to the shaft shall not exceed the allowable values defined under American Welding Society Inc.'s Structural Welding Code - Steel, AWS D 1.1 - 81 for a minimum fatigue life of 25 years. The stress range is defined as the peak-to-trough magnitude of stress fluctuations. In the case of stress reversal where the rotating biological contactor shaft applies, the stress range shall be computed as the numerical sum (algebraic difference) of maximum repeated tensile and compressive stresses, or the sum of shearing stresses of opposite direction at a given point, resulting from changing conditions of load. The stress range shall be determined using calculated dead loads, torsion loads, and live loads corrected for buoyancy using actual media percent submergences and the appropriate AWS projected curve category for the

tubular structures, as outlined in AWS D 1.1 - 81, Chapter 10, Section 10.7. The live loads shall be based on a bio-mass thickness of 0.125" for the standard density contactors and 0.075" for the high density contactors. The most important of all is that the manufacturer shall submit the design calculations to the consulting engineers at the time of the shop drawing approval to substantiate compliance.

Failures associated with the media have also been reported. An equal distribution of flows to various trains of the rotating biological contactor system should help alleviation of developing a thick layer of bio-mass on the contactor media and help preservation of the media stiffness. The manufacturers for the sake of business survival should improve the media durability and resistance to temperature as well.

CONCLUSIONS

The rotating biological contactor process has demonstrated its capability in removal of soluble BOD and oxidation of ammonia. When the metabolic rates are high, soluble BOD and insoluble BOD can be removed concurrently. However, when the metabolic rates are low, soluble BOD becomes a preferred carbonaceous component for metabolisms.

With respect to the microbial responses to the nitrogenous compounds, ammonia is readily oxidized or utilized by the microorganisms. While, organic nitrogen cannot be catabolized to an appreciable extent in the presence of ammonia in excess of the amount required for the metabolic needs. As nitrification takes place, oxidation of 1 mg/l of ammonia nitrogen consumes about 8 mg/l of alkalinity. When temperatures stay above 10 ^{o}C and other favorable environmental factors prevail, a complete oxidation of ammonia can be expected. On the other hand, when temperatures fall below 10 ^{o}C regardless of other environmental conditions, a complete oxidation of ammonia cannot be achieved.

In cognizance of the process limitations, cautions must be exercised in evaluations of its treatability toward removal of insoluble BOD and oxidation of organic nitrogen. It deems necessary to conduct a pilot plant study and determine if the rotating biological contactor is an applicable process for the treatment of industrial wastewaters.

The primary treatment is not a prerequisite in conjunction with the rotating biological contactor process, but the capability of primary clarifiers in removal of insoluble BOD and organic nitrogen is too great to be ignored. The rotating biological contactor process in line with the primary treatment

definitely improves the efficiency of the over-all plant performance.

The mixed liquor suspended solids generated from the rotating biological contactor process settle rapidly. A detention time of 30 minites proves to be adequate for the sedimentation purpose. In considerations of flow fluctuations, it is recommended that a detention time of 30 minutes be provided to accommodate the peak flow rate entering the secondary sedimentation process. This unique settling characteristic will result in cost savings for the construction of secondary clarifiers.

As wastewaters enter the rotating biological contactor process in a direction perpendicular to the shaft, the mixed liquor in each compartment represents a completely mixed system. The metabolic response to a certain substrate component in each compartment should follow zero order kinetics under a continuous flow condition. A great effort has been made to develop models for the design and operation guidance. Before a specific model is used for engineering applications, the model's practical implications and built-in limitations must be fully understood.

It is known to all that the metabolic activities are substantially high at the upstream stages, while, substantially low at the downstream stages. A good engineering practice requires the following considerations: (1) how to maximize the over-all performance efficiency, (2) how to minimize the unexpected organic over-loading condition, (3) how to prevent occurrence of the oxygen deficit condition, and (4) how to increase an additional redundant capability at a minimum cost. These ideal goals can be accomplished by promoting parallel treatment schemes through flow distributions to as many trains of contactors as possible, and by planning future expansions in phases as the need arises.

There is no doubt that the rotating biological contactor is one of the viable alternatives for the treatment of wastewaters. The past history in many instances has not proved its structural dependability. Manufacturers are urged to make all necessary improvements so that the technological reputation can be built on an unshakable foundation.

Engineers are indebted to their clients for the fiduciary reward in expectations of the service being rendered with the highest degree of professionalism. Responsibilities and obligations must be fulfilled at both the design and construction stages. The shaft design should meet the minimum requirements as outlined in AWS D 1.1 - 81, Chapter 10, Section 10.7. The live load should be calculated on the basis of a bio-mass thickness of 0.125" for the standard density unit and 0.075" for the high density one.

REFERENCES

1. Wu,Y.C.,Smith,E.D.,and Hung,Y.T.," Modeling of Rotating Biological Contactor Systems ", Biotechnology and Bioengineering, Vol. 12, pp. 2055-2064, 1980

2. Clark,J.H.,Moseng, E.M., and Asano,T.," Performance of a Rotating Biological Contactor Under Varying Wastewater Flow ", Journal Water Pollution Control, Vol. 50, pp. 896- 911, 1978

3. Autotrol Wastewater Treatment Systems - Design Manual, Autotrol Corporation, 1979

THE KINETICS OF ROTATING BIOLOGICAL CONTACTORS AT TEMPERATURES: 5°C, 15°C, AND 20°C

Abraham Pano. Culp-Wesner-Culp Consulting Engineers, Denver, Colorado.

E. Joe Middlebrooks. Newman Chair Professor, Department of Agricultural Engineering, Clemson University, Clemson, South Carolina.

INTRODUCTION

Rotating biological contactors (RBC) treating municipal wastewater have been shown to be efficient in carbon and ammonia nitrogen removal (1,2,3). In recent years in the U.S., the use of the RBC process has increased mainly because of the simplicity of operation and the low power consumption.

The design of RBC systems has been based primarily on empirical relationships between the pollutant removal efficiency and the hydraulic loading rates based on the total surface area of the RBC. Presently, the design hydraulic loading rates are adjusted by a safety factor for wastewater temperatures below 12.8°C (55°F) (4). The employed safety factor generally varies according to the RBC manufacturer recommendations, because of lack of established kinetic constants associated with RBC substrate removal at different temperatures. Also there is little information available concerning the effects of staging on the kinetic constants associated with RBC substrate removal.

The existing data from RBC studies generally indicate that the kinetics for carbonaceous substrate removal and

ammonia nitrogen removal are first order, with substrate limiting phenomenon (1,5,6,7,8,9).

Several studies (10,11,12) employed Monod kinetics to describe carbonaceous substrate removal in fixed film reactors. Kornegay and Andrews based their model on a constant amount of active attached biomass (10). Clark, Moseng and Asano (11) used 70 percent of the total attached biomass to determine the kinetic constants for Monod growth kinetics. Mikula (12) based his kinetic model on the total attached biomass and the biomass in suspension. Other investigators developed conceptual models (13,14,15) incorporating fundamentals of substrate and oxygen diffusion and biological reaction. Friedman and his co-workers (16,17) used a mass transport model to determine the kinetic constants of substrate removal in an RBC unit. Also ammonia nitrogen removal in RBC units was described either by Monod growth kinetics (18), or by mass transport models (19). Some of the studies mentioned above were conducted with synthetic substrate (10,16,17,18) and others at fluctuating wastewater temperature (11,12).

The general objective of this study was to determine the kinetics of carbon and ammonia nitrogen removal as a function of temperature in an RBC system treating domestic wastewater.

The specific objectives were:

1. To develop kinetic models for different processes associated with carbonaceous and ammonia nitrogen removal in the first and following stages of an RBC system.
2. To determine the kinetic constants for each process at each stage and each temperature.
3. To determine the effect of temperature on the kinetic constants.

MATERIALS AND METHODS

Four experimental rotating biological contactor (RBC) units were operated from late October, 1979, until mid-July, 1980, in the laboratories of Utah State University, Logan, Utah (20). The study was conducted in three consecutive phases at three different temperatures of 5°C, 15°C, and 20°C. Each phase was started with "clean" RBC units (without biomass). Table I contains a summary of the detailed dimensions of the RBC units employed during the three phases of the study.

Comminuted wastewater was collected at the Hyrum, Utah, wastewater treatment plant, and hauled to the laboratory for

TABLE I. SUMMARY OF THE DIMENSIONS OF THE RBC EXPERIMENTAL UNITS

Phase	1	2,3
Parameter		
Number of stages	4	4
Number of discs/stage	4	4
Discs diameter, cm	37.5	39.0
Inflation factor	1.37	1.37
Side discs diameter, cm	22.9	22.9
Total surface area/stage, m^2	1.375	1.474
Water volume/stage, liter	6	7
Submergence, %	33.3	35.5
Rotational speed, rpm	16	16

use as the influent to the RBC units. The wastewater was stored in a refrigerated tank with the temperature controlled at 2°C.

The experimental units were operated continuously at constant influent flow rates, constant wastewater percentage and constant temperature. The influent wastewater was maintained at a constant temperature, and the experimental RBC units were located in a constant temperature room to maintain the desired water temperature through the four stages of the RBC units. A schematic diagram of the experimental apparatus is shown in Figure 1.

Table II contains a summary of the operating conditions used during the study. Table III contains a summary of the mean liquid temperatures in the various stages of the four experimental units. There was a gradual decline in the liquid temperature due to evaporation heat losses as the wastewater flowed through the RBC units.

Table IV contains a summary of the mean pH values and dissolved oxygen concentrations measured in the various stages of the four experimental units. An examination of Table IV shows the units were operating as an aerobic biological system.

The influent to the system and the effluent from each stage was monitored by collecting 24-hour composite samples at 20-minute intervals during the period of steady-state operation. Temperature, dissolved oxygen and pH values were measured on grab samples.

The ampule technique (21) was used to measure both total and filtered COD. Nitrogen compounds (Kjeldahl, nitrate and

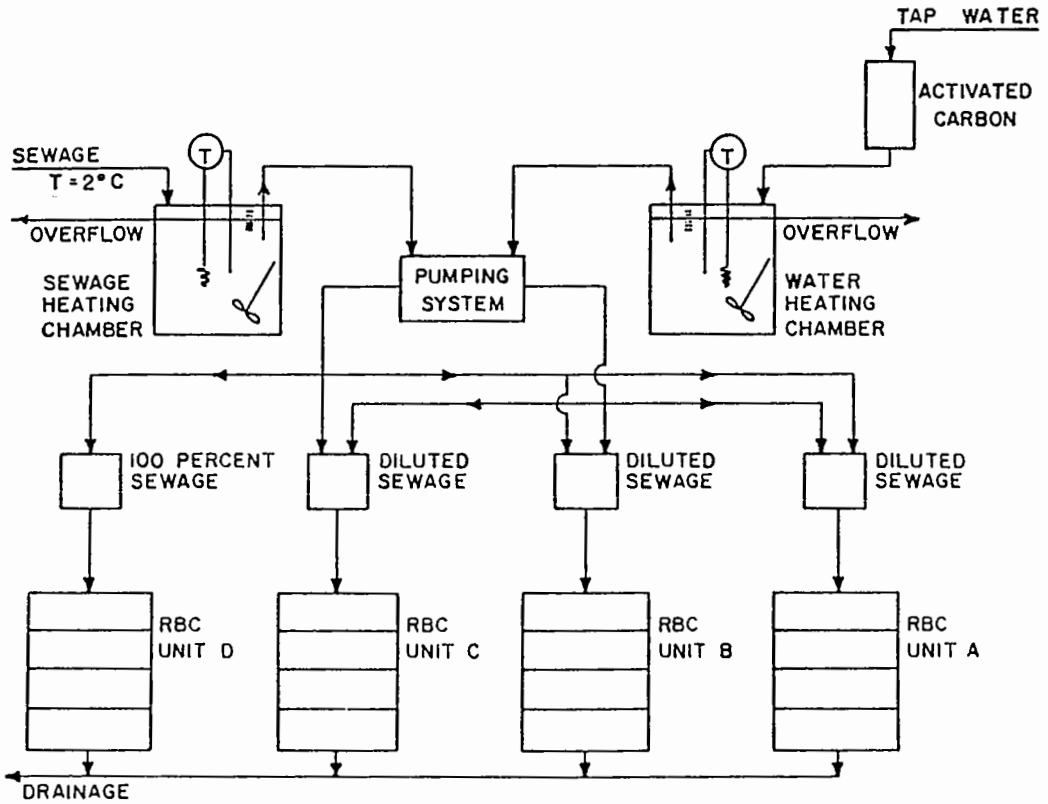

Figure 1. Schematic diagram of experimental apparatus.

TABLE II. SUMMARY OF MEAN STEADY-STATE OPERATING CONDITIONS

Temperature, °C	Parameter	Unit A	Unit B	Unit C	Unit D
5	Hydraulic loading rate, $m^3/m^2/day$ (gpd/sq ft)	0.049 (1.2)	0.048 (1.2)	0.050 (1.2)	0.051 (1.3)
	Organic loading rate, $gCOD/m^2/day$	5.76	4.13	7.08	8.90
	Influent COD concentration, mg/L	118.4	85.6	142.0	173.3
	Influent NH_4-N concentration, mg/L	13.34	9.69	15.98	20.27
15	Hydraulic loading rate, $m^3/m^2/day$ (gpd/sq ft)	0.050 (1.2)	0.052 (1.3)	0.051 (1.3)	0.053 (1.3)
	Organic loading rate, $gCOD/m^2/day$	3.98	7.50	9.88	13.92
	Influent COD concentration, mg/L	79.3	144.5	192.6	265.2
	Influent NH_4-N concentration, mg/L	7.70	14.76	22.30	29.79
20	Hydraulic loading rate, $m^3/m^2/day$ (gpd/sq ft)	0.048 (1.2)	0.048 (1.2)	0.049 (1.2)	0.050 (1.2)
	Organic loading rate, $gCOD/m^2/day$	6.92	9.73	12.51	13.97
	Influent COD concentration, mg/L	145.5	202.3	256.7	281.9
	Influent NH_4-N concentration, mg/L	10.00	13.00	17.50	22.30

TABLE III. SUMMARY OF THE OBSERVED MEAN AND RANGE OF VALUES FOR THE LIQUID TEMPERATURE (^{o}C) IN THE VARIOUS STAGES OF THE FOUR EXPERIMENTAL RBC UNITS

Phase	I		II		III	
Stage	Mean	Range	Mean	Range	Mean	Range
First	16.3	16.0-16.7	20.8	20.5-21.2	5.9	5.4-6.1
Second	15.4	15.0-15.6	20.3	20.0-20.7	5.1	4.7-5.3
Third	14.7	14.4-15.1	19.7	19.1-20.1	4.5	4.2-4.8
Fourth	14.4	14.1-14.8	19.3	18.6-19.8	4.1	3.8-4.5
Overall	15.2		20.0		4.9	

TABLE IV. SUMMARY OF MEAN PH VALUES AND DISSOLVED OXYGEN CONCENTRATIONS (MG/L) IN FIRST AND FOURTH STAGES OF THE RBC UNITS

Temperature, ^{o}C	5				15				20			
Unit	First Stage		Fourth Stage		First Stage		Fourth Stage		First Stage		Fourth Stage	
	pH	DO	pH	DO	pH	DO	pH	DO	pH	DO	pH	DO
A	7.97	7.6	8.22	9.4	7.80	5.0	8.00	7.8	7.95	3.6	8.13	6.9
B	8.03	8.8	8.23	9.6	7.70	3.9	7.90	7.5	7.98	2.9	8.08	6.5
C	8.00	7.9	8.25	8.8	7.70	3.6	7.73	7.9	7.95	2.4	8.05	6.5
D	8.03	7.4	8.27	8.3	7.73	2.7	7.68	7.1	7.80	1.9	7.98	5.8

nitrite) were measured with a Technicon Auto Analyzer II (22,23,24). Other analytical methods employed in this study were conducted according to Standard Methods (25). Four to five samples were collected for each stage effluent during the steady-state period. The influent was generally sampled ten times during a steady-state period.

At the end of each phase, the total amount of biomass attached to the discs in each stage was measured by weighing the discs and biomass after drying at 105°C and weighing the clean dried discs. Several samples were taken from the dried biomass to determine the VSS fraction as outlined in Standard Methods.

PROCESS PERFORMANCE

Attached Biomass

In each phase of the study after a week of operation, a thin layer of growth covered the discs in the first stages. Generally in the second week some biomass sloughing was observed in the first stages, and within a few days a new biofilm was built-up. After 3 to 4 weeks of operation, the discs in the first stages were covered with a thick, dark brown or grey biofilm, and further detectable changes in its appearance were not observed. The structure of the biofilm in the first stages seemed to be spongy, rather than a smooth structure. A filamentous growth in these stages may have been the reason for this type of structure.

In the successive stages, the discs were covered with a thinner biofilm layer and were relatively smooth in appearance. In the experiments at temperatures of 15°C and 20°C, the color of the biomass was tan-brown. In the experiment at 5°C, the biomass in the second through the fourth stages had a black-brown appearance. The tan color observed at 15°C and 20°C was probably due to growth of nitrifiers in these stages. Figures 2, 3 and 4 show the variation in the quantity of attached biomass in the four stages of the RBC units at 5°C, 15°C, and 20°C, respectively. In all three phases, there was a successive decrease in the quantity of biomass attached to the discs from the first to the fourth stages. At lower organic loading rates and higher temperatures, there was a sharp decline in the quantity of attached biomass following the first or second stages. At lower organic loading rates and higher temperatures, less substrate and less unstabilized

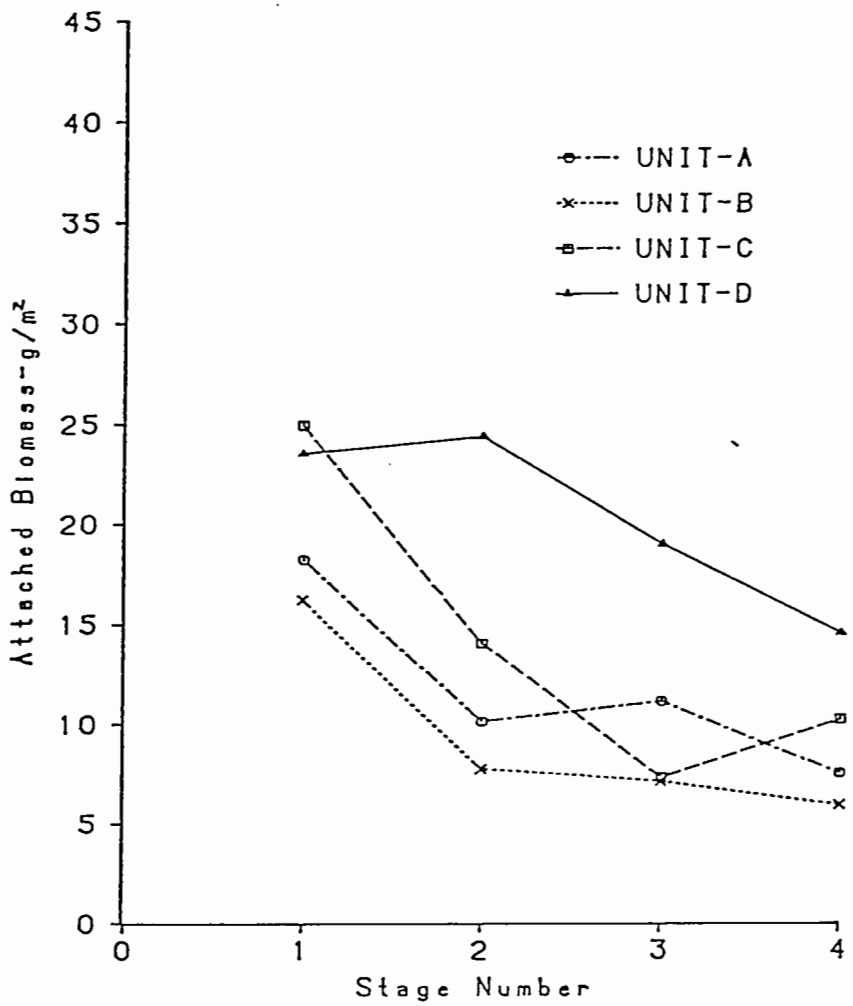

Figure 2. Attached biomass in the four stages of the RBC units operating at 5°C.

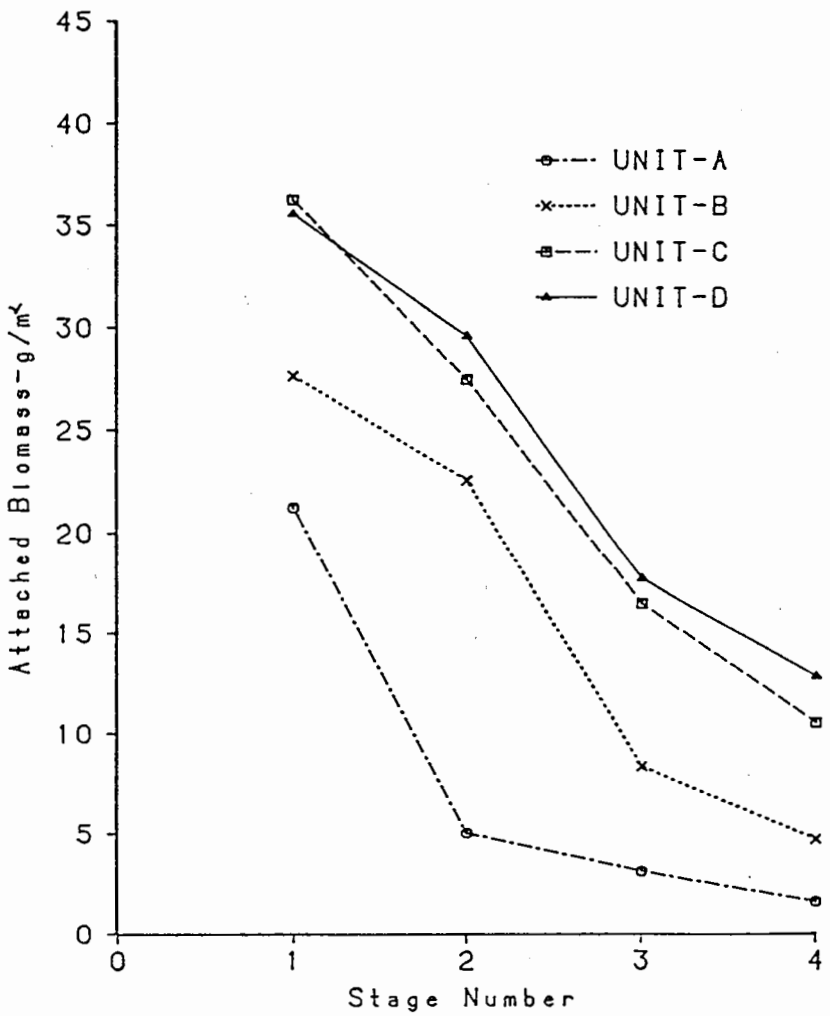

Figure 3. Attached biomass in the four stages of the RBC units operating at 15°C.

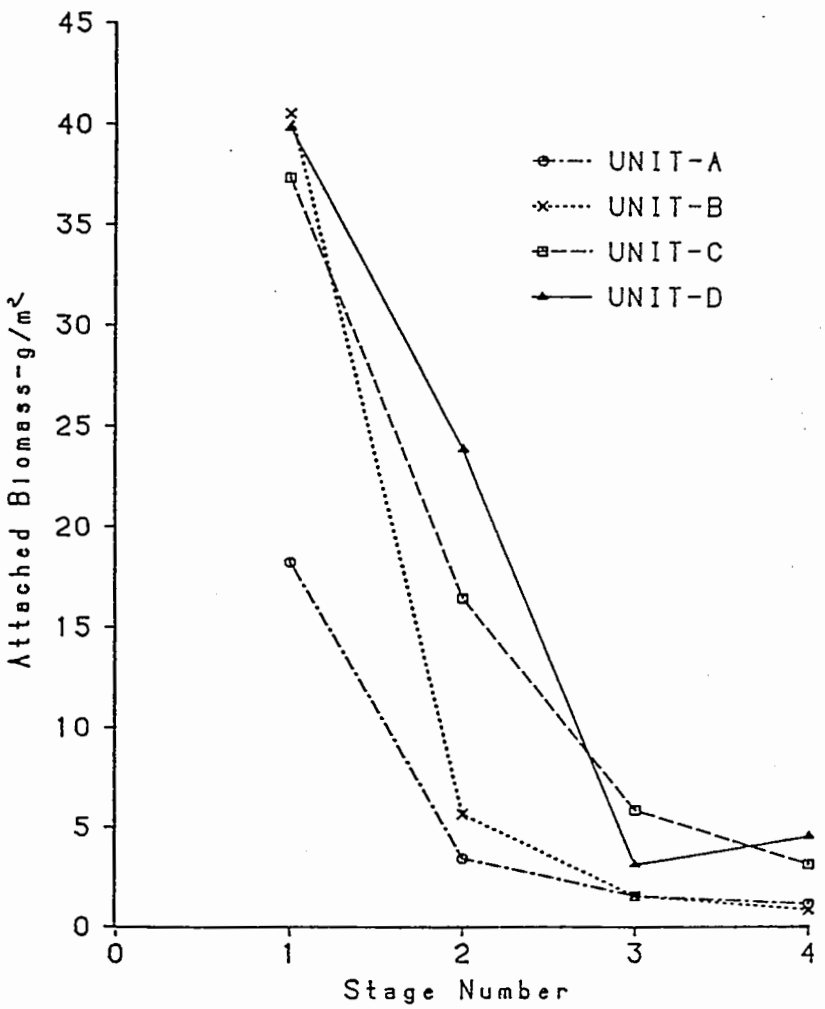

Figure 4. Attached biomass in the four stages of the RBC units operating at 20°C.

sloughed biomass were available to establish attached growth in these later stages.

Carbon Removal

Figures 5, 6, and 7 show the mean steady-state mixed liquor filtered COD concentrations when operating the RBC units at 5, 15 and 20°C, respectively. An analysis of Figures 5, 6 and 7 show that for the first stages of the units, the removal of filtered COD (influent filtered COD minus stage filtered COD) increased when the influent filtered COD was increased. This observation supports the contention that the removal of filtered COD can be described by a substrate limiting reaction. As the temperature was increased, the removal of the filtered COD increased, even beyond 15°C, which is contrary to the results reported by others (1). In stages two through four, there was further removal of filtered COD in the higher loaded units, but the removal rate per stage was much less than in the first stage. There was an inconsistent pattern of filtered COD removal in stages two through four, probably attributable to cell lysing. In the last stages of the RBC units, the differences in the effluent filtered COD for each of the four units were much smaller than those observed in the first stages.

Figures 8, 9, and 10 show the mixed liquor particulate (total-filter) COD, when operating the RBC units at 5°C, 15°C, and 20°C, respectively. As shown in Figures 8, 9, and 10, particulate COD removal occurred as the wastewater flowed through the stages, but with an irregular pattern of decline, due to the instability of the attached biomass of the last stages.

The removal of the influent particulate COD in the first stage, although the mixed liquor contained sloughed biomass, implies that the influent particulate COD is available substrate, as well as the soluble COD. Table V shows the substrate removal efficiencies when considering total COD as the available influent substrate and the filtered COD as the remaining substrate in the effluent from the RBC units.

A linear relationship exists between the overall removal of COD in terms of grams of COD removal per unit area and the influent substrate loading rate. The slopes of the relationships increased as the temperature increased: 0.811, 0.897, 0.976 for 5°C, 15°C, and 20°C, respectively (all are significant at a level of 0.01). The increase in slope shows the temperature dependency of the substrate removal performance.

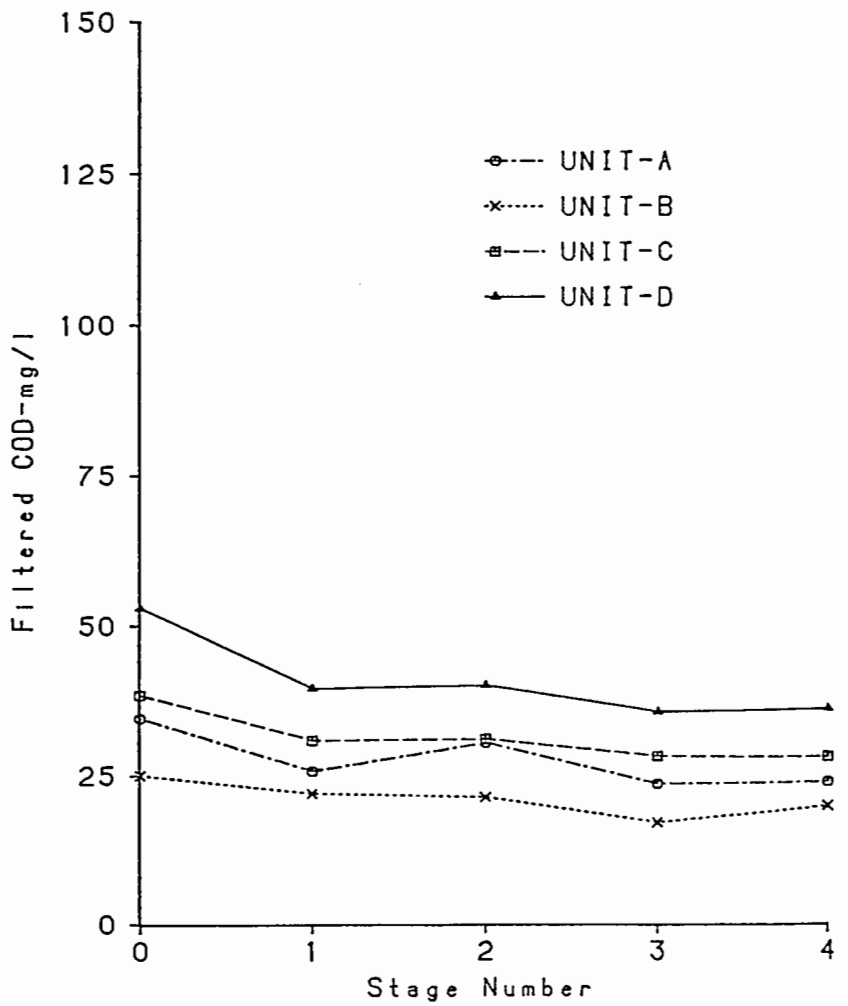

Figure 5. Mean steady-state mixed liquor filtered COD concentrations in the four stages of the RBC units operating at 5°C.

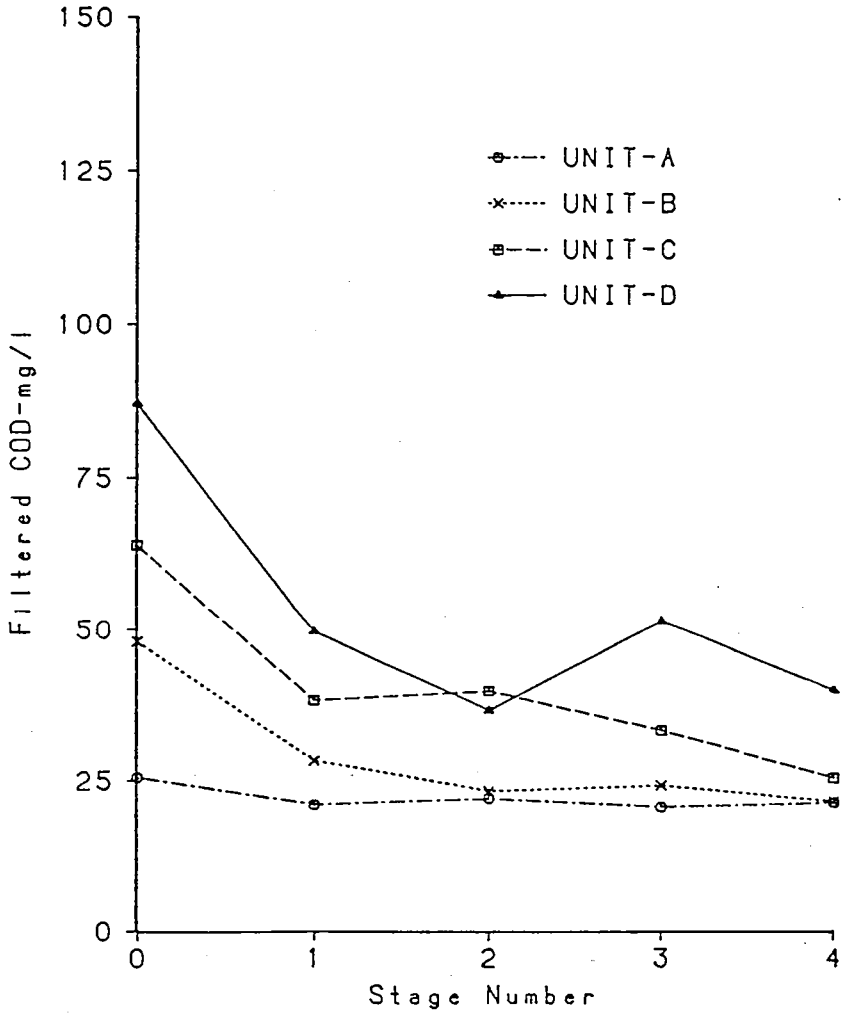

Figure 6. Mean steady-state mixed liquor filtered COD concentrations in the four stages of the RBC units operating at 15°C.

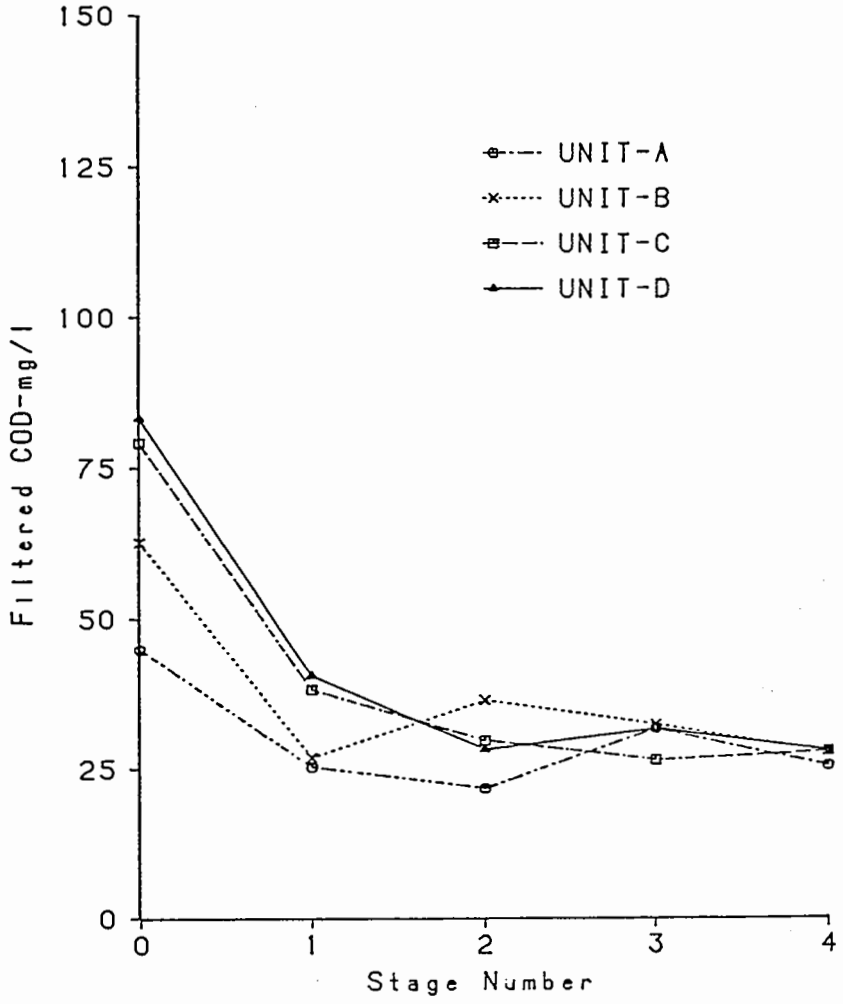

Figure 7. Mean steady-state mixed liquor filtered COD concentrations in the four stages of the RBC units operating at 20°C.

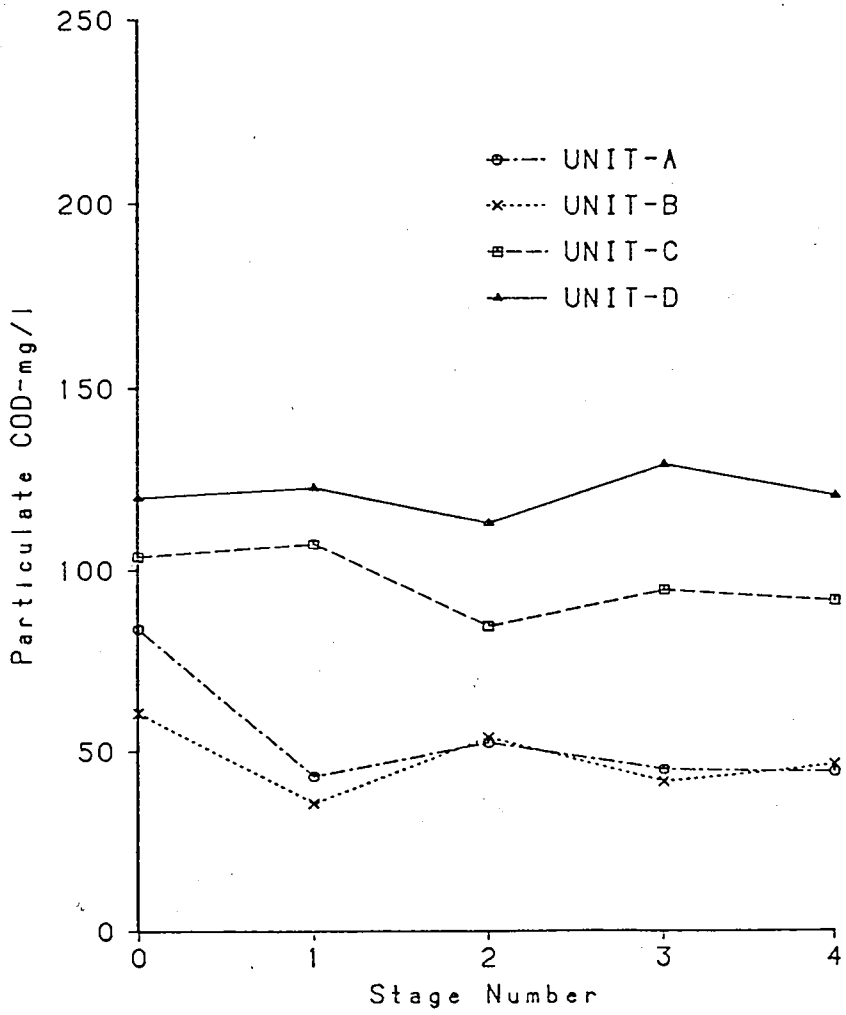

Figure 8. Mean steady-stage mixed liquor particulate COD concentrations in the four stages of the RBC units operating at 5°C.

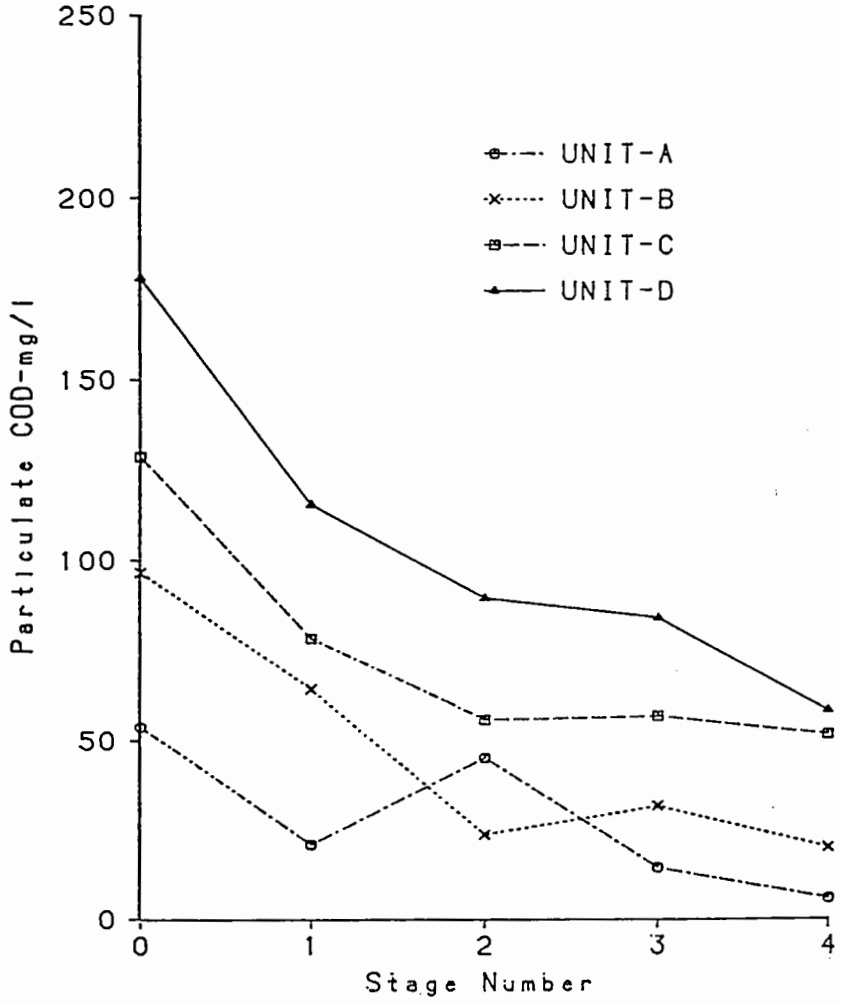

Figure 9. Mean steady-state mixed liquor particulate COD concentrations in the four stages of the RBC units operating at 15°C.

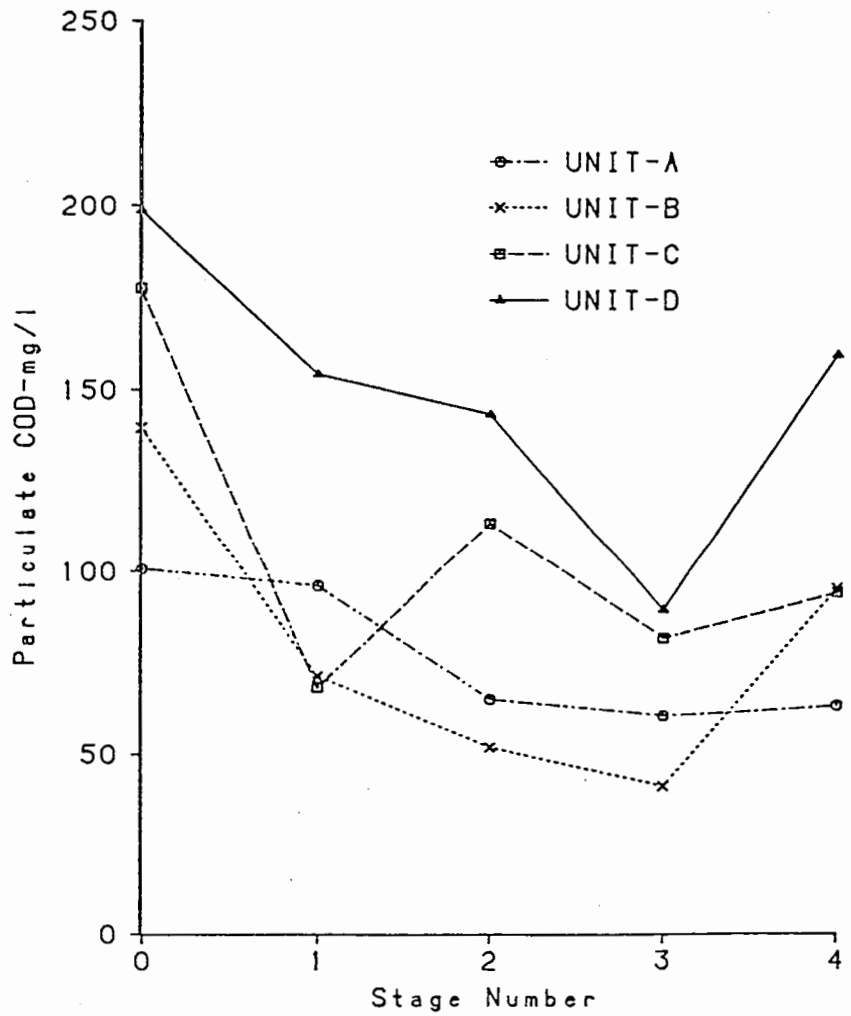

Figure 10. Mean steady-state mixed liquor particulate COD concentrations in the four stages of the RBC units operating at 20°C.

TABLE V. SUMMARY OF SUBSTRATE (COD) REMOVAL EFFICIENCIES (%)

Temperature, °C	5		15		20	
Unit	First Stage	Overall	First Stage	Overall	First Stage	Overall
A	78.3	79.8	73.5	73.1	82.7	82.5
B	74.4	76.9	80.4	85.1	86.8	86.3
C	78.3	80.2	80.1	86.8	85.2	89.1
D	77.2	79.2	81.3	85.0	85.7	90.0

A similar relationship was obtained with a full-scale RBC plant treating municipal wastewater at Kirksville, Missouri (26). The substrate concentration was measured as BOD_5, and the slope of the relationship was 0.893 based upon data collected over a period of two years.

The mixed liquor VSS production in terms of mg per mg COD removed was 0.50, 0.38, and 0.38 for 5°C, 15°C, and 20°C, respectively. The increase in sludge production at lower temperatures was probably due to lower decay rates. The increase of sludge production at lower temperatures was observed also in other studies (1).

Ammonia Nitrogen Removal

Figures 11, 12, and 13 show the mean steady-state mixed liquor ammonia nitrogen concentration when operating the RBC units at 5°C, 15°C, and 20°C (first period).* At 5°C, there was no ammonia removal in the system. Analyses of Figures 12 and 13 show that, generally, in the first stages there was limited ammonia nitrogen removal, except in Unit A, which was receiving the lowest organic loading rate. Significant ammonia nitrogen removal occurred in the second stages, and proceeded in the following stages in the units receiving high organic loading rates. The declining ammonia nitrogen removal rates were observed in the stages containing low concentrations of ammonia nitrogen and indicate substrate limiting conditions. In the region where substrate was not limiting, the decline in ammonia nitrogen removal followed a straight line, and the lines for the different units were generally parallel. At a temperature of 20°C, the slopes of these lines

* During the experiments conducted at 20°C, significant changes in the influent ammonia concentrations necessitated dividing this period for ammonia removal analysis.

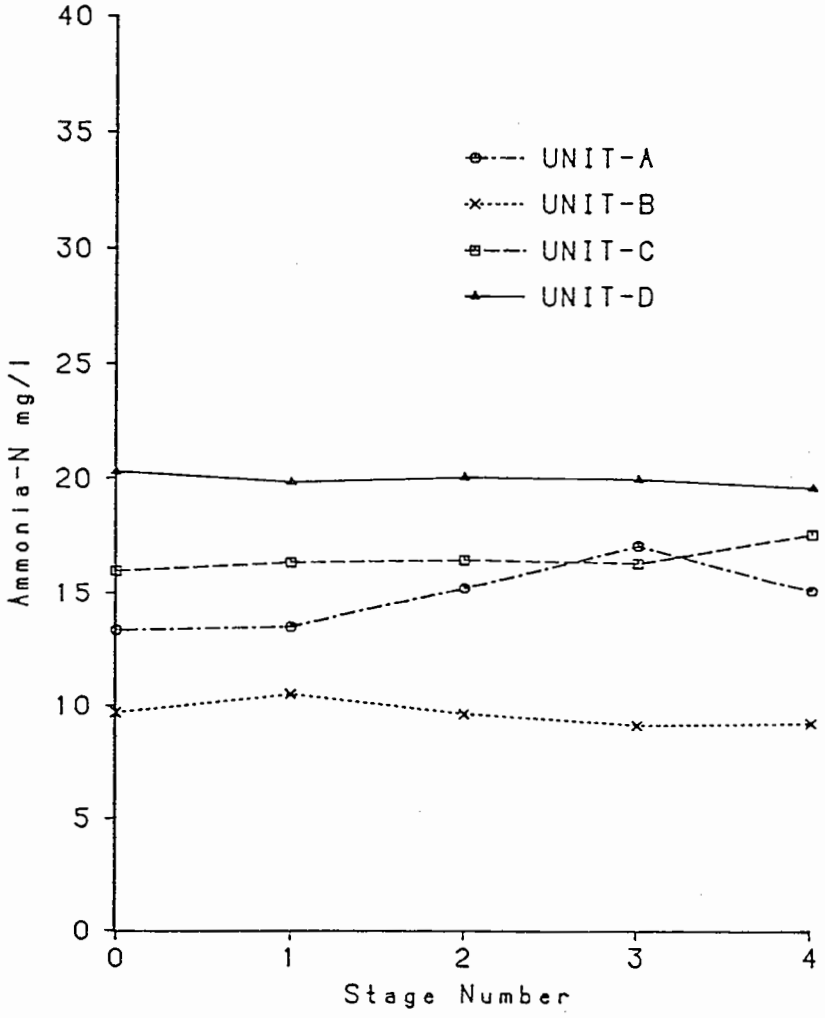

Figure 11. Mean steady-state mixed liquor ammonia nitrogen concentrations in the four stages of the RBC units operating at 5°C.

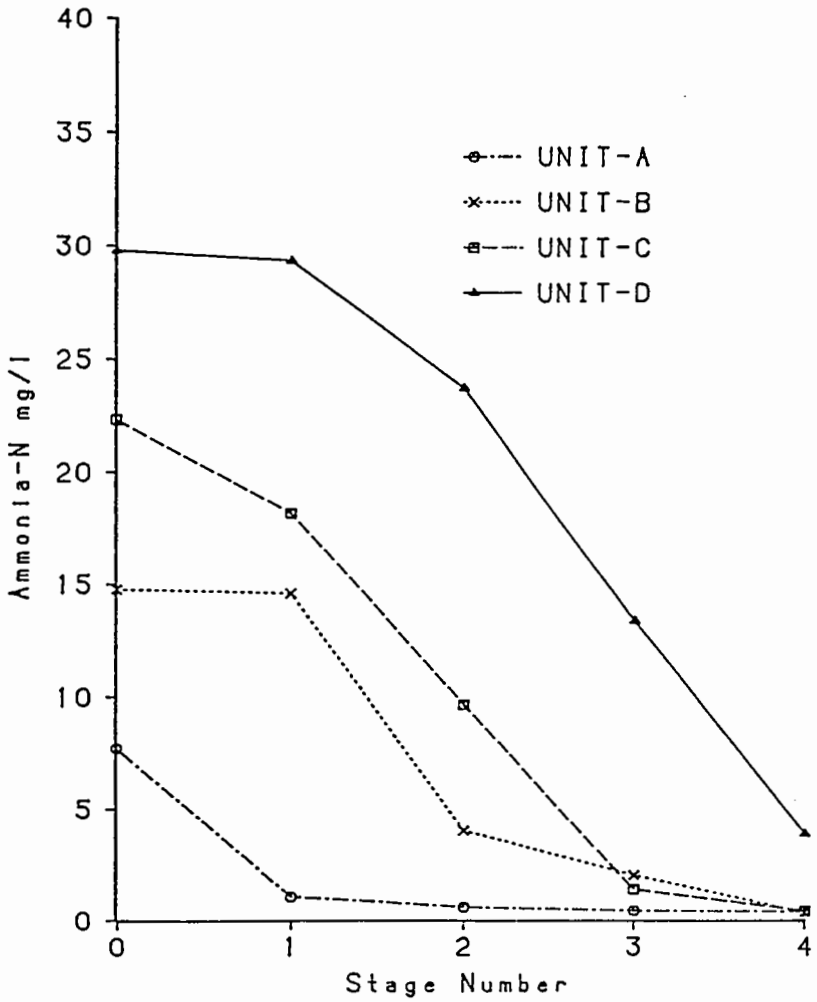

Figure 12. Mean steady-state ammonia nitrogen concentrations in the four stages of the RBC units operating at 15°C.

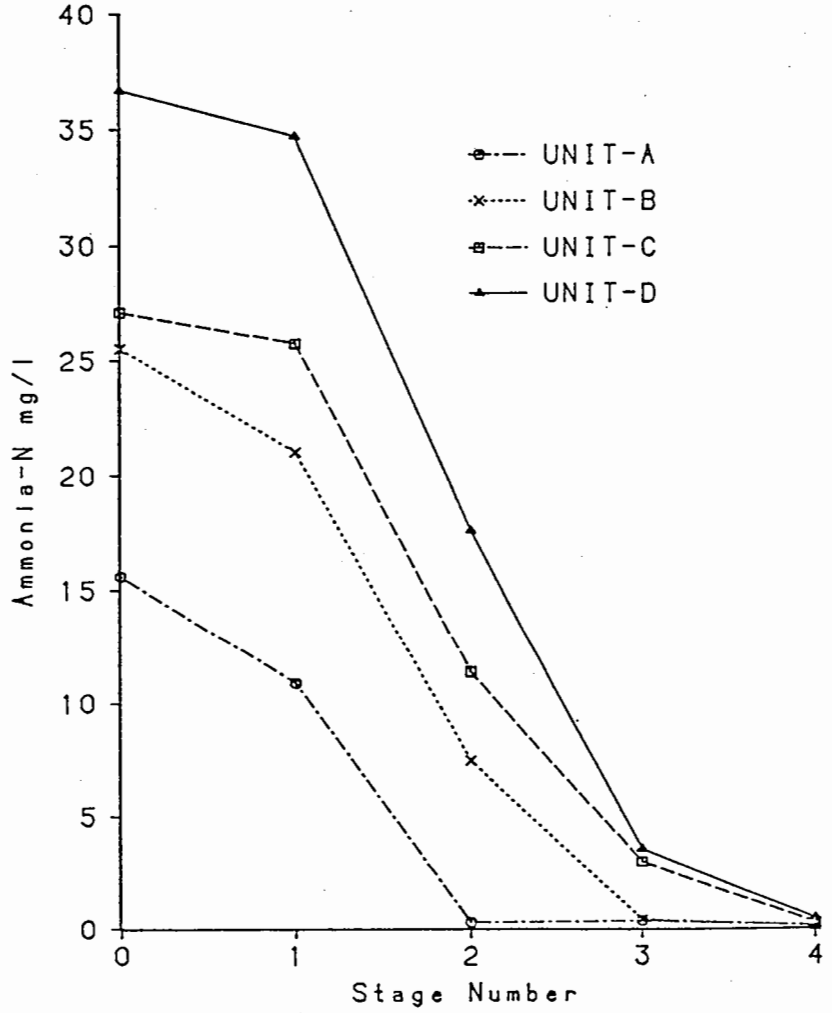

Figure 13. Mean steady-state ammonia nitrogen concentrations in the four stages of the RBC units operating at 20°C (first period).

were greater than the slopes of the lines at 15°C, emphasizing the effect of temperature on ammonia nitrogen removal rates beyond 15°C. Based upon the above observations, it appears that ammonia nitrogen removal could be described by Michaelis-Menten enzyme kinetics.

Table VI presents a summary of the overall ammonia nitrogen removal efficiencies at 15°C and 20°C. Only the second sampling period data were considered from the 20°C experiments, because an adequate number of data were not available during the first sampling period. The results in Table VI show that 98 to 99 percent ammonia nitrogen removal was obtained at organic loading rates up to 10 to 12.5 g COD/m^2/day. The removal efficiency decreased by approximately 10 percent at organic loading rates of 14 g COD/m^2/day. The percentage removal of ammonia nitrogen was higher at 20°C (Table VI).

Nitrate and nitrite nitrogen data revealed that 90 percent of ammonia removal in the system occurred through nitrification. The remaining portion of ammonia removal probably occurred because of stripping and assimilation.

TABLE VI. SUMMARY OF ORGANIC AND AMMONIA NITROGEN LOADING RATES AND THE AMMONIA-N REMOVAL EFFICIENCY

Temperature, °C	Unit	Organic Load g COD/m^2/d	Ammonia-N Load g N/m^2/d	Removal Efficiency Percent
15	A	3.984	0.387	94.8
	B	7.496	0.766	97.6
	C	9.875	1.143	98.1
	D	13.916	1.563	86.9
20	A	6.915	0.362	99.0
	B	9.734	0.520	98.0
	C	12.513	0.663	99.4
	D	13.971	0.786	90.5

KINETIC MODEL DETERMINATION

Carbon Removal

General

The first stages of the RBC units performed differently from the other stages as shown earlier, and the first stages

were considered separately in the analyses. The distinguishing features of the first stages were that they were receiving raw wastewater, while the influent to the other stages contained sloughed biomass from the preceding stages and unconsumed substrate. This difference in substrate affects the processes taking place in the RBC stages. The major processes that can be related to the biomass in the first stages are the carbonaceous substrate removal and endogenous respiration of the attached growth. In the following stages, the attached biomass is associated with several processes, i.e., stabilization of reattached biomass, nitrification, and exogenous substrate consumption.

The determinations of the kinetic constants were based upon the mean steady-state values of the parameters measured for each unit. The concentrations of the pollutants in the effluent were independent of the fluctuations in the influent concentrations; therefore, the mean concentrations from each unit were utilized in the calculations.

The following assumptions were made in the development of the kinetic model:

1. The available substrate in the influent to the first stage is the total COD.
2. The particulate material in the mixed liquor is sloughed biomass. Consequently the available exogenous substrate in the mixed liquor is the filtered COD.
3. The substrate consumption reaction takes place only in the attached growth.
4. The kinetics of substrate removal in the second, third, and fourth stages can be expressed by a common model.

First Stage Substrate Removal Kinetics

A mass balance of the biomass in the first stage yields the following equation under steady-state conditions:

$$\text{biomass produced} - \text{sloughed biomass} - \text{decay} = 0 \qquad (1)$$

In mathematical terms the equation can be written as follows:

$$Y\,Q(S_0-S_1) - Q\,X_1 - k_d A_1 \bar{X}_1 = 0 \qquad (2)$$

where

Y = yield coefficient, g VSS produced per g COD consumed

Q = influent flow rate, m^3/day
S_0, S_1 = influent and first stage effluent substrate concentration, mg/L-COD
X_1 = first stage effluent VSS, mg/L
$\overline{X}_1$ = first stage attached biomass g VS/m^2
A_1 = first stage discs area, m^2
k_d = decay coefficient, day $^{-1}$

A mass balance of the substrate in the first stage yields the following equation under steady-state conditions:

$$\text{substrate consumed} - \text{reaction} = 0 \qquad (3)$$

In mathematical terms Equation 3 can be written as follows:

$$Q(S_0-S_1) - A_1 r = 0 \qquad (4)$$

where

r = reaction rate, g COD/m^2/d

The reaction rate r in Equation 4 can be expressed using several kinetic models. The three models used in this study are summarized below:

1) Monod growth kinetics, incorporating the total attached biomass.

$$r = \frac{k\,\overline{X}_1 S_1}{K_s + S_1} \qquad (5)$$

k is defined as $\hat{\mu}/Y$

where

k = maximum reaction rate, day $^{-1}$
$\hat{\mu}$ = maximum specific growth rate, day^{-1}
K_s = half saturation constant, mg/L COD

2) Monod growth kinetics, incorporating a constant amount of active biomass.

$$r = \frac{k_a S_1}{K_S + S_1} \qquad (6)$$

where

k_a = maximum reaction rate, g COD/m^2/d

3) Mass transport model (15,16,17)

$$r = \frac{k_m S_1^2}{K_M + S_1} \qquad (7)$$

where

k_m = maximum reaction rate, g COD/m^2/ mg/L-COD/d

K_M = constant, mg/L COD

The reaction rate expressions (Equations 5, 6, and 7) were substituted into Equation 4, and the resulting equations were rearranged in the following format, to carry out linear regression analyses.

$$\left[Q(S_0-S_1)/A_1\bar{X}_1\right]^{-1} = \frac{K_s}{k}\frac{1}{\bar{S}_1} + \frac{1}{k} \qquad (8)$$

$$\left[Q(S_0-S_1)/A_1\right]^{-1} = \frac{K_s}{k_a}\frac{1}{S_1} + \frac{1}{k_a} \qquad (9)$$

$$\left[Q(S_0/S_1 - 1)/A_1\right]^{-1} = \frac{K_M}{k_m}\frac{1}{S_1} + \frac{1}{k_m} \qquad (10)$$

To carry out linear regression analyses for determining yield and decay constants, Equation 2 was rearranged as follows:

$$\frac{QX_1}{A_1\overline{X}_1} = Y\,\frac{Q(S_o-S_1)}{A_1\overline{X}_1} - k_d \qquad (11)$$

Table VII summarizes the results obtained from linear regression of Equation 11, and Table VIII summarizes the results obtained from linear regression of Equations 8, 9, and 10.

TABLE VII. SUMMARY OF THE RESULTS OF THE LINEAR REGRESSION ANALYSES OF THE DATA USED TO CALCULATE YIELD COEFFICIENTS AND DECAY RATES

Parameter	5°C	15°C	20°C
Yield coefficient, mg VS/mg COD	0.66	0.80	0.63
Decay rate, day^{-1}	0.07	0.22	0.26
Regression coefficient	0.998*	0.934	0.950
Significance level	0.05	0.10	0.05

*Based on three units; B, C, D.

The data from Unit A at 5°C was excluded from the analysis because the flow was changed during the experiment, and the unit did not approach steady-state conditions. Table VII emphasizes that the optimum growth and yield occurs at 15°C. Muck and Grady (27), using activated sludge mixed culture, observed an optimum in yield coefficient at 20°C. The difference in optimum temperature might be because of the different types of cultures growing in these systems.

Consistent results were obtained with Equation 8, which was derived from Equation 5, yielding reasonable values for the kinetic constants for all the temperatures (Table VIII). The mass transport model (Equation 7) produced reasonable values only with the data obtained at 5°C and 15°C. At these temperatures, the values for K_m were 20.8 mg/L and 42.5 mg/L, which are close to those obtained by Friedman et al (16,17). At 20°C the mass transport model resulted in a

TABLE VIII. SUMMARY OF THE KINETIC CONSTANTS FOR CARBONACEOUS SUBSTRATE REMOVAL IN THE FIRST STAGES[a]

Temperature, °C	5			15			20		
Equation	R	$\frac{k}{k_a}$	$\frac{K_s}{K_M}$	R	$\frac{k}{k_a}$	$\frac{K_s}{K_M}$	R	$\frac{k}{k_a}$	$\frac{K_s}{K_M}$
$Q(S_0-S_1) = A_1 \frac{k \bar{X}_1 S_1}{K_s+S_1}$	0.965	2.85	61.6	0.950	7.76	262.2	0.999	9.44	276.4
$Q(S_0-S_1) = A_1 \frac{k_M S_1^2}{K_M+S_1}$	0.893	1.12	20.8	0.886	1.80	42.5	-0.808	0.98	-5.8
$Q(S_0-S_1) = A_1 \frac{k_a S_1}{K_s+S_1}$	0.986	-41.0	-93.6	0.965	-36.3	-81.6	0.983	174	111.5

[a]For 5°C and 20°C only the data from units B, C, and D were used, for 15°C the data from units A, B, C, and D were used.
R = Correlation coefficient.

negative value for K_m, as shown in Table VIII. The negative value may have occurred, among other reasons, because at high temperatures the kinetics are described by substrate limiting conditions and not diffusion. Applying fixed biomass quantities with Monod growth kinetics as expressed in Equation 6 resulted in negative values for the reaction rates k_a at temperatures of 5°C and 15°C. At 20°C the reaction rate was determined to be 174 grams/m^2/d and the half saturation constant was 111.5 mg/L.

Clark, et al (11), reported values of $\hat{\mu}$, K_s and Y of 4.4, 431 and 0.96, respectively. These values were based on soluble BOD, and obtained from experiments conducted at uncontrolled temperature conditions. These values were calculated from an equation similar to Equation 5, except that only 70 percent of the total attached biomass was applied as active biomass. Considering that assumption, the $\hat{\mu}$ values from their studies and this study are comparable. The Y and K_s values differ significantly from the values obtained in this study, probably because of the differences in substrate and the fact that Clark et al (11) did not incorporate a decay factor in their equations.

The temperature relationship for k_d and k was obtained by using Equations 12 and 13:

$$(k_d)_T = (k_d)_{20}\, \theta_d^{T-20} \tag{12}$$

$$(k)_T = (k)_{20}\theta_s^{T-20} \tag{13}$$

where

$(k_d)_T$, $(k)_T$ = decay rate and reaction rate at temperature T (°C), day $^{-1}$

$(k_d)_{20}$, $(k)_{20}$ = decay rate and reaction rate at temperature 20°C, day^{-1}

Table IX summarizes the values obtained from linear regression analyses.

The temperature factor of 1.09 obtained with the Equations 12 and 13 is similar to the typical value of 1.08 for the trickling filter process (28).

The experimental data, as discussed previously, showed that the mass of attached growth was dependent upon the organic loading rate and could be defined by a saturation

TABLE IX. TEMPERATURE DEPENDENCY OF MAXIMUM REACTION RATE AND DECAY RATE

Parameter	Equation 12	Equation 13
Correlation coefficient	0.989	0.990
Significance level	0.1	0.1
k_{20}, day^{-1}		9.5
θ_s		1.09
$(k_d)_{20}$, day^{-1}	0.27	
θ_d	1.09	

function. A saturation type relationship was developed for the first stages that can be used for a given temperature to predict the quantity of attached biomass.

$$\overline{X}_1 = \frac{k_x M_1}{K_x + M_1} \tag{14}$$

where

X_1 = the quantity of attached biomass in the first stage per unit surface area, g VS/m^2
M_1 = organic load per first stage surface area, g COD/m^2/day
k_x = constant, g VS/m^2
K_x = constant, g COD/m^2/day

A regression analysis of Equation 14 in its linear (Eq. 15) form resulted in values as summarized in Table X.

$$\frac{1}{\overline{X}_1} = \frac{K_x}{k_x}\frac{1}{M_1} + \frac{1}{k_x}, \tag{15}$$

TABLE X. SUMMARY OF THE FIRST STAGE ATTACHED BIOMASS CONSTANTS

Temperature, °C			
Constant	5.9	16.3	20.8
k_x	46.15	52.54	58.50
K_x	31.07	23.77	23.77

The values of k_x were related to the temperature using the relationship shown in Equation 16. The correlation coefficient obtained from a linear regression analysis was 0.986.

$$(k_x)_T = 56.9(1.015)^{T-20} \qquad (16)$$

$$T = °C$$

$$(k_x)_T = g\ VS/m^2$$

Carbon Removal Kinetics in Stages 2-4

As discussed previously, the last stages of some units revealed instability. To compensate for this instability, all three stages were considered as one reactor where common reactions were taking place. Equation 17 was used to describe substrate removal as a function of temperature and influent substrate concentration to the second stages.

$$Q(S_1 - \overline{S}) = \sum_{i=2}^{4} A_i (k_L)_{20}\, \theta_L^{T-20} S_1^{\,n} \qquad (17)$$

where

Q	=	influent flow rate, m^3/d
S_1	=	first stage substrate concentration, mg/L
$\overline{S}$	=	the mean substrate concentration in the second through the fourth stages, mg/L
A_i	=	total available surface area/stage, m^2
$(k_L)_{20}$	=	reaction rate at 20°C, g COD/m^2/d
θ_L	=	temperature factor
T	=	temperature, °C
n	=	apparent reaction order

Multiple regression analysis with seven steady-state values (where substrate removal occurred) resulted in a regression coefficient of 0.986, which is significant at the 0.01 level. The values obtained were:

n	=	reaction order = 0.763
$(k_L)_{20}$	=	reaction rate at 20°C - 0.0444 g COD/m^2/d
θ_L	=	temperature factor = 1.11

The apparent reaction order 0.763 obtained for the second through fourth stages is in agreement with the apparent reaction order of 0.5-1.0 resulting from mass transport models for attached growth (14). The temperature factor of 1.11 is approximately the same as the temperature factor in the first stages.

Ammonia Nitrogen Removal Kinetics

A mass balance of ammonia nitrogen at stage i yields the following equation, at steady-state conditions.

$$Q\ C_{i-1} - Q\ C_i = A_i r \tag{18}$$

where

Q	=	flow rate, m^3/day
C	=	ammonia nitrogen concentration, mg/L
A	=	surface area of discs, m^2
r	=	reaction rate, grams/m^2/day

The reaction rate, r, can be expressed by the following kinetic models:

(a) Monod growth kinetics

$$r = \frac{k_N\ C_i}{K_N + C_i} \tag{19}$$

where

k_N	=	maximum reaction rate, grams/m^2/day
K_N	=	half saturation constant, mg/L

(b) Caperon and Meyer kinetics (29)

$$r = \frac{k_N(C_i - C_{min})}{K_N + (C_i - C_{min})} \tag{20}$$

where

C_{min} = minimum ammonia nitrogen concentration below which ammonia nitrogen removal does not occur (it is related to minimum intercellular stored nutrient necessary to sustain growth).

To carry out linear repression analyses, Equations 19 and 20 were rearranged as follows:

$$\left[\frac{Q(C_{i-1}-C_i)}{A_i}\right]^{-1} = \frac{K_N}{k_N}\frac{1}{C_i} + \frac{1}{k_N} \quad (21)$$

$$\left[\frac{Q(C_{i-1}-C_i)}{A_i}\right]^{-1} = \frac{K_N}{k_N}\frac{1}{(C_i-C_{min})} + \frac{1}{k_N} \quad (22)$$

To avoid large errors with the independent variable $1/C_i$ in Equations 21 and 22, where possible, data from stages with nitrogen concentrations less than 1 mg/L were not used in the linear regression analyses.

Figure 14 shows the measured concentrations of ammonia nitrogen in the RBC units operating at 15°C and the regression line calculated using the kinetic constants obtained from the linear regression analyses. The lower part of the prediction curve does not pass through the measured data, indicating that there may be a threshold concentration of approximately 0.4 mg/L-N below which ammonia nitrogen removal does not occur. Using Equation 22 with a C_{min} of 0.4 mg/L, better correlation was obtained as shown in Figure 15.

Figure 16 shows a plot of the data collected at 20°C and the curve plotted using the kinetic constants obtained from a linear regression of Equation 21. The plot of Equation 21 deviates from measured data points at the higher concentrations of ammonia nitrogen. Regression analyses of the Monod growth equation in the linearized form does not necessarily

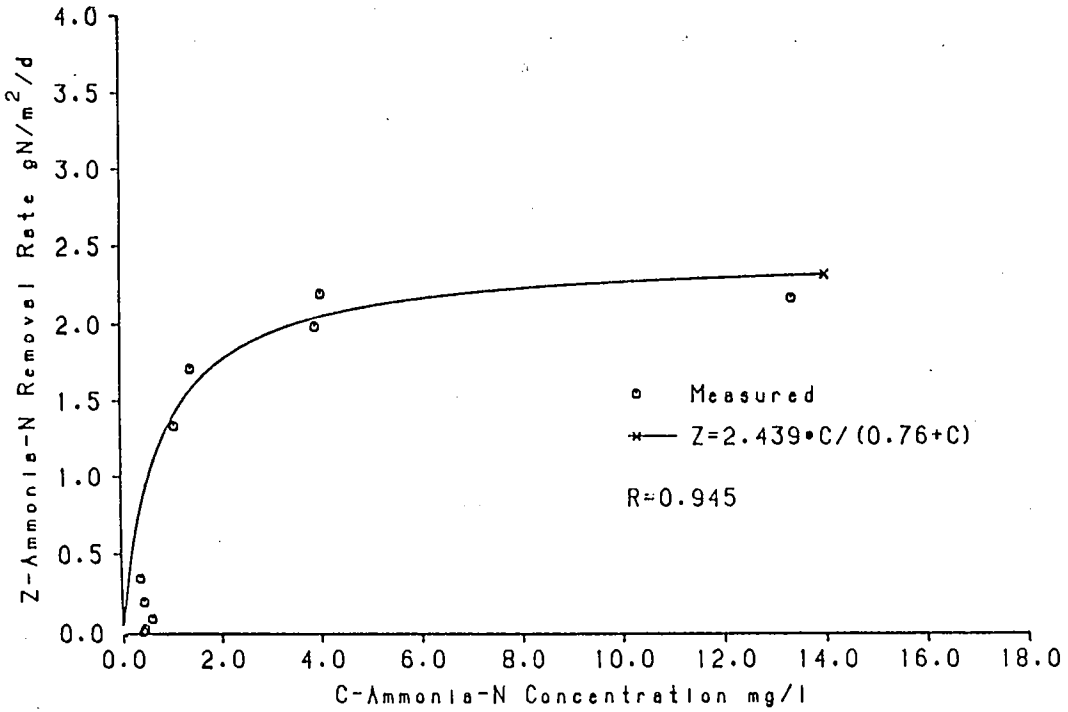

Figure 14. Relationship between the ammonia nitrogen removal rate and the ammonia nitrogen concentration at 15°C.

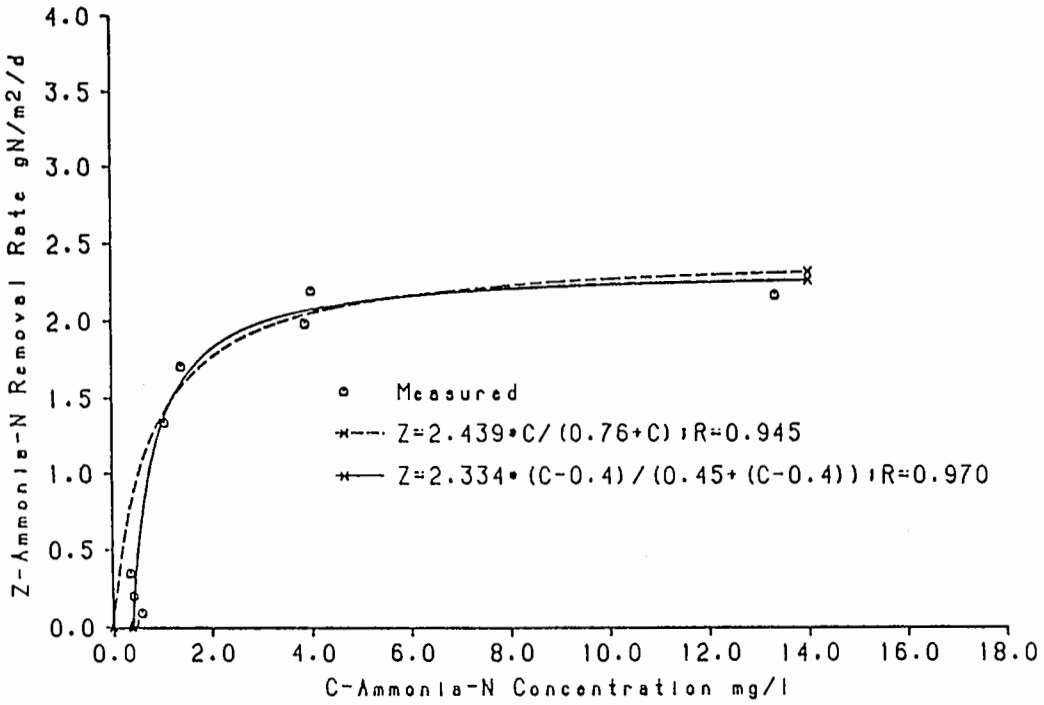

Figure 15. Comparison of the two predictive equations showing the relationship between ammonia nitrogen removal rate and the ammonia nitrogen concentration at 15°C.

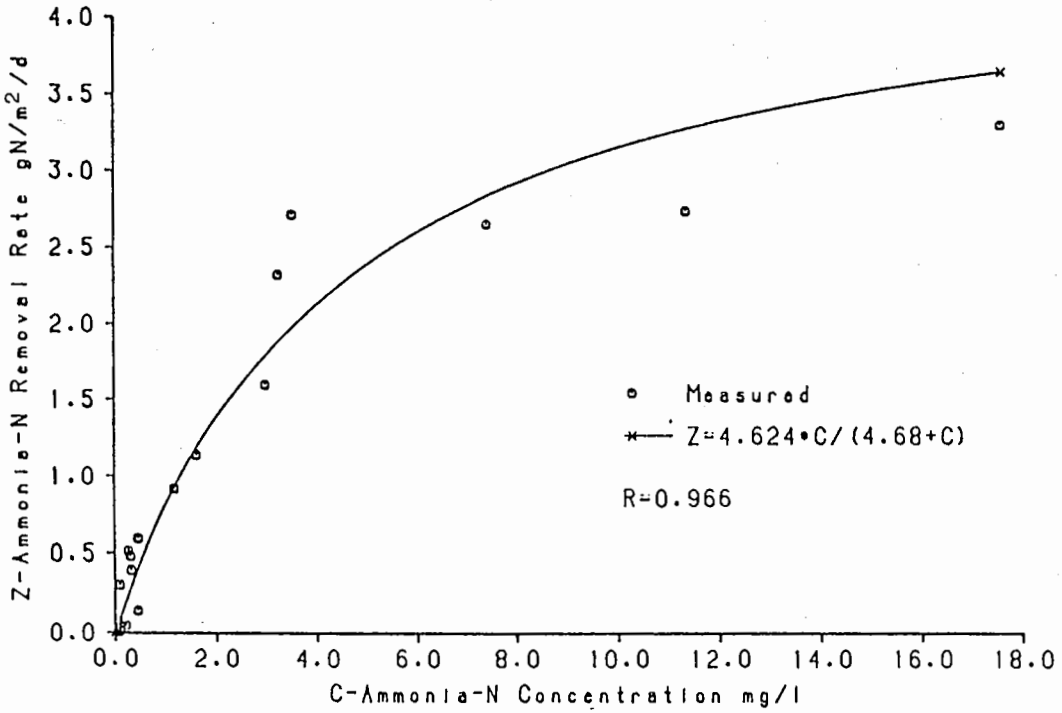

Figure 16. Relationship between ammonia nitrogen removal rate and the ammonia nitrogen concentration at 20°C.

provide the best fit for the Monod growth equation, although it is the best fit of the linearized form. The reason for deviation is that the low and medium concentrations have more impact than the high concentrations on the determination of the intercept and the slope.

An attempt was made to improve the fit of the theoretical expression and the measured data by choosing the pair of kinetic constants which yield the minimum sum of squares (SSQ) between the predicted and observed values.

Values of k_N in the range 3.00 to 4.60 g $N/m^2/day$ and K_N values from 1.0 to 4.6 mg/LN were evaluated. The minimum SSQ was obtained using the values of k_N = 3.74 $g/m^2/day$ and K_N = 2.8 mg/L.

Figure 17 shows the curve plotted using Equation 19 with the values obtained from linear regression and with the values obtained from non-linear fit analysis.

Table XI summarizes the kinetic constants for ammonia nitrogen removal.

Table XI. SUMMARY OF THE KINETIC CONSTANTS FOR AMMONIA NITROGEN REMOVAL

			k_N	K_N	C_{min}
Temperature	N	R	$g/m^2/d$	mg/L	mg/L
15	5	0.97	2.334	0.45	0.40
20	8	0.97	3.74	2.80	0.00

N - Number of observations
R - Correlation coefficient

The results obtained from the experiments conducted at 15°C show that a minimum concentration of 0.4 mg/L was necessary to maintain growth, while at 20°C this minimum concentration was not required. A possible explanation is that at higher temperatures the mass transport resistance decreases and, as a result, the requirement for stored material is less.

The kinetic constants obtained in this study for ammonia nitrogen removal are comparable with values obtained with synthetic substrate. Saunders et al (18) reported K_N values of 0.18 to 1 mg/L, and Ito and Matsuo (8) reported k_N value of 4 $g/m^2/day$.

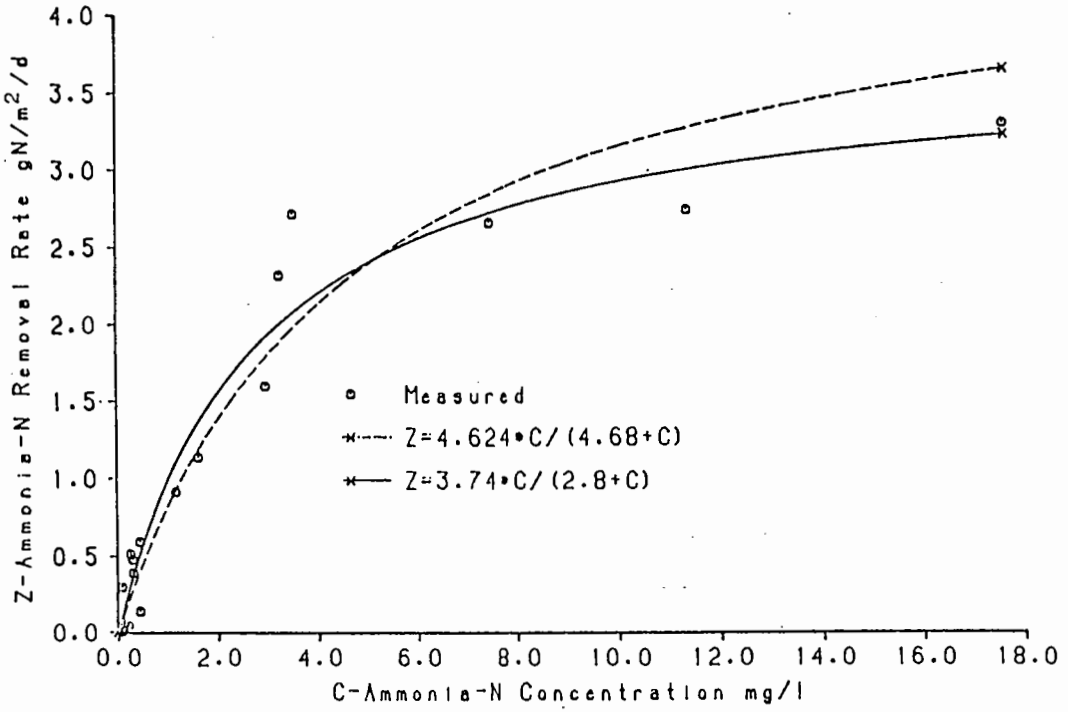

Figure 17. Comparison of the two predictive equations showing the relationship between ammonia nitrogen removal rate and the ammonia nitrogen concentration at 20°C.

The effect of temperature on the maximum reaction rate, k_N, can be expressed as follows:

$$(k_N)_T = (k_N)_{20}\ \theta_N^{T-20} \qquad (23)$$

where

$(k_N)_T$ $(k_N)_{20}$ = maximum reaction rate at T and 20°C, $g/m^2/d$
θ_N = temperature factor
T = temperature, °C

The temperature factor, θ_N, derived using Equation 23 is 1.1, is in agreement with temperature relationship developed for nitrification (30).

The inhibition of nitrification in the first stages was related to organic loadings and resulted in an equation with a correlation coefficient of 0.971 (significance level = 0.01):

$$f_1 = 1.43 - 0.1M;\ 4.3<M<14.3 \qquad (24)$$

where

f_1 = the ratio of measured ammonia removal to the theoretical ammonia removal without inhibition
M = overall organic load, g $COD/m^2/d$

The overall ammonia nitrogen removal (*) in four-stage RBC units can be expressed with Equation 25.

$$\frac{Q(C_O-C_4)}{\sum_{i=1}^{4} Ai} = k_N \left[\frac{f_1 C_{1M}}{K_N+C_{1M}} + \frac{C_2}{K_N+C_2} + \frac{C_3}{K_N+C_3} + \frac{C_4}{K_N+C_4} \right] \qquad (25)$$

* At lower temperatures of 15°C and below, C_{min} of 0.4 mg/L should be incorporated in equation, as shown in Equation 20.

where

C_{1M} = first stage ammonia nitrogen concentration at simulated maximum nitrification

C_0, C_2, C_3, C_4 = ammonia nitrogen concentration in influent, stage 2, 3, and 4, respectively

ENGINEERING SIGNIFICANCE

The steady-state kinetic models developed in this study for the RBC process treating domestic wastewater and the kinetic constants determined as a function of temperature provide a rational design approach for the RBC process. The mathematical expressions presented provide a basis for the calculation of the required RBC surface area to meet prescribed effluent standards for carbonaceous substrate, and ammonia nitrogen concentrations at temperatures ranging from 5°C to 20°C.

Design curves developed in this study for carbonaceous substrate removal in a four-stage RBC process at 20°C are presented in Figure 18. The corresponding temperature correction curves are presented in Figure 19.

To estimate the ammonia nitrogen concentration in the effluent, design curves based on the results of this study are presented in Figures 20 and 21 for an influent COD concentration of 300 mg/L and for temperatures of 15 and 20°C, respectively. For other influent COD concentrations, similar design curves can be developed using the equations presented herein.

When the RBC system is designed primarily to remove carbonaceous substrate, a different configuration of RBC staging can treat significantly higher loading rates than the conventional design, without bringing the first stage to anoxic-anaerobic conditions. The configuration can incorporate four shafts in three stages, with two of the shafts serving as the first stage, i.e., removal of the baffle between the first and second stages in the conventional configuration. A design example is presented below:

Assume that a design flow rate of 3800 m^3/d (1 mgd) of domestic wastewater with a primary effluent COD concentration of 300 mg/L COD and ammonia nitrogen of 30 mg/L, is to be treated with a RBC system to a degree that will produce a final effluent of 45 mg/L COD, or 85 percent removal.

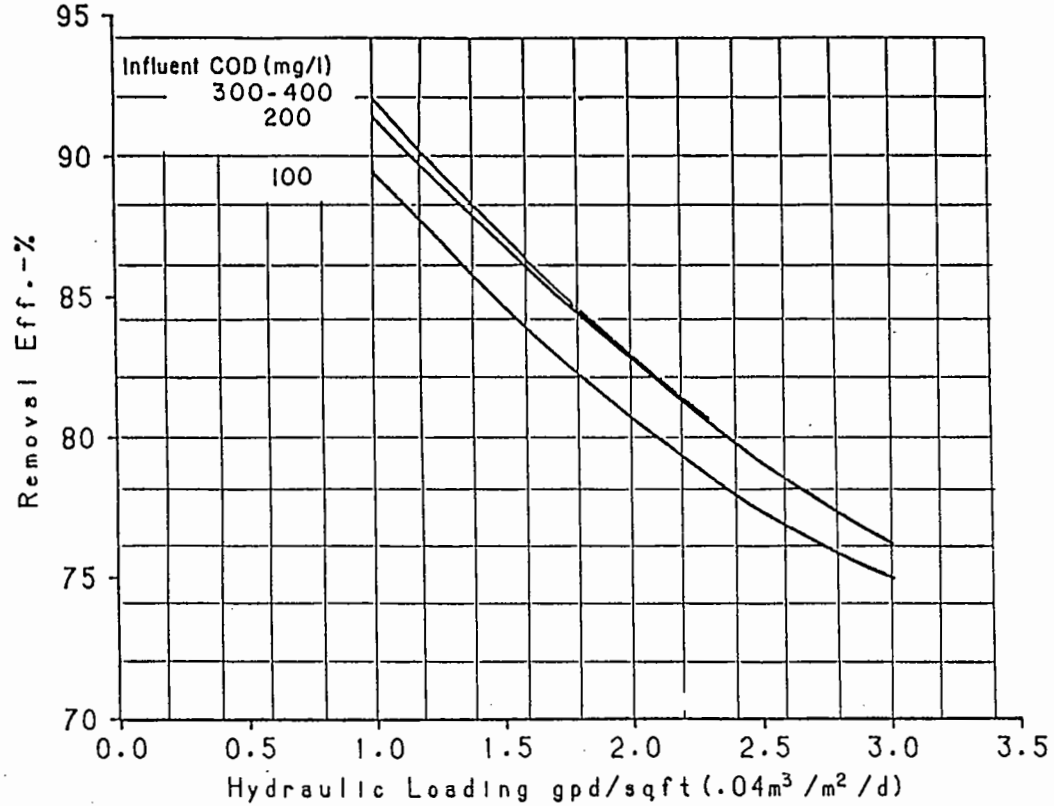

Figure 18. Design chart for COD removal in domestic wastewater treatment at 20°C.

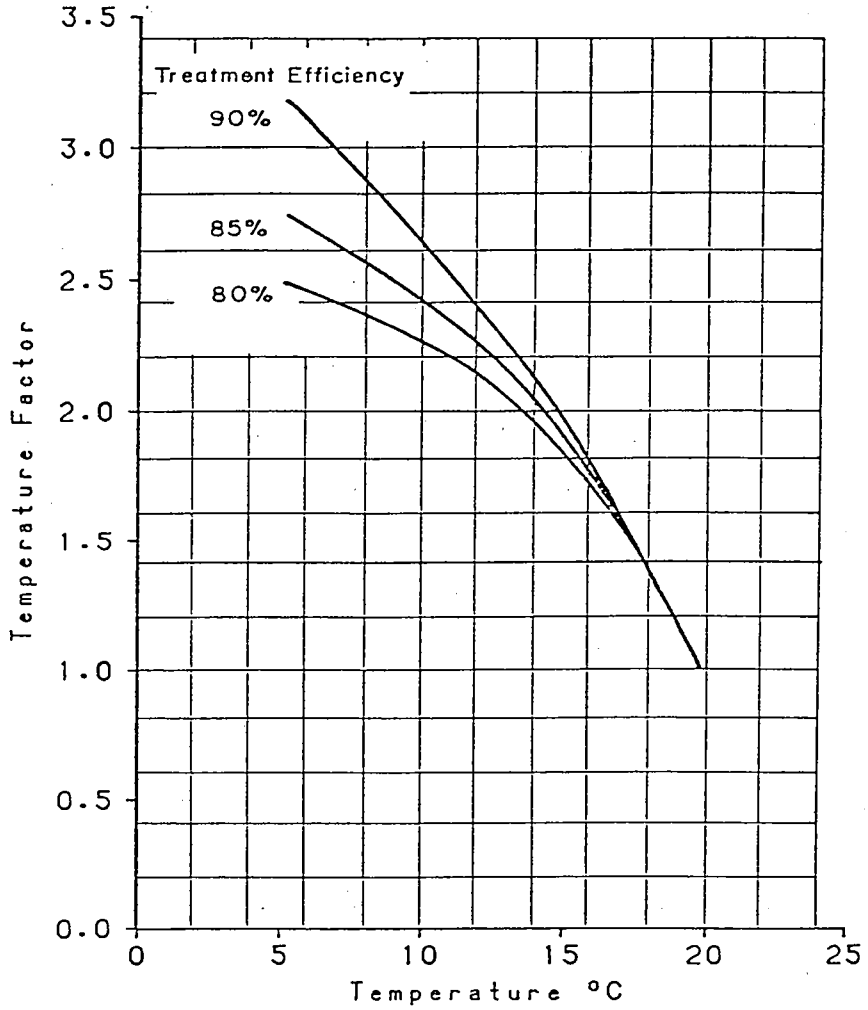

Figure 19. Temperature correction for COD removal.

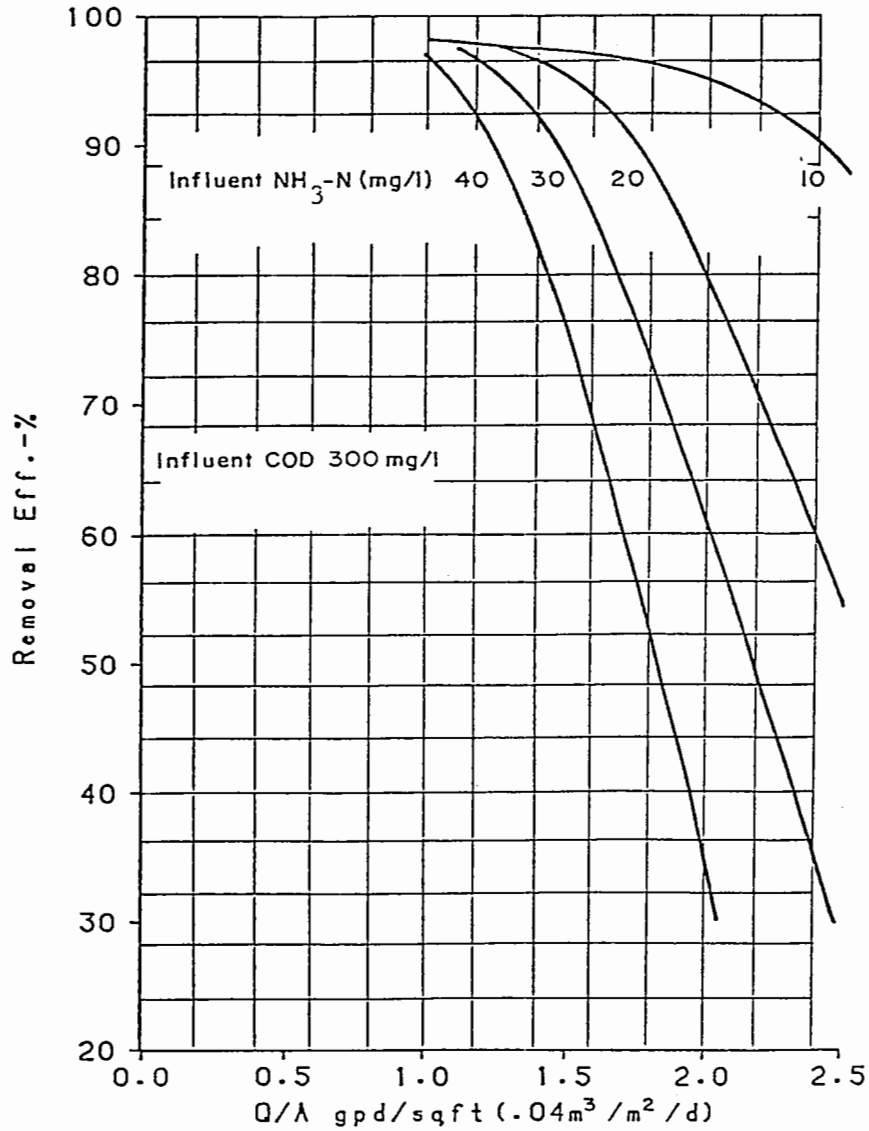

Figure 20. Design chart for ammonia nitrogen removal at 15°C.

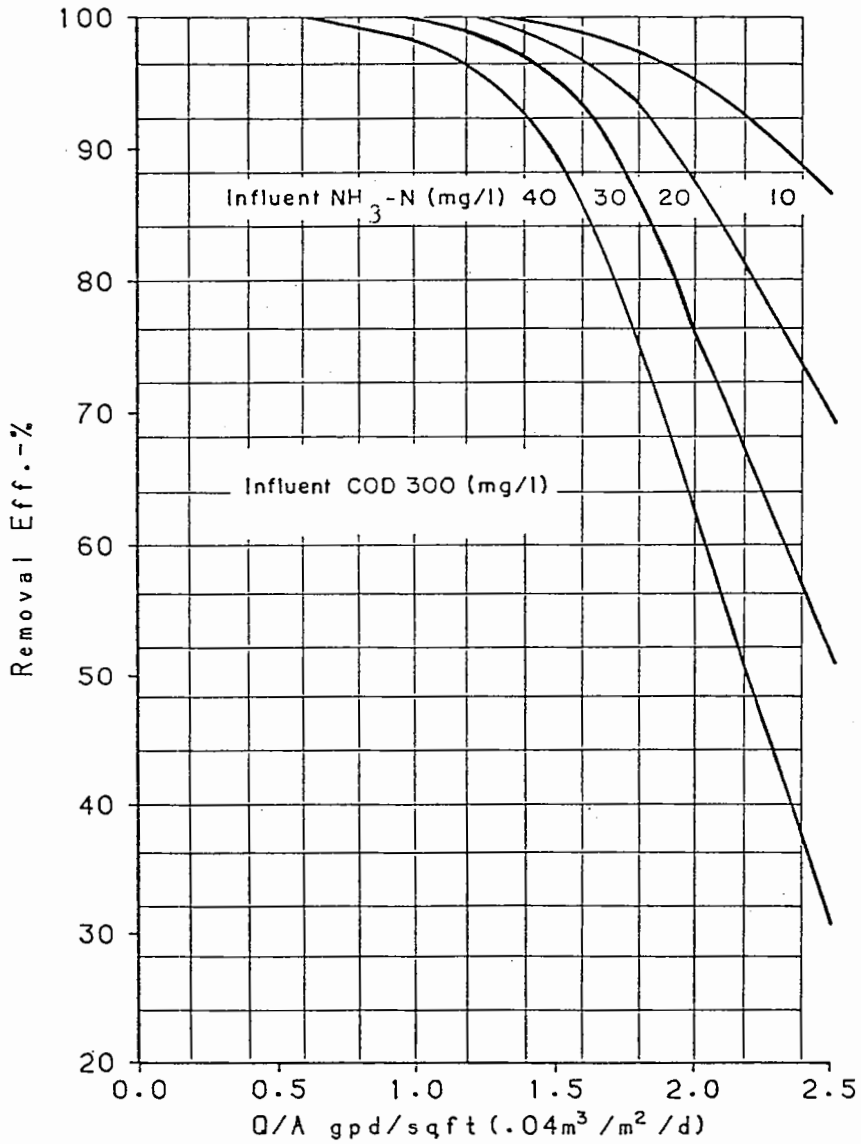

Figure 21. Design chart for ammonia nitrogen removal at 20°C.

Design winter temperature is 5°C, and summer temperature is 15°C.

1. Conventional Design 4-stage RBC: from Figure 18, the hydraulic load is found to be 0.07 $m^3/m^2/d$ (1.75 gpd/sq ft) at 20°C. At 5°C, the temperature factor is 2.7 (Figure 19). To meet the required effluent quality at 5°C, the designed hydraulic loading rate will be 1.75/2.7; i.e., 0.65 gpd/sq ft. The required total effective contactor area will be 1.5 x 10^6 sq ft. The ammonia nitrogen removal efficiency during the summer will be about 98 percent (Figure 20), i.e., the effluent will contain about 0.6 mg/L NH_3-N.

2. Three-stage RBC, first stage contactor area, 40 percent of the total RBC surface area.*

Based on the equations and kinetic constants presented in this study for first stage and the later stages of RBC, the total surface area required will be about 1 x 10^6 sq ft to meet the effluent requirements at 5°C. The organic loading rate to the first stage will be about 30 g/COD/m^2/day, which will assure aerobic conditions at summer temperatures.

The ammonia nitrogen concentrations in the effluent at summer conditions will be about 4.7 mg/L in this RBC configuration.

CONCLUSIONS

Carbonaceous Substrate Removal

1. Carbon removal in RBC units was influenced by temperature and organic loading rate. The overall removal efficiencies in this study were 80 percent, 85 percent, and 90 percent for 5°C, 15°C, and 20°C, respectively.

2. Majority of carbon removal occurred in the first stages. The COD removals in the first stages were 77 percent, 80 percent, and 85 percent for 5°C, 15°C, and 20°C, respectively.

3. The kinetics for carbon removal in the first stages can be described by Monod growth kinetics.

4. The temperature factor for the carbon removal reaction rate and the decay rate is 1.09.

5. The kinetics for carbon removal in the last stages can be described by variable order kinetics (in this study, 0.763), and a temperature factor of 1.11.

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* The current common design employs shafts of 100,000 sq ft in the first stages, and 150,000 sq ft in the last stages.

6. The kinetic constants determined in this study can be used to design RBC systems (minimum DO of 2 mg/L in the first stages) for carbon removal in a temperature range of 5°C to 20°C.

7. For low temperature design, providing more surface area in the first stages can reduce significantly the total RBC area required.

Ammonia Nitrogen Removal

1. Ammonia nitrogen removal in RBC units was influenced by temperature and organic loading rate. The overall ammonia removal ranged from 87 percent to 98 percent at 15°C, and from 91 percent to 99 percent at 20°C. At 5°C, there was no ammonia removal. As the influent organic loading rates increased, the overall ammonia removal decreased.

2. The kinetics for ammonia nitrogen removal can be described by Monod growth kinetics. At 15°C, the model incorporated a minimum concentration of 0.4 mg/L, below which ammonia removal did not occur.

3. The temperature factor for ammonia removal reaction rate was 1.10.

4. The inhibition of ammonia removal in the first stages was proportional to the organic loading rates.

5. The resulted kinetic constants in this study can be used to predict ammonia nitrogen removal in RBC systems within a temperature range of 5° to 20°C.

REFERENCES

1. Antonie, R.L., 1976, Fixed Biological Surfaces - Wastewater Treatment, CRC Press, Inc., West Palm Beach, Florida, 200 p.
2. Banerji, S.K., 1980, ASCE Water Pollution Management Task Committee Report on Rotating Biological Contactor for Secondary Treatment, Proc. 1st National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1:31-52.
3. Smith, E.D., and J.T. Bandy, 1980, A History of the RBC Process, Proc. 1st National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1:11-26.
4. Chesner, W.H. and T.T.I. Ionnone, 1980, Current Status of Municipal Wastewater Treatment with RBC Technology in the U.S., Proc. 1st National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1:53-70.
5. Torpey, W., et al, 1971, Rotating Discs with Biological Growth Prepare Wastewater for Disposal or Reuse, JWPCF 43 (11): 2181-2188.
6. Pescod, N.B., and J.V. Nair, 1972, Biological Disc Filtration for Tropical Waste Treatment, Water Res. (G.B.), 6 (12): 1509-1523.
7. Malhotra, S.K., T.C. Williams, and W.L. Morley, 1975, Performance of a Bio-Disc Plant in a Northern Michigan Community, Presented at the 48th Annual Conf., Water Poll. Contr. Fed., Miami Beach, Florida, Oct. 5-10, 1975, 29 p.
8. Ito, K., and T. Matsuo, 1980, The Effect of Organic Loading on Nitrification in RBC Wastewater Treatment Processes, Proc. 1st National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 2:1165-1175.
9. Zenz, D.R., et al, 1980, Pilot-Scale Studies on the Nitrification of Primary and Secondary Effluents Using Rotating Biological Discs at the Metropolitan Sanitary District of Greater Chicago, Proc. 1st National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 2:1221-1246.
10. Kornegay, B.H., and J.F. Andrews, 1968, Kinetics of Fixed Film Biological Reactors, JWPCF 40 (11):R460-R468.

11. Clark, J.A., E.M. Moseng, and T. Asano, 1978, Performance of a Rotating Biological Contactor Under Varying Wastewater Flow, JWPCF 50 (5): 896-911.
12. Mikula, W.J., 1979, Performance Characteristics and Kinetics of Substrate Removal in the Treatment of a Cheese Processing Wastewater with a Rotating Biological Contactor, M.S. Thesis, Utah State University, Logan, Utah, 195 p.
13. Hansford, G.S., et al, 1978, A Steady-State Model for the Rotating Biological Disc Reactor, Water Res. (G.B.), 12:855-868.
14. Rittman, B.E., and P.L. McCarty, 1978, Variable-Order Model of Bacterial Film Kinetics, J. Env. Eng. Div., ASCE 104 (EE5):889-900.
15. Schroeder, E.D., 1977, Water and Wastewater Treatment, McGraw-Hill Book Co., Inc., New York, pp. 288-312.
16. Friedman, A.A., R.C. Woods, and R.C. Wilkey, 1976, Kinetic Response of Rotating Biological Contactors, Proc. 31st Ind. Waste Conf., Purdue Univ., Ann Arbor Science Publishers, Inc., Ann Arbor, Michigan, pp. 420-423.
17. Friedman, A.A., L.E. Robbins, and R.C. Woods, 1979, Effect of Disc Rotational Speed on Biological Contactor Efficiency, JWPCF 51 (11):2678-2680.
18. Saunders, F.M., R.L. Pope, and M.A. Cruz, 1980, Effects of Organic Loading and Mean Solids Retention Time on Nitrification in RBC Systems, Proc. 1st National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1:409-432.
19. Watanabe, Y., M. Ishiguro, and K. Nishidome, 1980, Nitrification Kinetics in a Rotating Biological Disc Reactor, Prog. Water Tech (G.B.), 12:233-251.
20. Pano, A., 1981, The Kinetics of Rotating Biological Contactors, Treating Domestic Wastewater, Ph.D. Dissertation, Utah State University, Logan, Utah.
21. Oceanography International, 1978, Chemical Oxygen Demand (Standard Ampule Method), Federal Register Vol. 43, No. 45, Tuesday, March 7, 1978, Oceanography International Corporation, College Station, Texas.
22. Technicon Industrial Systems, 1977, Total Kjeldahl Nitrogen, Industrial Method No. 376-75W/B, Technicon Industrial Systems, Division of Technicon Instruments Corp., Terrytown, New York.

23. Technicon Industrial Systems, 1973a, Nitrite in Water and Seawater, Industrial Method No. 161-71W, Technicon Industrial Systems, Division of Technicon Instruments Corp., Terrytown, New York.
24. Technicon Industrial Systems, 1973b, Nitrate and Nitrite in Water and Wastewater, Industrial Method No. 100-70W, Technicon Industrial Systems, Division of Technicon Instruments Corp., Terrytown, New York.
25. American Public Health Association, 1976, Standard Methods for the Examination of Water and Wastewater, 14th Ed., Washington, D.C.
26. Dupont, R.R., and R.E. McKinney, 1980, Data Evaluation of a Municipal Installation, Kirkville, Missouri, Proc. 1st National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1:205-234.
27. Muck, R.E., and C.P.L. Grady, 1974, Temperature Effects on Microbial Growth in CSTR's, J. Env. Eng. Div., ASCE 100 (EE5):1147-1163.
28. Metcalf and Eddy, Inc., 1979, Wastewater Engineering, Treatment Disposal and Reuse, 2nd Ed., McGraw-Hill Book Company, New York, 920 p.
29. Caperon, J. and J. Meyer, 1972, Nitrogen-Limited Growth of Marine Phytoplankton I and II. Deep-Sea Res. (G.B.), 19:601-632.
30. Parker, D.S., R.W. Stone, and R.J. Stenquist, 1975, Process Design Manual for Nitrogen Control, USEPA, Technology Transfer, October, 1975.

KINETICS AND SIMULATION OF NITRIFICATION IN A ROTATING BIOLOGICAL CONTACTOR

Yoshimasa Watanabe. Department of Civil Engineering, Miyazaki University, Miyazaki 880, Japan

Kiyoshi Nishidome. Department of Civil Engineering, Kagoshima Technical College, Hayato 899-51, Japan

Chalermraj Thanantaseth. Department of Chemical Engineering, King Mongkut Institute of Technology, Thonbuli Campus, Bangkok, Thailand

Masayoshi Ishiguro. Department of Civil Engineering, Miyazaki University, Miyazaki 880, Japan

INTRODUCTION

A steady-state kinetics for fixed-biofilm reaction has been developed and applied to the denitrification and the nitrification processes in rotating biological contactors (1,2, 3). The proposed kinetics can be described as a process of molecular diffusion with a simultaneous zero-order biochemical reaction. The proposed kinetics has adequately explained all experimental results of nitrification in a partially submerged rotating biological contactor (RBC), but it cannot be strictly applicable to a partially submerged RBC process in which the biofilm alternately rotates into water and air. The partially submerged RBC has no steady-state substrate concentration pro-

file within the biofilm, even though the concentration of the substrate in the bulk water is the steady state.

In this paper, the authors report the results of a computer simulation of nitrification in a partially submerged RBC process to find out the reasoning behind the application of the steady-state kinetics. An analysis of the experimental data on combined carbon oxidation-nitrification in the same process is also included. All fluxes in this paper are expressed on the basis of the submerged disk surface area.

APPLICATION OF STEADY-STATE KINETICS TO A PARTIALLY SUBMERGED RBC PROCESS

Modification of Steady-State Kinetics

The proposed kinetics can be applied to nitrification in a fully submerged biofilm process, summarized below. At steady state, the transfer rate of ammonia to the biofilm surface through the diffusion layer is equal to that at the biofilm surface. Thus, the ammonia flux to the biofilm surface can be expressed by Eq. 1, if the amount of ammonia used for cell synthesis of the nitrifying bacteria is negligibly small compared to that nitrified by the same bacteria.

$$\frac{D_A}{L_d}(C_{bA} - C_{sA}) = F_A \tag{1}$$

Therefore, the relationship between bulk and surface ammonia concentrations is,

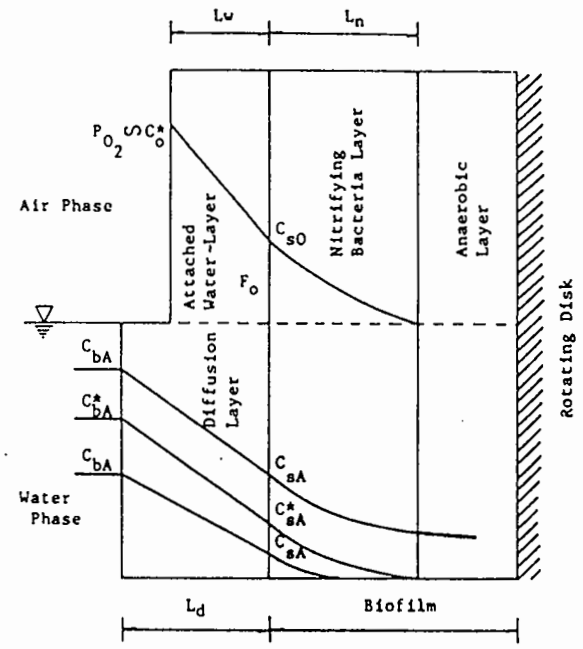
$$C_{bA} = C_{sA} + \frac{F_A}{D_A/L_d} = C_{sA} + \frac{F_A}{K_{dA}} \tag{2}$$

Ammonia flux at the biofilm surface (F_A) is represented by Eq. 3 for partial ammonia penetration and by Eq. 4 for complete ammonia penetration.

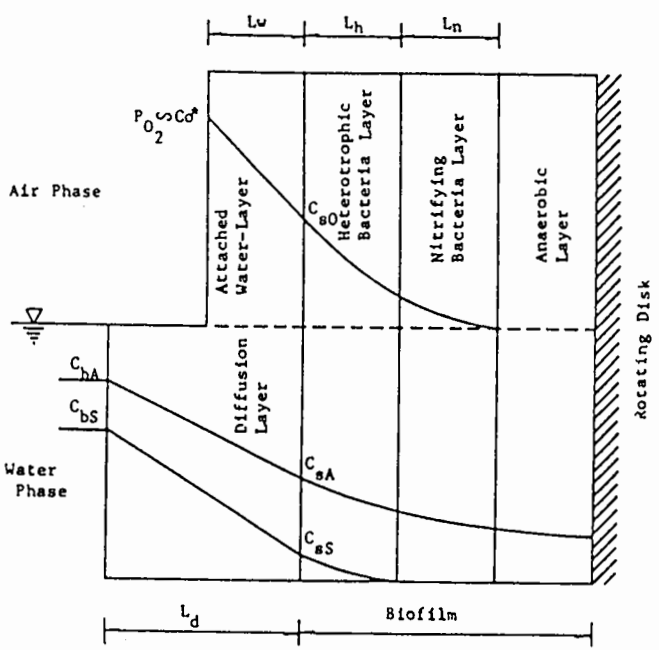
$$F_A = \sqrt{2D_A R_n C_{sA}} \qquad C_{sA} \leq C_{sA}^* \tag{3}$$

$$F_A = F_{A,max} = \sqrt{2D_A R_n C_{sA}^*} = R_n L_n \qquad C_{sA} \geq C_{sA}^* \tag{4}$$

However, the proposed steady-state kinetics would not be completely applied to a partially submerged RBC process for the following reasons. A steady-state substrate concentration profile within the biofilm cannot be assumed, even though the bulk substrate concentration is the steady state, since the biofilm alternately rotates into the air and



(a) Pure nitrification



(b) Combined carbon oxidation-nitrification

Fig. 1 Biofilm Model

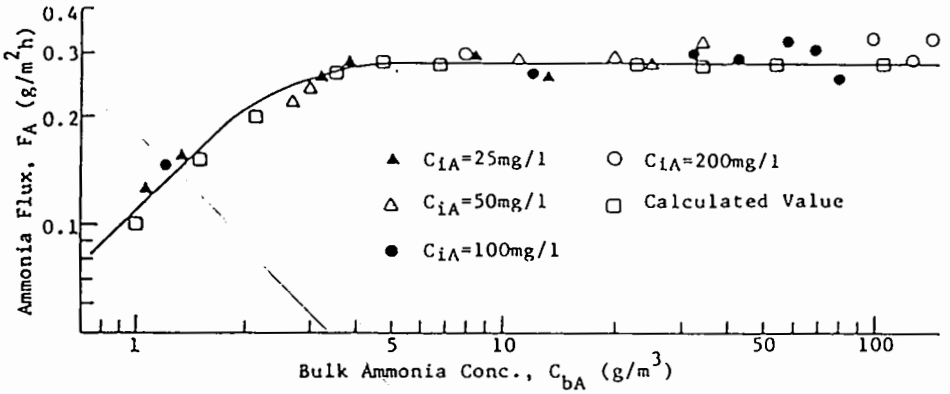


Fig. 2 Relationship between bulk ammonia concentration and ammonia flux
(Disk diameter=30cm, Disk rotating velocity=7.5rpm
Water temp.=23.5°C)

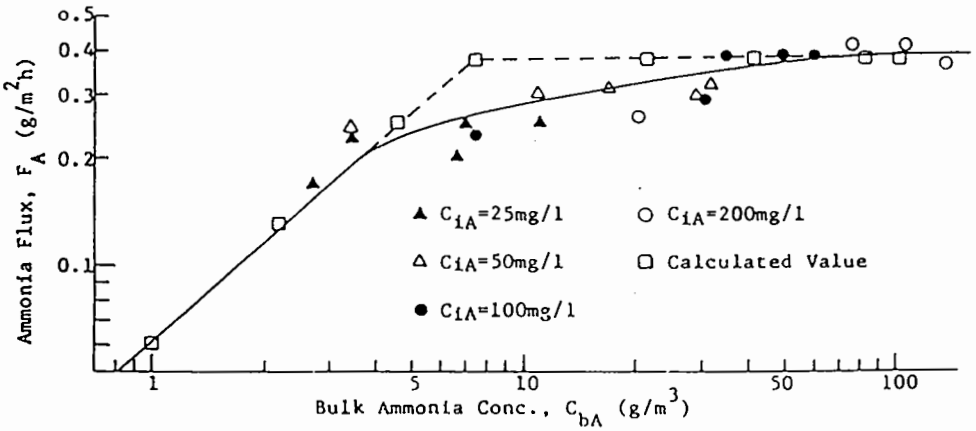


Fig. 3 Relationship between bulk ammonia concentration and ammonia flux
(Disk diameter=30cm, Disk rotating velocity=3rpm
Water Temp.=28.5°C)

the water. The authors (7) have developed a hypohesis about oxygen transfer which would be applicable to a partially submerged RBC. The hypothesis states that the oxygen transfer to the biofilm mainly occurs through the attached water-layer, during the time the biofilm rotates in the air. The nitrification biofilm model for a partially submerged RBC is shown in Fig. 1 (a). The penetration depth of oxygen (L_n) can be expressed as follows:

$$F_o = \frac{D_o(C_o^* - C_{so})}{L_w} \tag{5}$$

$$L_n = \frac{F_o}{R_o} = \frac{F_o}{4.33R_n} \tag{6}$$

Oxygen consumption for biological nitrification is 4.33 gO_2/g NH_4-N (4). Bintanja et al (5) proposed the following equation for the estimation of the thickness of the attached water-layer on the disk surface:

$$L_w = 0.98(\frac{\mu\omega r}{\rho g})^{1/2} \tag{7}$$

Employing Eq. 2 to 7, we can find the relationship between the bulk ammonia concentration and the ammonia flux.

Experimental Verification of the Modified Kinetics

The data obtained in a partially submerged RBC have previously been presented (2). Figs. 2 and 3 are examples of some of the data. The theoretical data calculated from Eq. 2 to 7 have also been plotted in Figs. 2 and 3. The experimental unit had a disk diameter of 30 cm, L_w equalled 42 μm at a disk rotating velocity of 7.5 rpm and water temperature of 23 °C (Eq. 7). These corresponded to the experimental conditions for the data shown in Fig. 2. Hartman (6) measured the attached water-layer thickness at about 40 μm in an actual RBC plant. As shown in the next section, the oxygen concentration at the biofilm surface (C_{so}) was estimated at about 2 mg/l. Therefore, the oxygen flux to the biofilm was calculated at 1.36 g O_2/m^2h from Eq. 5. Eq. 6 gave 60 μm as the value of L_n. Fig. 2 shows the $F_{A,max}$ was 0.27 g NH_4-N/m^2h. Eq. 4 then gave 52 μm as the L_n. The intrinsic nitrification rate (R_n) was determined as 5200 g/m^2h in the previous experiment (7). The penetration thickness of oxygen calculated from Eqs. 4 and 6 were almost the same. Therefore, the proposed hypothesis on oxygen transfer was confirmed.

COMPUTER SIMULATION OF NITRIFICATION IN A PARTIALLY SUBMERGED

RBC PROCESS

Model Development

The biofilm attached to a partially submerged RBC rotates alternately into the air and water. In the air phase, oxygen is supplied to the biofilm from the air, but there is no ammonia transport to the biofilm. In the water phase, ammonia diffuses into the biofilm from the bulk water. A computer simulation to identify the change in the ammonia and oxygen profiles in the system was carried out based on the assumptions which had been made for the development of our steady-state biofilm kinetics, namely:

1. The bulk water is completely mixed,
2. Only molecular diffusion occurs through the diffusion layer,
3. Molecular diffusion with a simultaneous zero-order biochemical reaction occurs within the active biofilm.

Fig. 4 illustrates the biofilm system divisions consisting of the attached water-layer, the diffusion layer, and the biofilm. The disk surface was divided into n small sectors each with an area equal to ΔA. The biofilm,the attached water-layer, and the diffusion layer were divided into sub-layers, each of them ΔZ thick.

The basic equation of the simulation was Fick's Second Law of Diffusion

$$\frac{\partial C_A}{\partial t} = D_A \frac{\partial^2 C_A}{\partial Z^2} \tag{8}$$

Eq. 8 was directly applied to the attached water-layer and the diffusion layer, but a biochemical reaction term had to be added to take into account the substrate uptake within the biofilm as follows:

$$\frac{\partial C_A}{\partial t} = D_A \frac{\partial^2 C_A}{\partial Z^2} - R_n \tag{9}$$

The difference form of Eq. 9 is

$$C_{A(n+1,i)} = K(C_{A(n,i-1)} - 2C_{A(n,i)} + C_{A(n,i+1)}) + C_{A(n,i)} - R_n \Delta t \tag{10}$$

$$K = \frac{D_A \Delta t}{(\Delta Z)^2} \tag{11}$$

where the subscript n refers to the number of Δt time and the

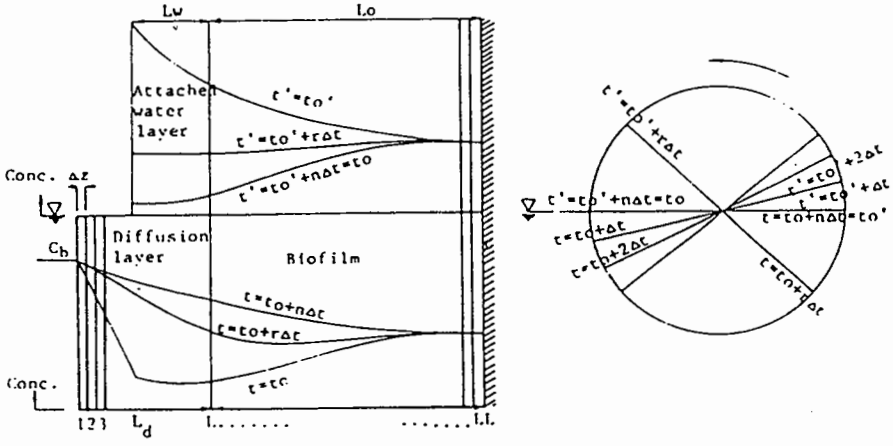


Fig. 4 Divisions of the biofilm system

Table 1 Simulation conditions

Parameter	Run 1	Run 2
Biofilm Thickness (L_o)	900 μm	900 μm
Attached-water layer thickness (L_w)	50 μm	50 μm
Diffusion layer thickness (L_d)	80 μm	80 μm
Nitrification rate (R_n)	7300 g/m^3h	5200 g/m^3h
Water Temperature	28.5 ^oC	23.5 ^oC
Disk rotational velocity	7.5 rpm	7.5 rpm
Diffusion coefficient of NH_4^+	2.0 cm^2/day	1.8 cm^2/day
Diffusion coefficient of O_2	2.4 cm^2/day	2.1 cm^2/day

subscript i refers to the concentration reference plane. The ammonia flux to the elemental biofilm at any time can be obtained as follows:

$$F_{A,n} = \frac{D_A}{\Delta Z}\left(C_{A(n,1)} - C_{A(n,2)}\right) \tag{12}$$

The average flux for all elements in the water phase at any time is shown in Eq. 13

$$\bar{F}_A = \frac{1}{t}\sum_{n=1}^{n=n} \Delta t \cdot F_{A,n} = \frac{D_A}{n\Delta Z}\sum_{n=1}^{n=n}\left(C_{A(n,1)} - C_{A(n,2)}\right) \tag{13}$$

Results and Discussion

The thickness of the diffusion layer, the intrinsic nitrification rate, and the relationship between the bulk concentrations of ammonia and dissolved oxygen were obtained in the previous experiment (2). The thickness of the attached water-layer was changed to match the simulated results with the experimental results. The thickness of 50 μm gave the best fit in both cases (Table 1). Fig. 5 shows the changes in the simulated concentration of ammonia and oxygen in the elemental biofilm with varying detention times for air and water phases. As shown in Fig. 5, the profile changes depended on the detention time in each phase, even for the steady-state conditions of bulk ammonia and dissolved oxygen concentrations. The dotted line represents the steady-state ammonia concentration profile predicted by the modified kinetics. Fig. 6 shows the average ammonia flux as a function of detention time in the bulk water. Fig. 7 shows the comparison between the simulated average flux and the flux obtained in the experiment. The simulation results based on the conditions shown in Table 1 compared favorably with the experimental data. Fig. 8 shows the effect of the attached water-layer thickness on the average flux. The thinner thickness gave a higher flux because of the high oxygen flux. However, the attached water-layer thickness was naturally determined as formulated in Eq. 7. Fig. 9 shows the effects of the dissolved oxygen concentration on the average flux. DO concentration was also naturally set at a level depending on the operational condition. Therefore, Figs. 8 and 9 show how to change the average flux, if the attached water-layer thickness and DO concentration are artificially controlled. Fig. 10 shows the average flux change with the disk rotating velocity at a fixed disk peripheral velocity of 7 m/min. With the disk at a fixed peripheral velocity, the average flux increased with the increase of the disk rotation-

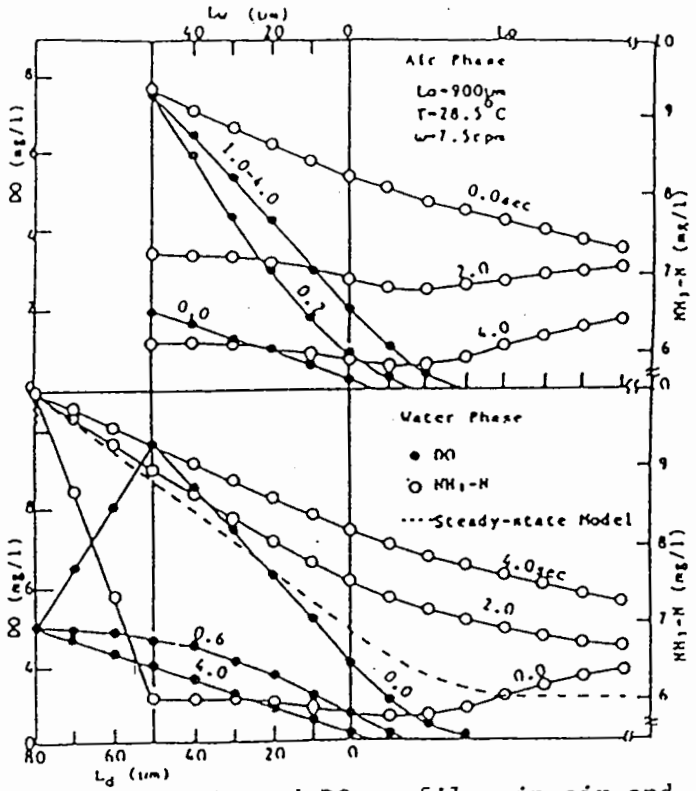


Fig. 5 Ammonia and DO profiles in air and water phases

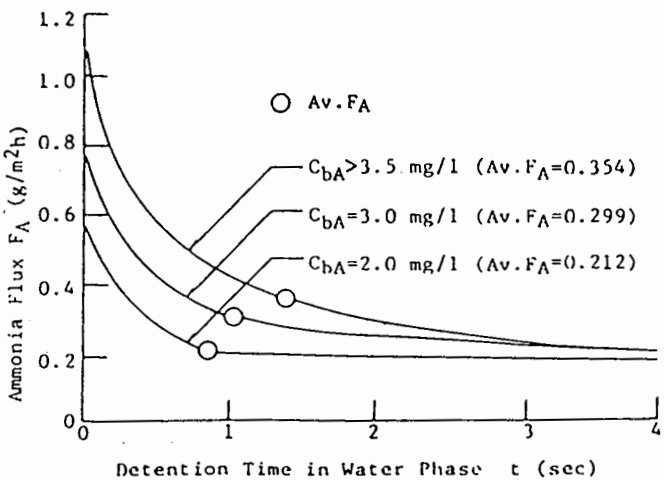


Fig. 6 Relationship between detention time in water and ammonia flux

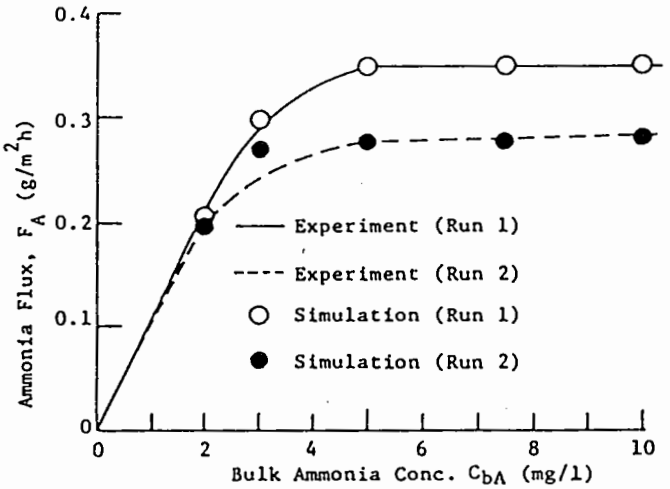


Fig. 7 Comparison of simulated flux with experimental flux

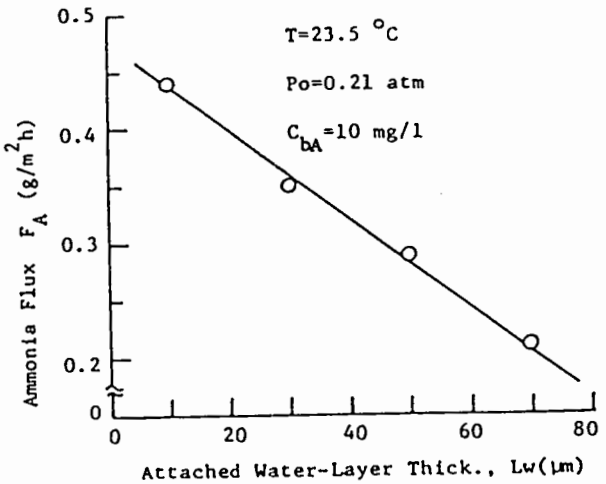


Fig. 8 Effect of attached water-layer thickness on ammonia flux

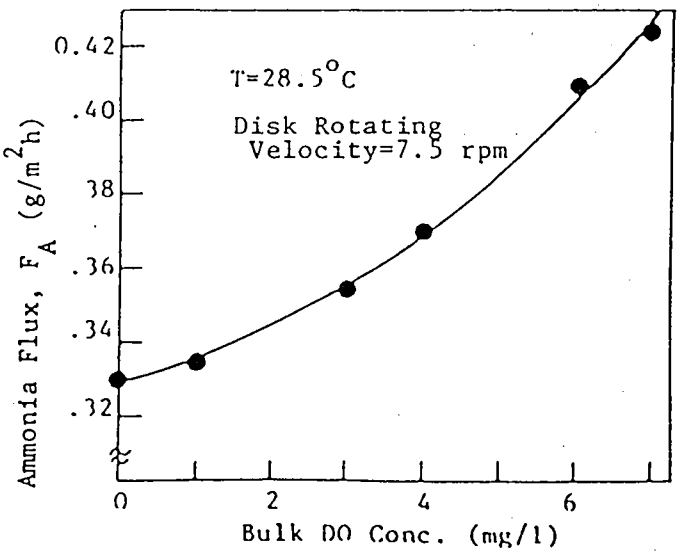


Fig. 9 Effect of DO concentration on ammonia flux

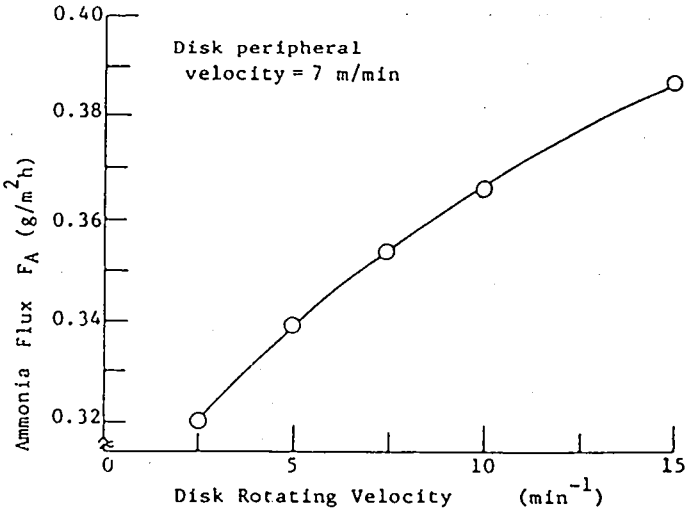


Fig. 10 Effect of disk rotating velocity on ammonia flux

al velocity under oxygen limiting, but the increment of the flux was very small compared with that of the disk rotational velocity.

The simulation study had clearly shown that the nitrification rate predicted by the modified kinetics would be equal to the average nitrification flux of each elemental biofilm. This fact provided the reasoning behind the application of the modified steady-state kinetics to the nitrification process in a partially submerged RBC. We concluded that the amount of ammonia nitrified within a biofilm rotating in the air phase can diffuse into a biofilm rotating in the water phase.

COMBINED CARBON OXIDATION-NITRIFICATION IN A PARTIALLY SUBMERGED RBC

Experimental Procedure

Two units with disk diameter of 30 cm were used for the experiment. Unit 1 consisted of 13 polywood disks mounted 2 cm apart on a horizontal shaft and a trough with a volume of 15 liters. Unit 2 consisted of 15 polywood disks, 2 cm apart on a horizontal shaft and a trough with a volume of 18 liters. The direction of the flow in both reactors was perpendicular to the rotating shaft. The residence time distribution of the water in the two reactors without biofilm perfectly coincided with that of a single completely mixed-flow reactor. The experimental variables were (a) hydraulic loading, (b) influent BOD_5 concentration, and (c) the type of organic matter (glucose and tapioca starch). Water temperature and pH were in the range of 23 ^{o}C to 27 ^{o}C and 7.8 to 8.2, respectively. The disk rotating velocity and the influent ammonia concentration were fixed at 8.5 rpm and 45 mg/l, respectively. Experimental conditions are summarized in Table 2. In each Run, samples were collected two or three times a day until the system reached a steady state.

Results and Discussion

(a) Effect of Organic Oxidation on Nitrification

The reduction of ammonia in combined carbon oxidation-nitrification process consists of a reduction due to nitrification and one due to cell synthesis, since the specific growth rate of heterotrophic bacteria is normally much higher than that of autotrophic nitrifying bacteria. Therefore, the ammonia utilized for the cell synthesis of heterotrophic bacteria cannot be neglected. The authors (7) have already evaluated the ammount of ammonia which would be utilized due to cell synthesis of heterotrophic bacteria at about 10 % of the BOD_5 reduction. Therefore, the total ammonia flux to the biofilm can be expressed by Eq. 14.

Table 2 Experimental conditions

Type of organic carbonaceous substrate	Run number	BOD_5/N	Hydraulic loading l/m^2h	Unit	Water Temp. °C
Glucose	1-1	0.6	5.1	2	27
	1-2	0.6	7.6	1	
	1-3	0.6	10.2	2	
	1-4	0.6	14.2	2	
	1-5	0.6	15.3	1	
	2-1	1.2	5.1	2	25
	2-2	1.2	7.6	1	
	2-3	1.2	10.2	2	
	2-4	1.2	15.3	1	
	2-5	1.2	21.2	1	
	3-1	2.8	3.9	1	25
	3-2	2.8	5.1	2	
	3-3	2.8	7.6	1	
	3-4	2.8	10.2	2	
	3-5	2.8	14.2	2	
Starch	4-1	2.7	3.4	2	25
	4-2	2.7	5.1	2	
	4-3	2.7	6.8	1	
	4-4	2.7	9.1	2	
	4-5	2.7	12.7	1	
	4-6	2.7	15.3	1	
	5-1	0.7	5.1	2	23
	5-2	1.4	5.1	2	
	5-3	2.7	5.1	2	
	5-4	4.0	5.1	1	
	5-5	5.8	5.1	1	

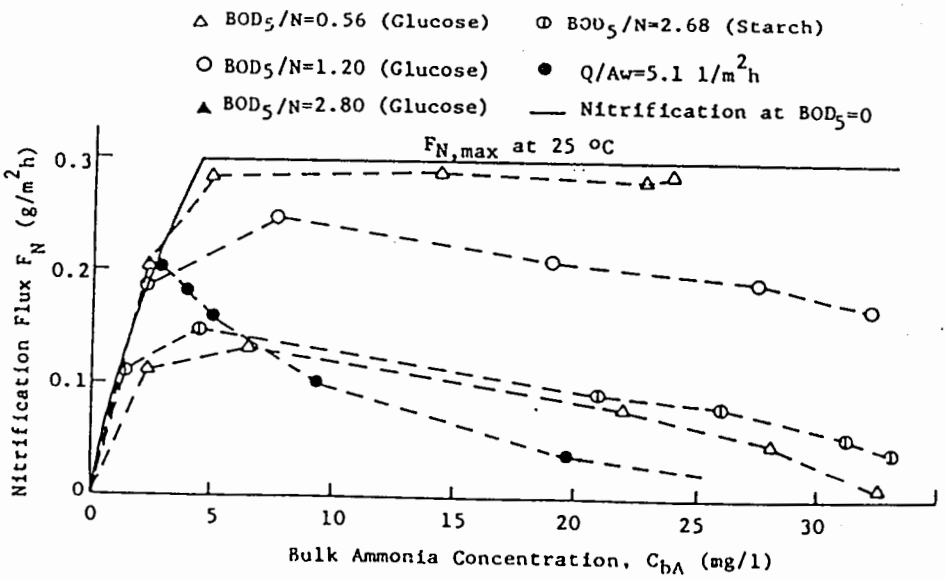


Fig. 11 Relationship between bulk ammonia concentration and nitrification flux

$$F_A = F_N + 0.1F_B \quad (14)$$

The relationship between the nitrification flux and the bulk ammonia concentration was obtained as shown in Fig. 11 by using measured values of F_A and F_B, and Eq. 14. Fig. 11 clearly shows the effects of organic carbon oxidation on nitrification. Carbon oxidation was also influenced by nitrification as shown in Fig. 12. The data represented by triangles were calculated by the kinetics for starch oxidation without nitrification. The data represented by circles were obtained in the experiment. In Fig. 12, it can be seen that the reduction of starch flux due to simultaneous nitrification was negligibly small until the bulk starch concentration reached about 40 mg/l (the corresponding BOD_5 was about 20 mg/l). When the bulk starch concentration increased beyond about 40 mg/l, the reduction of starch flux became remarkable, because the inner part of the aerobic biofilm mainly consisted of nitrifying bacteria. The results shown in Figs. 11 and 12 would come from the following hypothesis:

1) Most of the heterotrophic bacteria exist in the outer part of the aerobic biofilm while most of the nitrifying bacteria grow in the inner part of the aerobic biofilm as shown schematically in Fig. 1 (b). However, this would only be true when the specific growth rate of heterotrophic bacteria is much higher than that of nitrifying bacteria.

2) Both heterotrophic and nitrifying bacteria are aerobic and the amount of oxygen supplied to the biofilm would be almost the same under fixed operating conditions, independent of the aerobic bacteria composition under oxygen limiting.

Based on the above hypotheses, the following relationships applicable to a combined carbon oxidation-nitrification process in a partially submerged RBC have been developed.

$$F_o = 4.33\ F_{N,max} = 4.33\ F_N + F_B \quad (15)$$

Oxygen flux (F_o) is shown in Eq. 5. In addition, BOD_5 flux can be used to express the oxygen flux caused by the heterotrophic bacteria, if organic matter produces a straight increase of BOD against incubation time, i.e., the oxygen uptake rate is assumed to be constant independent of the residual organic concentration. This was almost true in the case of glucose and starch used in our experiment. Their BOD_5 per unit mass were 0.71 g O_2/g glucose and 0.55 g O_2/g starch. Both values were experimentally obtained (7). In dimensionless form, Eq. 15 becomes,

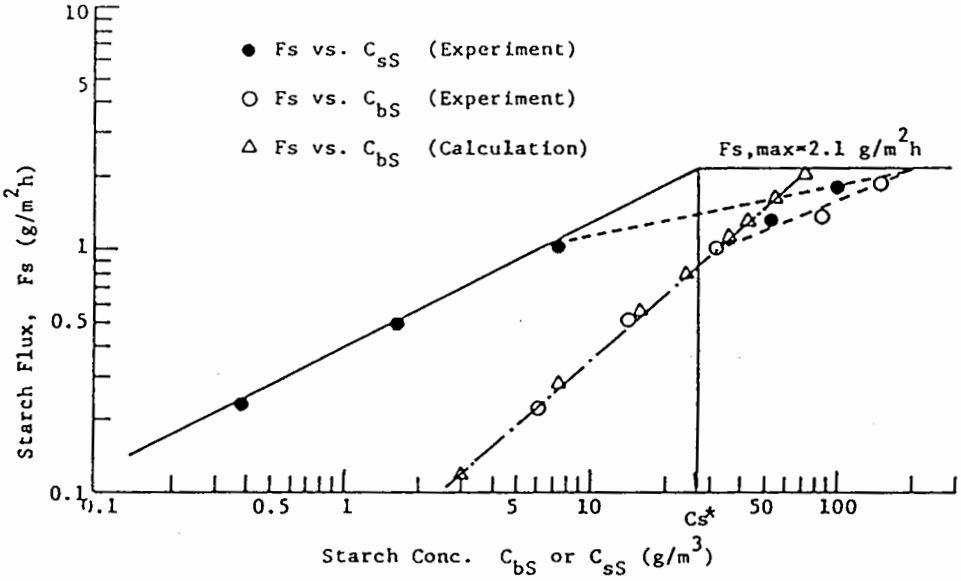


Fig. 12 Logarithm plots of starch flux

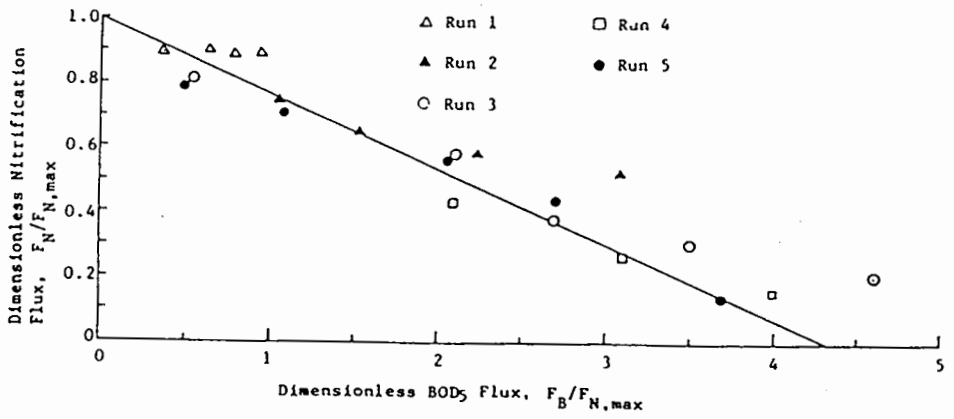


Fig. 13 Plots of BOD_5 flux and nitrification flux (straight line shows Eq.16)

$$\frac{F_N}{F_{N,max}} = 1 - \frac{1}{4.33}\frac{F_B}{F_{N,max}} = 1 - 0.23\frac{F_B}{F_{N,max}} \qquad (16)$$

Introducing Eq. 14 into Eq. 16 gives the relationship between the ammonia flux and the BOD_5 flux. In dimensionless form, it is expressed by Eq. 17.

$$\frac{F_A}{F_{N,max}} = 1 - 0.13\frac{F_B}{F_{N,max}} \qquad (17)$$

The maximum BOD_5 flux ($F_{B,max}$) is obtained when F_N is equal to zero, i.e., F_A is equal to 0.1 F_B.

$$F_{B,max} = 4.33\ F_{N,max} \qquad (18)$$

Figs. 13 and 14 show the experimental verification of Eqs. 16 and 17, respectively. In Runs 1 to 4 where the hydraulic loading increased at a fixed influent BOD_5 to NH_4-N ratio, the filamentous bacteria grew on the biofilm surface with the increase of hydraulic loading. Organic oxidation by the filamentous bacteria was not considered in the biofilm kinetics, therefore, the obtained BOD_5 and ammonia flux were higher than the predicted flux. As a result, most of the data in Runs 1 to 4 were slightly higher than the predictions.

(b) Comparison of Predicted Values with Existing Data

Fig. 15 shows the relationship between the bulk BOD_5 and the corresponding BOD_5 flux obtained in Run 5. Circles in Fig. 15 show BOD_5 flux without simultaneous nitrification as calculated by our kinetic model, explained below. Under operating conditions in Run 5,

$$F_{B,max} = 4.33\ F_{N,max} = 1.33\ g/m^2h\ ,$$

The molecular diffusion coefficient of starch was estimated by Fig. 16, because our kinetics states that the overall rate constant (K*) is equal to the mass transfer coefficient (K_d) when the gradient of the curve relating the bulk substrate concentration to the overall rate constant becomes zero. The diffusion layer thickness can be calculated, using the curve for pure nitrification, i.e., nitrification without simultaneous carbon oxidation.

$$K_{dA} = 0.1\ m/h = \frac{D_A}{L_d}$$

$$L_d = D_A/K_{dA} = (7.5 \times 10^{-6}\ m^2/h)/(1 \times 10^{-1}\ m) = 75 \times 10^{-6}\ m$$

The diffusion layer thickness was estimated at 75 μm on a disk of 30 cm diameter, rotating at 7.5 rpm, in water at a tempera-

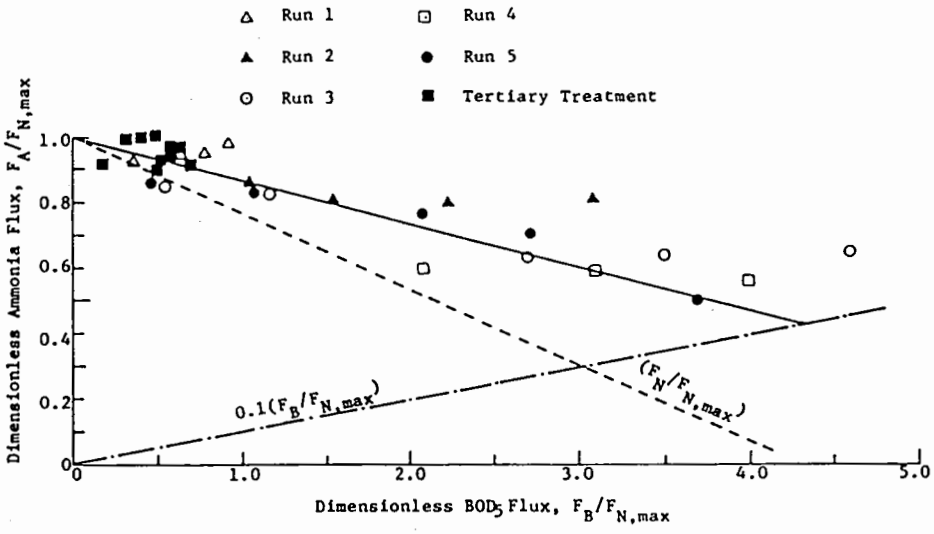


Fig. 14 Plots of BOD_5 flux and ammonia flux (straight line shows Eq.17)

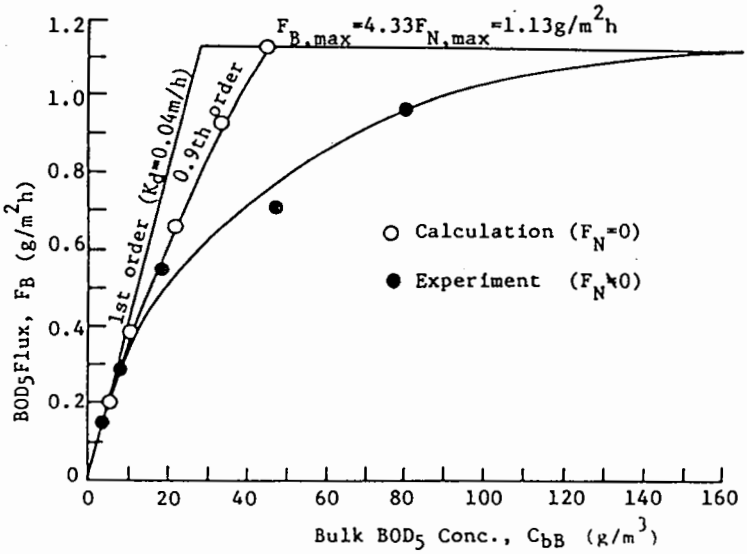


Fig. 15 BOD_5 flux vs. Bulk BOD_5 concentration(23 ℃)

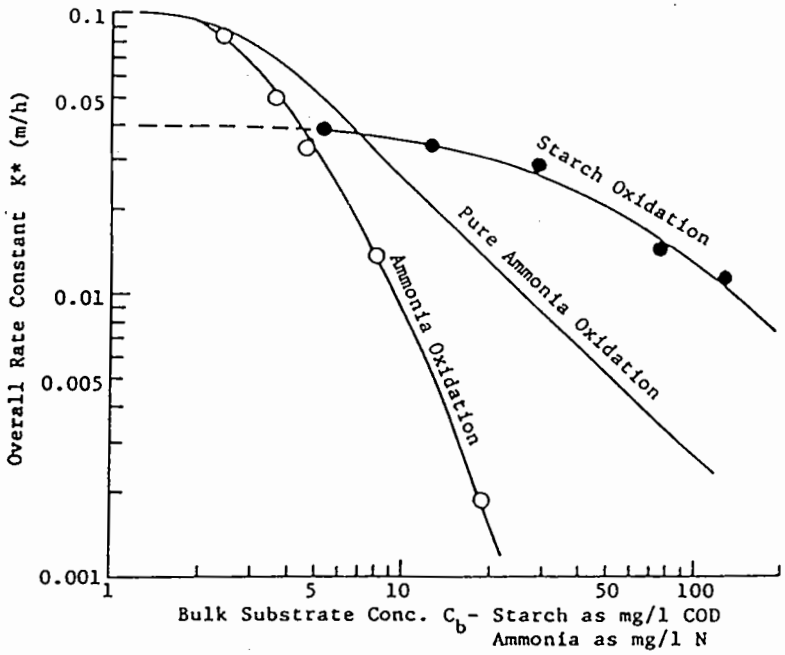


Fig. 16 Overall rate constant vs. bulk substrate concentration

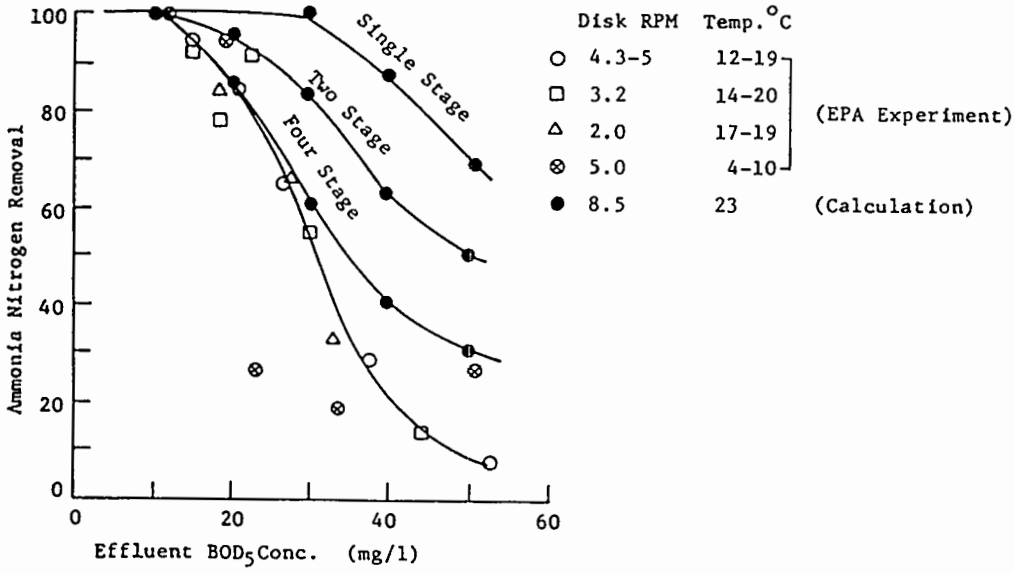


Fig. 17 Comparison of calculated value with USEPA data

ture of 23.5 °C. The thickness would be inversely proportional to the root of the disk peripheral velocity (3), therefore, in Run 5, it was estimated at about 70 μm. The molecular diffusion coefficient of starch was calculated as follows:

$$D_s = K_s^* L_d = (4 \times 10^{-2} \text{ m/h})(70 \times 10^{-6} \text{ m}) = 2.8 \times 10^{-6} \text{ m}^2/\text{h}$$

D_S equals 2.4×10^{-6} m^2/h, using Wilk and Chang's Equation (9). We employed 2.5×10^{-6} m^2/h as the molecular diffusion coefficient of starch in the calculation. Then, the intristic starch oxidation rate was determined as 3×10^4 g/m^3h, using the half-order plots shown in Fig. 12.

The ammonia flux at 23°C for any bulk BOD_5 can be predicted by using the curve in Fig.15 and Fig.14. The caluculated relationship between the effluent BOD_5 and the percent ammonia removal for a multi-stage completely mixed-flow RBC with the same total disk surface area is shown in Fig.17 along with US EPA data(10). The caluculation was made for an influent ammonia concentration of 30 mg/l and an influent BOD_5 concentration of 150 mg/l. USEPA data were collected in a two-stage RBC in which the direction of flow was parallel to the rotating shaft. The average influent Kjeldahl nitrogen and BOD_5 concentrations were 28.9 mg/l and 147 mg/l,respectively.

SUMMARY AND CONCLUSIONS

A modified biofilm kinetics for a partially submerged RBC process was developed. The proposed kinetics was applied to the nitrification process with and without simultaneous carbon oxidation. A computer simulation of nitrification based on the assumptions for the model development produced an average ammonia flux of the elemental biofilm rotating in the water phase that was almost equal to the flux calculated by the modified kinetics. This provided the reasoning behind the application of the modified steady-state kinetics to a partially submerged RBC in which the biofilm alternately rotates into the water and air.

In a partially submerged RBC, the oxygen transfer to the biofilm mainly occurs when the biofilm rotates in the air phase, while the biofilm rotating in the water phase adsorbs the substrates. The oxygen transfer rate through the attached water-layer or the oxygen flux to the biofilm rotating in the air can be shown by the following equation:

$$F_o = \frac{D_o(C_o^* - C_{so})}{L_w}$$

The maximum nitrification flux without simultaneous carbon oxidation is represented by the following equation:

$$F_{N,max} = \frac{F_o}{4.33\ R_n}$$

The authors considered that the amount of oxygen transported to the biofilm or consumed within the biofilm would be almost the same under fixed operating conditions, independent of the aerobic bacteria composition under oxygen limiting. Based on the above hypothesis, we proposed the following equations for nitrification flux and for ammonia flux in combined carbon oxidation-nitrification in a partially submerged RBC. Nitrification flux (F_N):

$$\frac{F_N}{F_{N,max}} = 1 - 0.23\frac{F_B}{F_{N,max}}$$

Ammonia flux (F_A):

$$\frac{F_A}{F_{N,max}} = 1 - 0.13\frac{F_B}{F_{N,max}}$$

The relationship between the effluent BOD_5 concentration and the percent ammonia removal calculated by the above equation and the experimental data almost coincided with that obtained in USEPA experiments.

Acknowledgment

The authors would like to express their appreciation and thanks to Japan International Cooperation Agency (JICA) for financial assistance.

NOMENCLATURE

Symbol	Dimension	Description
C_A	g/m^3	Ammonia concentration within biofilm
C_{bA}	g/m^3	Bulk ammonia concentration
C_{sA}	g/m^3	Ammonia concentration at biofilm surface
C_{sA}^*	g/m^3	Critical ammonia concentration at biofilm surface
C_{bS}	g/m^3	Bulk starch concentration
C_{sS}	g/m^3	Starch concentration at biofilm surface
C_{iA}	g/m^3	Influent ammonia concentration

C_o^*	g/m^3	Saturation concentration of oxygen
C_{sO}	g/m^3	Oxygen concentration at biofilm surface
D_A	m^2/h	Molecular diffusion coefficient of ammonia
D_o	m^2/h	Molecular diffusion coefficient of oxygen
D_s	m^2/h	Molecular diffusion coefficient of starch
F_A	g/m^2h	Ammonia flux
F_B	g/m^2h	BOD_5 flux
F_N	g/m^2h	Nitrification flux or ammonia flux due to nitrification
F_o	g/m^2h	Oxygen flux
K_d	m/h	Mass transfer coefficient
L_d	m	Diffusion layer thickness
L_n	m	Oxygen penetration thickness in nitrifying biofilm
L_w	m	Attached water-layer thickness
R_n	g/m^3h	Intrinsic nitrification rate
R_o	g/m^3h	Intrinsic oxygen uptake rate
R_s	g/m^3h	Intrinsic starch oxydation rate
μ	g/cm.sec	Absolute viscosity of water
ω	rps	Disk rotating velocity
r	cm	Disk radius

REFERENCES

1. Watanabe Y. and Ishiguro "Denitrification Kinetics in a Submerged Biological Disk Unit," Progress in Water Technology,Vol.10, Nos.5/6, 1978, pp.187-195
2. Watanabe Y. et. al. "Nitrification Kinetics in a Rotating Biological Contactor," Progress in Water Technology,Vol.12, 1980, pp.233-251
3. Watanabe Y. et. al. "Simulation of Nitrification and Its Dynamics in a Rotating Biological Contactor," Progress in Water Technology, Vol.14, 1982
4. Wezernak C.T. and Cannon J.J. "Oxygen-Nitrogen Relationships in Autotrophic Nitrification," Applied Microbiology, Vol.15, No.5, 1967, pp.1211-1215
5. Bintanja H.H.J. et. al. "Oxygen Transfer in a Rotating Disc

Treatment Plant", Water Research, Vol.9, 1975, pp.1147-1153
6. Hartman H. "Untersuchung über die Biologische Reinigung von Abwasser mit Hife von Tauchtropfkrorpern, Kommissionsverlag R. Oldenbourg München, 1960
7. Watanabe Y. and Thanantaseth C. "A Study on Purification Mechanism of Rotating Biological Contactor (III)", submitted to Journal of Japan Sewage Works Association
8. Ishiguro M. and Watanabe Y. "A Study on Tertiary Treatment of Municipul Sewage by Rotating Biological Contactor (II), Journal of Japan Sewage Works Association, Vol.14, No.152. 1977, pp.32-41
9. Welty J.R. et.al. "Fundamentals of Momentum, Heat and Mass Transfer, John Wiley and Sons Inc., 1969, pp.463-465
10. USEPA "Application of Rotating Disc Process to Municipal Wastewater Treatment", Water Pollution Control Research Series, 17050 DAM 11/71

PART IV: CONCEPTS AND MODELS

SELECTION AND OPTIMIZATION PROTOCOLS FOR ATTACHED GROWTH BIOLOGICAL PACKED COLUMNS

Sheldon F. Roe, Jr., P.E., Manager, Technical Market Research, The Munters Corporation, Fort Myers, Florida

Edward B. Hanf, Vice President, Director Sales and Marketing, The Munters Corporation, Fort Myers, Florida

1.0 INTRODUCTION

Attached growth has long been used for water treatment in packed columns. Now the idea of producing fuels, foods, or chemicals by similar techniques (anaerobic digestors for methane or columns for ethanol) promises an exciting future in this field.

Our reference point is cooling towers, SO_2 scrubbers, chemical absorption-desorption and tube settlers, in addition to various attached growth mechanisms. Many of these columns share common problems, but they also share common advantages for optimization and for adaptation to new processes and systems.

Too often we hear the comment "I tried your fixed film system and it didn't make any difference." Most likely the system was not operating at capacity and, indeed, the attached growth didn't make any difference. It is the purpose of this paper to provide guidelines for process selection and optimization.

Topics of discussion include: trade-offs, column characteristics, operating characteristics, solids handling, process selection, operating analysis, and recommendations. The objective is to design the process to fit the biocolumn rather than to adapt the biocolumn to the process.

2.0 DISCUSSION

2.1 Efficiency, Through-put, Pressure Drop Trade-off

One common characteristic of columns is the trade-off between efficiency, through-put and pressure drop as illustrated in Figure 1. Traditionally, high efficiency, high through-put, and low pressure drop are the ideals we are looking for, while, on the other hand, low efficiency, low through-put and high pressure drop are what we most seek to avoid.

Between these extremes there exist twelve combinations of high and low efficiency, through-put and pressure drop which we must consider. Let us start with the most forgiving combinations.

Traditionally, high efficiency and low pressure drop (sometimes called HELPD Packings) are the ideal combination, allowing for some trade-off as far as through-put is concerned. The combination of high efficiency and high through-put, on the other hand, allows for flexibility regarding pressure drop. While the combination of high through-put and low pressure drop permits a tolerable trade-off on the efficiency.

At the other extreme, low through-put and high pressure drop represents the least forgiving combination and can only be compensated for by a high efficiency. Similarly, low efficiency and low through-put don't offer much opportunity for compensation. Listed below are combinations which do, on occasion, offer viable trade-offs. In each case, the third component of the trade-off or some other un-named factor must compensate for the undesirable aspect as listed.

1. High efficiency and high pressure drop
2. Low efficiency and high pressure drop
3. Low efficiency and low pressure drop
4. Low efficiency and high through-put
5. High efficiency and low through-put
6. High through-put and high pressure drop
7. Low through-put and low pressure drop

For the purpose of this discussion, pressure drop and pumping head are used interchangeably since both represent resistance to flow and operating costs.

The above is a somewhat over simplified view of complex real-life situations. But let us throw in another factor here which comes with attached growth. This is fouling. We will use the same approach as above, but now with the extra factor added. (see Figure 2)

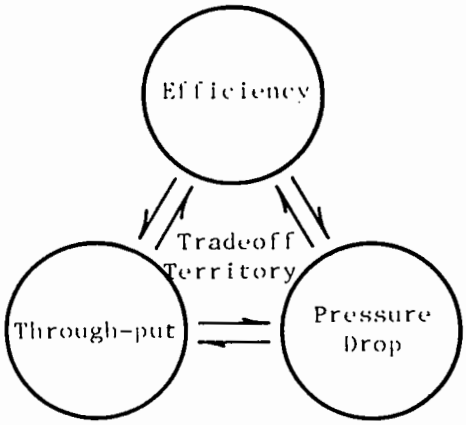


Figure 1

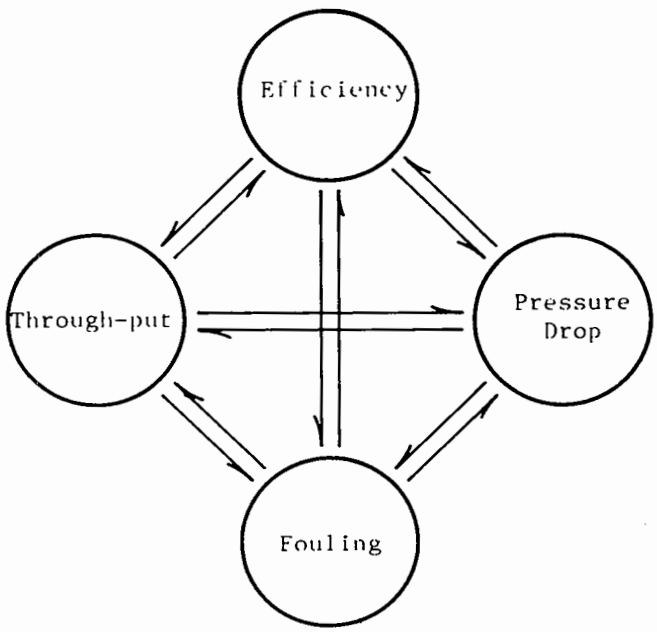


Figure 2. Tradeoff Territory

2.2 Definition of Column Characteristics

An arbitrary list of 8 parameters which we think are important for biocolumns is given in Figure 3 (a matrix of these parameters with themselves). The shaded areas define the subject of principle interest in this paper. As the light areas indicate we will talk very little about time, biochemistry, energy, or economic analysis. The areas are subdivided as shown in Figure 4. This results in a far more complex matrix which demonstrates not only combinations of parameters but combinations within parameters. And it will, in fact, be desirable to discuss some of these internal combinations (for example 2.3 with 2.1 - the combination of an entrained bed with a fixed bed). The purpose of the matrix is to ensure that all possible alternatives are evaluated in the selection of a given process.

After considering the above, an intensive look at the operating characteristics of an individual column is in order. The morphology (shape) of the column and packing material are important. These are considered in the next two sections.

2.3 Operating Characteristics of an Individual Column--Column Morphology

The interaction of flow and velocity is illustrated in Figure 5. For a given through-put, a counterflow column can have either of two extreme shapes. In Case No. 1 the column is long and slender, operating at high velocity. The pressure drop of the fixed bed per $\Delta\ell$ of the column length must be low or the pressure drop of the total column will be prohibitively expensive. The efficiency per length of column may be low, but the column can be extended in length to compensate for this low efficiency per unit of length.

The shape of the packing material in a fixed bed in such a column of course influences the length required for a suitable efficiency. Here, large edge effects may be expected because the substance flowing through the column would rather go to the walls than through the center of the column. Redistribution may be necessary to counteract this. Opportunities for channeling in this column of Case No. 1 are minimal when compared to the column of Case No. 2. Flow conditions can be important; i.e., under conditions of a low Reynolds Number separation can take place. But with a high Reynolds Number the column can actually be a mixer.

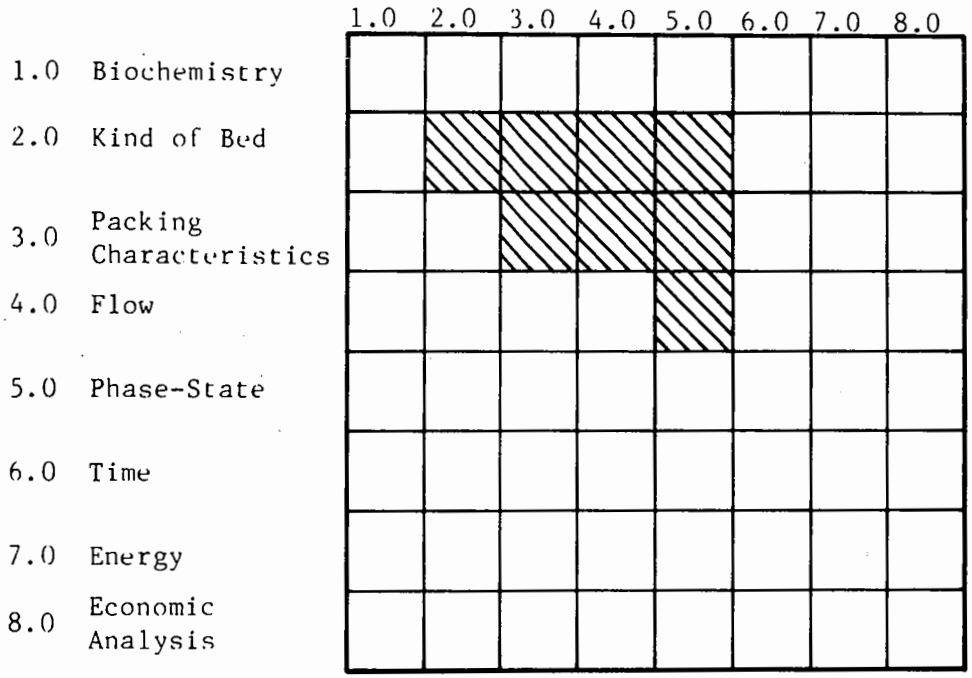


Figure 3. Biocolumn Matrix

2.0 KIND OF BED
- 2.1 FIXED
- 2.2 FLUIDIZED
- 2.3 ENTRAINED
- 2.4 MOVING-MECHANICAL
- 2.5 DEGRADABLE

3.0 PACKING CHARACTERISTICS
- 3.1 ORDERED SHAPES
  - 3.1.1 TUBES
  - 3.1.2 SHEET
  - 3.1.3 OTHER-DIRECTION ORIENTATION
- 3.2 RANDOM SHAPES
  - 3.2.1 SPHERICAL
  - 3.2.2 FIBERS
  - 3.2.3 OTHER-BLOCKS
- 3.3 SURFACE AREA
  - 3.3.1 MACRO
  - 3.3.2 MICRO
  - 3.3.3 MOLECULAR
- 3.4 STEADY STATE-CHANGES
  - 3.4.1 DEGRADABLE

4.0 FLOW
- 4.1 FLOW DIRECTION
  - 4.1.1 CROSSFLOW
  - 4.1.2 COUNTERFLOW
  - 4.1.3 COCURRENTFLOW
- 4.2 FLOW CHARACTERISTICS
  - 4.2.1 VELOCITY
  - 4.2.2 GRAVITY
- 4.3 INTERMEDIATE FEEDS, RECYCLE, STAGING

5.0 PHASE-STATE
- 5.1 SOLIDS
  - 5.1.1 SAND
  - 5.1.2 COAL
  - 5.1.3 BIOSOLIDS
  - 5.1.4 FIXED FILM
  - 5.1.5 PLUGGING
- 5.2 GAS
  - 5.2.1 BUBBLES-FOAMING
  - 5.2.2 CONTINUOUS PHASE
- 5.3 LIQUID
  - 5.3.1 LIQUID FILM
  - 5.3.2 LIQUID DROPS
- 5.4 REACTANT VS. CARRIER

6.0 TIME
- 6.1 SOLIDS RETENTION TIME
- 6.2 LIQUID RETENTION TIME
- 6.3 GAS RETENTION TIME

7.0 ENERGY
- 7.1 MATERIAL-ENERGY BALANCE
- 7.2 HEAT
- 7.3 KINETIC
- 7.4 CHEMICAL REACTION

Figure 4. Matrix Detail

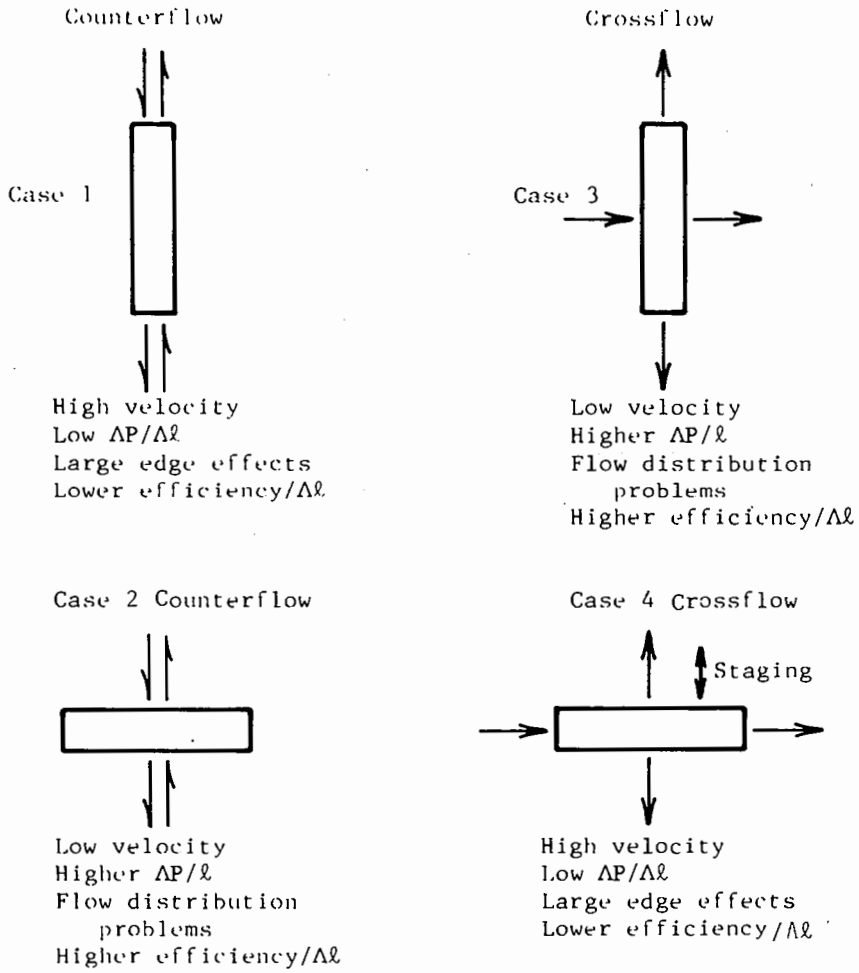


Figure 5. Gross Morphology of Biocolumns

In case No. 2 we have a column of large flow area and low velocity. Rather than edge effect, here the problem is one of obtaining equal flow distribution over the large area. Since this column is operating at a lower Reynolds Number, it may be a better separator than mixer. A plug or piston-type flow may be more difficult to obtain in a large column and may present problems in scaling up into larger diameters. For example: how does one get plug flow in a fixed bed 100 feet in diameter and 6 feet high?

The crossflow versions of this problem are illustrated in Cases 3 and 4 of Figure 5. In Case 4, staging offers a means of successively making counterflow in a horizontal direction.

2.4 Handling Solids--Packing Morphology

Handling solids in a fixed bed with a continuous liquid phase presents an important problem in controlling fouling. An analysis of the surface area yields the conclusion that the shape of the surface is as important as the amount of area. This is illustrated in Figure 6 where the distribution of area with the column axis is presented. The right-hand side of the horizontal axis in Figure 6 represents area perpendicular to the flow. This area must be kept to a minimum to prevent solids build up. Because this area is generally small for all cases described (in some cases only 10 or 20% of the column cross section), differences in solids build-up can be experienced which are by orders of magnitude.

The left-hand of the horizontal axis represents surfaces parallel to flow. From the biofilm standpoint, these surfaces would more closely approach laminar flow while the surfaces toward the right-hand of the axis represent progressively higher shear rates. However, this is not absolutely true because the distribution along the column axis must also be considered.

It is also interesting to define the parameters in Figure 6 in terms of the phase, with the following possibilities of contacting the shape of the surface:

1. Continuous gas phase contacting a liquid film.
2. Continuous gas phase contacting liquid drops to a liquid film.
3. Continuous gas phase contacting solid particles on a liquid film.

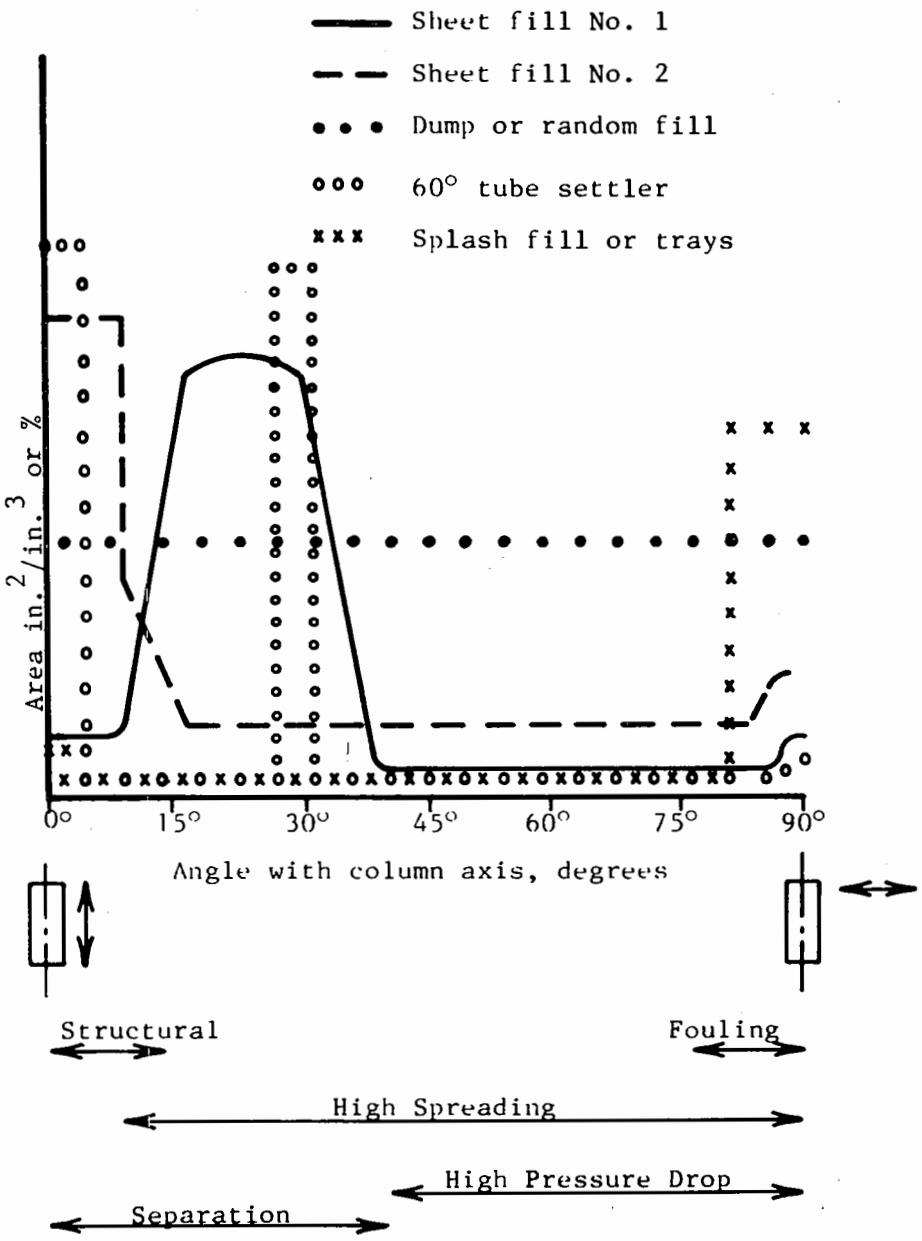


Figure 6. Area Orientation of Fill in a Tower--Packing Morphology

4. Continuous liquid phase contacting gas bubbles.
5. Continuous liquid phase contacting solid particles.
6. Continuous liquid phase contacting a biofilm combined with gas bubbles or solid particles as in 4 or 5 above.

After considering the morphology of columns (Figure 5) and the morphology of packings (Figure 6), the protocol proceeds to a selection decision tree and an intensive analysis of column operating parameters in the next two sections.

2.5 Selecting a Process

Each of the expanded categories in Figure 4 have been arranged by coded number in a decision analysis tree. This, in combination with a matrix as in Figure 3, yields one means for selecting a process. The interdependence of alternates is not accounted for in this system. The selection decision in Figure 7 is a fixed bed 2.1 (Figure 4), sheet packing (3.1.2), crossflow (4.1.1), and fixed film (5.1.4). Admittedly, this is arbitrary and the selection sequence can be changed to emphasize important criteria first. Figure 4 also straddles the fence between pure logic and existing columns such as: fluidized beds with sand or coal, woven fabric, glass fiber discs, glass beads, Raschig rings, sheet packings, bricks, corn stalks, or chunks of foam.

2.6 Analyzing Operation of a Column

Another step to be considered is given in Figure 8. This is a more intensive analysis of the given column. Indeed, this is the subject of the usual operational study of a column. Each column has its own operating characteristics or range of values for acceptable operation depending on the value analysis in the trade-off of efficiency, through-put, and pressure drop.

In addition to the parameters named in Figure 8, several other common sense limitations should be considered including: dissolved gas limitation, film transfer limitation, substrate limitation, and fixed film solids vs. suspended solids. Processes such as gas bubble release, settling, coagulation, or liquid drop mechanics must be analyzed on a microstructure basis.

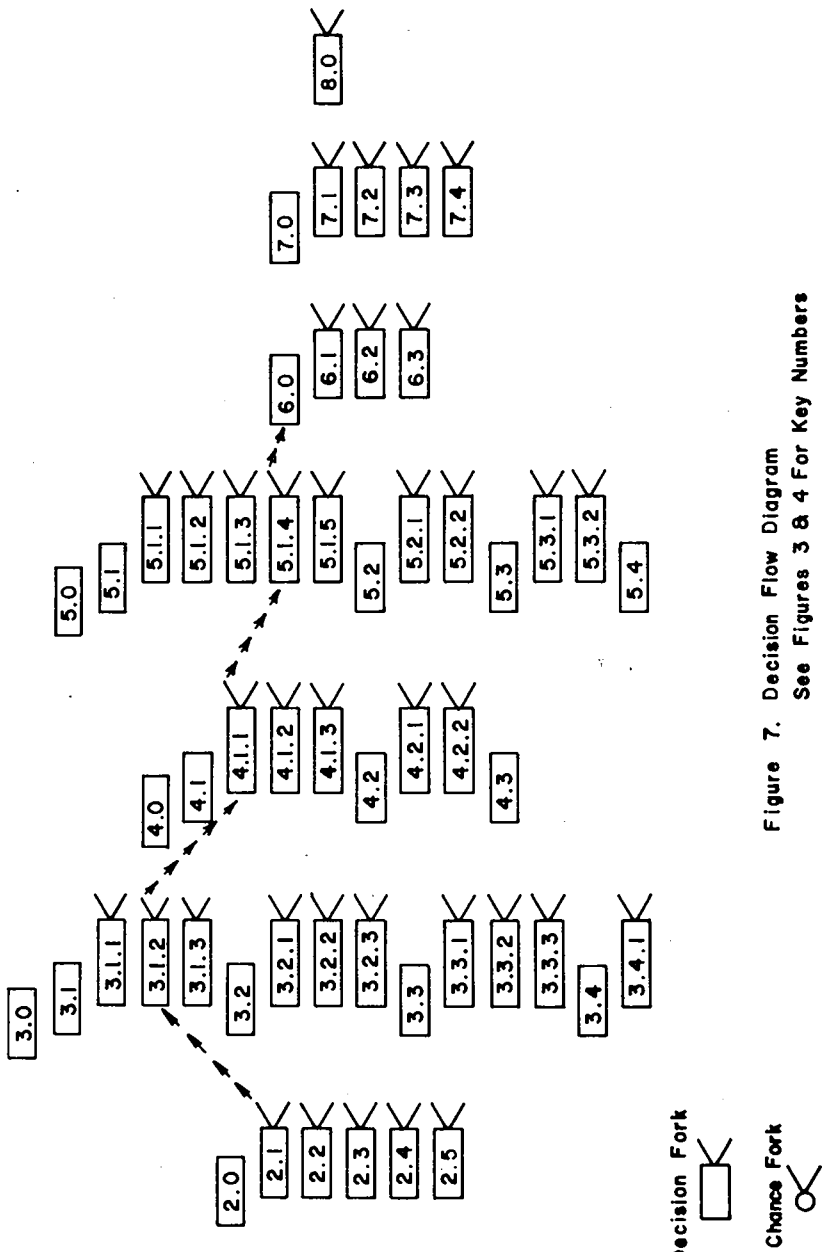


Figure 7. Decision Flow Diagram
See Figures 3 & 4 For Key Numbers

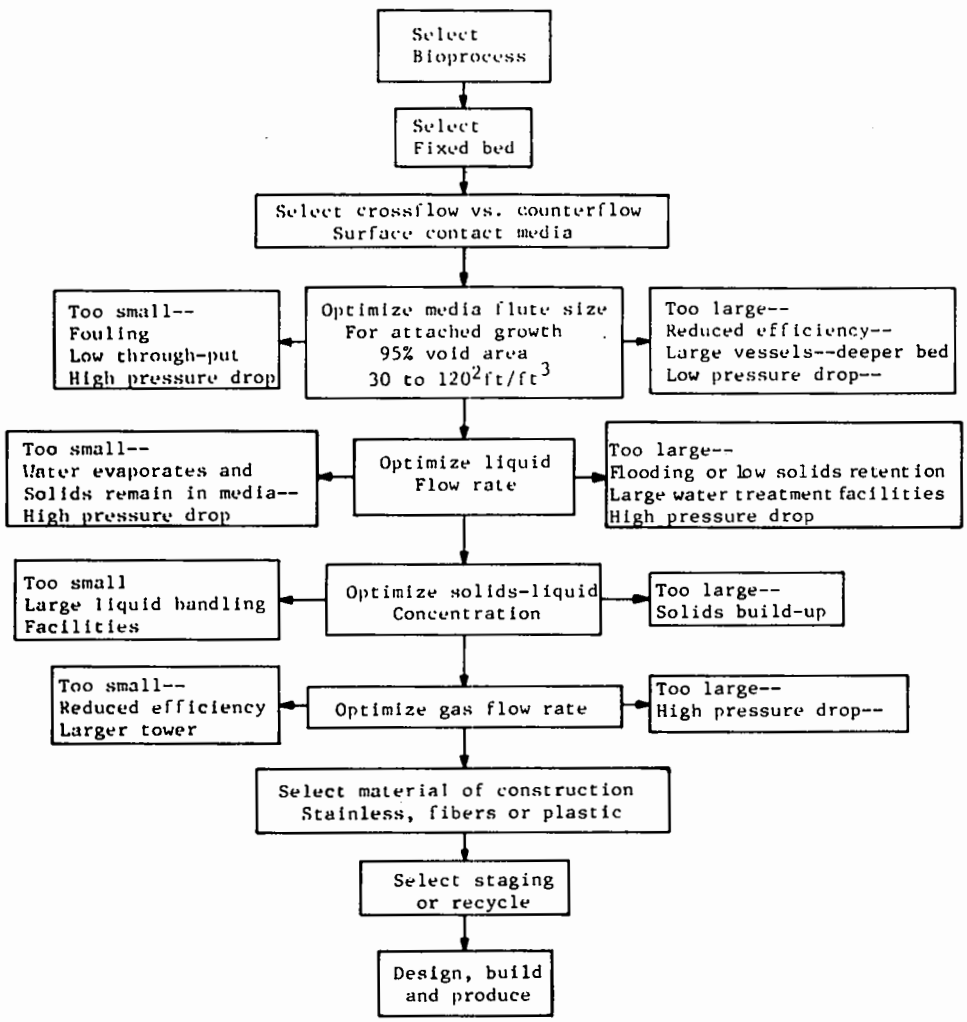


Figure 8. Optimization of Surface Contact Media

3.0 RECOMMENDATIONS

Selection and optimization of biological columns is complex but may contribute as much to the success of an operation as the biochemical process itself. Fouling, channeling, and low efficiency are to be avoided. However, with proper selection through optimization, the contribution of a column can be as important as the biochemical process in the success of the many new processes on the horizon.

Several alternatives to conventional processes are suggested below:

1. The combination of entrained or fluidized beds with fixed beds where particles in the continuous liquid phase contribute surface area and the fixed bed contributes operating stability of attached growth, flow modification, or a separation characteristic.
2. Trickle beds with gases other than air - e.g. nitrogen or carbon dioxide.
3. Crossflow trickle beds.
4. Crossflow columns with fixed beds and liquid continuous phase for release of carbon dioxide along the length of the column.
5. Staging in both crossflow and counterflow where higher surface area fixed bed is utilized at the latter stages.
6. Microstructures such as glass or cellulose fibers utilized in fixed bed for immobilization while maintaining adequate passages to prevent fouling.
7. Moving mechanical beds which operate at non-steady conditions where the fixed film is subjected to alternating conditions as catalysts and regeneration in conventional chemical processing.
8. The stability (resistance to operating upsets) of attached growth beds should be utilized more fully: a. to treat toxic chemicals, b. for pure cultures (not biological soups) to produce chemicals. This stability, known for years in water treatment, might be compared to immobilization in biotechnology. For monocultures sterilization could be a solvable problem. Sterilization could be considered the antonym of operating stability.

MODELING OF BIOLOGICAL FIXED FILMS -- A STATE-OF-THE-ART REVIEW

C. P. Leslie Grady, Jr. Department of Environmental Systems Engineering, Clemson Univeristy, Clemson, South Carolina.

INTRODUCTION

Although fixed-film biological processes found early application to wastewater treatment their use declined with the development and wide-scale adoption of the activated sludge process. There were many reasons for this, ranging from trivial to well-founded, but the net result was that for many years fixed-film processes (most notably the trickling filter) were relegated primarily to the treatment of low flow domestic wastewater or to the pretreatment of industrial wastewater. During the early 1960's however, with the advent of plastic media for trickling filters, there was a resurgance of interest in fixed-film reactors. This interest was stimulated in the late 60's and early 70's by the development and commercialization of rotating disk reactors which provided many of the benefits of trickling filters without some of the disadvantages. Finally, in the late 1970's and early 1980's fluidized bed biological reactors moved from the laboratory to the field, thereby opening up a whole new area for application of biological fixed films. Because of these advances in

process development and because fixed-film reactors are generally less energy intensive than activated sludge, engineers are now employing fixed-film biological processes in a host of new applications with a great deal of success.

Concurrently with the new developments in fixed-film reactors has come a renewed interest in their modeling. There are at least two reasons for this. One is that models are the basic tools of engineering which facilitate the design process. The other is that models help us achieve a better understanding of something by guiding our analysis of it. Modeling and experimentation are interdependent, with each providing input to and taking information from the other. Consequently, as we have learned more about fixed-film processes we have been able to develop better models which have helped us to see new applications and to develop better methods for design.

Mathematical models may be divided into two categories: empirical and mechanistic. Empirical models simply relate operating input and output variables to each other and make little pretense of representing individual phenomena. Such "black box" descriptions are quite useful for design from pilot plant data and have found wide use in environmental engineering. Many of the models for biological fixed-film processes fall into this category. Mechanistic models, on the other hand, express the influences and interrelationships of individual mechanistic phenomena in a manner which allows the investigator to discover how the system might respond under unexplored conditions. Thus one might argue that the primary purpose of a mechanistic model is to further understanding. This additional understanding will be of direct benefit to the practitioner, however, because it is the nature of practice to apply knowledge to areas in which no prior experience exists. Mechanistic models have broader utility than empirical ones. Consequently this review will be limited to models of that type.

Mechanistic models of biochemical processes generally are developed by application of reactor engineering principles, i.e., they combine expressions representing the intrinsic kinetic and transport events with mass balance equations describing the characteristics of the particular physical system under consideration. Consequently, simulation with such models give insight into the basic events occurring within a process as well as the influence that the system configuration has upon the outcome of those events. When we examine fixed-film biological processes we see that in spite of the

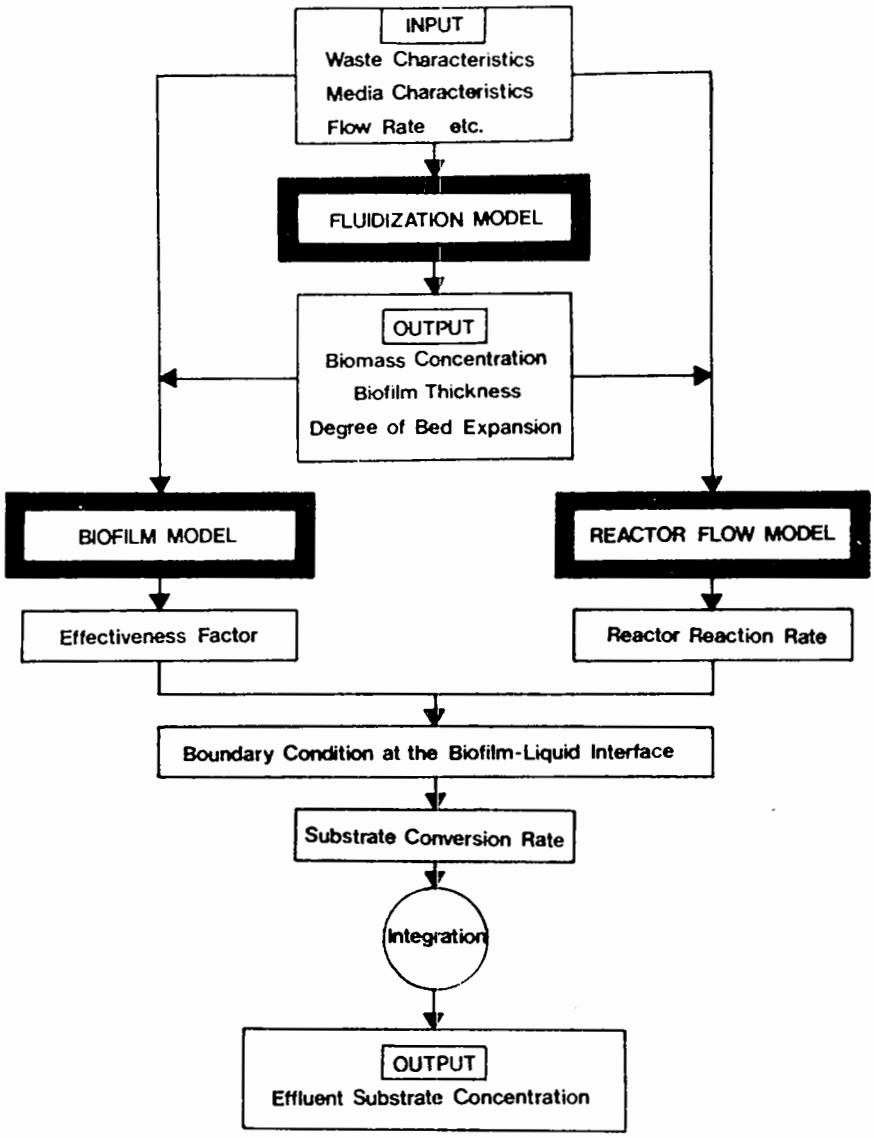


Figure 1. Flow diagram of model for a fluidized bed biological reactor illustrating interfacing of biofilm model with model of the physical characteristics of the reactor. (From Shieh and Mulcahy (1).).

diversity in physical configuration, all contain biofilms. Consequently, we might model these processes by developing a mechanistic model for the biofilm and then interfacing it with appropriate models for each of the physical process configurations. Figure 1 adapted from Shieh and Mulcahy (1) illustrates the application of this approach to a fluidized bed biological reactor. Similar flow diagrams could be developed for other fixed-film reactors but they would differ in the way in which the biofilm submodel is interfaced with the other system submodels. It therefore follows that a prerequisite to successful modeling of fixed-film biological processes is a realistic model for the biofilm. Do we have one? How have researchers sought to develop one? Is there concensus in the approach that is being taken? The purpose of this paper is to address questions such as those.

THE BIOFILM

The first step in the development of a mechanistic model for a system is its reduction to its essential components. Figure 2 is a schematic of the essential components required to model a biofilm. As shown there, organisms in a biofilm with density or concentration X_f grow attached to a solid support. In a trickling filter that support is either rock or plastic, in a rotating disk reactor it is plastic, and in a

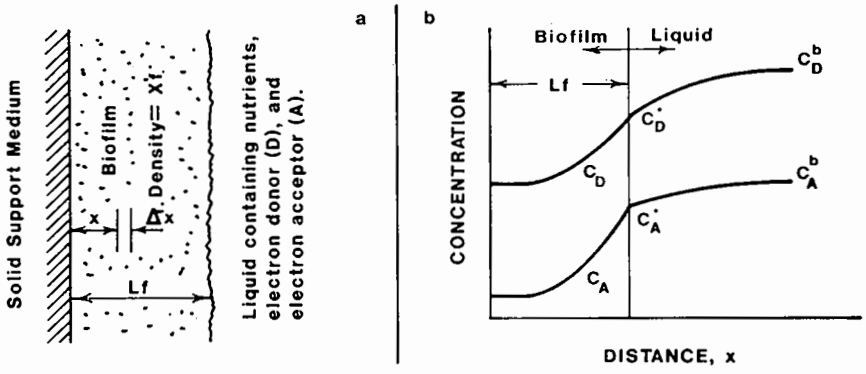


Figure 2. Schematic diagram of biofilm.

fluidized bed biological reactor it is sand, coal, or some other granular material. In the first two reactor types the radius of curvature of the support is large with respect to the biofilm thickness so that the support may be considered to be flat. In the third this may not be true, so spherical particles are generally assumed, although it has been shown that the solutions for particles are similar to those for slabs when a suitable characteristic dimension is chosen (2). Growth continues until some thickness L_f is attained, with the method of control of that thickness being a function of the type of process being considered. Adjacent to and permeating the biofilm is a liquid layer whose total thickness depends upon the type of process, and in some cases, upon the time within the process. Growth of the organisms is dependent upon the transport of an electron donor, an electron acceptor and nutrients through the liquid layer and into the film. Generally, nutrients are provided in excess, so the electron acceptor and donor are the only constituents considered. Since it is possible for transport of either the donor or acceptor to limit the rate of growth of the organisms in the biofilm, knowledge of their concentrations in the bulk liquid (C_D^b or C_A^b), at the interface (C_D^* or C_A^*), and in the biofilm (C_D or C_A), is quite important. The relationship between C^b and C^* is influenced by the nature of the process and represents a place where the biofilm model must be interfaced with the process model. Furthermore, the change of C_D or C_A with depth in the biofilm is influenced by the relative concentrations of the electron donor and acceptor at the interface, the thickness and physical properties of the biofilm, and the kinetics and stoichiometry of the biochemical reactions. Ths first two groups of characteristics are process dependent, and thus represent additional connections with the process model. The last group is an intrinsic characteristic of the transformations being carried out in the reactor.

From consideration of the essential characteristics illustrated in Figure 2 it can be deduced that development of a biofilm model requires knowledge in the following areas: (a) transport of materials in the liquid phase; (b) characteristics of the biofilm, including its thickness, density, and composition; (c) transport and reaction within the biofilm; and (d) techniques for solving the resultant equations. In the following sections each of these areas will be reviewed in depth.

TRANSPORT OF MATERIALS IN THE LIQUID PHASE

There is abundant evidence that the rate of transport of materials from the bulk liquid to the biofilm:liquid interface can be an important determinant of the performance of a fixed-film process. The references cited below to demonstrate this phenomenon should be viewed as representative of the broader body of literature rather than as all-inclusive. This same caveat should also be applied to the remainder of this article because inclusion of all literature dealing with the mechanistic modeling of fixed-films was beyond the scope of this endeavor.

The clearest evidence for external mass transport and the necessity for its inclusion comes from microprobe measurements of the dissolved oxygen profile up to and through a biofilm. Bungay and his coworkers have been the primary utilizers of this technique and Figure 3 from their most recent work (3) clearly demonstrates that the oxygen concentration at the biofilm:liquid interface can be appreciably less than that

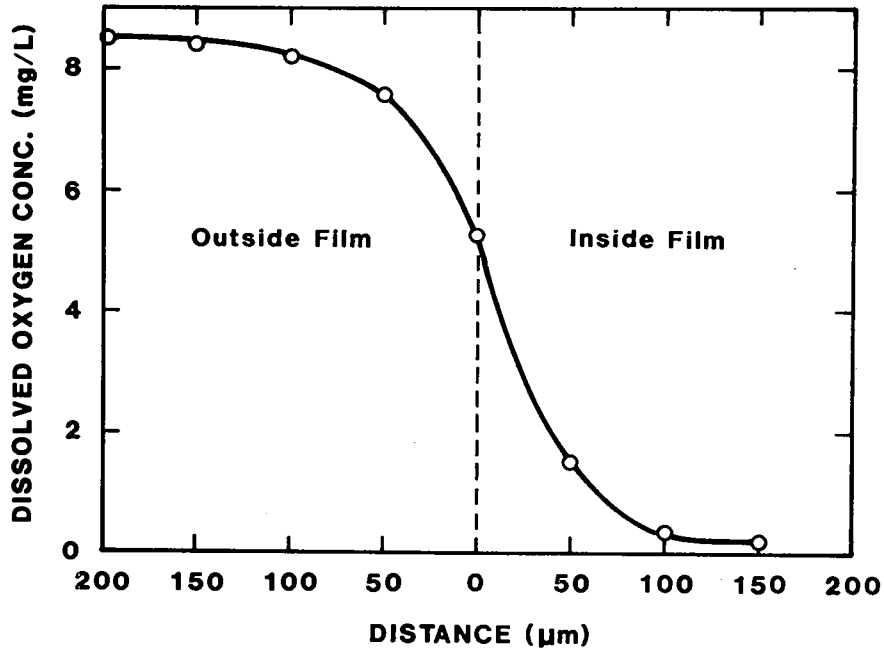


Figure 3. Oxygen concentration profile up to and through the biofilm on a rock from a trickling filter. Data from Chen and Bungay (3).

in the bulk liquid. This particular study was conducted with slime-covered rock media removed from a trickling filter and placed into an experimental apparatus which allowed the maintenance of liquid velocities similar to those found in field installations. The key point to note is that reaction rates calculated by using the bulk liquid concentration in the intrinsic rate equation would be in error because the concentration at the biofilm:liquid interface was lower than the bulk concentration. Similar findings have been reported by others (4).

Indirect evidence for the importance of external mass transport has been obtained by observing changes in the reaction rate when the fluid velocity past a biofilm is changed. The effects of only external mass transport may be isolated by using an extremely thin biofilm so that mass transfer effects in its interior are minimized. This was done by LaMotta (5) using a rotating annular reactor. He found that the reaction rate increased until the velocity past the biofilm was around 0.8 $m.s^{-1}$ but that thereafter it was constant. This suggests that at higher velocities the possible transport rate of reactants to the biofilm exceeded the maximum reaction rate whereas at lower velocities transport limited the reaction rate. With thicker films the transport of reactants into the biofilm makes calculation of the rate constant for external transport more difficult, but from a qualitative point of view it is still possible to show that the rate will increase with increasing velocity until some limiting point is reached. This has been done by Trulear and Characklis (6) in a reactor similar to that employed by LaMotta (5) and by Castaldi and Malina (7) in a rotating tube. Although Trulear and Characklis (6) found that the reaction rate became independent of velocity at a value of 0.93 $m.s^{-1}$, a value remarkably close to that of LaMotta (5), it should not be concluded that velocities in this range will always make external mass transfer effects unimportant. Rather, it will depend upon both the concentrations of reactants in the bulk liquid and the potential reaction rates in the biofilm. Consequently, values well above or below that value may be required.

Even though external mass transfer effects can be important, a number of workers have concluded that they were not significant in their systems and therefore have excluded them from their biofilm models (8-13). Howell and Atkinson (8) modeled sloughing in a trickling filter and stated "It is reasonable to assume that the liquid phase diffusional resistances in the packing units are negligible..." and reference

El Amin (14) for evidence regarding that assertion. Shieh and coworkers (9,10) based their decision to ignore external mass transfer effects upon calculations of expected reaction rates in the presence and absence of them. Since the maximum difference was no more than 7 percent for the expected reaction conditions they concluded that the error was not large enough to justify the additional mathematical complexity which the inclusion of external mass transfer would introduce. Jansen and Kristensen (11) used a rotating annular reactor similar to that employed by LaMotta (5) and thus were able to adjust the rotational speed to make external mass transport limitations insignificant. Andrews and Tien (12) and Wang (13) simply assumed that external mass transport limitations were insignificant without giving their reasoning. It should be noted, however, that the physical system they were using was similar to that of Shieh and coworkers (9,10) and thus the reasoning of the latter workers may be valid in this case as well.

Consideration of all available evidence suggests that external mass transport limitations should be considered when developing biofilm models unless it can be specifically shown that the effects are negligible for the entire range of conditions under considerations. This will generally require some way of predicting these effects and thus the ability to model external mass transport effects is important regardless of whether those effects are ultimately incorporated into the biofilm model.

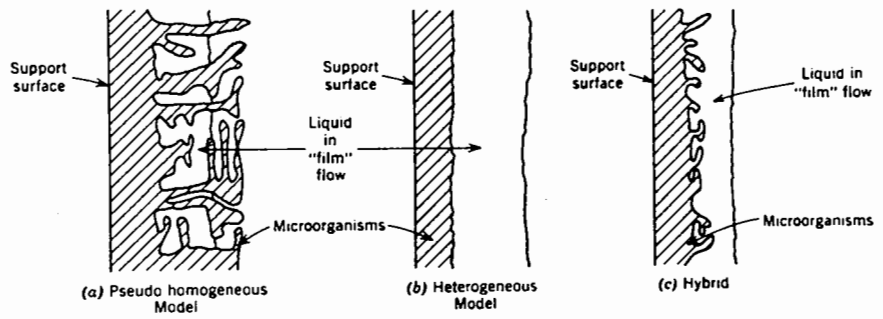


Figure 4. Characterization of the biofilm:liquid interface. (From Atkinson and Howell (17)).

Before external mass transport can be modeled it is necessary to consider the nature of the biofilm:liquid interface. Figure 4, taken from the work of Atkinson and Howell (17) shows three ways in which that interface might be viewed. In the pseudohomogeneous view (4a) the liquid in film flow is considered to move through the biofilm so that no clear liquid film exists. While this concept might characterize trickling filters it would not accurately depict other fixed-film processes. The heterogeneous view, on the other hand, depicts a clear interface between the liquid and the biofilm, so external mass transport occurs in a totally liquid layer. With regard to the true conformation, Atkinson and Howell (17) state: "While the true description of the

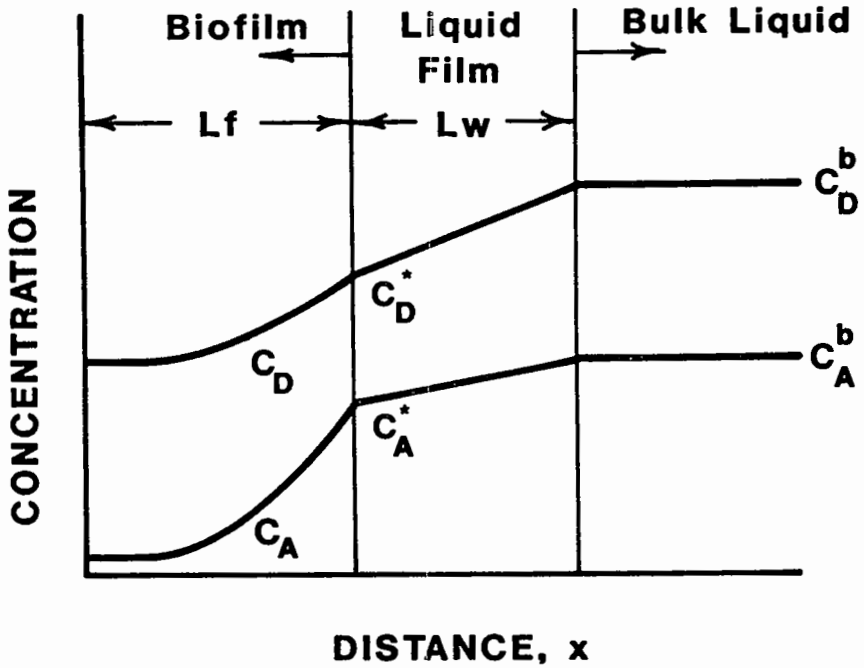


Figure 5. Idealized biofilm and stagnent liquid layer illustrating concentration profiles of the electron donor (D) and the electron acceptor (A).

geometry of the slime surface is probably a hybrid between the pseudohomogeneous and heterogeneous systems, visual observations of slime layers and the experiments of Atkinson et al. (18), suggest that the heterogeneous system is to be preferred." Consequently, most researchers have adopted that view and their models treat the biofilm:liquid interface as if it were analogous to the interface between a flowing fluid and a solid support. In contrast, Williamson and McCarty (19,20) have depicted the liquid side of the interface as containing two components: one adjacent to the biofilm which cannot be removed by mixing and one whose thickness depends upon the turbulence in the liquid phase and approaches zero at very high velocity. The former, which they considered to be approximately 60μm thick, would always present a resistance to external mass transport. They do state, however (19): "Whether such a layer exists in all biofilms is currently unknown." The implications of the existance of such a layer are quite important to modeling, however, and will be discussed more below.

The usual way to depict the necessity for external mass transport is to imagine a hypothetical stagnant liquid film or boundary layer of thickness L_w between the biofilm liquid interface and the bulk liquid phase as shown in Figure 5. All resistance to the transport of materials from the bulk liquid to the biofilm is then assumed to occur in that layer. Two methods of modeling that transport are commonly used.

One method assumes that transport across the liquid layer is by molecular diffusion, with diffusivity D_w. Consequently, the flux, or mass of substrate transported per unit area per unit time, is given by

$$N = \frac{D_w}{L_w} (C^b - C^*) \qquad (1)$$

Because the diffusivity is an intrinsic characteristic of the mateial being transported (the fluid is assumed to be water) the thickness of the liquid layer, L_w, becomes the parameter which must be evaluated before Eq. 1 can be used to depict the rate of transport of the reactants up to the biofilm. On the other hand, the value of L_w will depend upon the physical configuration of the particular reactor being employed and the fluid velocity within it. Consequently, equations relating L_w to those features represent one place in which the biofilm model interfaces with the process model.

As discussed above, Williamson and McCarty (19,20) found that the stagnant liquid layer consisted of two layers which they termed L_1 and L_2:

$$L_w = L_1 + L_2 \qquad (2)$$

The thickness of the outer layer, L_1, is dependent upon the level of turbulence and can be reduced to zero with adequate mixing. The thickness of the inner layer, L_2, is dependent upon the physical characteristics of the biofilm and was considered to be constant. They used the correlations of Welty et al. (21) to relate L_1 to the fluid and physical characteristics of flow inside pipes, over flat plates, and through packed solids. A similar approach was used by Famularo et al. (22) and subsequently by Mueller et al. (23) to model rotating disk reactors. In this case L_2 was estimated from consideration of the depth of surface irregularities in the biofilm and L_1 was calculated from the relationship of Levich (24) for the thickness of liquid entrained on a flat plate.

Other investigators have also used Eq. 1 to estimate external mass transport but they did not incorporate a fixed layer, L_2. Even though Williamson (19,20) originally proposed the incorporation of L_2 into L_w, in a later investigation he and Meunier (25) stated "From the review of various formulas for this parameter (L_w), it appears that the expression proposed by Snowdon and Turner (26) for flow past particles most closely models conditions in packed and expanded bed biofilm reactors." The nature of this expression is such tht L_w will appraoch zero as the velocity of flow gets large. Since no mention was made of L_2 one is uncertain as to why it wasn't considered and whether the authors have concludedd from further work that it is unimportant. Mulcahy et al. (11) also used the equation of Snowdon and Turner (26) when they determined that external mass transport resistances were unimportant for dentrification in fluidized beds. Since that determination was based on computations of utilization rates with and without a stagnant layer one must wonder if the same conclusion would have been reached if a permanent layer had been included. To add further uncertainty to the importance of the layer L_2, McCarty, who advised Williamson's original work (9,20), has since coauthored papers with Rittmann (27,28) which did not include such a layer. In one of these papers (27) they state "Many empirical formulas for evaluating L in porous media were reviewed, and the one presented by Jennings (29) was felt most appropriate...".

However, in the other paper they appear to have used the relationship of Welty et al. (21) as used originally by Williamson (19). Whether L_2 was included was not stated.

The other approach to modeling external mass transport is to use a mass transfer coefficient, k, such that

$$N = k(C^b - C^*) \qquad (3)$$

Comparison of this equation to Eq. 1 reveals that k is equivalent to D_w/L_w and consequently the comments made in the preceding paragraphs are equally applicable here. Investigators using this approach have tended to use empirical equations for the mass transfer coefficient taken from the chemical engineering literature. For example, Grady and Lim (30,31) used the correlations of Mixon and Carberry (32) (for flow over flat plates), Wilson and Geankoplis (33) (packed beds) and von Karman as given by Levich (24) (rotating disk) in their work. Dahodwala et al. (34) used the relationship of Brian and Hales (35) to estimate k for gently stirred suspended particles. Finally, Mueller et al. (36) used the relationship of Charpentier (37) for mass transfer for clean packed beds for their trickling filter model. The fact that different models must be used for different types of reactors demonstrates again how the biofilm model may be interfaced with the process model. The significant fact about all of these relationships is that they were determined for clean, nonporous material. Given the nature of the biofilm, however, one must wonder how applicable they really are for situations with high turbulence. While the mass transfer coefficient for transfer to a nonporous material might become quite large in turbulent flow, is it possible that the coefficient for biofilms will approach some maximum value due to a permanent stagnant layer like L_2? This is an area that needs further investigation because the available evidence is not clear, yet the implications of decisions made from the models are quite important.

CHARACTERISTICS OF BIOFILM

Substrate removal in any heterogeneous environment is the result of interaction between the rates of transport and the intrinsic rates of reaction. Intrinsic rates of biological reactions are generally expressed on the basis of a unit of

biomass, e.g., the intrinsic rate of substrate removal is expressed as the mass of substrate removed per unit time per unit of biomass and the intrinsic rate of biomass growth is expressed as the amount of biomass formed per unit time per unit of biomass present. Before considering the form of the intrinsic rate expressions, consideration should be given to the mass of organisms likely to be found in the biofilm. Since the surface area available for biofilm colonization and development is generally a physical characteristic of the type of reactor being modeled, the mass of biofilm in the reactor becomes a function of the thickness and density of the biofilm. Furthermore, it is possible that not all of the organisms present in a biofilm wil be capable of utilizing all substrates entering it; e.g., if organic matter and ammonia nitrogen were both present only the heterotrophic organisms would be capable of oxidizing the organic matter whereas only the autotrophs could oxidize the ammonia nitrogen. Consequently, the composition of the biofilm may also be an important determinant of the potential reaction rates.

Biofilm Thickness

When considering biofilm thickness it is important that a distinction be made between the total film thickness and the active film thickness. In a review of 10 papers in which biofilm thicknesses were measured, Atkinson and Fowler (2) found that the total film thickness was between 0.07 and 4.0 mm. They divided the biofilms into two groups, however, to better reflect the growth conditions. When the films were subjected to mechanical or hydrodynamic control, the thickness was generally less than 0.2 mm. When the films were uncontrolled, however, they were as thick as 4.0 mm, though it has been asserted that in turbulent flow systems, biofilm thickness seldom exceeds 1 mm (38). Thicker, uncontrolled films are not likely to have greater substrate removal rates than thin films because diflusional resistances within the film limit the amount that is actually contributing to substrate removal. This amount is termed the active layer and two types of evidence for its existence have been gathered.

The earliest evidence was based upon observations of changes in the rate of substrate removal as the depth of biofilm increased in a reactor with a fixed surface area. Those observations indicated that the rate of substrate consumption

increased as the biofilm depth increased up to a limiting depth of 70-100 μm; after that the removal rate was independent of depth (39, 40). The depth at which substrate consumption reached a maximum value was defined as the active depth. Trulear and Characklis (6) observed a similar phenomenon, although they also found that the active depth increased as the substrate concentration in the liquid phase increased.

Shortly after the observation that the substrate removal rate increased with depth up to a maximum, Bungay et al. (41) used a microprobe technique to determine oxygen profiles within a film. Their results indicated that respiration ceased at depths of 50-150 μm, depending upon the substrate concentration is the medium. This is consistent with the interpretation that only the organisms in the active layer are contributing to substrate removal. Similar observations have been made by Hoehn and Ray (4) and by Chen and Bungay (3) as shown in Figure 3. The latter workers also found, however, that at low bulk substrate concentration, the oxygen concentration in the biofilm reached a constant value at some depth, thereby demonstrating that the active layer may be defined by depletion of either the electron donor or the electron acceptor.

It is now generally accepted that the active thickness is a result of transport limitations within the biofilm. Only when the film is very thin, when the electron donor and acceptor concentrations are very high, or when the rates of transport are large in relation to the reaction rates will the active film thickness approach the total film thickness. For many biofilm reactors these circumstances will not exist, with the result that the total film thickness (and by extension the total amount of biomass) has no impact upon reactor performance. If one knew in advance that the total film thickness was in excess of the active film thickness then the system could be accurately modeled with any arbitrarily assumed thickness because the differential equations depicting transport and reaction within the biofilm would automatically show a cessation of substrate removal when either the electron donor or acceptor was exhausted. The need to know the film thickness arises, however, when the potential active film thickness exceeds the thickness that could actually exist under the given physical circumstances because then the extent of reaction will be limited by the actual film thickness.

Prediction of the biofilm thickness within a fixed-film reactor is the least developed of all of techniques needed for adequate modeling, primarily because relatively little funda-

mental study has been devoted to the factors governing biofilm development. Some of the best experimental work on biofilm development has been done by researchers interested in biofouling of heat exchangers, pipes, etc., and the reader is referred to the review by Characklis (38) in this regard. Basically a biofilm will continue to increase in thickness as long as the rate at which the microorganisms are growing exceeds their rate of loss by decay and by attrition. In a highly turbulent regime, attrition will be relatively constant and appreciable so that, as mentioned earlier, biofilm thicknesses seldom exceed 1000 μm. Even in less turbulent regimes, however, steady state biofilms can develop when the available substrate concentration is low because then cell decay will balance the growth. Unfortunately, the general

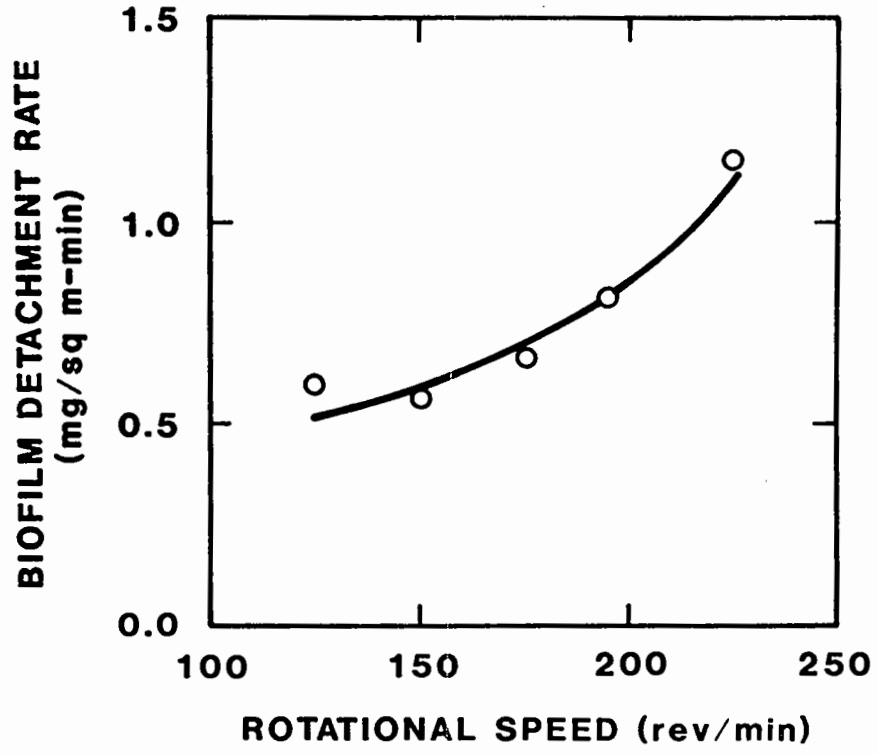


Figure 6. Effect of rotational speed in an annular reactor on the detachment rate of a biofilm with a mass of 150-160 mg. (From Trulear and Characklis (6)).

situation for most biofilm reactors in wastewater treatment is one in which continual attrition is not sufficient to balance the net growth, with the result that the film grows until conditions develop near the support:biofilm interface which cause adhesion to be lost and the film to slough away. This results in a continually dynamic state for the reactor, which makes analysis particularly difficult. Consequently, Atkinson and Fowler (2) have suggested the application of positive control over biofilm thickness and fluidized bed biofilm reactors represent one reactor configuration within which such control can be practiced. In that type of reactor the height of the bed is functionally related to the thickness of the biofilm on the particles so that maintenance of a constant bed height by the removal and cleaning of particles results in a maximum known film thickness (11,42,43,44).

The majority of the biofilm reactors used in wastewater treatment are of a configurtion which prevents positive control of biofilm thickness. This means that the film will either reach a natural steady state in which growth is just balanced by decay and attrition losses or it will increase

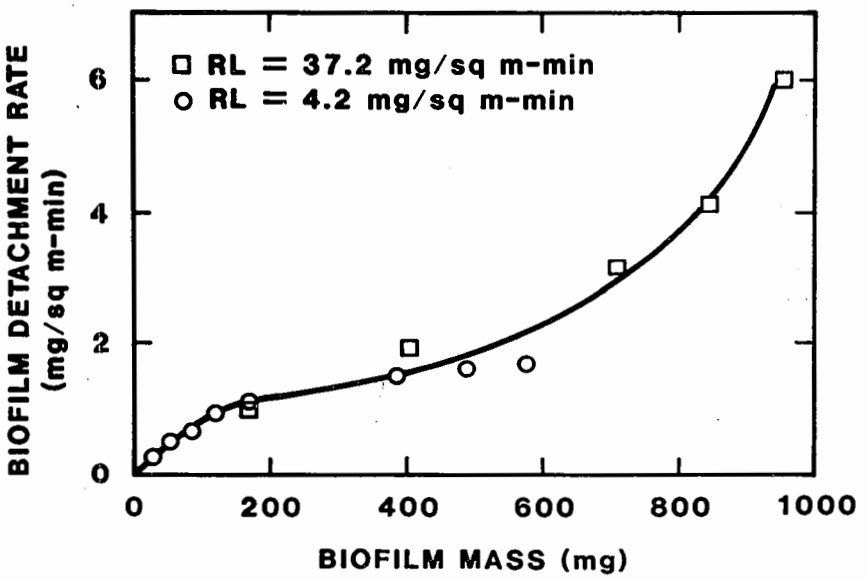


Figure 7. Effect of biofilm mass (proportional to thickness) in an annular reactor rotating at constant speed on the detachment rate of biofilm. R_L refers to the organic loading applied. (From Trulear and Characklis (6)).

continually until sloughing occurs. Knowledge of the conditions controlling which of those conditions exist is required for accurate modeling. Unfortunately, relatively few studies have been done on factors affecting attrition and thus the data are limited. The most complete study is that reported by Trulear and Characklis (6) who grew fixed-films in an annular reactor consisting of two concentric cylinders, one stationary and the other rotating. The rotational speed determined the shear stress developing at the biofilm:liquid interface and the biofilm detachment rate increased as the rotational speed increased, as shown in Figure 6 (6). This suggests that the rate of biofilm detachment increases as the shear stress at the interface increases. Furthermore, as shown in Figure 7, the detachment rate also increases as the biofilm mass increases (6). This suggests a mechanism whereby films of different thickness can be attained in a reactor with a fixed shear stress as substrate is applied at various rates.

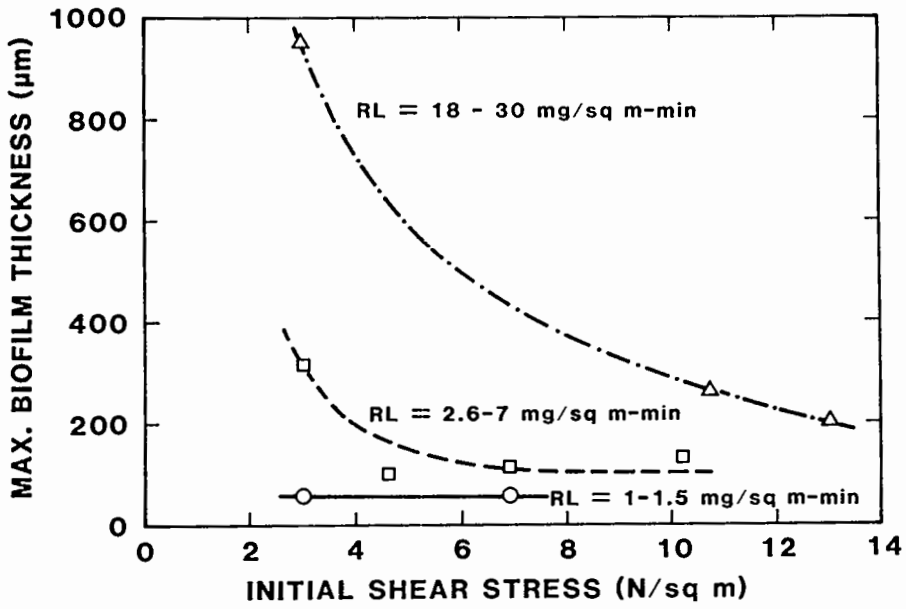


Figure 8. Effect of fluid shear stress and organic loading (R_L) on the maximum biofilm thickness attained in an annular reactor. (Data from Zelvar (45) as presented by Characklis (38)).

At low substrate application rates (K_L) the rate of new biofilm growth would be low and would be balanced by the attrition rate when a relatively low mass (or thickness) was present. At higher substrate application rates, however, the film would be growing faster and thus a greater mass (thickness) would develop before the detachment rate balanced the accumulation rate. Evidence for this may be seen in Figure 8 (38) which is from the work of Zelvar (45). The curve in Figure 7 also indicates that the detachment rate approaches a very large value at a high biofilm mass, thereby suggesting that the fluid shear stress can limit the maximum quantity of attached biofilm in a turbulent flow regime (45).

Since the biofilm mass is a function of the thickness, these results suggest that the detachment rate associated with a given shear stress will increase as the thickness increases, thereby allowing the mechanism discussed above to be modeled. Unfortunately, things are not that simple because Trulear and Characklis (6) have also shown that biofilm density is a function of the applied substrate loading rate. This means that the thickness associated with a given mass was greater at a lower loading, thereby suggesting that the detachment rate cannot be expressed as a simple function of thickness but must be expressed in terms of both thickness and density. What functional forms should such models take? Why does this relationship exist? Were the results influenced by the type of reactor employed? These and many other questions must be answered before truly mechanistic models can be written depicting biofilm thickness. This has been recognized by Characklis who listed such questions in a recent review (38). One can only hope that work is continuing on such matters.

Because until recently relatively little was known about biofilm detachment, the vast majority of the models for biofilms have assumed a constant film thickness consistent with the type of process under consideration. Such an assumption will probably work well for a fluidized bed biofilm reactor because the biofilm thickness can be controlled and because the fraction of particles removed to control thickness is small. Furthermore, as long as the recycle ratio is kept above 2.0, the bed may be considered to be completely mixed with respect to the soluble compounds (46), and thus all

particles experience the same reactant concentrations. It may not be a good assumption for other fixed-film processes, however, because the concentration gradients within them could cause wide variations in film thickness. This could prevent full films from developing near the outlet. Thus, in general, it would be better to have some way of estimating the biofilm thickness.

The major attempt to model biofilm thickness is that of Rittmann and McCarty (27,47) who have done so by assuming that a steady state biofilm is one in which growth would just be balanced by cellular decay so the observed yield would be zero. Thus there is no explicit term for attrition or detachment in their model. Rather, "it is assumed that the total amount of biofilm mass is just equal to that which can be supported by the substrate flux. The steady-state-biofilm thickness can then be computed by equating the available and maintenance energy rate..." (47). This assumption is probably a reasonable one for the situation for which they developed their model, i.e., for very low substrate concentrations such as in ground water recharge. The model did a reasonable job of tracking the substrate concentration profile through a small tower even when intrinsic parameters were utilized, although it did not do as well tracking the biofilm thickness (27). This may have been due to their assumed constant biofilm density, however. The value of this model comes from its ability to predict the minimum substrate concentrtion attainable in a fixed-film system. It has been applied, however to a broad range of reactor configurations which would operate with electron donor and acceptor concentrations much different from the ones for which it was developed (48). Arcuri and Donaldson (49) criticized the basic assumption of the model, stating that other mechanisms of cell loss would be important in most biofilm reactors. This criticism certainly appears to be valid.

Recognizing that the steady-state biofilm concept is limited to a particular situation, Rittmann (50) extended it to incorporate detachment by shear stress. His analysis was based upon the data of Trulear and Characklis (6) and incorported the concept that the detachment rate depended upon the film thickness and mass as well as upon the shear stress. He asserted that the basic steady-state biofilm model could be employed for a broad range of cases by recognizing that the biomass decay rate, b, could be replaced by a combined factor, b', which includes both decay and attrition by fluid shear.

From analysis of the data of Trulear and Characklis (6) he concluded that the attrition portion of b' would be a function of the shear stress alone for biofilms less than 30 μm thick but would be a function of both the shear stress and the thickness for thicker films. This approach appears to be quite reasonable, given the limited data available. As pointed out previously, however, there is a need for more work on the subject since Characklis himself raises questions about how the detachment rate changes with fluid shear stress and biofilm thickness (38). When the researcher who develops data indicates that more needs to be known about the relationships involved, it could be argued that the development of mathematical functions depicting those relationships is premature. Furthermore, it should be recalled from the previous discussion that the detachment rate is an apparent function of the biofilm density as well. Rittmann (50) did not incorporate this, thereby giving another reason for viewing his relationships with caution.

Andrews and Tien (14) have developed a model for biofilm growth on activated carbon particles that is similar in concept to the steady-state biofilm model of Rittmann and McCarty. Although they state that their decay term "accounts for both the basal metabolism (cell maintenance energy) of the bacteria and for wash-off of cells from the film" it is assumed to be a constant and is not a function of film thickness, biomass density, turbulence, etc. Thus the biofilm thickness aspect of their model is essentially the same as that of Rittmann and McCarty's model and the comments made about it are equally applicable.

In contrast to the steady-state approach taken by Rittmann and McCarty, Howell and Atkinson (8) modeled biofilm thickness from the dynamic point of view, i.e., they modeled sloughing. In their model they allowed the film thickness to increase over time by assuming that no continual detachment occurred so substrate removal would result in accumulated cell mass. As the thickness increased the substrate concentration profile changed until eventually the concentration in the interior of the film was too low to sustain the cells, thereby allowing lysis to occur, leading to sloughing. Recognizing that there is a certain amount of randomness associated with sloughing, they arranged the model to take that randomness into account. They then applied their model to a trickling filter and investigated the time-dependent performance. Because of the sloughing the effluent substrate concentration always varied in a dynamic manner showing that the nature of the bio-

films in such reactors is in part responsible for their dynamic behavior.

Because of the importance of film thickness to the proper modeling of fixed-film reactors it is important that accurate (both conceptually and mathematically) models be developed. As seen, however, there are still many questions to be answered. A reasonable start has been made but in the opinion of this author, a much greater effort is still required. As will be seen in later sections, many sophisticated solution techniques have been applied to biofilm models, but almost all have been applied to films of constant, arbitrary thickness. Thus it would appear that the questions regarding biofilm thickness should be resolved before more effort is expended on new, general, biofilm models.

Biofilm Density

Because the rates of reaction are a function of the mass of microorganisms present, the density must be coupled with the thickness and area to allow computation of the reaction

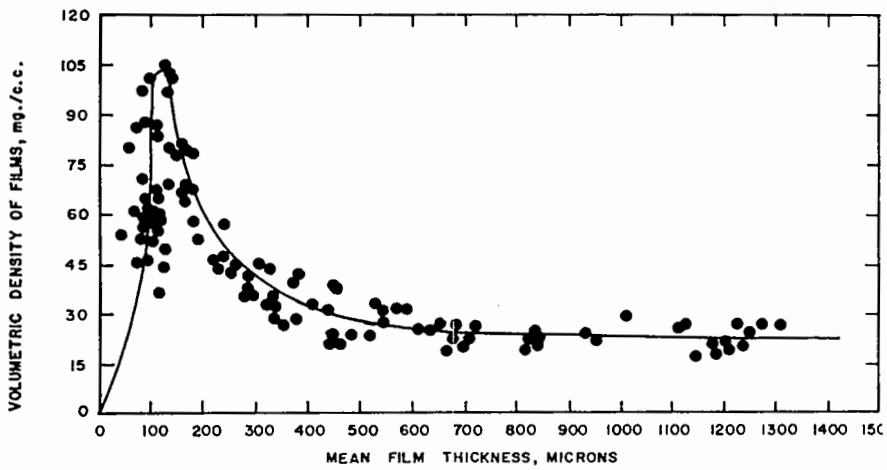


Figure 9. Effect of biofilm thickness on the density of the biofilm growing on a rotating drum. (From Hoehn and Ray (4)).

rate. Although it has generally been assumed that the density is constant and independent of film thickness, there is evidence that this is not the case. The first to discover that the density of a biofilm depends upon its thickness were Hoehn and Ray (4) who obtained the results shown in Figure 9. There it can be seen that the density reached a maximum value at a thickness consistent with the active film thickness. They postulated that the changes in density were due to variations in the microbial populations within the film. The maximum density was thought to represent the tight packing which would exist in the aerobic layer whereas the lower densities were thought to be due to the lysis of cells in the anaerobic region. Shieh et al. (44) and Mulcahy and LaMotta (51) have observed similar reductions in density with increasing thickness, although they observed no region of increasing density. The latter authors developed empirical equations which were subsequently used to calculate biomass concentrations in their model for the fluidized bed biofilm reactor (10,51). Trulear and Characklis (6) also observed changes in biofilm density, but their observations cause one to ask whether thickness is the correct parameter to which to relate density. This is because the density in their films approached a maximum value as the substrate loading rate was increased at a constant shear stress. It was seen previously that the film thickness associated with a given shear stress was also a function of the substrate loading, thereby raising the question of whether it is the thickness or the loading (i.e. the net growth rate) that influences the density. Furthermore, within the limits of film thickness in their studies, no decrease in biofilm density was observed.

There are many factors which could be responsible for changes in density, from the lysis postulated by Hoehn and Ray (4) to the changes in culture morphology observed by Trulear and Characklis (6), and more fundamental research will be required to delineate the mechanism and its importance. In the mean time it is unclear whether changes in density should be incorporated into models and if so, how they should be formulated. If the observed decreases in density are due to lysis or culture changes in the interior (i.e., nonactive) zones, should they be included in models in which substrate removal only occurs in the active layer? In other words, if the sole purpose of the biofilm density in the model is to obtain the reaction rate, should a constant value be used? If, on the other hand, the density is required to predict other factors, such as the bed height in a fluidized bed

model, then should a variable density be used? These and other questions must be answered before more exact models can be developed.

Biofilm Composition

Although there have been several observations of changes in microbial composition with a biofilm (4,6,38,52,53) most models treat the biofilm as if it were homogeneous throughout. This is because most models have sought to predict the fate of one constituent such as organic matter or ammonia nitrogen. Nevertheless, as Alleman and Veil (52) have noted "...fixed-film communities likely include discrete microbial strata, with divergent metabolic and diffusional characteristics. Appropriate refinements to these models may therefore be necessary to insure their validity and utility". While it may be some time before this is necessary for models simply depicting the removal of soluble organic matter, it is already necessary for models which depict the fate of both carbon and nitrogen in biofilms. One such model is that of Mueller et al. (23) which predicts the amount of carbon removal, nitrification, and denitrification in an RBC. Carbon removal is assumed to occur by aerobic metabolism as long as sufficient oxygen is present in the film, but will occur by denitrification when the oxygen concentration gets sufficiently low in the presence of a carbon source and nitrate nitrogen. Nitrification occurs as long as sufficient ammonia nitrogen and oxygen are present and the ratio of heterotrophs to autotrophs at any depth within the film is set equal to the ratio of their growth rates.

The work of Bryers (54) represents the most ambitious attempt at modeling spatial profiles within biofilms. His model can predict the profiles of heterotropic, Nitrosomonas spp. and Nitrobacter spp. within biofilms housed in a CSTR receiving a constant input. It also considers substrate profiles for NH_4^+, NO_3^-, NO_2^-, O_2 and acetate. A finite element technique is used to integrate dual substrate limiting rate expressions over both time and distance, thereby showing changes within a biofilm as it builds up.

In general, however, relatively little work has been performed to assess the composition of biofilms in general, much less to look at how they might change with depth within a given film. One might imagine, however, that such information

would be very useful to the development of a better understanding of how individual substrate components might behave in a fixed-film reactor.

TRANSPORT AND REACTION WITHIN BIOFILM

Once the electron donor and acceptor have been transported up to the biofilm:liquid interface they must be carried into the biofilm where the reactions will occur. These events occur simultaneously and thus the concentration profiles of the two constituents in the film will reflect their relative rates. Since the reactive capability of the biofilm (i.e., its overall average reaction rate) depends upon the nature of those concentration profiles, the heart of any biofilm model is the conceptualization and mathematical expression of the simultaneous transport and reaction events. Consider for the moment the biofilm depicted in Figure 2a. Even though the cells are held together in a complex geometric arrangement and have some sort of spatial distribution, the majority of the models assume that they are uniformly distributed throughout the film. Because of the gelatinous character of the biofilm matrix it is thought that convective transport contributes little to the movement of reactive constituents within the film and that the electron donor and acceptor reach the organisms by diffusion, which is characterized by Fick's law:

$$N = D_e dC/dx \tag{4}$$

Unlike Eq. 1, in which the diffusivity was given as D_w, the free diffusion coefficient in water, the coefficient is given as an effective diffusivity D_e, which reflects the fact that the diffusion in the biofilm will generally be retarded because it must occur through the gelatinous matrix. If a mass balance on a reactive constituent is then performed around a differential element of steady-state film of constant microbial composition, the result is an equation which is almost universally used to model reactions within biofilms:

$$-D_e A \left.\frac{dC}{dx}\right|_x + D_e A \left.\frac{dC}{dx}\right|_{x+\Delta x} + rA\Delta x = 0 \tag{5}$$

in which A is the total surface area normal to the direction of diffusion, x. The key elements which must be inserted into Eq. 5 are the effective diffusivity, D_e, and the reaction

rate expression, r, because these determine the nature of the resulting concentration gradient and overall reactivity.

Diffusion

Given the importance of diffusion to the modeling of fixed-film reactors there is surprisingly little agreement about how the presence of the biofilm influences the diffusivity. This is due in part to the fact that the character of a particular biofilm depends upon the type of organisms growing in it (55) but it is probably also due to the many different techniques that have been used to estimate the coefficient.

The most direct method of estimating D_e is to measure a concentration gradient through a biofilm and to couple it with the flux of material into the film to allow direct computation of D_e. This has been done by Bungay and associates for oxygen diffusion into laboratory-grown (56) and actual trickling filter films (3). The coefficient for laboratory-grown films was approximately 80 percent of the value in water whereas the coefficient for field-grown films was about 35 percent. It is likely that the differences in the values reflect differences in the cultures residing in the two films.

Another direct technique is to place a film in a special chamber which allows a component to diffuse through it, thereby allowing measurement of the flux. Then from knowledge of the film thickness the diffusivity may be estimated. Three investigators have used this technique (20,55,57). Williamson and McCarty (20) measured the diffusivities of ammonia, nitrite, nitrate, and oxygen through films which had been formed by filtration of dispersed nitrifying bacteria onto supporting membrane filters. The values were all in excess of 80% of the values in water. Matson (55) and Pipes (57) grew mixed cultures of bacteria on glucose in completely mixed reactors, concentrated then by centrifugation, and formed them into films by spreading them onto a template with a spatula. The biofilm was then sandwiched between two membrane filters prior to placement in the diffusion apparatus. Pipes (57) grew his organisms at different carbon-to-nitrogen ratios and found that the diffusivity of glucose ranged from 6 to 60 percent of the value in water, depending upon the growth conditions. Matson (55) not only varied the carbon-to-nitrogen ratio but also varied the specific growth rate. He found that both parameters influenced the diffusivity and that the value for glucose ranged from 10 to 30 percent of the value in water

whereas the value for oxygen ranged from 20 to 100 percent. Since the experimental reactors displayed different macroscopic characteristics it was speculated that the most important factor determining the diffusional characteristics could have been the particular microbial species in residence.

A third technique that has been employed for estimating D_e is to grow biofilms in a fixed film reactor, measure its performance under a variety of conditions, and then evaluate D_e by curve fitting the model under consideration to the experimental data. Using a fluidized bed Andrews and Tien (14) found the effective diffusivity of valeric acid to be 34 percent of the value in water whereas Wang (15) (in the same lab) found it to be 67 percent. Wang (15) also found the effective diffusivity of oxygen to be approximately 10 percent of the free diffusion value, although he stated that the uncertainty associated with the number was expected to be high. Mulcahy et al (10,58) calculated D_e for nitrate for cells growing on a rotating disk reactor and found it to be approximately 50 percent of the value in water whereas Jansen and Kristensen (13) found that it varied from 30 to more than 100 percent for films grown in a rotating annular reactor, depending upon film thickness. Although it was apparent that the value of D_e increased as the film thickness increased, the finding of values in excess of the free diffusivity suggested errors in the estimation of the reaction rate constants which were ultimately used to calculate the diffusivity. Furthermore even though the variation in D_e with film thickness is consistent with the variations in biofilm density discussed earlier, these results illustrate the dangers in computing coefficients from assumed models.

Because of the difficulties associated with direct measurement of the diffusivity, most modeling studies have used assumed values. Because of their previous work (20), subsequent studies by Williamson and his students (19,59,60) and by McCarty and his students (27,28,47,48,61) have assumed effective diffusivities of 80 percent of the free values for a large number of substances. The modeling work done at Manhatten College (22,36) assumed a similar value based upon that same work as well as upon an analysis of the oxygen profiles developed by Whalen et al (62). Harris and Hansford (63) assumed that the effective diffusivity of glucose was equal to the value in water because of the results of Atkinson and Daoud (64) and Atkinson and How (65). They also used a value equal to that in water for oxygen, but this time their justification was that the spread in the reported values made

it impossible to obtain a reasonable estimate.

From the preceding discussion it is apparent that there is no consensus concerning the effects of biofilms upon the diffusive transport of reactive species. Furthermore, it appears that this lack of consensus is due to variations within the biofilms caused by growth conditions, predominant microbial populations, thickness, etc. Thus, while there can be no doubt that additional well defined and controlled studies are needed, perhaps the most logical approach to modeling at the present time is to just include the diffusivity in one of the dimensionless groups that must be evaluated during experimentation (66).

Reaction Rate Expressions

Fixed-film processes are generally used for one of three purposes: to remove soluble organic matter, to convert NH_4^+-N to NO_3^--N (nitrification) and to convert NO_3^--N to N_2 (denitrification). In some cases more than one of these may be accomplished in a single reactor, but in all cases two soluble, transporting components are necessary for the reactions to occur. These are an electron donor and an electron acceptor. In processes focusing on the removal of soluble organic matter, that organic matter serves as the electron donor and oxygen serves as the electron acceptor. (In anaerobic fixed-film reactors some other constituent will serve as the electron acceptor. The situation is complicated, however, by the nature of the microbial interactions involved and thus it will not be considered herein). When nitrification is the objective, NH_4^+-N serves as the electron donor and oxygen again serves as the acceptor. Some nitrification models seek to also account for the production and subsequent oxidation of NO_2^--N to NO_3^--N but the bulk consider only the oxidation of NH_4^+-N. Finally, when denitrification is the objective, NO_3^--N serves as the terminal electron acceptor and some form of organic matter generally serves as the donor; the focus is generally on the fate of the NO_3^--N, however. Consideration of these processes suggests that they can be generalized by writing the reaction rate expressions in terms of the concentrations of the electron donor, C_D, and the electron acceptor, C_A. That approach will be taken herein.

It is now widely recognized that in the most general case the reactions within a biofilm may be controlled by the

concentrations of both the electron donor and the electron acceptor. If the concentration of one is much higher than the other however, then only one constituent controls. This latter situation has been assumed to exist by most modelers, and thus most of the models have been written in terms of only one constituent. Thus let us first examine the basic single-substrate rate equations and then expand them to the two-substrate case, which will serve as a more general model.

Cell growth and substrate oxidation are generally considered to be coupled reactions, i.e., substrate removal occurs because of cell growth. The proportionality constant is the true growth yield, Y_g. Furthermore, the rate of cell growth is proportional to the cell concentration or density within the film, X_f:

$$r_g = \mu X_f \tag{6}$$

where μ is the specific growth rate, T^{-1}. Likewise with the substrate removal rate:

$$-r_s = q\ X_f \tag{7}$$

where q is the specific substrate removal rate, T^{-1}, which is related to the specific growth rate by

$$q = \mu/Y_g \tag{8}$$

These definitions assume that all substrate utilization is channeled into cell synthesis and that cell maintenance needs are met by decay. Another approach would be to assume that a portion of the substrate was channeled directly into cell maintenance. Although there are differences in the fundamental mechanisms employed by the two models both yield the same result and can be considered to be equivalent (30). In the few models where cell maintenance energy needs have been considered, the growth/decay concept has been employed. Consequently, it will be used here as well.

A multitude of models could be (and have been) written to depict the relationship between the specific growth rate of bacteria and the concentration of a single limiting nutrient, since all such models are strictly empirical (30). Consequently, this review will be limited to the two most widely used ones: Monod (68)

$$\mu = \frac{\mu_m C}{K + C} \tag{9}$$

Blackman (69)

$$\mu = \frac{\mu_m C}{2K} \text{ for } C < 2K; \qquad \mu = \mu_m \text{ for } C \geq 2K \tag{10}$$

In these models μ_m is the maximum specific growth rate and K is the saturation constant. A plot of these models is depicted in Figure 10. There they are shown in dimensionless form:
Monod

$$\frac{\mu}{\mu_m} = \frac{C/K}{1+C/K} \tag{11}$$

Blackman

$$\frac{\mu}{\mu_m} = \frac{1}{2}\frac{C}{K} \text{ for } C/K < 2; \qquad \frac{\mu}{\mu_m} = 1 \text{ for } C/K \geq 2 \tag{12}$$

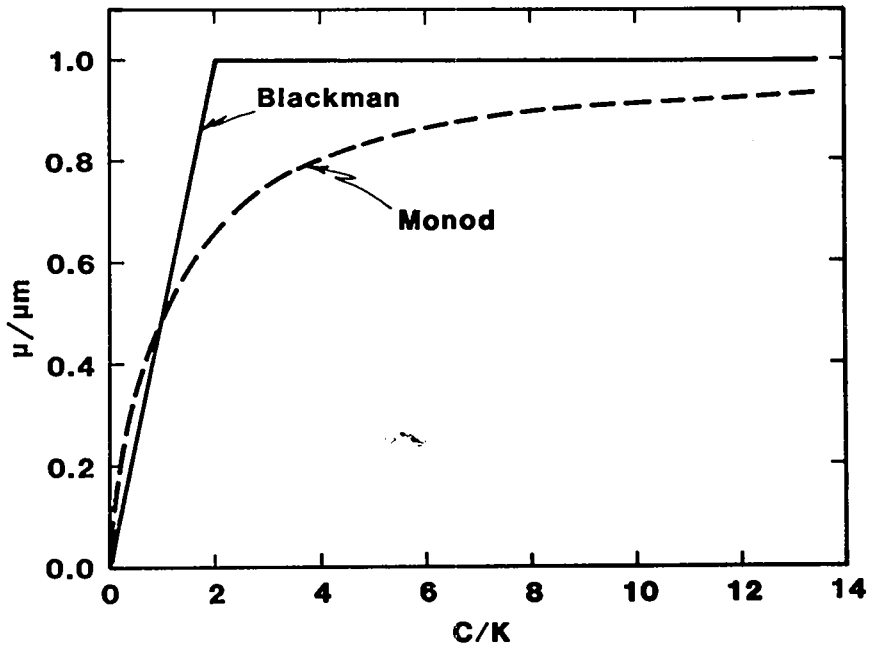


Figure 10. Dimensionless plots of Monod and Blackman kinetic models for substrate limited growth of bacteria. (Adapted from Badar (67)).

The Monod model has been widely used to depict the removal of soluble organic matter and the oxidation of NH_4^+-N in fixed-film reactors under the assumption that the electron acceptor is present in unrestricting concentration. The Blackman model has found extensive use in the modeling of denitrification in fixed-films when the concentration of both the electron donor and acceptor are high.

As discussed previously, because of the concentration gradients within the biofilm, it is likely that the concentrations of both the electron donor and the electron acceptor could limit the rates of reaction. Thus, it would be desirable to have a general model which handles all types of doublesubstrate limitation. As Bader (67) has pointed out, however, "this becomes rather difficult since two separate schools of thought exist about the nature of growth with two limiting substrates, and there is insufficient experimental data to support either school. In fact, it is doubtful that sufficient experimental evidence will be developed in the near future". Thus a review of the current state-of-the-art of fixed film modeling must incorporate the two philosophies, which have been labeled noninteractive and interactive.

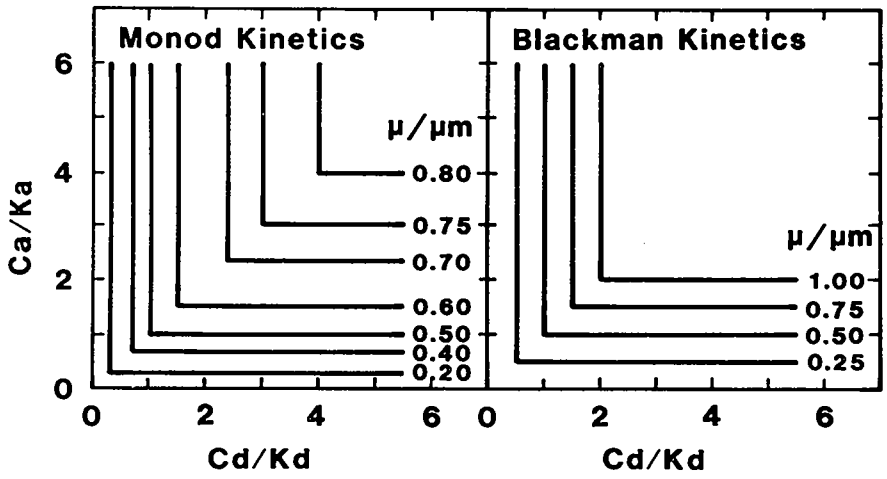


Figure 11. Plots of lines of constant dimensionless specific growth rate (μ/μ_m) as a function of the dimensionless concentrations of electron donor (C_D/K_D) and electron acceptor (C_A/K_A) for noninteractive models using Monod and Blackman kinetics. (Adapted from Bader (67)).

A noninteractive model is based upon the concept that the specific growth rate of an organism can only be limited by one substrate at a time. Therefore, the specific growth rate will be equal to the lowest rate that would be predicted from the separate single-substrate models. For the Monod model, this may be written:

$$\frac{\mu}{\mu_m} = \frac{C_D/K_D}{1+C_D/K_D} \quad \text{for} \quad \frac{C_D}{K_D} < \frac{C_A}{K_A}$$
$$\frac{\mu}{\mu_m} = \frac{C_A/K_A}{1+C_A/K_A} \quad \text{for} \quad \frac{C_A}{K_A} < \frac{C_D}{K_D} \qquad (13)$$

where the subscript D refers to the electron donor and A refers to the acceptor. Similar equations may be written for Blackman kinetics. Graphs of constant dimensionless specific growth rate (μ/μ_m) as a function of dimensionless substrate concentrations (C_D/K_D and C_A/K_A) are shown for the two types of kinetics in Figure 11.

An interactive model is based upon the assumption that if two substrates are present in less than saturating concentrations, then both must affect the overall specific growth rate of the organism. One type of interactive model may be constructed by multiplying two single-substrate limited models

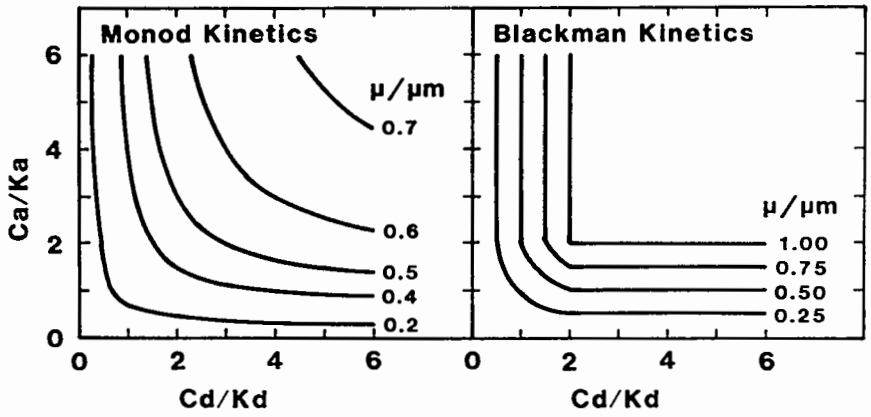


Figure 12. Plots of lines of constant dimensionless specific growth rate (μ/μ_m) as a function of the dimensionless concentrations of electron donor (C_C/K_D) and electron acceptor (C_A/K_A) for interactive models using Monod and Blackman kinetics. (Adapted from Bader (67)).

together, as shown below for the "double Monod model":

$$\frac{\mu}{\mu_m} = \left(\frac{C_D/K_D}{1+C_D/K_D}\right)\left(\frac{C_A/K_A}{1+C_A/K_A}\right) \tag{14}$$

A similar equation could be written for Blackman kinetics. Plots of these models are shown in Figure 12. Howell and Atkinson (70) have utilized a slightly simpler model proposed by Bright and Appleby (71) and have found that for parameter values likely in the processes employed in wastewater treatment there would be little difference in the results obtained with it and with the double Monod model.

To summarize, there are six categories of models which could be used to depict the specific growth rate, μ, of the microorganisms in the biofilm: single-substrate, Monod (SSM); single-substrate, Blackman (SSB); noninteractive double-substrate, Monod (NDSM); noninteractive double-substrate, Blackman (NDSB); interactive double-substrate, Monod (IDSM); and interactive double-substrate, Blackman (IDSB).

The complete reaction rate expression for removal of the electron donor in a biofilm may be obtained by combining Eqs. 7 and 8, yielding

$$-r_s = \mu X_f/Y_g \tag{15}$$

and then substituting the appropriate equation for μ into the result (i.e., Eq. 14). The complete reaction rate expression for net growth of cells must combine two loss terms with Eq. 6 for growth:

$$r_x = \mu X_f - bX_f - r_a \tag{16}$$

where the term bX_f represents the decay of cells for maintenance purposes and the term r_a is the loss by attrition. The rate constant b has been taken to have a fixed value or has been considered to be a function of the concentration of the electron acceptor (22):

$$b = \frac{b'C_A}{K_A+C_A} \tag{17}$$

The rate of utilization of the electron acceptor will be related stoichiometrically to the rates of oxidation of the electron donor for cell synthesis and the decay of the cell mass:

$$-r_A = \frac{\mu X_f}{Y_{gA}} + \frac{bX_f}{Y_b} \tag{18}$$

where Y_{gA} is a conversion factor relating the mass of cells formed to the amount of electron acceptor utilized for cell synthesis and Y_b is a factor relating the mass of cells lost to decay to the mass of electron acceptor utilized for decay. If oxygen is the electron acceptor, Y_{gA} will be related to the true growth yield, Y_g (mg cells/mg COD removed) by:

$$Y_{gA} = \frac{Y_g}{1-\beta Y_g} \tag{19}$$

where β is the oxygen equivalence of the cell material (often taken as 1.25 mg O_2 or COD/mg cells or 1.44 mgO_2 or COD/mg volatile solids) (30). Y_b is just given by

$$Y_b = 1/\beta \tag{20}$$

When NO_3^--N serves as the electron acceptor, Y_{gA} is given by:

$$Y_{gA} = \frac{2.86Y_g}{1-\beta Y_g} \tag{21}$$

and Y_b by

$$Y_b = 2.86/\beta \tag{22}$$

where Y_g and β are on a COD basis. It is assumed in all of the above that biodegradable COD is used to express the concentration of the electron donor. If NH_4^+-N served as the electron donor appropriate conversion factors would be required in all of the equations.

Having delineated the equations for the reactions in the biofilm we now have established a framework which can be used to categorize fixed-film models. The following questions should be asked in the categorization:

(1) Which type of reaction rate expression is used?
(2) If a single-limiting substrate model is employed, what is the limiting material, the electron donor or the electron acceptor?
(3) Is the utilization of the nonlimiting material calculated?
(4) Is cell decay included in the electron acceptor balance?
(5) Is the film thickness an input to the biofilm model (either by assumption or by interfacing with the process model) or is it an output from the biofilm model?
(6) Is a distination made between thick and thin films? (A thin film is one in which both reactants penetrate to the support:biofilm interface). Such a distinction is necessary in establishing the boundary conditions and solution techniques for some models but is not needed with others.

Table I presents a review of some recent models in terms of these questions. Examination of the table reveals that many models have similar characteristics. When two or more references appear on the same line, the models in them are very similar, both with respect to the rate equations employed and the details involved. When two lines have the same entries, the models on them have similar characteristics but differ materially from one another in one or more ways which are not reflected by the questions. Not surprisingly, single-substrate models constitute the bulk of those which have appeared in the literature, primarily because of the evolutionary nature of research. It is now known that both the electron donor and the electron acceptor can be important determinants of the performance of a fixed-film process and thus the more recent models have sought to account for the effects of both. However, both interactive and noninteractive double-substrate limited models have been employed, reflecting the two philosophies discussed earlier. It appears to this author, however, that use of the noninterative model is less straight-forward because it requires identification of the limiting substrate within the biofilm before the correct equation can be chosen. Furthermore, it is possible for the limiting substrate to change with depth. Such complications do not exist with the interactive models, although they require solution of simultaneous differential equations. Many of the models require that the film thickness be an input,

Table I. CHARACTERISTICS OF BIOFILM RATE EQUATIONS

Model No.	Type of Reaction Rate Expression[1]	Limiting Material[2]	Consideration of Nonlimiting Material Utilization?[3]	Cell Decay in Electron Acceptor Balance?[3]	Film Thickness[4]	Distinction between Thick and Thin Films?[5]	References
1	SSM	ED	No	N/A	Input	Yes	17, 72
2	SSM	ED	No	N/A	Input	N/N	30, 31
3	SSM	ED	No	N/A	Output	N/N	8
4	SSM	ED	No	N/A	Output	Yes	70
5	SSM	ED	No	N/A	Input	Yes	28
6	SSM	ED	No	N/A	Output	Yes	47
7	SSM	ED	No	N/A	Input	Yes	61
8	SSM;SSB	ED	No	N/A	Input	N/N;No	73
9	SSB(1)	ED	No	N/A	Output	N/N	14
10	SSB(1)	ED	Yes	No	Output	N/N	15
11	SSB	EA	No	No	Input	Yes	11
12	SSB(0)	EA	No	No	Input	Yes	9,10,12,51
13	NDSM	ED,EA	N/A	No	N/N	Yes	19
14	NDSM	ED,EA	N/A	No	Input	Yes	25, 59
15	NDSB(0)	ED,EA	N/A	No	Input	Yes	13
16	IDSM	ED,EA	N/A	No	Output	Yes	70
17	IDSM	ED,EA	N/A	No	Input	N/N	63
18	IDSM	ED,EA	N/A	Yes	Input	N/N	22,23,36

[1] SSM, single-substrate, Monod; SSB, single-substrate, Blackman, (0) = zero order only, (1) = first order only; NDSM, noninteractive double-substrate, Monod; NDSB(0), noninteractive double-substrate, Blackman, zero order only; IDSM, interactive double-substrate, Monod.

[2] ED = electron donor; EA = electron acceptor

[3] N/A means that the question is not applicable to the particular reaction rate expression employed.

[4] Is the value of the film thickness an input or an output? N/N means that it is not necessary to know the thickness to proceed.

[5] N/N means that it is not necessary to make a distinction to solve the equations employed.

either by assumption or by interfacing with the process model. Others, however, provide the film thickness as an output from the biofilm model, either by involving some type of steady-state assumption or by solving the dynamic case. With some of the models knowledge of the film thickness is required to establish the solution technique which will be employed. Prior knowledge of the film thickness is particularly critical when a zero-order rate expression is employed because the point of substrate disappearance in the film must be identified to have the proper boundary condition for solution of the differential equation.

In summary, it is apparent from Table I that there are many ideas about how the rate equations for biofilms should be written. No doubt those ideas will continue to develop and change as we learn more about the processes.

SOLUTION TECHNIQUES

In the preceding sections we have established that transport of materials both up to and through a biofilm can have a significant effect upon the rates of reaction achieved by that film. This has an important impact upon the way that models of reactors containing biofilms must be solved. Consider for the moment a plug-flow reactor with biofilm distributed along its length. In modeling that reactor our primary interest will be in the change in substrate concentration axially within it; in other words we want to know how reactor length influences performance. We realize, however, that more than one substrate concentration exists at any axial position within the reactor. Returning to Figure 5 we see that the concentration in the bulk fluid is higher than that existing at the biofilm:liquid interface, and furthermore, that the concentration at the interface exceeds the concentration surrounding the organisms within the film itself, all because of the necessity for transporting materials from the bulk fluid through the biofilm. This means that the model for our reactor must combine mass balance equations in the axial direction with mass balance equations in a direction perpendicular to that axis (i.e., like Eq. 5). The terms in these equations must reflect both reaction and transport. In the axial direction, transport will be primarily by fluid flow and reaction will be limited to that caused by organisms being carried with the fluid. In the direction perpendicular to flow, transport will be by eddy diffusion in the liquid film and by diffusion within the biofilm. The bulk of the reaction, however, will

be caused by the organisms within the biofilm. Although the complexity of the resulting equations will depend upon the particular characteristics of the process being modeled, it will be necessary to solve the two equation sets simultaneously.

Three approaches have been used for solving the equations in the overall process model: direct, indirect and with an effectiveness factor. In the direct approach appropriate numerical techniques are employed to solve the two sets of equations simultaneously. As a consequence, the entire set of equations must be solved every time the model is employed to investigate a new condition. In the indirect approach, Eq. 5, which depicts reaction and transport within the biofilm, is solved for various concentrations of reactants in the bulk liquid and the result is expressed as the flux of material into the biofilm (which is equal to its removal rate from the bulk fluid) as a function of the bulk fluid concentrations, transport characteristics, etc. This flux relationship can then be used during solution of the process equations in an iterative manner, i.e., the concentration leaving the control volume by fluid flow is assumed and the corresponding reaction rate is determined from the flux:bulk concentration relationship. The flux is then used to calculate the concentration leaving the control volume and the procedure is repeated until the two concentrations agree. Even though an iterative procedure is utilized, the indirect technique is more efficient because the second order differential equation resulting from Eq. 5 need only be solved once to establish the flux:bulk concentration relationship and then that relationship can be used with any process model. The effectiveness factor approach is similar to the indirect approach but results in a somewhat more general solution. The effectiveness factor is defined simply as the ratio of the actual, observed reaction rate in the presence of mass transport limitations to the theoretical rate in their absence (i.e., the intrinsic reaction rate) (30). As such it becomes a correction factor that can be applied to the reaction rate as calculated from the intrinsic kinetics at the bulk substrate concentration, thereby converting that rate into the actual rate occurring in the presence of mass transport limitations both up to and through the biofilm. The second order differential equation resulting from Eq. 5 is solved with the appropriate boundary conditions to obtain the concentration gradients through the biofilm associated with various bulk substrate concentrations. The average reaction rate is then obtained by integrating over the entire biofilm depth and is divided by the intrinsic reaction

rate at the corresponding bulk substrate concentration to get the value of the effectiveness factor. By doing this for a large number of conditions, effectiveness factors can be obtained as a function of the various parameters describing transport to and through the film. Once this effectiveness factor relationship has been determined it can be used with any type of process model. Details of the procedures required to obtain effectiveness factor relationships are given elsewhere (30, 66). Like the indirect technique, once the effectiveness factor relationship is known, the second order differential equation arising from Eq. 5 need not be solved again to solve the process model. Rather the approach to the model solution is quite similar to the indirect approach discussed above. Let us now categorize the models in Table I in terms of the solution technique employed.

Direct Technique

Six of the models in Table I were solved by direct techniques: numbers 4, 9, 10, 16, 17, and 18. Howell and Atkinson (70) used both a single-substrate Monod model (#4) and an interactive double-substrate (electron donor and electron acceptor) Monod model (#16) to determine the active film thickness associated with various concentrations of substrate and oxygen at the biofilm:liquid interface. When the effects of both the electron donor and acceptor are being considered an equation like Eq. 5 must be written for each component, with appropriate reaction rate expressions substituted into each. For this more general case, an interactive model was utilized. Taking the limit as Δx approaches zero yields two second-order differential equations which must be solved simultaneously. These form a two point boundary value problem which is inconvenient to solve. However, by regarding the film thickness as an unknown and by assuming coupling between the removal of the electron donor and the electron acceptor (i.e., b in Eq. 19 was set equal to zero) Howell and Atkinson were able to convert the problem into an initial value problem which could be readily solved. Solutions were then obtained for a number of interface concentrations, yielding graphs which showed how those concentrations influenced the active film thickness. They also solved the equations for the situation where only the elector donor was rate limiting by setting K_A equal to zero, thereby making the reaction rates zero-order with respect to the concentration of the electron acceptor.

Andrews and Tien (14) used first-order substrate-limited kinetics (#9) to model biofilm growth and adsorption in a CSTR containing activated carbon granules. The electron donor (and adsorbate) was valeric acid and the electron acceptor was nitrate. A direct solution was used because the assumption of first-order kinetics and the absence of external mass transfer resistance made it possible to obtain an explicit solution to the second-order differential equation which results from Eq. 5.

Wang (15) extended the work of Andrews and Tien (14) by considering biofilm growth and adsorption in a fluidized bed reactor. Although the kinetics were again taken to be first order with respect to the concentration of electron donor alone, the utilization of electron acceptor was accounted for by stoichiometry (assuming no decay). In addition, the presence of two electron acceptors (oxygen and nitrate) was accounted for so that the biofilm was divided into two regions, aerobic and anoxic. As a result the system model contained a large number of simultaneous equations which were solved numerically.

Harris and Hansford (63) incorporated their biofilm model (#17) into a process model for a vertical biofilm with a thin liquid film flowing over it. Their biofilm model was written in terms of both the elector donor and acceptor with an interactive double substrate Monod equation like Eq. 14, resulting in two simultaneous second-order differential equations. The equations were directly coupled, however, because b in Eq. 19 was assumed to be zero. Only the situation without recirculation of fluid around the film was considered by breaking the vertical biofilm up into a number of sequential sectors. Starting with the first sector the concentration of electron donor entering in the liquid phase, C_{D1}^b, was known and the concentration leaving (C_{D2}^b) was assumed. By knowing the liquid flow rate through the sector it was then possible to calculate the substrate removal rate which must equal the flux of substrate into the biofilm. The flux of electron acceptor was then calculated from stoichiometry. Knowing the fluxes, C_D^b avg, C_A^b, and the external mass transfer coefficients made it possible to calculate the concentrations at the biofilm: liquid interface, C_D^* and C_A^*. These provided one set of boundary conditions for the simultaneous differential equations which were solved numerically to obtain the fluxes. These fluxes were compared to the external fluxes and the procedure was repeated until they agreed. The known output from the first sector then became the input into the second sector

and the procedure was repeated on down the vertical face of the biofilm. The results were then given as concentration profiles down the reactor.

Mueller et al. used the sector technique to model the performance of an RBC (22,23,36) as well as a trickling filter (36). They also used an interactive double-substrate Monod model (#18) but unlike the others they took cell decay into account when writing their rate equation for the electron acceptor. To simplify the determination of the concentration gradient into the biofilm they also broke it up into sectors. The biofilm model was coupled with the process model by equations depicting external mass transfer, diffusion, etc., and the entire system model was solved by finite-difference techniques. This biofilm model is one of the most complete, taking into account carbon oxidation, nitrification, and denitrification. Perhaps as a consequence, less detail has been provided in the literature about the solution techniques employed.

A review of the models that have been solved by direct techniques reveals that all of the interactive double- substrate Monod models fall into this category. This is probably because of the large number of parameters required, which make it more expeditious to obtain a direct solution for a specific application than to try to develop the dimensionless groups required for either the indirect or the effectiveness factor approaches. The use of the direct technique limits their flexibility, however, and requires a relatively large effort to obtain a solution for a new situation. The other models which were solved by direct techniques also were quite complicated but in one case an explicit solution was possible. Nevertheless, it appears that direct solutions to complete process models have been limited to specific problems for which general solutions are difficult.

Indirect Technique

In contrast to the direct technique in which the biofilm and process models are solved together, the indirect technique employs a generalized solution form for the biofilm model to arrive at specific solutions for particular process models. The generalized solution for the biofilm model often takes the form of a family of curves, although simplified equations have also been employed. The process model is then solved by using the generalized biofilm model in an iterative fashion. Four of the biofilm models in Table I (numbers 6, 7, 13, 14) were solved directly to arrive at generalized solutions which

could subsequently be used to solve a number of process models.

One of the earliest biofilm models to be solved in a manner which makes it available for use in the indirect technique is that of Williamson and McCarty (19) (#13 in Table I). Because it is a noninteractive double-substrate Monod model, solutions are presented for only one limiting constituent. Selection criteria are provided for determining which constituent (ED or EA) is rate limiting, although solution is restricted to the situation where a single constituent is limiting throughout the entire film depth. When Eq. 5 is applied to a single limiting constituent one second order differential equation results. To solve the equation they made use of the fact that the concentration and concentration gradient of the limiting constituent approach zero at a depth corresponding to the active film depth. They used a Runge-Kutta finite difference technique starting at an interior point where the concentration of limiting constituent was set equal to a small, nonzero value. Computation then proceeded in small steps toward the biofilm surface, with the concentration and gradient of the limiting constituent being calculated at each step. When the concentration equaled or slightly exceeded a preset interface concentration, C*, the computation was stopped and the flux was calculated as the product of the effective biofilm diffusivity and the concentration gradient just inside the biofilm. The results were presented as graphs of active film thickness and limiting constituent flux as a function of C*. Plots were prepared for five different values of K (including zero) and each plot contained seven curves for different values of the group $\mu_m X_f D_e / Y_g$. These curves can be used to solve any fixed-film process model. To incorporate external mass transfer effects, the flux and the bulk substrate concentration are assumed and the value of C* is determined. Using C* and the appropriate graph, the internal flux is determined and compared to the assumed value. If they agree, the flux is correct and the removal rate associated with the known bulk concentration is known. If not, a new flux is assumed and the procedure repeated. While this solution technique makes it possible to model processes without recourse to complex numerical techniques, the indirect solution provided by the graphs is limited in the number of parameter values considered. Furthermore, one must determine beforehand whether the electron donor or acceptor is limiting. These limit the model's utility. Nevertheless, this model served a useful purpose as the starting point from which other

models have been developed.

Williamson has continued to work with the noninteractive double-substrate Monod model with his latest effort being that with Meunier (25) (Model #14 in Table I). The basic biofilm model is similar in concept to the preceding one but the solution approach is different. Using the technique of Chung (59), the second order differential equation arising for Eq. 5 was integrated once by assuming that the concentration approaches zero within the film, thereby giving an equation for the substrate concentration gradient. Multiplication of the value of the gradient at the biofilm:liquid interface concentration, C*, by the effective biofilm diffusivity, D_e, results in the flux associated with C*. This flux can ultimately be expressed in terms of the bulk concentration, C^b, through knowledge of the mass transfer characteristics. This technique is only applicable when the concentration of the limiting component approaches zero within the biofilm and thus the solution is limited to what are called "thick" or "deep" biofilms. As seen earlier, most practical wastewater systems fall within this category. Furthermore, the solution is only valid when a single constituent is limiting throughout the entire film. Because of this; and because there will be some range of bulk fluid concentrations over which the limiting constituent changes within the biofilm, Meunier and Williams (25) have expressed their biofilm model solutions in the form of operating diagrams which can be used to solve specific process models (60). These operating diagrams show substrate flux as a function of C_A^b/C_D^b, the ratio of the bulk fluid concentrations of electron acceptor and electron donor. In developing the diagrams, C_D^b is fixed and the biofilm model is solved for various C_A^b concentrations. This can be done both for the region where the electron donor limits throughout the biofilm (which gives a single value of the flux for the fixed C_D^b value) and for the region where the electron acceptor limits throughout the film (which gives a flux value for each value of C_A^b). Two curves are obtained when these values are plotted on the operating diagram and these curves are connected by extrapolation to obtain the flux in the region where the limiting constituent changes within the biofilm. This must be done for a number of C_D^b values to generate the complete operating diagram. Because specific values for the kinetic and mass transport parameters must be assumed to generate the operating diagrams, each diagram is specific for a given biofilm process. Once it has been generated, however, the performance of that process can be evaluated

under a large number of conditions without resolving the differential equations. Furthermore, because the concentrations of both the electron donor and acceptor are incorporated into the operating diagram, no further consideration need be given to which is limiting while utilizing the diagram.

Rittmann and McCarty (61) also used the integration technique of Williamson and Chung (59) to solve a single-substrate Monod biofilm model (#7). The same general solution approach was utilized but because only one limiting constituent was considered they were able to present their results in dimensionless form, thereby increasing the generality of their plot of flux versus bulk substrate concentration. The parameters in their plot were effective diffusivity and active depth, both in dimensionless form. As a consequence their curves can be applied to any combination of kinetic and mass transfer parameters. The major limitation, however, is that they are limited to thick films because of the use of the integration technique. From inspection of their curves they developed simplified equations to depict them, thereby facilitating their use in the solution of a broad range of models for processes which contain thick biofilms.

The majority of the biofilm models have been solved for a biofilm thickness which is either assumed or is a coupling point with the process model. Rittmann and McCarty (47), however, extended the solution techniques of the previous model to one for a steady-state biofilm, i.e., one in which cell growth is just balanced by decay (#6 in Table I). In a steady-state situation there is a unique film thickness associated with each bulk substrate concentration. When that thickness is "deep", the concentration of substrate reaches zero at some interior point. When it is "shallow" a finite substrate concentration remains at the support:biofilm interface. Two solution techniques were utilized to generate the plot of flux versus bulk substrate concentration, depending upon whether the film was deep or shallow. Using the steady-state assumption, the film thickness, L_f, was calculated for an assumed flux and the deep film technique (61) was used to get the bulk substrate concentration associated with that flux. Because of the need for growth to balance decay in a steady-state biofilm there will be some minimum bulk substrate concentration, C_{min}^b required to maintain a steady-state film. When that concentration is reached the flux into the film will be zero and no film will be maintained. This means that film thickness will vary from zero at C_{min}^b to some maximum determined by the maximum substrate concentration. Furthermore

this means that the flux into a shallow, steady-state biofilm will vary from zero at C^b_{min} to the deep value at some higher bulk substrate concentration. The fluxes associated with C^b values between C^b_{min} and that for the thinnest "deep" film were calculated in the following manner. First, a value for the flux was assumed and the corresponding film thickness was calculated from the steady-state assumption. This film thickness was then divided into a finite-difference grid and the steady-state concentration profile in the biofilm was solved for (subject to the boundary condition that there be no flux into the solid surface) by an implicit, finite difference technique. To start the routine the value of C* was taken to be the value that gives the deep solution for the flux. The profile was then used to get the average reaction rate within the film by numerical integration. This average reaction rate was compared to the initially assumed flux and if they did not agree the procedure was repeated by assuming a new value for C*. When the two fluxes agreed, knowledge of the external mass transfer characteristics and C* allowed computation of the bulk concentration C^b associated with the flux. Repetition of this procedure resulted in a plot of flux versus C^b which was continued until it intersected the plot for the deep biofilm. The plot was made in dimensionless coordinates which incorporated all kinetic and mass transfer parameters except the external liquid film thickness, which was employed as a parameter. As in their previous model (61), they then developed a simplified equation to facilitate use of the model for solving various process models. There is a unique curve associated with each decay rate since it determines the steady-state film thickness. It should be recalled that the major criticism of this model was the assumption that decay is the only mechanism removing the biofilm, but that Rittmann (50) has shown how removal by shear stress may be incorporated, thereby allowing additional interfacing with a broader range of process models.

Examination of the models that fall into this category reveals that both noninteractive double-substrate and single Monod models have been employed. As we will see in the next section, single Monod models can be handled just as well, if not better, by the effectiveness factor technique because it allows more parameters to be included and simplifies the solution technique somewhat. Thus one must question whether the indirect technique is the best to use. This is particularly true for the noninteractive double-substrate Monod model because the oeprating curves developed were unique for a given

set of kinetics and mass transfer parameters. If those plots could be arranged as dimensionless plots their utility would be extended. Whether this can easily be done is not yet apparent.

Effectiveness Factor Technique

The majority of the models in Table I (#1,2,3,5,8,11,12) have been presented with effectiveness factor techniques and all are single substrate models. The first worker to apply this approach to the modeling of fixed-film biological reactors was Atkinson and his book (66) should be consulted for the details of how the effectiveness factor curves were developed. Generally, however, numerical techniques are used to solve the biofilm model directly and the results are used to determine the effectiveness factor as a functon of various dimensionless groups reflecting the kinetic and mass transport characteristics of the system. Atkinson and his coworkers have limited their effectiveness factors to transport within the biofilms so that the substrate concentration at the biofilm:liquid interface must be known or must be calculated from knowledge of the external mass transfer resistance. Such an effectiveness factor is called an internal effectiveness factor (30).

Atkinson and Howell (17) used the internal effectiveness factor technique to model substrate removal in a trickling filter with single-substrate Monod kinetics. A mass balance was written over a liquid element prependicular to the reactor axis, resulting in a first-order ordinary differential equation which equates the flux to the biofilm:liquid interface with the flux into the biofilm. The mass transfer coefficient approach (Eq. 3) was used to model the flux to the biofilm and the Monod equation in terms of the interface substrate concentration, C*, was multiplied by the internal effectiveness factor to compute the flux into the film. Algebraic manipulation allowed the differential equation to be rewritten in terms of C*, thereby giving an integral equation relating C* to the axial position in the reactor. Numerical solution then gave C* as a function of axial position and knowledge of the flux and the external mass transfer coefficient at each position allowed computation of the bulk concentration, C^b. Through the dimensionless groups the effectiveness factor is given as a function of both the film thickness and C* so these dependencies had to be accounted for during the numerical solution. Although this approach could be used directly by other investigators to model trickling filters under a broad

range of conditions, Atkinson and Howell (17) used their model to investigate a variety of limiting conditions and to write simplified analytical procedures for those cases, thereby facilitating computations.

Even though Atkinson and Howell (17) only used their solution technique for a trickling filter, Rittmann and McCarty (28) used it to develop relationships between the bulk substrate concentration and the substrate removal rate by biofilms of any thickness. Their results were presented as plots of dimensionless flux (removal rate) as a function of dimensionless bulk substrate concentration with dimensionless film thickness as a parameter. By so doing, they used the internal effectiveness factor technique to develop information which could be used in the indirect technique with bulk substrate concentrations.

Howell and Atkinson (8) also used the internal effectiveness factor technique to model sloughing in a trickling filter. In this case, however, they assumed that external mass transport was not limiting so the interface substrate concentration was equal to the bulk concentration, thereby simplifying the solution. The filter was modeled as a series of completely mixed elements and a dynamic equation was used in which film thickness within an element was allowed to increase with time. Film growth and substrate removal were modeled by the Monod equation with the bulk substrate concentration and the effectiveness factor. Integration was performed over a fixed time interval, thereby allowing the film thickness in each element to increase. At the end of each interval, a sloughing criterion was checked in each element and within each one meeting it, the film was sloughed, leaving a new thin film thickness. Integration again proceeded forward in time until the next time interval, when the criterion was again checked in each element. The results were used to investigate how sloughing introduces variation into the performance of a trickling filter.

Grady and Lim (30,31) have also used the effectiveness factor approach with the single-substrate Monod model, but unlike the previous examples they used an overall effectiveness factor which accounts for both internal and external mass transport limitations. The overall effectiveness factor was derived by Fink et al. (74) for immobilized enzyme catalysts and was solved by a transformation which permitted the rewriting of the two-point boundary value problem as an initial value problem. In this case the effectiveness factor was given as a function of a modified Thiele modulus (which

relates the maximum reaction rate to the maximum internal diffusion rate) with the Sherwood number (which relates external transport to internal transport) as a parameter. Although the general solution was presented in graphical form, empirical equations were given for certain regions to facilitate numerical analysis. They then developed models for both trickling filters and RBC's in which the substrate removal rate was expressed as a function of the bulk substrate concentration and the overall effectiveness factor. Since both the Thiele modulus and the Sherwood number depend upon the biofilm thickness that dimension serves as a link with the process model.

Jennings et al. (73) numerically solved the second order differential equation resulting from Eq. 5 for both the single-substrate Monod and the single substrate Blackman models. The reaction rates obtained were divided by the intrinsic rates for the two rate expressions to develop curves of overall effectiveness factors as functions of a number of variables. Their intent was to see how those variables influenced the effectiveness of the biological reactions and thus no attempt was made to develop an all-inclusive effectiveness factor plot like that developed by Fink et al (74). Nevertheless, the results were very useful in determining the conditions likely to maximize reaction rates. They were subsequently used to model a submerged filter.

Finally La Motta and coworkers (9-12,51) have used the effectiveness factor approach extensively in their modeling of fluidized bed biofilm reactors. In all cases, however, only an internal effectiveness factor was used, under the assumption that external mass transfer resistance was not important. Single-substrate Blackman kinetics was employed which enabled the development of explicit equations representing the effectiveness factor for both zero-order and first-order kinetics. These equations were then coupled with the intrinsic reaction rates in the process model to allow prediction of performance under a large number of conditions. Figure 1 illustrated this coupling.

Effectiveness factor techniques have a long history in the field of heterogeneous catalysis and have been beneficial in the modeling of fixed-film biological reactors containing a single limiting component. They are particularly advantageous where a large range of parameter values are likely to be encountered and can be easily coupled with process models through the biofilm thickness and the external mass transfer coefficient. Consequently, they appear to be more broadly applicable than the indirect technique for which graphs have

only been given over a restricted range of parameter values. No application of them has been made to double-substrate limited models, however, and for that situation more progress has been made with the direct and indirect techniques. There is no theoretical reason why effectiveness factors could not be developed for interactive double-substrate limited models, although they are likely to be complex and may not be amenable to two dimensional plots like those used for single substrate models. Nevertheless, the general utility of the effectiveness factor approach to the modeling of complex processes is sufficient to encourage the development of overall effectiveness factors for interactive double-substrate models. Perhaps the work that is underway in the modeling of double-substrate limited immobilized enzmes will provide guidance in the way to approach the problem (34,75).

CRITIQUE AND RECOMMENDATIONS

Having reviewed the characteristics of a number of biofilm models one question remains: How good are they? This is a difficult question to answer. When used in the simulation of various fixed-film processes, all give results which are qualitatively similar to observed performance. Furthermore, when the parameters are calibrated for a particular situation (i.e. reactor type, flows, nature of electron donor and acceptor, etc.) all do a reasonable job of tracking experimental data. Thus in one sense all of them are good for at least the limited situations for which they were derived. It will be recalled, however, that the purpose of this review was to evaluate mechanistic models and mechanistic models should be capable of predicting performance outside of our experience. How well will the models do that? To answer that we must look again at each of the component parts and ask how good they are.

First, consider transport in the liquid phase. It is evident that external mass transport limitations can and do occur and thus any mechanistic model of broad utility must include them. If they happen to be insignificant in a particular process application this insignificance will be reflected in the model solutions if the model is properly constructed. It makes no difference whether external transport is modeled with a diffusivity and a stagnant film thickness (Eq. 1) or with a mass transfer coerfficient (Eq. 3) since both lead to the same result. What is unknown, however, is the fate of the external mass transfer resistance as turbulence

becomes large. With the exception of the original work of Williamson and McCarty (19,20) and its subsequent use by Famularo et al. (22) and Mueller et al. (23), all models have treated the biofilm:liquid interface as if it were analogous to the interface between a flowing fluid and a solid support. Is this an accurate picture? Or does external resistance to mass transfer continue to exist even at high velocities because of the pseudohomogeneous character of the interface? If the latter is true, there are likely to be few circumstances in which the interface substrate concentration is equal to the bulk fluid concentration, thereby making a basic assumption of many of the models invalid. This is an area needing further study and is perhaps one to which microprobe technology could be applied with beneficial results.

Another important link between the biofilm model and the process model is the biofilm thickness, because that thickness is an important determinant of the concentration profiles which develop within the biofilm. Unfortunately, it is still unclear what controls that thickness. Rittmann and McCarty (47) have presented the concept of a steady-state biofilm in which cell growth is just balanced by cell decay and this appears to be a useful concept for biofilms growing in environments with low substrate concentrations, such as in aquifers receiving recharge by treated effluents. Such a situation is unlikely, however, in other environments so Rittmann (50) has extended the concept to a film in which loss is by attrition as well as by decay. How then does one handle the attrition rate? The work of Trulear and Characklis (6) and Zelvar (45) have shown that the rate depends both upon fluid shear stress and the mass of biofilm present. To be useful for modeling purposes it would be better to relate the attrition rate to thickness rather than mass but this can only be done directly if the density is constant. Evidence by Hoehn and Ray (4), Mulcahy and LaMotta (51) and Trulear and Characklis (6), however, all suggest that the biofilm density is influenced by the thickness, but both the mechanism and the functional relationship are unclear. Thus while it is apparent that the thickness of a biofilm will be determined by a balance between growth and loss by attrition and decay, it is not apparent how the rate of attrition should be modeled. More fundamental experimental work is needed in this area.

Almost all biofilm models assume steady-state biofilms of some sort. However, sloughing is a well known phenomenon although its mechanisms are unclear. Only Howell and Atkinson (8) have attempted to model sloughing, but their model

includes a number of simplifying assumptions, including one which limits biofilm loss to sloughing alone. Nevertheless, their work indicates that the irregular loss of biofilm by sloughing can have a major impact upon performance. The magnitude of that impact, however, will depend upon the frequency with which sloughing occurs (which will depend upon the net growth rate at the biofilm) and the thickness of the film left after sloughing. If the remaining film thickness is greater than the usual active thickness then the impact of sloughing would be small, whereas if it were smaller, the impact would be larger. Our ability to model this phenomenon depends upon knowledge of the attrition rate discussed in the preceding paragraph and the characteristics of the remaining film. Very little work has been done on the latter. Thus it appears that a good deal more experimental work is needed before this important aspect of fixed-film reactors can be adequately modeled.

The variation of density with thickness was discussed above with regard to its importance to the modeling of attrition. Such variations are also important because they influence the quantity of biomass present within the biofilm. As seen in Eqs. 6 and 7 the rates of cell growth and substrate removal both depend upon the amount of biomass present. While the majority of models assume that the density is independent of depth so the mass is directly proportional to thickness, the evidence cited above has shown that this is not the case. This constitutes an important weakness in most existing models. The key question, however, is whether changes in density occur within the active film thickness or only in the regions beyond which no significant transport occurs. If the latter case exists it may be adequate to model the reaction rate expressions with a constant density term. If the former is true, it will be necessary to use a variable density to accurately reflect the reaction rates. Again, additional experimental work is needed to resolve this.

The two main determinants of the concentration profiles within the biofilm are the rates of transport and reaction. Although considerable effort has been expended on evaluations of diffusion coefficients within biofilms there is little consensus in the literature regarding the magnitude of the retardent effects which may be attributed to the slime material within the film. This is a major weakness of current modeling efforts. There appears to be two possible causes for these variations: experimental techniques and variations in film microbial composition. As discussed in detail earlier

almost every investigator has a unique way of measuring rates of diffusion within films. Many of these require formation of an artificial film and it would appear that the exact conditions existing during formation of a film would determine its diffusive characteristics. Thus it is not surprising that diffusivities measured in films formed by filtration (20) differ from these measured in films formed by spreading (55,57). Futheremore, there is evidence that diffusivities in laboratory films are higher than those in field films (3,56). It appears that the more direct the technique for measuring the diffusivity and the fewer the assumptions involved in its computation, the more likely the values are to be correct. This suggests that microprobe techniques offer the best potential for determination of how various physical factors affect internal diffusivities. Certainly more work is needed in this area.

From the review of the results obtained with the various reaction rate models in Table I there can be no doubt about the fact that the transport and utilization of both the electron donor and the electron acceptor are important to the performance of a fixed-film reactor. This suggests that unless evidence to the contrary is overwhelming, double-substrate limited models should be employed. However, as seen in Table I, two-thirds of the listed models are single substrate models. Thus, unless care is taken to ensure that they are only applied in circumstances where only one component is limiting throughout, these models are likely to give predicted performance which is not in conformance with reality. With regard to the dual-substrate limited models the literature is divided as to whether they should be interactive or noninteractive. Furthermore, as pointed out by Bader (67), there is not yet sufficient evidence to allow conclusive determination of which is of the more general utility. Nevertheless, consideration of the circumstances under which each type of model is likely to be valid (67) and evaluation of the data of Ryder and Sinclair (76) suggests that an interactive model is more likely to be correct for situations in which electron donor and electron acceptor are the two limiting components. When this is coupled with the fact that a noninteractive model produces discontinuities in the solution (i.e., regions of limitation must be identified a priori), it would appear that an interactive model should be employed unless there is conclusive evidence that the noninteractive model is mechanistically more accurate. As far as the form that the model should take (Monod or Blackman) there is no conclusive evidence in either direction. The arguments for each are the same in this con-

text as they have been for cell growth in general because the rate equations should reflect intrinsic kinetics. The main argument for one over another in that context has been one of mathematical convenience (30). If that same argument is applied here, one would favor Monod over Blackman kinetics because it is a continuous function which avoids discontinuties. Certainly more work on intrinsic kinetics under double-substrate limitation is needed to resolve this issue.

Finally, as far as solution techniques are concerned a number of numerical procedures have been employed in direct solutions and to develop the graphs or effectiveness factor charts for the other techniques. Direct solutions offer perhaps the most straight forward approach to modeling of a fixed-film reactor. They have the drawback, however, of being complicated and therefore of being unlikely to be used by anyone other than the developer. Thus for wide-scale study of fixed-film reactors it would appear that either the indirect or effectiveness factor approaches offer the most utility. Of those two, the effectiveness factor approach appears to be more useful because its dimensionless groupings allow more parameters to be considered simultaneously. Furthermore, since the kinetic and mass transfer coefficients and the film thickness are incorporated into the solution in a way which allows them to serve as links with the process model, an effectiveness factor solution to the biofilm model can be developed while the questions regarding these items are being resolved. Thus it appears to this author that the next step in the development of mechanistic biofilm models of broad utility in process modeling should be the development of effectiveness factor relationships for interactive double substrate limiting kinetics. Since the solutions to the complex two-point boundary value problems need be made only once, they can be made with few simplifying assumptions, even if the required numerical solution are not very efficient. Once complete effectiveness factor solutions are available, however, then extensive sensitivity analyses can be run, resulting ultimately in simplified effectiveness factor charts which have little likelihood of being incorrect because of unwarranted simplifications.

In conclusion, it is clear that we have not yet achieved a complete and general mechanistic model for biofilms which can be used to simulate the performance of a broad range of fixed-film processes. It should be recalled, however, that a major goal of mechanistic modeling is to increase understanding.

The fact that the large number of events occurring within biofilms is now widely recognized is evidence for the attainment of that goal. Compared to the situation which existed twenty years ago a great deal of knowledge has been obtained and a great deal of progress has been made. Today, we have a good idea of what we don't know and therefore we can design the experimental programs required to gain that knowledge. With the renewed interest in fixed-film processes evident today even greater energies can be brought to bear upon the problem and the remaining gaps in knowledge can be filled.

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REFERENCES

1. Shieh, W. K. and Mulcahy, L. T., Discussion of Fluidized Bed for Biological Wastewater Treatment, Journal of the Environmental Engineering Division, ASCE, Vol. 107, 293-295, 1981.

2. Atkinson, B. and Fowler, H. W., The Significance of Microbial Film in Fermenters, Advances in Biochemical Engineering, Vol. 3, 221-277, 1974.

3. Chen, Y. S. and Bungay, H. R., Microelectrode Studies of Oxygen Transfer in Trickling Filter Slimes, Biotechnology and Bioengineering, Vol. 23, 781-792, 1981.

4. Hoehn, R. C. and Ray, A. D., Effects of Thickness on Bacterial Film, Journal, Water Pollution Control Federation, Vol. 45, 2302-2320, 1973.

5. LaMotta, E. J., External Mass Transfer in a Biological Film Reactor, Biotechnology and Bioengineering, Vol. 18, 1359-1370, 1976.

6. Trulear, M. G. and Characklis, W. G., Dynamics of Biofilm Processes, Presented at the 53rd Annual Conference of the Water Pollution Control Federation, Las Vegas, Nev., Oct. 1980 and accepted for publication in the Journal.

7. Castaldi, F. J. and Malina, J. F., Jr., Velocity-dependent Reaction Rates in a Slime Reactor, Journal, Water Pollution Control Federation, Vol. 54, 261-269, 1982.

8. Howell, J. A. and Atkinson, B., Sloughing of Microbial Film in Trickling Filters, Water Research, Vol. 10, 307-315, 1976.

9. Shieh, W. K., A Suggested Kinetic Model for the Fluidized Bed Biofilm Reactor (FBBR), Biotechnology and Bioengineering, Vol. 22, 667-676, 1980.

10. Mulcahy, L. T., Shieh, W. K., and LaMotta, E. J., Kinetic Model of Biological Denitrification in a Fluidized Bed Biofilm Reactor (FBBR), Progress in Water Technology, Vol. 12 (Toronto), 143-157, 1980.

11. Mulcahy, L. T., Shieh, W. K., and LaMotta, E. J., Simplified Mathematical Models for Fluidized Bed Biofilm Reactor, Water - 1980, AIChE Symposium Series, Vol. 77, No. 209, 273-285, 1981.

12. Shieh, W. K., Mulcahy, L. T., and LaMotta, E. J., Unified Design Basis for the Fluidized Bed Biofilm Reactor, Paper presented at the Second World Congress of Chemical Engineering and the Ninth Congress of the Inter-American Confederation of Chemical Engineering, Montreal, Canada, Oct. 1981.

13. Janson, J. la C. and Kristensen, G. H., Fixed Film Kinetics - Denitrification in Fixed Films, Report No. 80-59, Dept. of Sanitary Engineering, Technical University of Denmark, 1980.

14. Andrews, G. F. and Tien, C., Bacterial Film Growth in Adsorbent Surfaces, AIChE Journal, Vol. 27, 396-403, 1981.

15. Wang, S.-C. P., The Interaction between Adsorption and Microbial Growth in Biological Activated Carbon (BAC) Processes, Ph.D. Thesis, Syracuse University, Syracuse, N.Y., 1981.

16. El Amin, M. A., A Design Procedure for the Packed Bed Biological Reactor, Ph.D. Thesis, University of Wales, U. K., 1974.

17. Atkinson, B. and Howell, J. A., Slime Holdup, Influent BOD, and Mass Transfer in Trickling Filters, Journal of the Environmental Engineering Division, ASCE, Vol. 101, 585-605, 1975.

18. Atkinson, B., Busch, A. W., Swilley, E. L. and Williams, D. A., Kinetic, Mass Transfer and Organism Growth in a Biological Film Reactor, Transactions, Institution of Chemical Engineers, Vol. 45, 257-264, 1967.

19. Williamson, K. and McCarty, P. L., A Model of Substrate Utilization by Bacterial Films, Journal, Water Pollution Control Federation, Vol. 48, 9-24, 1976.

20. Williamson, K. and McCarty, P. L., Verification Studies of the Biofilm Model for Bacterial Substrate Utilization, Journal, Water Pollution Control Federation, Vol. 48, 281-296, 1976.

21. Welty, J. R., Wicks, C. E. and Wilson, R. E., Fundamentals of Momentum, Heat and Mass Transfer, John Wiley & Sons, Inc., New York, N.Y., 1969.

22. Famularo, J., Mueller, J. A. and Mulligan, T., Application of Mass Transfer to Rotating Biological Contactors, Journal, Water Pollution Control Federation, Vol. 50, 653-671, 1978.

23. Mueller, J. A., Paquin, P. and Famularo, J., Nitrification in Rotating Biological Contactors, Journal, Water Pollution Control Federation, Vol. 52, 688-710, 1980.

24. Levich, V. G., Physiochemical Hydrodynamics, Prentice Hall, New York, N.Y., 1968.

25. Meunier, A. D. and Williamson, K. J., Packed Bed Biofilm Reactors: Simplified Model, Journal of the Environmental Engineering Division, ASCE, Vol. 107, 307-317, 1981.

26. Snowdon, C. B. and Turner, J. C. R., Mass Transfer in Liquid Fluidized Bed of Ion Exchange Resin Beads, Proceedings, International Symposium on Fluidization, 599-608, 1967.

27. Rittmann, B. E. and McCarty, P. L., Evaluation of Steady State Biofilm Kinetics, Biotechnology and Bioengineering, Vol. 22, 2359-2373, 1980.

28. Rittmann, B. E. and McCarty, P. L., Substrate Flux into Biofilms of any Thickness, Journal of the Environmental Engineering Division, ASCE, Vol. 107, 831-849, 1981.

29. Jennings, P. A., A Mathematical Model for Biological Activity in Expanded Bed Adsorption Columns, Ph.D. Thesis, University of Illinois, Urbana, IL, 1975.

30. Grady, C. P. L., Jr., and Lim, H. C., Biological Wastewater Treatment: Theory and Applications, Chapter 14, Marcel Dekker, Inc., New York, 1980.

31. Grady, C. P. L., Jr. and Lim, H. C., A Conceptual Model of RBC Performance, Proceedings: First National Symposium/ Workshop on Rotating Biological Contactor Technology, Vol. II, E. D. Smith, R. D. Miller and Y. C. Wu (Eds.), 829-859, 1980.

32. Mixon, F. O. and Carberry, J. J., Diffusion within a Developing Boundary Layer - A Mathematical Solution for Arbitrary Velocity Distribution, Chemical Engineering Science, Vol. 13, 30-33, 1970.

33. Wilson, E. J. and Geankoplis, C. J., Liquid Mass Transfer at Very Low Reynolds Numbers in Packed Beds, Industrial and Engineering Chemistry Fundamentals, Vol. 5, 9-14, 1966.

34. Dahodwala, S. K., Humphrey, A. E. and Weibel, M. K., Pore Diffusion Model for a Two-Substrate Enzymatic Reaction: Application to Galactose Oxidase Immobilized on Porous Glass Particles, Biotechnology and Bioengineering, Vol. 18, 987-1000, 1976.

35. Brian, P. L. T. and Hales, H. B., Effects of Transpiration and Changing Diameter on Heat and Mass Transfer to Spheres, AIChE Journal, Vol. 15, 419-425, 1969.

36. Mueller, J. A., Paquin, P. and Famularo, J., Mass Transfer Impact on RBC and Trickling Filter Design, Water-1979, AIChE Symposium Series, Vol. 76, No. 197, 250-270, 1980.

37. Charpentier, J. D., Recent Progress in Two Phase Gas Liquid Mass Transfer in Packed Beds, Chemical Engineering Journal, Netherlands, Vol. 11, 161-181, 1976.

38. Characklis, W. G., Fouling Biofilm Development: A Process Analysis, Biotechnology and Bioengineering, Vol. 23, 1923-1960, 1981.

39. Kornegay, B. H. and Andrews, J. F., Kinetics of Fixed-Film Biological Reactors, Journal, Water Pollution Control Federation, Vol. 40, R460-R468, 1968.

40. Tomlinson, T. G. and Snaddon, D. H. M., Biological Oxidation of Sewage by Films of Microorganisms, International Journal Air and Water Pollution, Vol. 10, 865-881, 1966.

41. Bungay, H. R., Whalen, W. J. and Sanders, W. M., Microprobe Techniques for Determining Diffusivities and Respiration Rates in Microbial Slime Systems, Biotechnology and Bioengineering, Vol. 11, 765-772, 1969.

42. Sutton, P. M., Langley, D. F., Warner, K. E. and Park, K., Oxitron System™ Fluidized Bed Wastewater Treatment Process: Application to High Strength Industrial Wastewater, Proceedings, 34th Industrial Waste Conference, 1979, Purdue University, Ann Arbor Science Publishers, Inc., Ann Arbor, Mich., 450-461, 1980.

43. Tsezos, M. and Benedek, A., A Method for the Calculation of Biological Film Volume in a Fluidized Bed Biological Reactor, Water Research, Vol. 14, 689-693, 1980.

44. Shieh, W. K., Sutton, P. M. and Kos, P., Predicting Reactor Biomass Concentration in a Fluidized Bed System, Journal, Water Pollution Control Federation, Vol. 53, 1574-1584, 1981.

45. Zelver, N., Biofilm Development and Associated Energy Losses in Water Conduits, M. S. Thesis, Rice University, Houston, TX., 1979.

46. Shieh, W. K., Comments on Biologically Active Fluidized Beds, Biotechnology and Bioengineering, Vol. 23, 2145-2147, 1981.

47. Rittmann, B. E. and McCarty, P. L., Model of Steady-State-Biofilm Kinetics, Biotechnology and Bioengineering, Vol. 22, 2343-2357, 1980.

48. Rittmann, B. E. and McCarty, P. L., Design of Fixed-Film Processes with Steady-State Biofilm Model, Progress in Water Technology, Vol. 12, (Toronto), 271-281, 1980.

49. Arcuri, E. J. and Donaldson, T. L., Biofilm Kinetics, Biotechnology and Bioengineering, Vol. 23, 2149, 1981.

50. Rittmann, B. E., The Effect of Shear Stress On Biofilm Loss Rate, Biotechnology and Bioengineering, (accepted).

51. Mulcahy, L. T. and LaMotta, E. J., Mathematical Model of the Fluidized Bed Biofilm Reactor, Report No. Env. E. 59-78-2 of the Environmental Engineering Program, Dept. of Civil Engineering, University of Massachusetts, Amherst, 1978.

52. Alleman, J. E. and Veil, J. A., Observations of Attached Biofilm Composition and Structure Using Scanning Electron Microscopy, Paper presented at the 1981 Annual Meeting of the American Institute of Chemical Engineers, New Orleans, LA, Nov. 1981.

53. Alleman, J. E., Veil, J. A. and Canaday, J. T., Scanning Electron Microscope Evaluation of Rotating Biological Contactor Biofilm, Water Research, Vol. 16 (in press), 1982.

54. Bryers, J. D., Swiss Federal Institute for Water Resources and Water Pollution Control, Dubendorf, Switzerland, Personal Communication, 1982.

55. Matson, J. V. and Characklis, W. G., Diffusion into Microbial Aggregates, Water Research, Vol. 10, 877-885, 1976.

56. Bungay, H. R. and Harold, D. M., Simulation of Oxygen Transfer in Microbial Slimes, Biotechnology and Bioengineering, Vol. 13, 569-579, 1971.

57. Pipes, D. M., Variations in Glucose Diffusion Coefficients through Biological Floc, M. S. Thesis, Rice University, Houston, TX., 1974.

58. Mulcahy, L. T., Shieh, W. K., and LaMotta, E. J., Experimental Determination of Intrinsic Denitrification Kinetic Constants, Biotechnology and Bioengineering, Vol. 23, 2403-2406, 1981.

59. Williamson, K. J. and Chung, T. H., Dual Limitation of Substrate Utilization Kinetics within Bacterial Films, Paper presented at the 49th National Meeting of the American Institute of Chemical Engineers, Houston, Texas, March 19, 1975.

60. Meunier, A. D. and Williamson, K. J., Packed Bed Biofilm Reactors: Design, Journal of the Environmental Engineering Division, ASCE, Vol. 107, 319-337, 1981.

61. Rittmann, B. E. and McCarty, P. L., Variable-Order Model of Bacterial-Film Kinetics, Journal of the Environmental Engineering Division, ASCE, Vol. 104, 889-900, 1978.

62. Whalen, W. J., Bungay, H. R. and Sanders, W. M., Microelectrode Determination of Oxygen Profiles in Microbial Slime Systems, Environmental Science and Technology, Vol. 3, 1297-1298, 1969.

63. Harris, N. P. and Hansford, G. S., A Study of Substrate Removal in a Microbial Film Reactor, Water Research, Vol. 10, 935-943, 1976.

64. Atkinson, B. and Daoud, I. S., Diffusion Effects with Microbial Films, Transactions, Institution of Chemical Engineers, Vol. 48, 245-254, 1970.

65. Atkinson, B. and How, S. Y., The Overall Rate of Substrate Uptake (Reaction) by Microbial Films - 2. Effect of Concentration and Thickness with Mixed Microbial Films, Transactions, Institution of Chemical Engineers, Vol. 52, 260-268, 1974.

66. Atkinson, B., Biochemical Reactors, Pion Limited, London, 1974.

67. Bader, F. G., Analysis of Double-Substrate Limited Growth, Biotechnology and Bioengineering, Vol. 20, 183-202, 1978.

68. Monod, J., The Growth of Bacterial Cultures, Annual Review of Microbiology, Vol. 3, 371-394, 1949.

69. Blackman, F. F., Optima and Limiting Factors, Annals of Botany, Vol. 19, 281-295, 1905.

70. Howell, J. A. and Atkinson, B., Influence of Oxygen and Substrate Concentrations on the Ideal Film Thickness and the Maximum Overall Substrate Uptake Rate in Microbial Film Fermenters, Biotechnology and Bioengineering, Vol. 18, 15-35, 1976.

71. Bright, H. and Appleby, M., The pH Dependence of the Individual Steps in the Glucose Oxidase Reaction, Journal of Biological Chemistry, Vol. 244, 3625-3634, 1969.

72. Atkinson, B. and Abdel Rahman Ali, M. E., The Effectiveness of Biomass Hold-up and Packing Surface in Trickling Filters, Water Research, Vol. 12, 147-156, 1978.

73. Jennings, P. A., Snoeyink, V. L. and Chian, E. S. K., Theoretical Model for a Submerged Biological Filter, Biotechnology and Bioengineering, Vol. 18, 1249-1273, 1976.

74. Fink, D. J., Na, T.-Y. and Schultz, J. S., Effectiveness Factor Calculations for Immobilized Enzyme Catalysts, Biotechnology and Bioengineering, Vol. 15, 879-888, 1973.

75. Engasser, J.-M. and Hisland, P., Diffusional Effects on the Heterogeneous Kinetics of Two-substrate Enzyme Reactions, Journal of Theoretical Biology, Vol. 77, 427-440, 1979.

76. Ryder, D. N. and Sinclair, C. G., Model for the Growth of Aerobic Microorganisms under Oxygen Limiting Conditions, Biotechnology and Bioengineering, Vol. 14, 787-798, 1972.

INVESTIGATION OF SOME PARAMETERS IN RBC MODELING

Khalil Z. Atasi, Department of Civil Engineering
University of Michigan

Jack A. Borchardt, Department of Civil Engineering
University of Michigan

INTRODUCTION

Although the RBC system has been studied and used on many full scale treatment plants, a great deal of research is still needed in order to better define its optimum design and operational characteristics. Implementation of this concept requires a more thorough knowledge of the kinetics of substrate utilization by a fixed-film in the form of a rotating disc system. Investigation and application of RBC kinetics has suffered from the inherent complications involved in simultaneous processes such as liquid film mass transfer, diffusion and reaction within the biofilm.

OBJECTIVE AND SCOPE

It is the purpose of this research to further study the kinetics of the RBC process for carbonaceous substrate removal using a synthetic sewage. The interest is to concentrate on the mechanism of the reaction and its rate order, both observed and intrinsic.

This paper details part of the experimental work of a long range project, the final results of which will be published at a later date. The overall objective is the development of a simplified, practical approach to design. In essence the extremely complicated kinetic expressions will be approached in three steps. Step one deals with interphase diffusion and surface reaction occuring in series. In this step, the problem is dealt with using an external effectiveness factor and a modified Damköhler number to describe the effect of the mass transfer phenomenon. This step, eventually will be related to the hydrodynamic characteristics of any specific support system. Step two deals with intraphase diffusion and those reactions which occur simultaneously within the film. The intraphase problem will be attacked by using an internal effectiveness factor and a modified Thiele modulus. This step will relate the effect of internal diffusion and the biochemical reaction taking place together inside the film. Finally, step three will relate the external mass transfer, the internal diffusion, and the substrate oxidation through the use of an overall effectiveness factor. The ultimate goal is, then, to use this technique for modeling the RBC process through the use of simple equations in terms of observable quantities. This paper will deal only with step one and the parameters involved with the definition of the kinetics external to the fixed film.

REVIEW OF PREVIOUS RELATED RESEARCH

Because more than one phase is involved, the RBC system is a heterogeneous system where mass transfer, molecular diffusion, and biological oxidation take place at the same time in parallel and/or in series. Accordingly, it is important to consider the above phenomena when studying the different factors that might affect the substrate utilization rate by the biological film that grows on this rotating surface. In a simple way, the phenomena that take place when a biological film is brought into contact with a liquid containing soluble substrate are as follows:

1. Transport of the soluble substrate from the bulk liquid to the surface of the biofilm (liquid-biofilm interface);
2. Internal transport of the soluble substrate through the biological film by diffusional processes;
3. Biological oxidation of the soluble substrate by the biomass in the biofilm;

4. Diffusion of part of the reaction products to the bulk of the liquid.

Any kinetic information gathered on the substrate removal mechanism under the effect of mass transfer and diffusion will neither give the true or intrinsic kinetics nor the true mechanism as these are only a part of the above mentioned effects in some combination.

Most of the research reported in the literature related to biofilm kinetics has been carried out on purely laboratory experimental equipment (4, 19, 21, 23, 24, 26, 31, 35). As a result these observations may not have a direct practical application as they can only with difficulty be transferred to prototype equipment.

Little data are available on the intrinsic and overall rate of substrate utilization within stages of an RBC process. Because of difficulties encountered when dealing with these problems in the RBC, most of the investigations were run using the previously mentioned experimental set up. A zero order intrinsic rate was assumed (6,23). Some assumed Monod kinetics using that concentration observed in the bulk liquid (21) while others used first order reaction without taking into account the mass transport phenomenon (3). Harremöes (15), in modeling the biofilm as a porous diffusion model, reached an interesting conclusion, namely that a first order heterogeneous reaction in a pore will lead to a first order reaction in terms of the bulk concentration.

In dealing with the biofilm growing in the RBC process, Kornegay (22) assumed a homogeneous system with Monod kinetics in terms of the bulk substrate concentration, with the same biokinetic constants for all the stages. The same assumption was used by others (12, 27, 28) but with the added conditions that the biokinetic constants change stagewise. Indeed, the authors of this paper do agree with the last point above, but disagree with an observed rate based on Monod kinetics written in terms of the measurable bulk substrate concentration unless the mass transport phenomenon is taken into account. At a later point, it will be shown that when a mass transfer phenomenon affects the substrate removal, the overall rate will no longer follow Monod kinetics. More specifically, it will be demonstrated that the observed rate is first order only when the intrinsic rate exhibits a first order (pseudo) mechanism.

Antonie(2), and Stover and Kincannon (30), concluded that the RBC process follows first order kinetics in terms of the substrate bulk concentration dealing with an RBC pilot unit of more than one stage collectively. Using the same reasoning, Harremöes (16) fitted the data presented by Pöpel (29) for a seven stage RBC pilot unit into an observed (bulk) fractional order (half order) and hence suggested that the intrinsic rate for BOD consumption is zero order. It is important to realize that an RBC plant with several stages in series behaves as a plug flow reactor although each stage is a complete mix reactor. Because of a varying biomass (flora) stagewise (32) and widely different bulk substrate concentrations, the use of an overall complete mix technique and the inference of a single kinetic expression for all stages collectively, is doubtfull or only very approximate. As a result, this research implies that a kinetic study should be carried out on each stage separately.

With respect to the bulk dissolved oxygen within an RBC reactor, several investigators have stressed the importance of keeping a minimum bulk D.O. (2 mg/l) to retain the process efficiency (9, 11, 18, 34). On the other hand, Hartmann(17) suggested that the bulk D.O may not affect the efficiency at all. It has been shown experimentally (7,33) that the RBC process efficiency can be improved by sealing the reactors and enriching them with pure oxygen. Because of the above conflicting findings, research seems to be warranted to further elucidate this point.

THEORETICAL BACKGROUND

The following will be a development of a concept of the observed rate of substrate utilization by a biological film. Some assumptions are made; these are:

1. that bacteria are uniformly distributed within a biofilm
2. that the biofilm thickness is uniform
3. that the bio-film is at steady-state. That is, the density of the biofilm does not change within any experimental run.
4. that the suspended solids in the bulk fluid can be neglected (they are too low in concentration to have any marked effect).
5. that the mass transport phenomenon can be handled by assuming a hypothetical liquid film. In any case the mass transfer coefficient k_m includes the convective and

diffusive mass transfer effects.

6. that the biofilm is considered to be an "equi-accessible" surface; this has also been called the "quasi-stationary" method as developed by Frank Kamenetskii (13)

7. that there is a single substance limiting growth; i.e. the main substrate providing the carbon

8. that the substrate removed is assumed to be consumed at the surface of the biofilm.

9. that the biofilm area is the same as the disc area.

Under the steady state, the assumption is made that the substrate cannot build up or accumulate at the surface of the film. As a consequence, the rate of substrate supplied by the mass transfer phenomenon must equal the rate of substrate utilization in the reaction at the interface. Assuming that Monod kinetics prevails at the surface, and denoting R as the overall or surface reaction rate, it can be stated that:

$$R = k_m(S_b - S_s) = k\,x\,\frac{S_s}{K_{sat} + S_s} \qquad (1)$$

(all terms are defined at the end of this paper)
Because S_s is unknown and can't be measured, it is more convenient to express the rate expressions in terms of observable quantities.

Solving equation (1) above for S_s :

$$S_s = \frac{1}{2}\left\{\left[S_b - K_{sat} - \frac{k_{max}}{k_m}\right] + \left[S_b - K_{sat} - \frac{k_{max}}{k_m}\right]^2 + 4K_{sat}S_b\right]^{0.5}\right\} \qquad (2)$$

where $k_{max} = k\,x$ is defined as the maximum surface reaction rate in analogy with Michaelis-Menten enzyme kinetics.

Substituting the value S_s as given by equation (2) into either side of equation (1), it can be shown that:

$$R = k_{max}\,\frac{\left(-K_{sat} + S_b - \frac{k_{max}}{k_m}\right) + \left\{\left[\left(S_b - K_{sat}\right) - \frac{k_{max}}{k_m}\right]^2 + 4K_{sat}S_b\right\}^{0.5}}{\left(K_{sat} + S_b - \frac{k_{max}}{k_m}\right) + \left\{\left[\left(S_b - K_{sat}\right) - \frac{k_{max}}{k_m}\right]^2 + 4K_{sat}S_b\right\}^{0.5}} \qquad (3)$$

From equation 3, it appears that Monod kinetics can become a pseudo first or zero order depending on the relative value of K_{sat} as compared to the substrate concentration. If K_{sat} is much larger than S_s, then equation (2) becomes:

$$S_s = \frac{k_m}{k_m + (k_{max}/k_{sat})} S_b \quad \text{(for pseudo-first order)} \tag{4}$$

Under this condition, and according to the value of k_m two regimes (13) might exist:

1. A kinetic regime, if $k_m >> (k_{max}/k_{sat})$, where the following prevails:

$$S_s \tilde{\simeq} S_b \tag{4a}$$

or

2. A mass transfer regime, if $km << (k_{max}/k_{sat})$, where the prevalent condition is:

$$S_s << S_b \tag{4b}$$

as a result, and for maximum efficiency, an RBC system should be operated under case (1) above. Since k_m is related, among other things, to the hydrodynamic characteristics of the system, the appropriate kinetic regime could possibly be attained depending on the design of the system.

For pseudo first order, equation (3) becomes:

$$R = k_o S_b \tag{5}$$

where,

$$1/k_o = 1/km + 1/(k_{max}/k_{sat}) \tag{6}$$

where k_o is the observed first order reaction rate.

So, in the presence of mass transfer resistance, and based on the above, the following can be concluded:

1. The rate expression, in terms of the observable bulk concentration, will not exhibit a Monod type mechanism, even when the intrinsic rate is assumed so (equation 3).
2. The observable rate will exhibit a first order mechanism, only when the intrinsic rate exhibits a pseudo-first order rate in S. In this case, it is additive (equation (5), (6)). This finding agrees with Harremöes (16).
3. When the intrinsic rate is zero order, then R is no longer influenced by the mass transfer phenomenon.

The above can be simplified by using a dimensionless concept (5, 10, 20). For this purpose, the following dimensionless numbers are defined:

$$\Psi = \frac{K_{sat}}{S_b}$$

$$s = \frac{S_s}{S_b}$$

$$Da = \frac{k_{max}}{k_m S_b} = \frac{\text{Maximum reaction rate}}{\text{Maximum mass-transfer rate}}$$

where Da stands for the Damköhler number. The magnitude of the Damköhler number indicates the significance of the mass resistance. Thus:

If Da >1 : a mass-transfer regime prevails
and if Da <1 : the reaction is rate limited

By substituting the above dimensionless numbers in the previous equations, the following results can be obtained:

For Monod kinetics:

$$s = \frac{\alpha}{2} \left(\frac{\alpha}{2} \left(1 + 4\Psi/\alpha^2 \right)^{0.5} - 1 \right) \qquad (7)$$

where $\alpha = Da + \Psi - 1$

For pseudo 1st order kinetics:

$$s = \frac{\Psi}{\Psi + Da} \qquad (8)$$

Finally, defining an external effectiveness factor η_e as the ratio of the reaction rate in the presence of mass transfer to the rate which would be obtained with no mass transfer resistance, that is when $S_s = S_b$, it can be shown:

For Monod kinetics:

$$\eta_e = s \quad \frac{\Psi + 1}{\Psi + s} \tag{9}$$

where s is given by equation (7)

For pseudo 1st order kinetics:

$$\eta_e = \frac{\Psi}{\Psi + Da} \tag{10}$$

In its application the external effectiveness factor acts like a correction factor. It provides for a decrease in the reaction rate due to the presence of mass transfer resistance. Accordingly, the reaction rate can be written as follows:

For Monod kinetics: $$k_{max} \frac{S_b}{K_{sat} + S_b} \eta_e \tag{11}$$

For pseudo 1st order: $$\frac{k_{max}}{K_{sat}} S_b \eta_e \tag{12}$$

where η_e is given by the appropriate equation.
This approach is much simpler than equation (3) and (5) and can prove to be advantageous if η_e can be represented graphically in terms of the relevant parameters such as Da and Ψ. But, before doing so, the Da number should be clarified. As can be seen, the Da number is not an observable quantity. Therefore, a new term must be introduced, $\overline{Da}$, which can be designated as an observable Damköhler number, such that:

$$\overline{Da} = \eta_e Da \tag{13}$$

by algebraic conversion, it can be shown that:

For Monod kinetics: $\overline{Da} = \frac{R}{k_m S_b}(1 + \Psi)$ (14)

For pseudo 1st order kinetics: $\overline{Da} = \frac{R}{k_m S_b}\Psi$ (15)

It becomes advantageous to relate analytically η_e to $\overline{Da}$. This has been done and it can be shown that:

For Monod kinetics: $\eta_e = \frac{(1 + \Psi - \overline{Da})}{1 + \Psi - (\overline{Da}/(1 + \Psi))}$ (16)

For pseudo-1st order kinetics $\eta_e = 1 - \frac{\overline{Da}}{\Psi}$ (17)

Figure (1) shows a computer plot of η_e as a function of $\overline{Da}$ and values of Ψ as a parameter, for pseudo 1st order reaction kinetics.

It should be remembered that $\Psi > 1$ as equation (17) holds only when the intrinsic rate is pseudo-first order in the substrate concentration. It is very clear from figure (4) that the kinetic regime becomes well defined at a value of $\eta_e = 1$. At this point the relationship becomes a horizontal line for values of $\overline{Da} < 1$. Contrariwise, the vertical lines resulting at $\overline{Da} > 1$ and $\eta_e << 1$ represent the mass transfer regime. Accordingly, an intermediate region exists between the above two regimes, where both mass transfer and kinetics affect the process. This region is evidenced by a drastic transition in the slope of the curves. It can be seen also, that any increase in $\overline{Da}$ above a certain critical value, would not have any impact on the results, since the overall reaction is mass transfer limited. By knowing $\overline{Da}$ and Ψ, one can find η_e so that the rate expression can be expressed by equation (12) in terms of the bulk substrate concentration S_b, which is an observable quantity.

EXPERIMENTAL EQUIPMENT, MATERIALS, AND METHODS:

The experiments detailed in this paper were run on the first stage of a pilot plant consisting of six stages of RBC. Each stage consisted of four discs 2 ft. diameter fabricated of ultra thin sheets of polyethylene with a sinusoidal surface configuration that generated a great deal of turbulence. This deformation increased the area

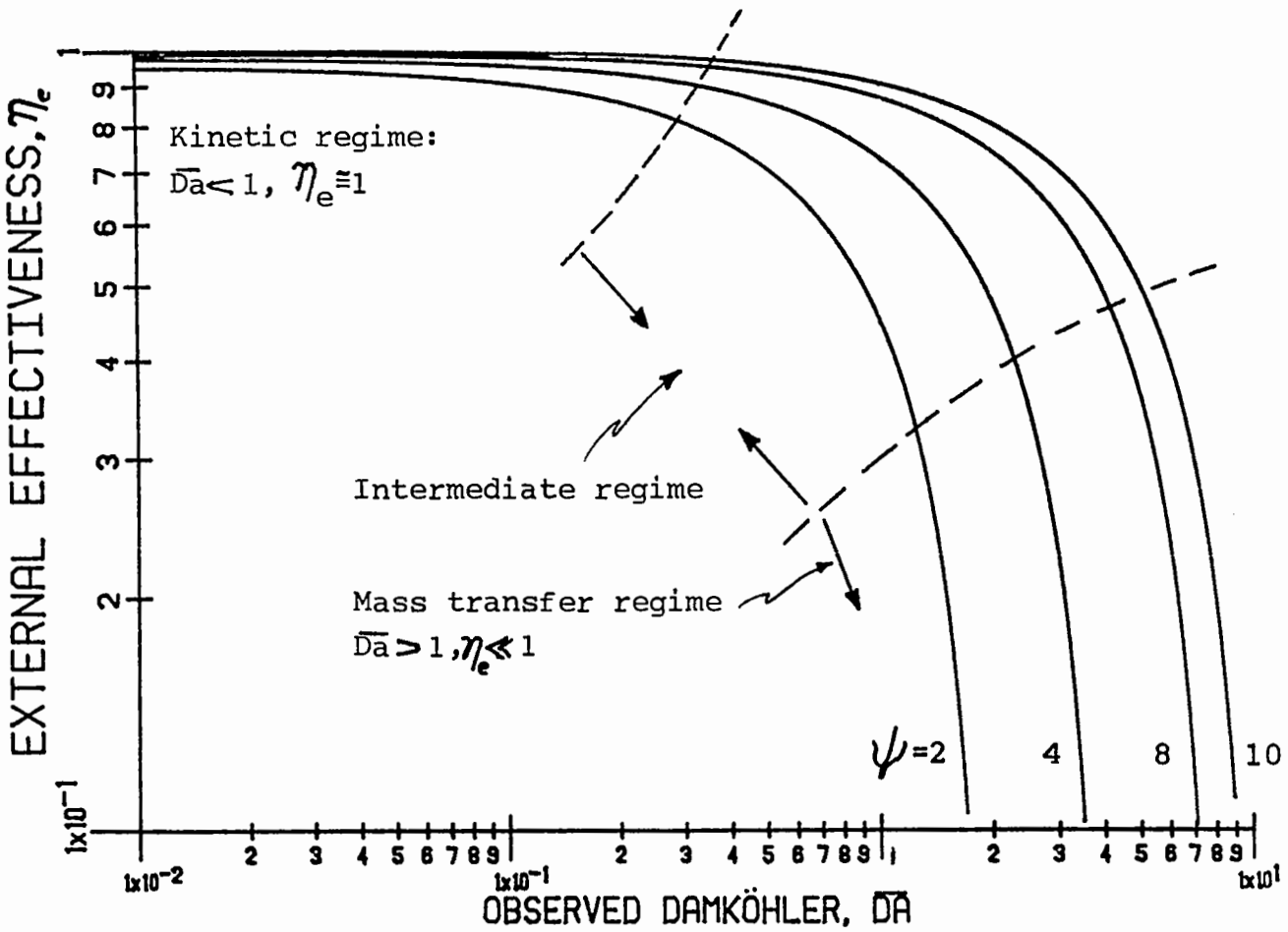


Figure 1. Biofilm external effectiveness as a function of the observable Damköhler number and ψ for first order reaction.

of the disc by an average of 50% over a flat disc turbulence. The sheets were heat formed spot welded, and provided by the FMC corporation. The rectangular tanks made of plexiglass, were 5 1/2" wide, 11" deep, and 28" long for each stage. A one inch diameter hole was drilled in the partition wall between stages for the flow of sewage. A concrete fillet with a triangular cross section of 9" x 9" coated with parafin wax, was slipped into each stage to avoid any possible shortcircuiting. The discs were mounted on a stainless steel shaft, 3/4" diameter, equipped with a sprocket and chain drive which was driven by an AC motor and speed controller to provide different rotational speeds. The discs were approximately 35% submerged. The liquid volume was about 18.5ℓ. The surface area provided by one stage (one module) was about 38 ft^2. Fig (2) shows the RBC reactor and Fig. (3) shows a detail of the rotating media. This configuration provided a ratio of growth surface area to liquid volume, a, equal to 1.92/cm.

The pilot plant was fed with a synthetic sewage (the formulation is shown in) Table (1).

Table 1. Composition of the synthetic sewage.

to 1 liter of tap water* add

Dextrin	184.94 mg
NH_4Cl	76.43 mg
Na_2HPO_4	15.85 mg
$MgSO_4 7H_2O$	8.2 mg
Beef Consomme**	1.05-2.1 ml

*Ann Arbor municipal tap water was used
**Campbell's condensed consomme (beef) soup
Campbell Soup Co., Camden, N.J. 08101

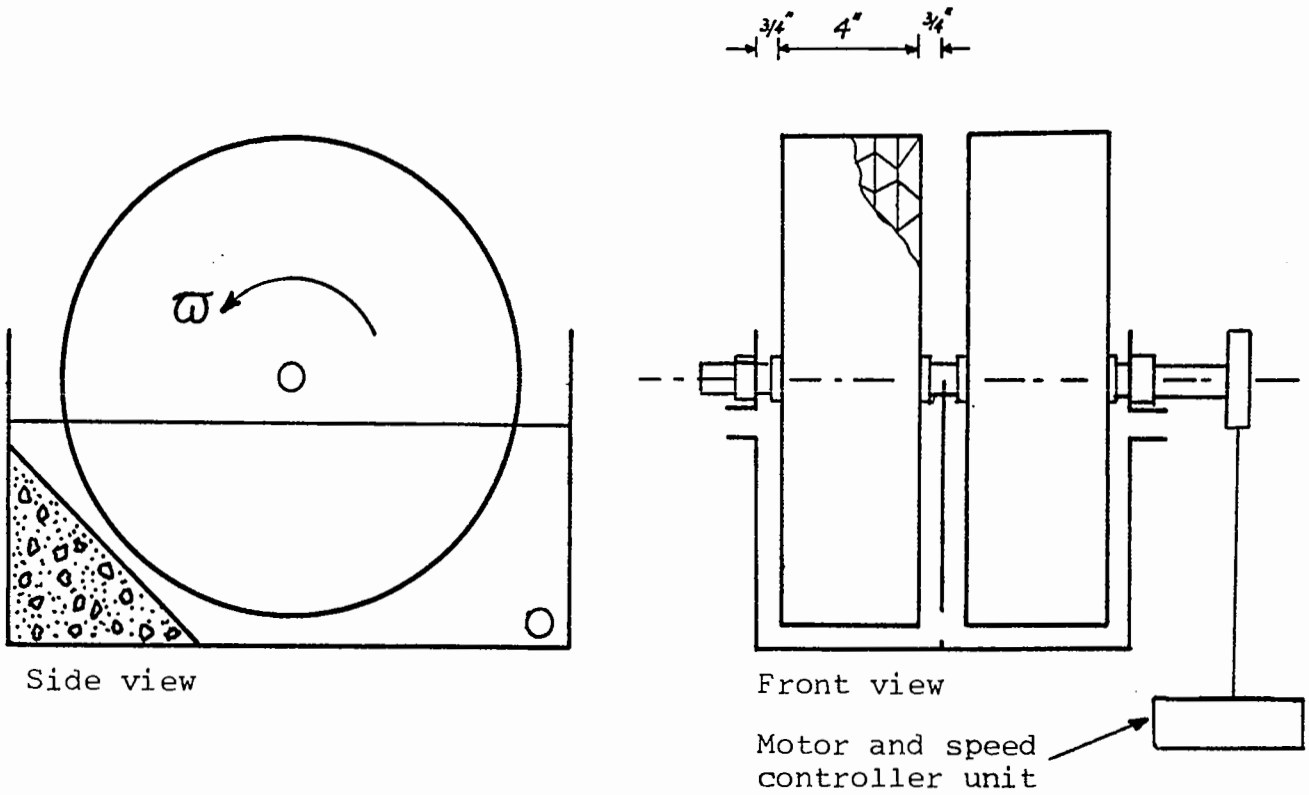


Figure 2. An overall view of a biological reactor disc unit with the FMC media.

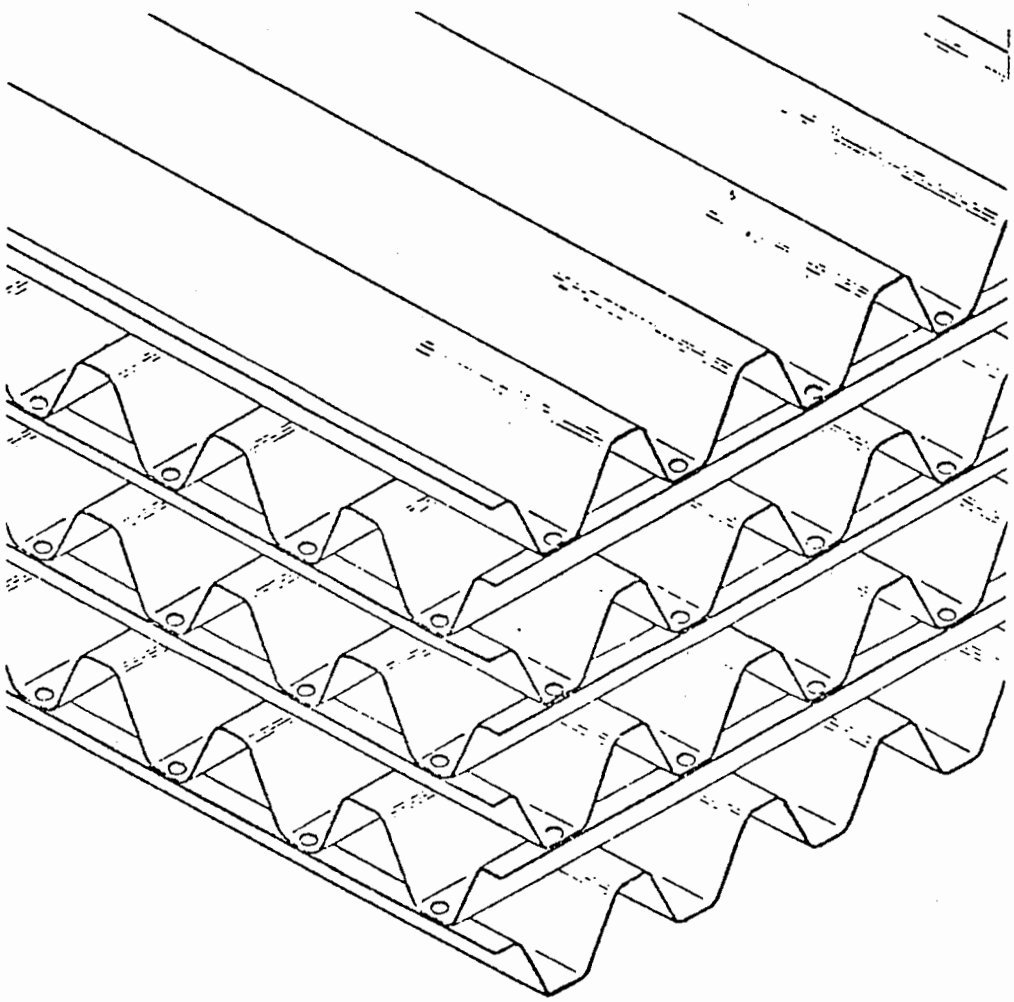


Figure 3. FMC media detail (high density polyethylene). Approximately 1/3 scale.

This synthetic sewage provided a waste with the following approximate strength:

$BOD_5 \cong 250 - 290$ mg/l

COD $\cong$ 300 - 350 mg/l

The above synthetic sewage can be concentrated and diluted to cover a wide range of organic loading at any specific hydraulic loading.

The influent and effluent was monitored daily with analyses being performed for suspended solids, volatile suspended solids, and soluble chemical oxygen demand. These analyses were conducted in compliance with Standard Methods (1975). Dissolved oxygen was measured with a D.O meter (YSI model 54ARC) and pH with a pH meter (Corning Model 12). Both the D.O. and pH were measured in situ as well as the flow and temperature. Samples were filtered (for TSS, VSS, COD) using glass fiber filter, Type A-E 47 mm diamter, manufactured by Gelman Instrument Co. The biofilm density was measured by a scraping technique from a measured area and the scraped biomass volume was measured volumetrically. Film density estimated in the scraped biomass gravimetrically.

EXPERIMENTAL PROCEDURES

Prior to the experimental work detailed in this paper, the pilot plant was operated for three months, approximately, under the following loading and other physical conditions:

Hydraulic Loading:	1.85 gpd/ft^2	(based on flat area)
Influent, COD:	305-325 mg/l	(BOD:250-268 mg/l)
Effluent of first stage, COD:	80 + 5 mg/l	(BOD:77-77 mg/l)
Rotational Speed:	8 rpm	(70-77) mg/l

The total suspended solids within any stage, had an average of about 200 mg/l and the maximum value found in that time span was 592 mg/l. On this basis, it was decided to neglect the effect of suspended solids in the substrate removal.

Because kinetics and rate studies are best performed on a batch mode, it was decided to run all the experiments on the first stage of this pilot plant, using the batch mode during each experiment. Before each test was undertaken, the pilot plant was running under steady state with a well established biofilm. Under such conditions, the biofilm density was 28.34 to 30 mg TS/ml and its thickness was about 1856-2325 μ. A point worth noting is that all the disc surfaces were covered by slime.

When running any batch kinetics test, the flow was stopped, and the reactor outlets were sealed by a rubber stopper. Then adding a precalculated volume (usually small) of a concentrated solution of the synthetic sewage, the test was started by collecting samples every five minutes for at least one hour. The organic content of each sample was measured by the COD test after filtering. Hence the COD values reflect only solubles. Total suspended solids were very low, as mentioned earlier, and neglected. Temperature in all the runs was about 21 $\pm$ 0.5°C. The dissolved oxygen within the run of any batch test never went below 2.7 mg/l. These tests were run at two different levels of initial substrate concentration; 80 mg/l COD (average) and 500 mg/l COD; rotational speed was varied from 4 to 10 rpm in an increment of 2 rpm.

To study the effect of dissolved oxygen concentration in the bulk liquid on the substrate utilization rate, it was decided to run several tests under a condition of zero D.O. in the bulk liquid. To remove the D.O. from the liquor, nitrogen gas was bubbled through the reactor during the batch operation. The nitrogen bubbles stripped the D.O. from the liquid phase. In any run, the liquid became void of D.O. within 15-30 minutes. But even with zero D.O. the nitrogen gas was kept flowing for an extra 45-60 minutes. Then, at time zero, a precalculated volume of a concentrated sewage was added and the test started while maintaining the flow of nitrogen gas throughout the run. D.O. was monitored continuously and was always 0.0 mg/l except for a couple of runs where a trace of D.O. was detected. Those tests with zero D.O. were all conducted with an initial substrate concentration of 516 mg/l COD on the average, and a rotational speed equals to 4 rpm.

The maximum resistance to mass transfer usually is exhibited when no turbulence at all exists. It was desirable to estimate this value so that the importance of rotation (turbulence) on the efficiency of any type of disc media would be demonstrated. To do so, a slide previously attached to the discs and covered with biomass was suspended in 2ℓ of substrate with an average of 530 mg/l COD. This system was controlled as far as temperature, D.O. and minimal mixing were concerned. This slide provided an area of 168 cm^2 approximately. A ratio of surface area to liquid volume, a_f, was calculated at 0.084/cm.

In order to measure the intrinsic reaction rate of the substrate uptake rate by the biofilm, two factors had to be fulfilled: (1) eliminate the resistance to mass transfer and (2) to minimize the internal diffusion problem (unless substrate is consumed at the surface layer). In the field of catalytic engineering, this has been achieved by using a semi-batch reactor. These units share a common feature of tremendous high fluid flow rates near the catalyst surface to minimize mass transfer resistance (generate a small Da. But, unfortunately, this requirement presents a limitation for the RBC system since it would result in a tremendous amount of shear and biomass sloughing at high rotational speeds. In addition, all rotational speeds above 10.5 rpm were impossible because there was excessive liquid loss from the reactor due to tremendous turbulent splashing. To overcome such problems, the following procedure was devised. Some slime was scraped from a measured area (457.8cm^2), then homogenized in a blender for a few seconds (less than 8 seconds) and then the dispersed biomass was suspended in a batch reactor. Oxygen was maintained by aeration. The air was provided for both D.O. (minimum was 9.8 mg/l) and to assure a complete mix regime. Samples were collected every 5 minutes for one hour. These samples were centrifuged and the supernatant filtered. The filtrate, was analyzed for organic carbon by the COD test. All biomass from the centrifugation and filtration process was returned to the reactor to minimize the loss of bio active solids. This test was an attempt to minimize the mass transfer resistance of the film as well as the internal diffusion if such effects existed. Because this biofilm was suspended, it had to be related in some way to the area of the disc. This was done by calculating a factor a_s, defined as the ratio of the area scraped to the batch volume. In this test the relationship was 0.23/cm. The TSS in such tests

(the suspended biofilm) measured an average of 4900 mg/1.

Table (2) below summarizes the tests and testing conditions under which these runs were performed.

Table 2. Tests Run and Testing Conditions

(all are batch reactors; Temp. = 21°± 0.5°C

Type	Initial Substrate Conc. COD, mg/l	$\bar{w}$, rpm	Bulk D.O. mg/l	Number of Runs
RBC, a=1.92/cm	515	4	2.7-2.8	4
	494	6	2.7-2.8	4
	538	8	3.0-3.1	4
	513	10	5.6-5.8	4
	120	4	3.8-4.0	4
	76	8	4.4-4.6	4
	516	4	0.0*	6
Slide a_f=0.084/cm	521	--	>10.2	4
Suspended Film a_s=0.23/cm	551		>9.8	4

*With perfusion of nitrogen gas

RESULTS AND DISCUSSION

After obtaining the analytical results from the batch studies by monitoring S ver time, an attempt was made to fit these data to several kinetic models:

Zero order	$-r_s = k$	(plot S_b ver t)
1st order	$-r_s = k\ S_b$	(plot Log S_b ver t)
2nd order	$-r_s = k\ S_b^2$	(plot $1/s_b$ ver t)

where the k's are not the same.

The results of all data without exception, fitted the 1st order model. In this paper it is impossible to show all the results but Fig (4) shows some typical plots. Table (3) shows the averages of the results using a calculated reaction rate constant.

It can be seen from these results that the observed rate of substrate utilization in this RBC reactor is 1st order in the measurable bulk substrate concentration, at least within the range of experimental data obtained. As a result:

$$R = k_o\ S_b \qquad (6)$$

k_o is the observed 1st order substrate reaction rate constant in units of LT^{-1}. This then is related to k_{eo} (where "e" indicates the natural logarithm base, and "o" indicates observed) as follows:

$$k_o = k_{e,o}/a$$

$k_{e,o}$ is calculated from the slope of the plot of LnS_b ver t by fitting the data into a linear regression model using the least square estimator. These results have an important implication. It was demonstrated earlier in this paper that for an observed rate to be first order, and because mass transfer is likewise 1st order, it follows that the intrinsic rate has to be first order (method of additive or combined resistance, Frank Kamemetskii (13). In fact it was also demonstrated that the intrinsic rate was a 1st order rate. These data also show that the

Table 3. Summary of Experimental Results

Type	Initial Substrate Conc., COD, mg/l	w, rpm	Bulk D.O., mg/l	$k_e o$, 1/day	$k_o = \frac{keo}{a}$ $\frac{cm}{day}$	k_m, cm/day
RBC,	515	4	2.7-2.8	35.83 ± 2.27	18.66	22.68
a=1.92 /cm	494	6	2.7-2.8	53.97 ± 2.57	28.11	38.36
	538	8	3.0-3.1	56.09 ± 5.13	29.21	40.43
	513	10	5.6-5.8	64.65 ± 2.69	33.67	49.51
	120	4	3.8-4.0	14.30 ± 2.51	7.45	
	76	8	4.4-4.6	23.89 ± 8.80	12.44	
	516	4	0.0	34.54 ± 3.46	17.99	
Slide	521	-	>10.2	1.24±0.039	14.76	
a_f = 0.084 /cm						

For the intrinsic reaction rate: (at an initial substrate concentration of 551 mg/l COD) from the plot (lnS) ver (t) get the slope: ke,i = 24.2 + 1.15, 1/day
and as related to surface area, where $a_s = \frac{457.8}{2000} = 0.23$ /cm

$$ki = \frac{ke,i}{a_s} = \frac{24.2}{0.23} = 105.22 \text{ cm/day}$$

*This is the slope of the plot in S ver t, where t = time

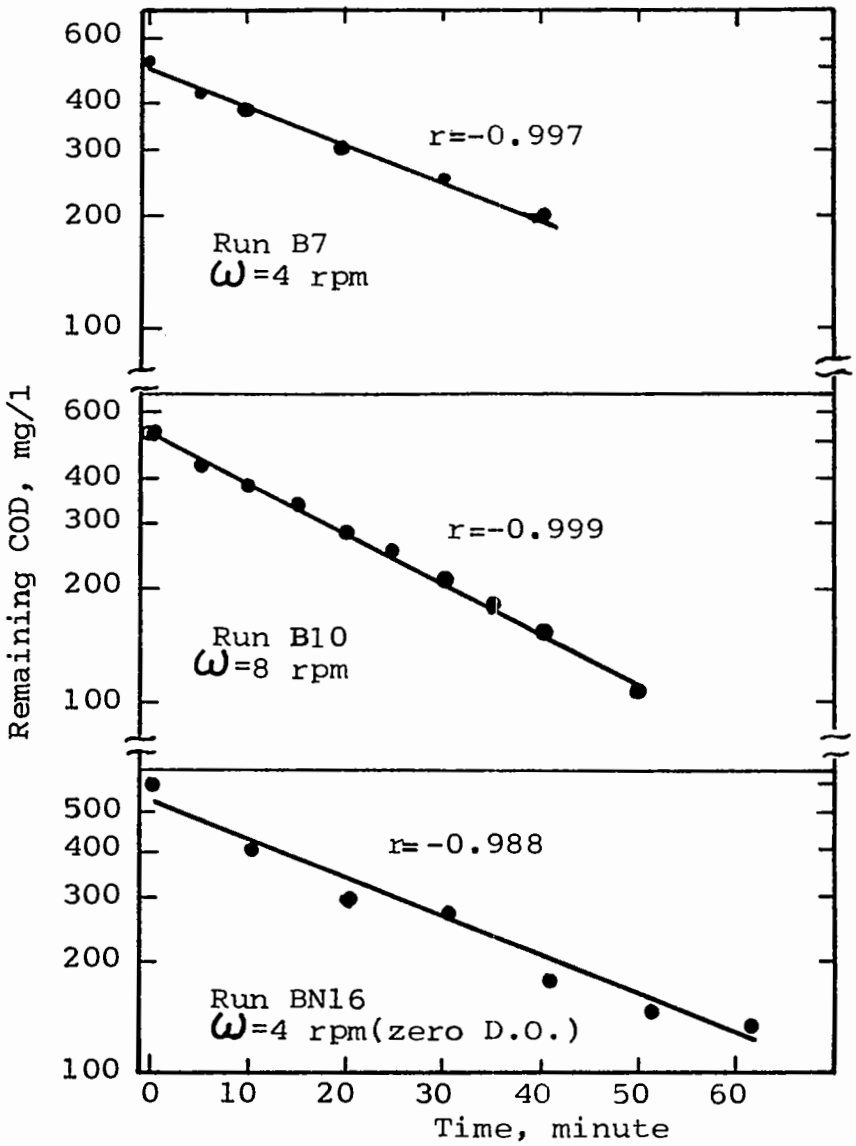


Figure 4a. Log remaining COD ver. time for first order kinetics model.

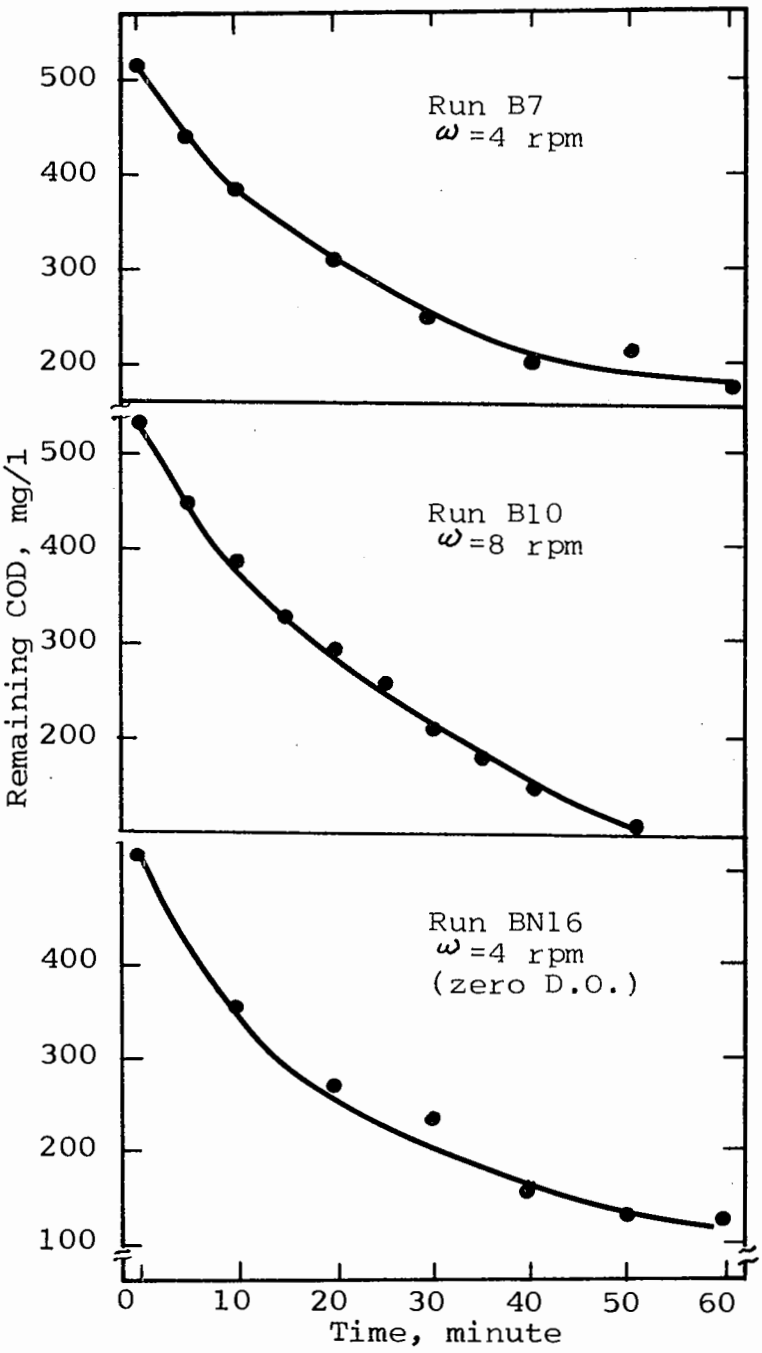


Figure 4b. Remaining COD ver. time for zero order kinetics model.

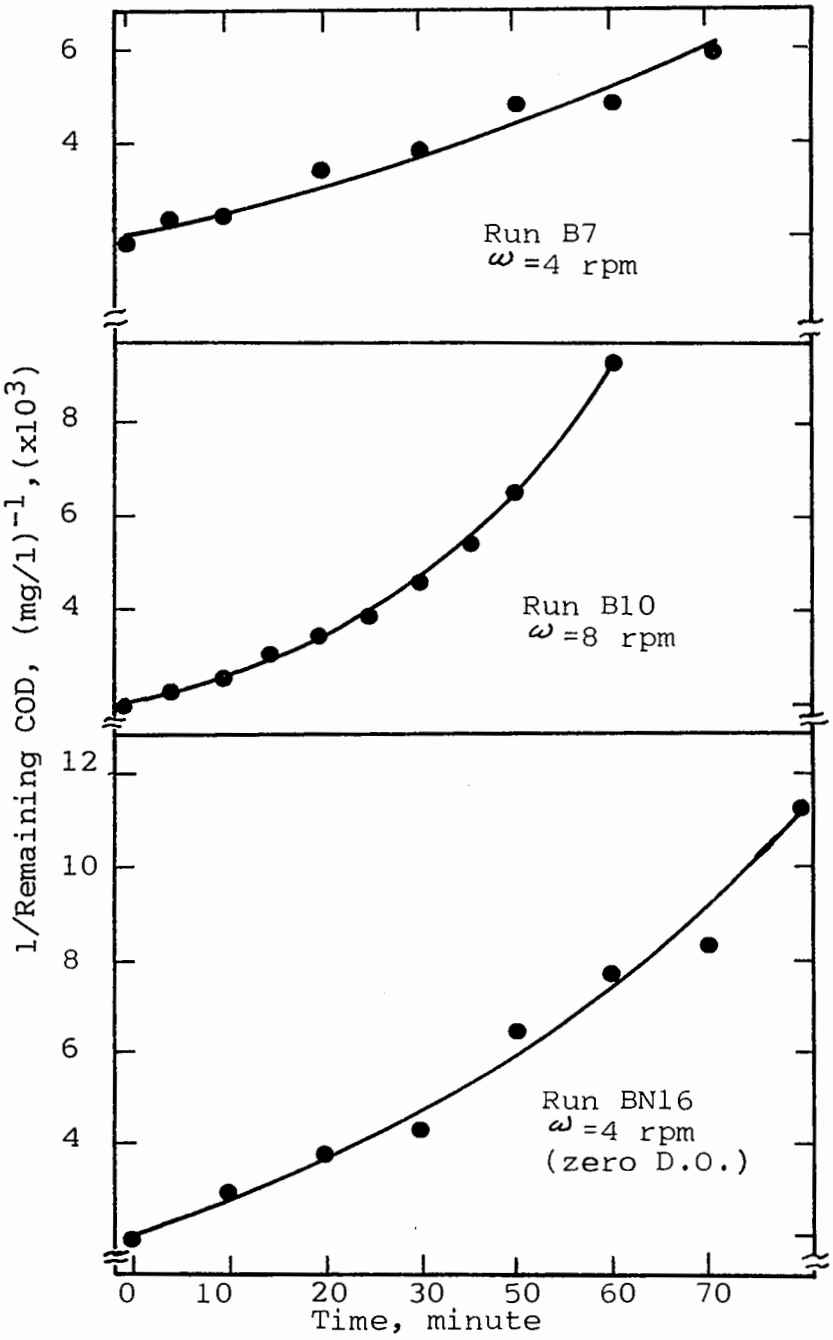


Figure 4c. 1/remaining COD ver. time for second order kinetics model.

observed rate is 1st order regardless of the rotational speed, within the range of speeds tested.

The observed reaction rate constant k_o increased with an increase in $\bar{w}$. Since the data of Table 3 likewise indicates that k_m is substantial which implies that mass transfer is significant, this result should be expected. Indeed, as w increases, the thickness of the diffusion boundary layer (different than Prandtl hydrodynamic boundary layer) should decrease. Levich (25) has demonstrated that the thickness of the diffusion layer, for a disc fully submerged in a liquid and rotating around its own axis is inversely proportional to the square root of $\bar{w}$. This author estimated that this layer thickness is about one tenth of Prandtt layer. As a result, increasing $\bar{w}$ should boost the efficiency of substrate removal but only up to a limit beyond which little improvement would result. From the negative point of view a high $\bar{w}$ would increase the sloughing rate, and would have a detrimental effect on biochemical removal. A representative, empirical equation relating the mass transfer coefficient to $\bar{w}$ has been developed:

$$k_m = 7.87 \quad (\bar{w})^{0.809} \qquad (13)$$

where k_m is in cm/day and $\bar{w}$ in rpm. Fig (5) shows this relationship. It is important to realize that this equation holds only for the media, used in this experiment, that it can not be used for extrapolation beyond $\bar{w}$ = 10 rpm, and finally that it should not be used for scale-up. One should expect that k_m would become independent of $\bar{w}$ beyond a certain value. This equation (or any similarly derived equation) would enable a designer to estimate the effect of media geometry as to whether or not the mass transfer regime would be eliminated at the minimum rotational speeds expected in the prototype assuming other parameters are not affected. Equation (13) does not include the k_m values calculated for the slide test or those from the low initial substrate concentration run for the following reason. The tests run using the slide and its attached biomass, attempt to simulate a condition where turbulent does not exist. Such a test does not have a practical value. However, it helps in showing the maximum transfer resistance. Beside that, it is clear that at 4 rpm the observed mass transfer rate did not show much increase, showing that the mass transfer resistance is still large.

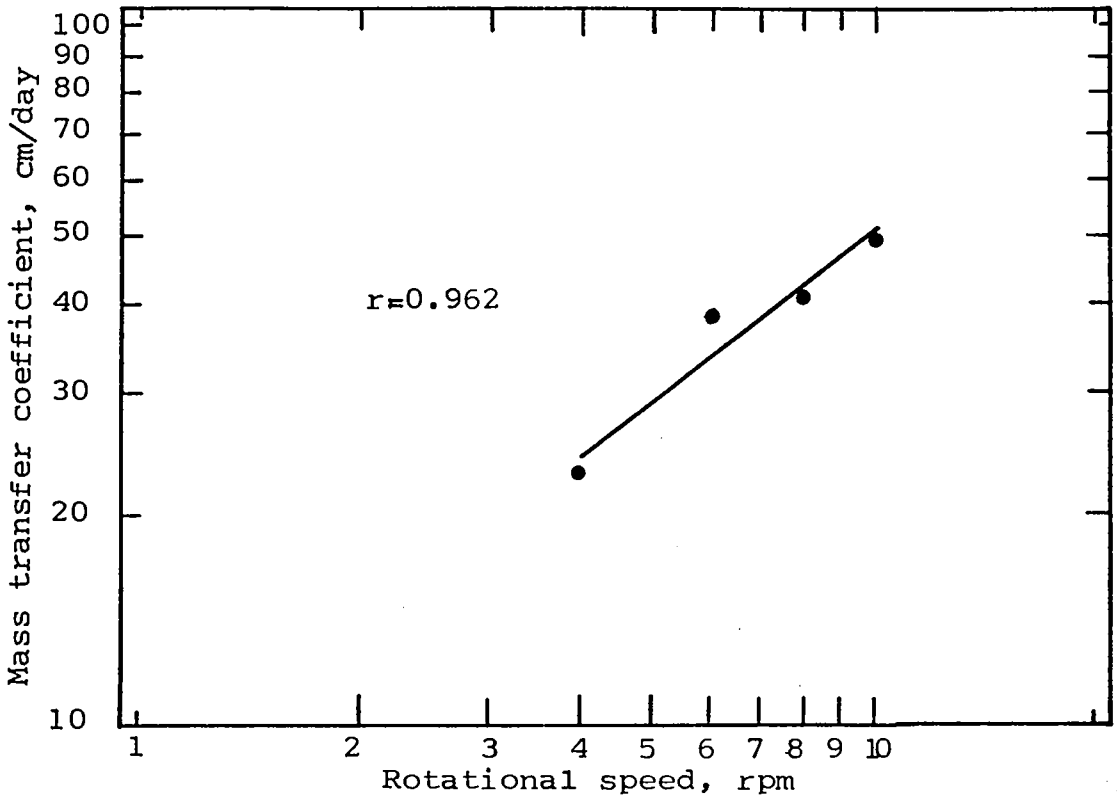


Figure 5. Empirical relationship between the mass transfer coefficient and the rotational speed

Where running the test in the RBC using a batch mode for two different initial substrate concentrations ($\bar{w}$ = 4 rpm, with COD: 120 mg/l and 515 mg/l) and (w = 8 rpm, with COD 76 mg/l and 538 mg/l), one might have expected some differences to be evidenced. In these cases differences were minimal and as far as the rate mechanism was concerned, its 1st order dependence on S_b was maintained. To make this point more clear, it should be recalled that the intrinsic rate is a biokinetic mechanism. Generally, the Monod equation is used:

$$-r_s = kx \frac{S}{K_{sat} + S} \cdot$$

where two extreme cases can be expected:

(1) low S: such that $K_{sat} >> S$

hence: $-r_s = \left(\frac{kx}{K_{sat}}\right) \cdot S$ (pseudo-first order in S)

(2) high S, such that $K_{sat} << S$, then:

$-r_s = kx$, (pseudo zero order in S)

Hence, for the observed rate to be first order in S_b would imply that the intrinsic rate be a pseudo-first order up to higher values of substrate concentration. Indeed experimental data have shown that the intrinsic rate is 1st order in S up to at least S_o = 550 mg/l (soluble COD). This implies that the Monod saturation constant, K_{sat}, has a very large value; that is K_{sat} >550 mg/l measured in COD for the carbon compound (Dextrin $(C_6H_{10}O_5)_n$. Accordingly this biofilm needs a very high substrate concentration (above 550 mg/l as COD) to reach half the maximum specific growth rate ($\hat{\mu}$). This high value for K_{sat} is much larger than those reported in the literature (Grieves, (14).

Fig (6) shows the effect of K_{sat} on μ in the Monod equation. Another point one might expect is that the observed rate constant, k_o, under the same $\bar{w}$ would have the

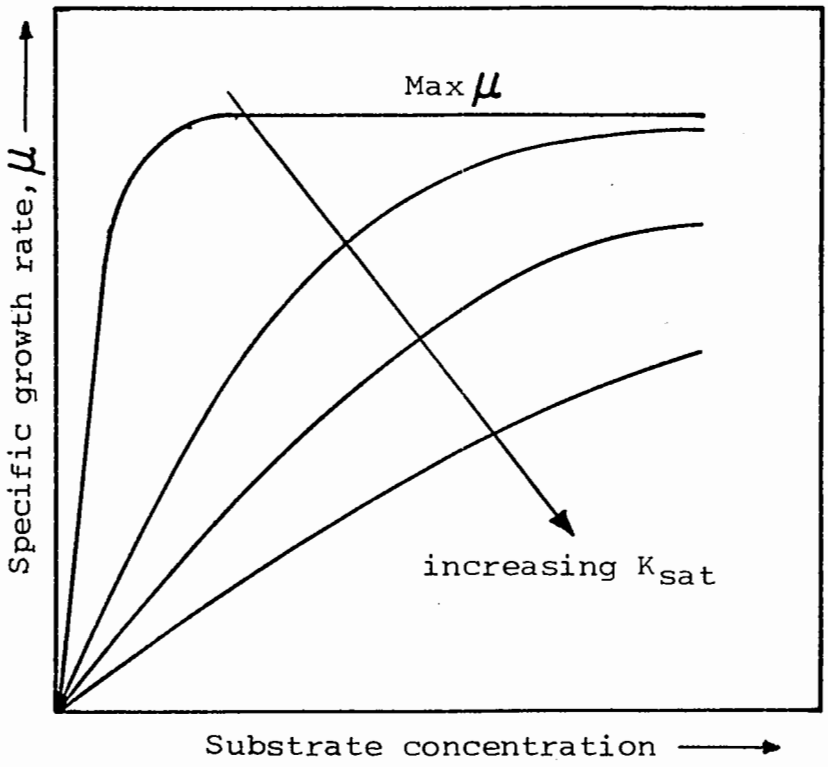


Figure 6. Relationship between the specific growth rate and the substrate concentration showing the effect of Monod saturation constant on Monod equation.

same value for two different initial substrate concentration (e.g., 120 and 516 mg/l COD @ $\bar{w}$ = 4 rpm). But it was not the case. Two possible justifications could be thought of: The first is that Monod saturation constant has changed, but this is very unlikely; the second is that sloughing occurred between these tests (there was 6 days lap time) and the biofilm interface texture has changed hence affecting the mass transfer coefficient.

One striking point was that the dissolved oxygen concentration in the bulk fluid did not affect the substrate removal rate. Even under zero D.O. conditions, the observed reaction constant for the 1st order mechanism was not affected. This contradicts the generally accepted concept in the biological wastewater treatment field that a minimum of 2 mg/l D.O. in the bulk fluid should be maintained for successful operation. This finding implies that the source of oxygen required by the biofilm for the oxidation of substrate is the surrounding atmosphere and that the bulk of the liquid plays little or no part in this two foot model. If this observation could be extrapolated to a prototype plant for BOD removal, it would not have to be operated at a certain rotational speed controlled by the bulk D.O. for a minimum value of 2 mg/l. Rather, $\bar{w}$ should be looked at as the frequency of this system at which any point in the biofilm should be exposed to the atmosphere for optimum removal efficiency under a given substrate strength and loading. Of importance too would be the additional effects of increasing k_m and decreasing the thickness of the diffusion boundary layer. The increase in efficiency gained by enclosing the RBC in an atmosphere enriched with pure oxygen or increasing the air pressure (Torpey et al., (33) and Bintanja et al (7) actually enhances the diffusion rate by increasing the partial pressure and the concentration gradient of the oxygen into the biofilm besides oversaturating the liquid film attached to the biofilm as it emerges from the tank.

CONCLUSIONS

1. Bulk D.O., as low as 0.0 mg/l, did not affect the substrate removal rate in this pilot plant work.

2. The observed rate was first order in the substrate bulk concentration only when the intrinsic rate was also pseudo-first order. Under this case the observed rate k_o is related to the mass transfer coefficient k_m and the surface intrinsic rate as follows $1/k_o = 1/k_m + 1/k_i$, where all values are in unit of length per unit of time.

3. The mass transfer coefficient km was related to the rotational velocity by the empirical formula:

$$k_m = 7.87 \quad (\bar{w})^{0.809} \qquad (13)$$

where km is in cm/sec and $\bar{w}$ in rpm. This is valid strictly for this pilot unit and for $\bar{w}$ values up to 10 rpm.

4. In this case, the Monod saturation constant, K_{sat}, had a larger value than 550 mg/l soluble COD when the carbon source was provided by dextrin $(C_6H_{10}O_5)_n$. This is higher than previously reported values in the literature.

5. It appears that a kinetic study should be done on each stage separately and not on the overall system as a single unit. This latter assumption can result in an error in detecting the rate order.

REFERENCES

1. Antonie, R.L., Kluge, D.L., and J.H. Mielke, "Evaluation of a Rotating Disk Wastewater Treatment Plant, " J. WPCF, 46, 498 (1974).
2. Antonie, R.L., Fixed Biological Surfaces-Wastewater. The Rotating Biological Contactor, CRC Press, Inc., (1976).
3. Atkinson, B., Busch, A.W., and G.S. Dawkins, "Recirculation, Reaction Kinetics, and Effluent Quality in A Trickling Filter Flow Model," J. WPCF, 35, 1307 (1963).
4. Atkinson, B., Swilley, E.L., Busch, A.W., and D.A. Williams, "Kinetics, Mass Transfer, and Organism Growth in a Biological Film Reactor," Trans. Inst. Chem. Engrs., 45, T257 (1967).
5. Bailey, J.E., and D. F. Ollis, Biochemical Engineering Fundamentals, McGraw-Hill Book Company, N.Y.1977.
6. Baillod, C.R., and W.C. Boyle, "Mass Transfer Limitations in Substrate Removal," J. San Engin. Div., ASCE, 96, SA2, 525 (1970).
7. Bintanja, H.H.J., Brunsmann, J.J., and C. Boelhouwer, "The Use of Oxygen in a Rotating Disc Process," Water Res., 10, 561 (1976)
8. Borchardt, J.A., "Biological Waste Treatment Using Rotating Discs," Biotech. & Bioengin Symp. No. 2, 131 (1971)
9. Borchardt, J.A., Kang, S.J., and T. H. Chung, "Nitrification of Secondary Municipal Waste Effluents by Rotating Bio-Discs," EPA-600/2-78-061, June 1978.
10. Carberry, J. J., Chemical and Catalytic Reaction Engineering, McGraw Hill Book Comp., N.Y. 1976.
11. Chesner, W.H., and A.H. Molof, "Biological Rotating Disk Scale-up Design: Dissolved Oxygen Effects," Prog. Water Tech., 9, 811 (1977).
12. Clark, J. H., Moseng, E.M., and T. Asano, "Performance of a Rotating Biological Contractor Under Varying Wastewater Flow," J.WPCF, 50, 896 (1978)
13. Frank Kamenetskii, D.A., Diffusion and Heat Transfer in Chemical Kinetics, Plenum Press, N.Y., 1969
14. Grieves, C.G., Dynamic and Steady-State Models for the Rotating Biological Disc Reactor, Ph.D., dissertation, Clemson University, 1972

15. Harremoes, P., "The Significance of Pore Diffusion to Filter Denitrification," J. WPCF, 48, 377 (1976).
16. Harremoes, P, "Biofilm Kinetics, in Water Pollution Microbiology, Vol. 2, edited by P. Mitchell, J. Wiley 1978.
17. Hartmann, H., "Untersuchung Liber die Biologische Reinigung von Abwasser mit Hilfe von Tauchtropfkorperanlagen" (Investigation of the Biological Clarification of Wastewater Using Immersion Drip Filters). Band 9 der Stuttgarter Berichte zur Siedlungswasserwirtschaft Kommissionsverlag Munich R. Oldenbourg 1960.
18. Hitdlebaugh, J.A., and R.D. Miller, "Operational Problems with Rotating Biological Contactors", J. WPCF, 53, 1283 (1981)
19. Hoehn, R.C., and A. D. Ray, "Effects of Thickness on Bacterial Film, " J. WPCF, 45, 2302 (1973)
20. Horvath, C., and J.M. Engasser, "External and Internal Diffusion in Heterogeneous Enzyme Systems", Biotech. and Bioengin, 16, 909(1974).
21. Kornegay, B.H., and J.F. Andrews, "Kinetics of Fixed Film Biological Reactors", 22nd Ind. Waste Conf., 1967, Part 2, Pudue U., 620 (1967)
22. Kornegay, B. H., "Modeling and Simulation of Fixed Film Biological Reactors for Carbonaceous Waste Treatment", in: Mathematical Modeling for Water Pollution Control Processes", Edited by T.M. Keinath and M.P. Wanieliste Ann Arbor Science, Publisher, Ann Arbor, MI (1975)
23. LaMotta, E.J., Evaluation of Diffusional Resistances in Substrate Utilization By Biological Films", Ph.D. Dissertation, University of North Carolina, Chapel Hill, 1974
24. LaMotta, E.J., "External Mass Transfer in A Biological Film Reactor", Biotech. and Bioengin, 18, 1359 (1976)
25. Levich, V.G., Physicochemical Hydrodynamics, Prentice-Hall, Inc., 1962.
26. Maier, W.J., Behn, V.C., and C.D. Gates, "Simulation of the Trickling Filter Process", J. San Engin, Div., ASCE, 93, SA6, 91 (1967)

27. Mikula, W.J., Reynolds, J. H., George, D.B., Porcella, D. B., and E. J. Middlebrooks, "A Kinetic Model for the Treatment of Cheese Processing Wastewater with A Rotating Biological Contactor", in Proc. 1st Nat'l Symp/Workshop on Rotating Biological Contractor Technology, 491 (1980). Edited by E.D. Miller, et. al.
28. Pano, A., Reynolds, J.H., and E.J. Middlebrooks, "The Kinetics of a Rotating Biological Contactor Treating Domestic Sewage", in: Proc. 1st Nat'l Symp/ Workshop on Rotating Biological Contactor Tech., 449 (1980) edited by E.D. Miller, et al.
29. Popel, F., "Leistung, Berechnungund Gestaltung von Tauchtropfkorperanlagen "(Estimating, Construction, and Output of Immersion Drip Filter Plants), Stuttgarter Berichte zur Siedlungswasserwirtschaft, 11, Kommissionverlag R. Oldenbourg, Munchen 1964.
30. Stover, E.L., and D.F. Kincannon, "Evaluating Rotating Biological Contactor Performance," Water & Sew. Works, 123, 88 (1976)
31. Tomlinson, T. G., and D.H.M. Snaddon, "Biological Oxidation of Sewage by Films of Microorganisms," Air and Water Pollut. Int. J., 10, 865 (1966).
32. Torpey, W.N., Heukelekian, H., Kaplovsky, A.J., and R. Epstein, "Rotating Disks with Biological Growths Prepare Wastewater for Disposal or Reuse", J. WPCF, 43 2181 (1971)
33. Torpey, W., Heukelekian, H., Kaplovsky, A.J., and L. Epstein, "Effects of Exposing Slimes on Rotating Discs to Atmospheres Enriched with Oxygen", in Adv. in Water Poll. Res., 405 (1972) Edited by S.H. Jenkins.
34. Welch, F.M., "Preliminary Results of a New Approach in the Aerobic Biological Treatment of Highly Concentrated Wastes", 23rd Indus. Waste Conf., 428 (1968) Purdue U.
35. Williamson, K.J., "The Kinetics of Substrate Utilization by Bacterial Films", Ph.D. Thesis, Stanford U., 1973.
36. WPCF, "Operation of Wastewater Treatment Plants", Manual of Practice No. 11, WPCF, 1976

NOMENCLATURE

Symbol	Definition	Unit
A	Disc or biofilm area	L^2
a	Ratio of disc area to bulk liquid volume	L^{-1}
a_f	as a above but for slide	L^{-1}
a_s	as a above but suspended film	L^{-1}
Da	Damköhler number	--
$\overline{Da}$	Observable Damköhler number	--
k_i	intrinsic rate constant (1st order)	LT^{-1}
k_m	Substrate mass transfer coefficient (liquid film)	LT^{-1}
k_o	Observed rate constant (1st order)	LT^{-1}
ke,o	k_o as measured from plot in S_b ver t; $ke,_o = ko \cdot a$	T^{-1}
ke,i	k_i as measured from plot in S_b vert; $ke,i = k_i \cdot a_s$	T^{-1}
k	Maximum specific substrate utilization rate in Monod equation	T^{-1}
k_{max}	Maximum reaction rate in Monod equation; $k_{max} = k.x$	$ML^{-2}T^{-1}$
K_{sat}	Monod saturation constant	ML^{-3}
R	Overall reaction rate	$ML^{-2}T^{-1}$
r_s	Substrate removal rate	$ML^{-2}T^{-1}$
S_b	Limiting Substrate concentration in bulk fluid	ML^{-3}
S_s	Limiting substrate concentrate at biofilm interface	ML^{-3}

Symbol	Definition	Unit
s	Ratio of S_s to S_b	--
V	Liquid volume in RBC reactor	L^3
X	Biofilm biomass density as TS or TVS	ML^{-2}
Y	Yield coefficient in Monod equation	--
η_e	External effectiveness factor	--
$\hat{\mu}$	Maximum specific growth rate in Monod equation	T^{-1}
$\bar{w}$	Disc rotational speed	rpm

ANALYSIS OF STEADY STATE SUBSTRATE REMOVAL MODELS FOR THE RBC

David E. Schafer. Camp, Dresser and McKee, Inc., Boston, Massachusetts.

James C. O'Shaughnessy. Department of Civil Engineering, Northeastern University, Boston, Massachusetts.

Frederic C. Blanc. Department of Civil Engineering, Northeastern University, Boston, Massachusetts.

INTRODUCTION

This paper evaluates three independently-derived steady state mechanistic substrate removal models for the rotating biological contactor (RBC), intended for use by design engineers. The three models evaluated are: 1) Kornegay's steady state model for carbonaceous waste treatment; 2) Schroeder's steady state RBC design equation; and 3) Grieve's Pseudo-homogeneous steady state model. Utilizing a common data base, each has been assessed with respect to model calibration, adequacy of fit, relative influence exerted by various design parameters, and limitations and restrictions.

MATHEMATICAL MODELING OF THE RBC

Several empirical models have been developed to predict the steady state substrate removal in RBC units (1,2,3). These models express stage-by-stage removal of substrate as a power function of major design variables such as hydraulic loading rate, influent concentration, retention time, surface area,

temperature, disc configuration and rotational speed.

To improve RBC process modeling, current research efforts are being focused upon the development of a mechanistic or deterministic model for substrate removal. Although proposed mechanistic models for wastewater treatment processes also possess empirical qualities, a "true" mechanistic model is defined as one which assists understanding and allows useful, though not necessarily exact, extrapolation over a wide range of operating conditions (4).

Mechanistic modeling of substrate uptake and cell growth in biological systems is highly complex. Even in the simplest biological reaction, a multiplicity of cellular reaction mechanisms take place. Adsorption, enzyme catalysis and diffusional processes represent major functional mechanisms which can control the uptake of a specific substrate (4).

BIOFILMS

Biologically, each RBC consists of a complex interrelated population of predominantly heterotrophic attached microorganisms. In general, this attached microogranism population will be comprised of aerobic, facultative and anaerobic bacteria (5). In addition, as indicated by Kornegay (6), a significant population of suspended microorganisms may also be present if the system is operated at a long hydraulic retention time.

With respect to substrate removal, the concept of an "active" microbial depth has been adopted by several investigators (6,7,8,9,10 and 11). This hypothesis divides the total microbial film thickness into two layers. The outermost layer, being in direct contact with the adhered liquid film, is termed the active layer. The "inactive layer", if present, is in direct contact with the support media.

Sanders (9) evaluated active depth in terms of the "critical" depth at which diffusion of oxygen within the slime layer becomes limiting. Tomlinson and Snaddon (10) have also suggested that the active layer consists of the aerobic microorganism zone. However, Atkinson and Davies (11), Kornegay (6) and Grieves (7) contend that the active depth should be defined with respect to the depth of penetration of a limiting nutrient.

Estimated values of active microfilm depths have ranged from 27 to 200μm. To date, no universally acceptable technique exists for the measurement of active depth in any fixed film system. Conceptually, substrate removal from the bulk liquid phase requires diffusion of metabolic reactants into the attached biofilm, metabolism by the organisms, and diffusion of

the metabolic by-products back through the biofilm and into either the bulk liquid or the atmosphere. Since relatively thick biofilms are employed, significant concentration gradients, resulting from mass transport resistances, can exist between the bulk liquid and the active microbial layer (12).

TREATMENT KINETICS

In 1950, Monod presented an initial mathematical analysis for cell growth based upon work with batch reactors (13). His hypothesis assumes that microorganism growth rate is dependent upon the concentration of a limiting substrate, which he tested using a completely-mixed continuous flow chemostate containing a dispersed culture of microorganisms. The versatility of the Monod kinetic relationships in fitting data normally obtained from a variety of wastewater treatment systems has made it a logical starting point for modeling the RBC process (6,7,14 and others).

STEADY STATE MODELS FOR THE RBC

Kornegay's Model

The mathematical algorithms proposed by Kornegay to simulate RBC system performance have been developed under the assumption that ultimate substrate removal is dependent upon microbial growth and that the entire mass of attached film is not considered active in the removal of organics. Additional assumptions are as follows (6):

1) complete mixing is achieved in the liquid volume;
2) organism decay is neglected;
3) maintenance energy is not included in explicit terms;
and 4) saturation or Monod function coefficients are assumed to remain constant during periods of dynamic operation.

Kornegay's approach to system performance under continuous flow conditions is illustrated in Figure 1 and expressed by the following steady state equation (6):

$$FC_o - FC_b = \frac{\mu_{max} 2N\pi(r1^2 - r2^2)Xd(C_b)}{Y(K_c + C_b)} \qquad (1)$$

where: μ_{max} is maximum specific growth rate of fixed film organisms; K_c is half saturation constant; Y is the apparent yield of fixed film organisms; F is influent flow rate; C_o is

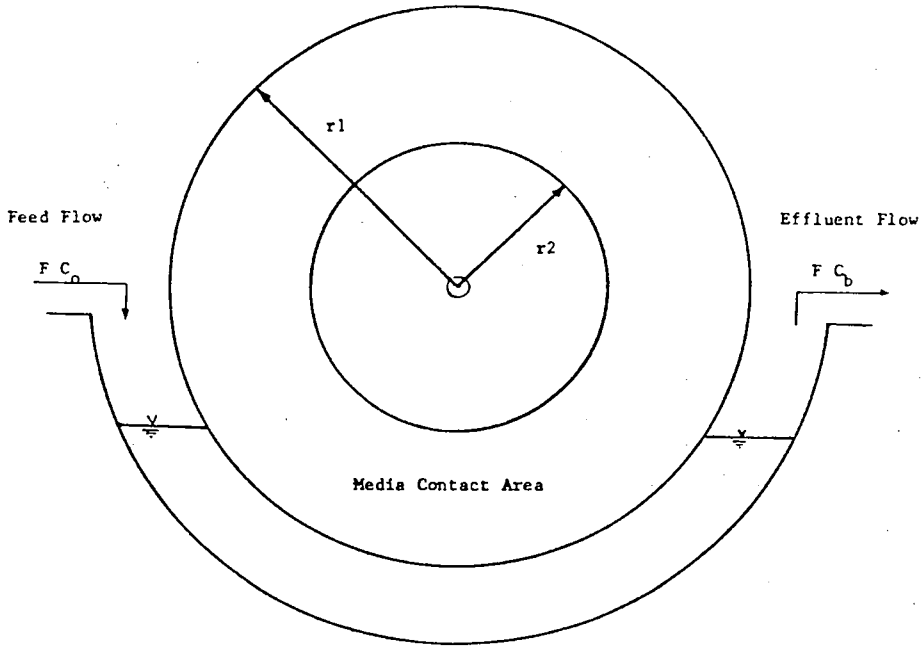


FIGURE 1: Definition Sketch for Kornegay's Substrate Removal Models

influent substrate concentration; C_b is reactor substrate concentration; N is the number of discs; r1 is total disc radius; r2 is unsubmerged disc radius; X is unit mass of the fixed microbial film; and d is active film depth.

Setting the area capacity constant, P, to $\mu_{max}Xd/Y$ and the contact area, A, to $2N\pi(r1^2 - r2^2)$, the removal equation for a single stage RBC in which suspended growth is negligible becomes (6):

$$F(C_o - C_b) = PA \frac{(C_b)}{(K_c + C_b)} \qquad (2)$$

Multi-stage operation can be evaluated by setting the influent concentration of the second reactor equal to the bulk liquid (i.e., effluent) concentration of the first reactor and performing successive iterations until all reactor concentrations are known.

Schroeder's Model

Schroeder's steady state design model for the RBC process is based upon a theoretical analysis of substrate utilization by microbial films conducted by Atkinson and his coworkers during the late 1960's and early 1970's. Schroeder has modified the Atkinson Model for use in municipal wastewater treatment applications, incorporating the following assumptions (14):

1) slime phase diffusion controls overall system performance;
2) no significant concentration gradients exist within the adhered liquid film while in the bulk gas phase;
3) mass transport through a differential element follows Fick's Law of Diffusion; and
4) A plug flow mode of operation is appropriate in modeling the RBC process.

Schroeder's approach to system performance under continuous flow conditions is expressed by the following steady state equation (14):

$$K\left(\frac{1}{C_b}-\frac{1}{C_o}\right) + \ln\frac{C_o}{C_b} = \frac{f\ K^*\ A_s\ \theta\ d}{V_L} \qquad (3)$$

where: K is the half saturation constant; C_b is bulk liquid substrate concentraton; C_o is influent substrate concentration; f is the proportionality factor; K* is the maximum specific growth rate; A_s is submerged disc area; θ is reactor hydraulic retention time; d is active biofilm depth; and V_L is liquid volume per disc.

Multi-stage operation is evaluated using a technique similar to the Kornegay approach.

Grieves' Model

Grieves combined both biological growth kinetics and mass transport resistances in the development of a dynamic substrate removal model for the RBC. Grieves adopted a more complex physical representation for the system than those developed by Kornegay or Schroeder by subdividing each disc into pie-shaped segments as detailed in Figure 2. Within each segment, substrate is assumed to be transferred across a biofilm-liquid film interface at a rate directly proportional to the concentration gradient between the two phases. Major model assumptions are as follows (7):

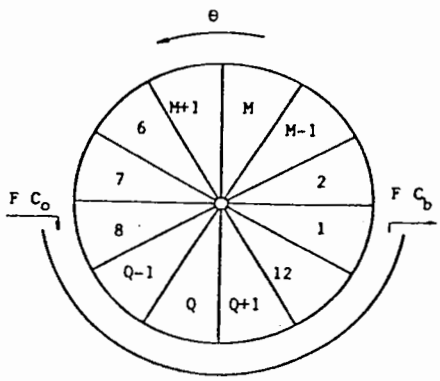


FIGURE 2: Grieves' Model-Details of the Elements Taken Around the Disc

1) there is complete mixing in the reactor, biofilm and liquid film;
2) as the disc leaves the bulk liquid phase, a stationary liquid film having an initial substrate concentration equal to that of the bulk liquid phase adheres to the microbial film;
3) substrate utilization by an individual microorganism in the biological film can be represented by a saturation or Monod function;
4) Monod coefficients are assumed to remain constant during periods of transient operation;
5) the mass of substrate consumed by the organisms for maintenance purposes is negligible when compared with that used for growth; and
6) substrate diffusion in the radial and circumferential directions is negligible when compared with diffusion into the biological film.

In the derivation of this model, Grieves has considered mass balances on substrate in: 1) any segment of the liquid film exposed on the disc; 2) the biological film exposed on the disc; 3) any segment of the biological film submerged in the bulk liquid; and 4) the reactor bulk liquid. With the additional assumption that first order removal kinetics are appropriate when substrate concentrations are relatively low, Grieves' steady state model takes the form (7):

$$\frac{C_b}{C_o} = \frac{1}{1 + \frac{2N}{F}\left[(P_1')(A_s) + F_f\left[1 - e^{-\{(P_1')(A_a)/F_f\}}\right]\right]} \qquad (4)$$

where: C_b is bulk liquid substrate concentration; C_o is influent substrate concentration; N is number of discs per reactor; F is influent flow rate; P_1' is $(K_L)(K_1')/(1+K_1')$; A_a is area of the disc in the air; F_f is liquid film flow rate; A_s is submerged disc area; K_1' is $\{(\mu_{max})(TF)^{n-1}(X)(\Delta z)\}/\{(Y)(K_c)(\eta)(K_L)\}$; TF is treatability; N is stage number; X is organism density in segment L,M: Y is organism yield coefficient; μ_{max} is maximum specific growth rate; K_c is saturation constant in Monod equation; η is the effectiveness factor; K_L is mass transfer coefficient; and Δz is active biofilm thickness.

PILOT PLANT OPERATION

The three steady state models were compared using simulation results and data collected in a pilot plant investigation conducted at the Yankee Greyhound Racing, Inc. dog track located in Seabrook, New Hampshire. Septic tank effluent characteristics (used as influent feed to the RBC unit) measured during the 60-day pilot plant study are summarized in Table I and listed in Tables II and III (15). High nitrogen and total and soluble BOD values indicate a waste strength roughly three times that of normal domestic sewage.

Table I

Septic Tank Effluent Characteristics Pilot Plant Influent Feed, Throughout the Testing Period (15)

Parameter	Range**
BOD_5	250 - 600
COD	350 - 750
Suspended Solids	50 - 200
NH_3-N	100 - 200
Organic-N	50 - 100
NO_3-N	<1.0
PO_4^{-3}	10 - 20
Grease and Oil	50 - 200
Alkalinity	250 - 500
pH	6 - 8

**All values except pH in mg/l.

Table II

Yankee Greyhound Inc. Dog Track Pilot Plant Data Summary

Run	Date	Temp. °C	Flow Rate*	Influent Conc.**	Stage 1**	Stage 2**	Stage 3**	Stage 4**
1	5-12	12.0	0.50	212	84	29	16	14
2	5-14	14.5	0.50	288	155	80	38	28
3	5-19	13.0	0.67	275	133	50	43	13
4	5-22	11.5	1.00	223	123	53	19	16
5	5-25	9.5	0.93	265	118	62	27	22
6	5-27	12.0	0.80	375	126	74	30	25
7	5-30	16.0	0.40	235	68	37	28	27
8	5-31	17.5	0.40	375	110	138	65	39
9	6- 2	14.0	0.37	515	143	83	45	37
10	6- 4	12.5	0.40	250	68	42	11	19
11	6- 6	12.5	1.82	305	102	59	36	12
12	6- 9	17.0	1.30	395	256	182	110	58
13	6-11	18.8	0.26	470	88	20	16	9
14	6-15	18.0	0.25	400	89	39	21	18
15	6-17	20.0	0.40	535	188	68	24	24
16	6-19	19.0	0.15	365	39	45	29	8
17	6-21	--	0.15	420	60	45	24	22
18	6-23	--	0.15	455	86	18	29	22

*gal/min
**mg/l

Table III

Pilot Study Process Loading Factors and Removal Efficiencies

Run	First Stage Loading Factors			Overall Unit Loading Factors		
	Hydraulic Loading*	Organic Loading**	% BOD Removal	Hydraulic Loading*	Organic Loading**	% BOD Removal
1	1.80	3.183	60	0.45	0.796	93
2	1.80	4.323	46	0.45	1.081	90
3	2.41	5.532	52	0.60	1.383	95
4	3.60	6.695	45	0.90	1.674	93
5	3.35	7.399	55	0.84	1.850	92
6	2.88	9.007	66	0.72	2.252	93
7	1.44	2.822	71	0.36	0.706	89
8	1.44	4.504	71	0.36	1.126	90
9	1.33	5.721	72	0.33	1.430	93
10	1.44	3.002	73	0.36	0.751	92
11	6.55	16.666	67	1.64	4.167	96
12	4.68	15.417	35	1.17	3.854	85
13	0.94	3.669	81	0.23	0.917	98
14	0.90	3.002	78	0.23	0.751	96
15	1.44	6.425	65	0.36	1.606	96
16	0.54	1.644	89	0.14	0.411	93
17	0.54	1.892	86	0.14	0.473	98
18	0.54	2.049	81	0.14	0.512	95

*gal/day-sq.ft.
**lb. sol. BOD/day-1000 sq. ft.

The pilot plant utilized was a 4-foot, 4 equal stage rotating biological contactor supplied by the Environmental Pollution Control Division of the George A. Hormel Company (EPCO-HORMEL) of Austin, Minnesota. This unit provided a total of 1600 square feet of polyethylene surface area for biomass growth and a liquid volume in the unit of approximately 100 gallons. This unit was fed continuously by a small submersible pump suspended between the floating scum layer and the bottom sludge deposit in one of the secondary septic tanks. The flow during the study ranged from 0.15 to 1.82 gallons per minute. During the course of the study, the pilot unit was operated at 4-stage detention times which ranged from 57 minutes to 695 minutes and overall organic loading rates ranging from 0.41 to 4.17 pounds of soluble 5 day BOD per day per 1000 square feet (15).

CALIBRATIONS/RESULTS/SENSITIVITY ANALYSES

Kornegay Model

Calibration of Kornegay's steady state substrate removal model involved the evaluation of two unknown kinetic parameters: K_c and P. These values are idealy developed by curve fitting of actual data. The rearrangement of Equation 2 as follows:

$$\frac{A}{(F)(C_o-C_b)} = \frac{K_c}{P}\frac{1}{C_b} + \frac{1}{P} \qquad (5)$$

should plot as a straight line having a slope of K_c/P and intercept $1/P$ when $1/C_b$ is plotted against the term on the left side of the relationship (6).

In the pilot plant study, approximately 60-80% of the total BOD reduction occurred in the initial stages of treatment. Estimates for the unknown parameters were therefore established based upon a least squares analysis of the first and second stage data, presented in Figure 3. An initial estimate for the area capacity constant, equal to the inverse of the Y-axis intercept, is 5.62 lb. BOD/day-1000 ft^2 (a value roughly equivalent to the mean first stage organic loading of 5.72 lb. BOD/day-1000 ft^2 applied during the waste treatability study). From the slope of the line, an initial estimate for the saturation coefficient was determined to be equal to approximately 150 mg/l.

Simulation results were obtained through rearrangement of Equation 5 into its quadratic form with respect to C_b such that:

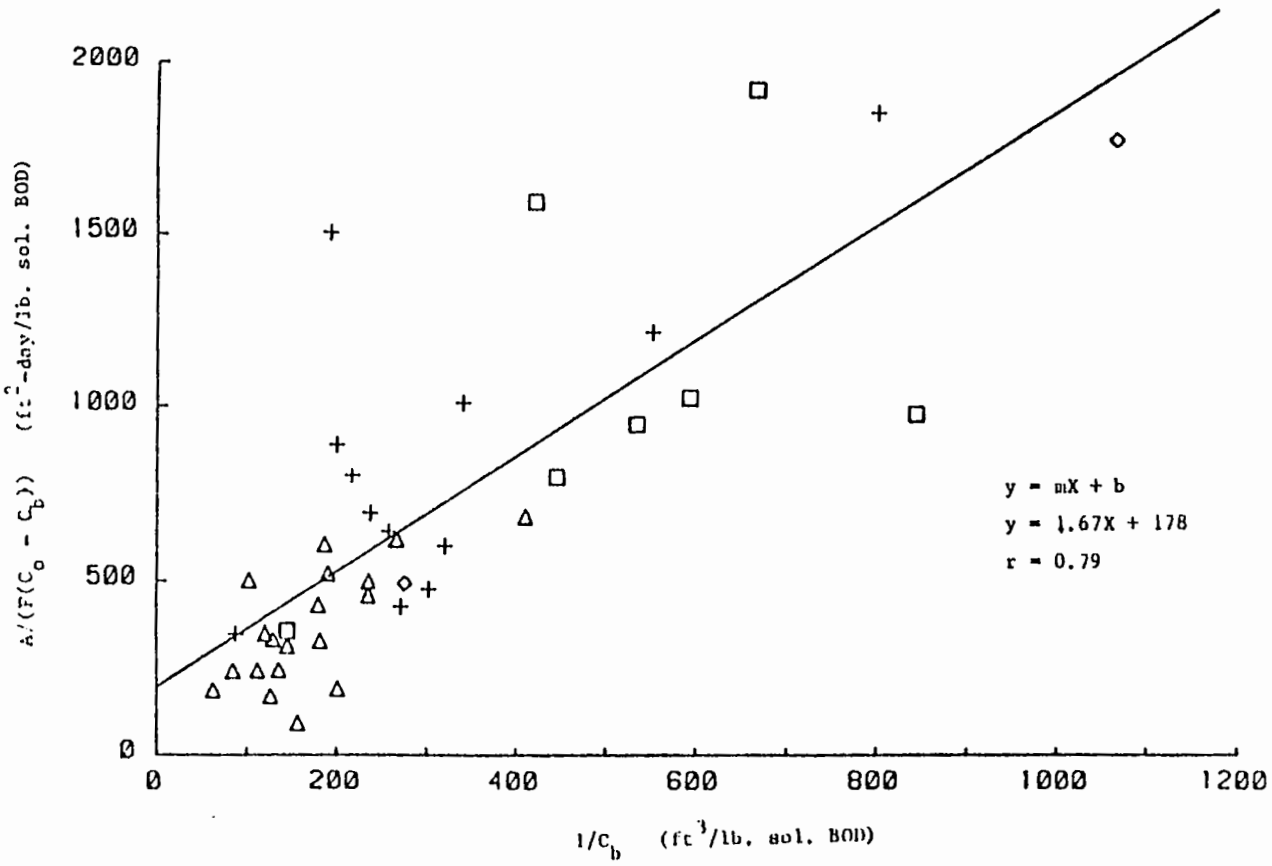


FIGURE 3: Kornegay's Model-Final Plot for Biological Parameter Estimation

$$C_b^2 + \frac{(A)(P) + (F)(K_c) - (F)(C_o)}{(F)} (C_b) - (K_c)(C_o) = 0 \qquad (6)$$

Equation 6 permitted easy calculation of successively-staged effluent concentrations for each reactor given an influent waste strength and flow rate.

In order to test the model's adequacy of fit, a statistical least sum of squared error analysis was selected to serve as the basis for final parameter calibration. Utilizing this approach, simulation results were equally weighted for each stage of treatment. However, because of the relative magnitude of the initial stage bulk liquid concentrations, numerical emphasis was focused upon the initial stages of treatment where the majority of organic removal occurred.

For the initial parameter estimates obtained from Kornegay's steady state model, the total sum of squared error for the 72 simulated values (4 stages on 18 testing dates) was relatively large, equaling 119,564. However, 44% of the total error resulted from poor simulation of Run 11, attributed to the model's pronounced response to flow rate variations.

Figure 4 depicts the model's sensitivity to biological parameter estimates with respect to single stage removal efficiency. This figure served as a guide for additional calibrations as it indicated the relative importance of each parameter within the given range of waste strengths and flow rates. By definition, model sensitivity with respect to the saturation constant varies with reactor bulk liquid concentrations, particularly with series treatment applications. However, as the magnitude of the slope of each line indicates model results are slightly more dependent upon the value of the area capacity constant, P.

Utilizing Figure 4, the initial parameter estimates were systematically adjusted in an effort to improve overall model simulation. By increasing the area capacity constant from its initial value of 5.62 to 6.50 lb. BOD/day-1000 ft^2 and decreasing the saturation constant in the Monod relationship from 150 to 135 mg/l, overall simulation results improved approximately 22%, with a total sum of squared error equal to 92,887. Again, poor simulation of Run 11 resulted in the generation of 39.5% of the total squared error.

Figures 5 and 6 illustrate single stage model response or sensitivity with respect to variations in organic loading resulting from an increase in either influent waste strength or flow rate. Single stage reactor response was selected for illustration because of the dampening effects produced by multi-

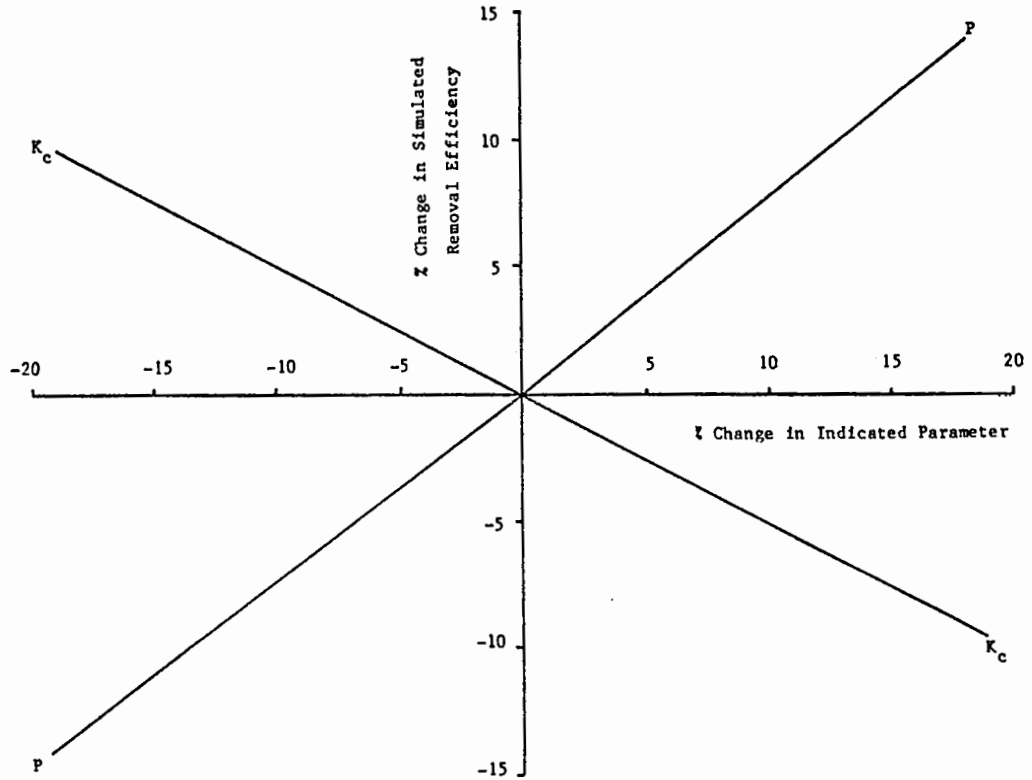


FIGURE 4: Kornegay Model-Biological Parameter Sensitivity

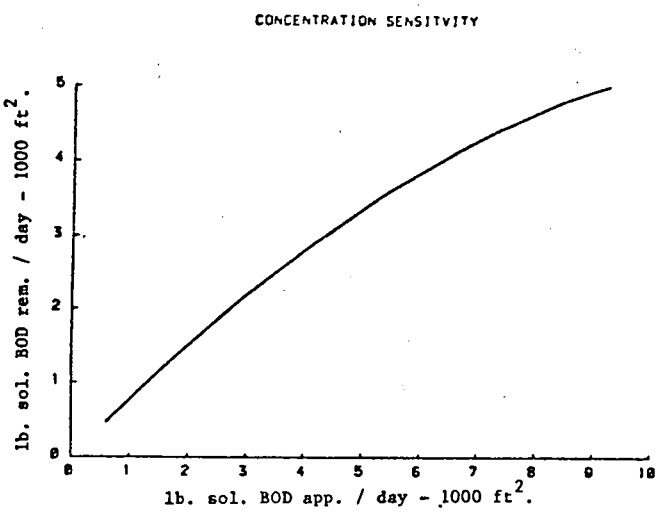


FIGURE 5: Kornegay Model-Single Stage Reactor Response at Constant Flow Rate

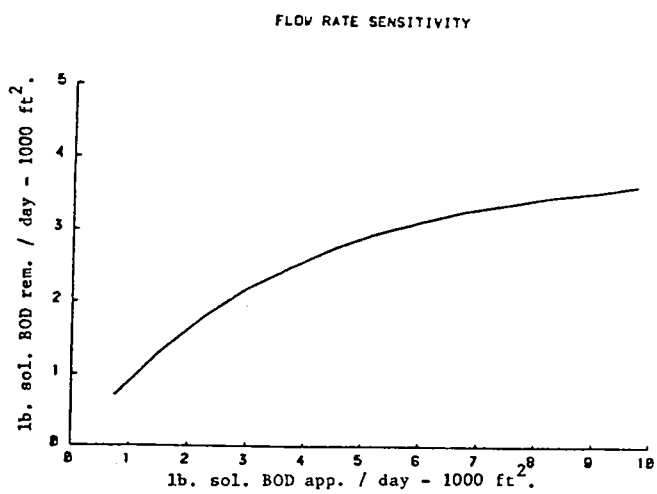


FIGURE 6: Kornegay Model-Single Stage Reactor Response at Constant Waste Strength

stage simulation.

Figure 5 was generated by utilizing the final values obtained at a constant flow rate of 0.40 gal/min. As shown, calculated first stage organic removal efficiency remained essentially constant below the recommended upper limit on organic loading of 5 lb. BOD/day-1000 ft^2. Above this value, predicted removal efficiency decreased due to the fixed nature assumed for the saturation coefficient.

Figure 6 indicates the model's response to flow rate variations and was generated using a constant influent waste concentration of 250 mg/l. Above an organic loading of 3.5 lb. BOD/day-1000 ft^2 (i.e., first stage hydraulic loading in excess of 1.5 gal/day-ft^2), first stage simulated removal efficiency decreased from approximately 70 to 35% as first stage organic loading approaced 10 lb. BOD/day-1000 ft^2 (i.e., as first stage hydraulic loading approached 4.7 gal/day-ft^2).

Schroeder Model

Calibration of Schroeder's steady state simulation model (Equation 3) required the evaluation of four unknown parameters: K, the saturation coefficient in the Monod equation; K*, the maximum removal rate constant; f, a proportionality factor; and d, the active biofilm depth. F, K* and d appear together in the group YK = (f)(K*)*d) and the reactor retention time, θ, is equal to the reactor volume, V, divided by the flow rate, F. Therefore, a modified relationship expressing steady state system performance can be stated as:

$$K\left(\frac{1}{C_b} - \frac{1}{C_o}\right) + \ln \frac{C_o}{C_b} = \frac{(YK)(A_s)(V)}{(V_L)(F)} \qquad (7)$$

However, unlike Kornegay's two-parameter relationship, Equation 7 does not permit rapid parameter estimation through standard curve-fitting techniques. Therefore, it was necessary to establish initial values for the unknown parameters from the available estimates contained in the relevant literature.

Assuming that each stage of the four stage unit acts as an independent reactor, sensitivity runs were performed by solving for YK in Equation 7 using measured BOD and flow rate values as well as values for the remaining system variables. By varying the value of the saturation coefficient throughout its recommended range, a new range for the variable YK was determined and extended from 0.00025 to 0.0005. Similarly, an appropriate

range in values for Schroeder's saturation coefficient was evaluated and found to lie between 5.0 and 10.0 mg/l.

Single stage results from this analysis are illustrated in Figure 7. Although restrictions similar to those indicated for Figure 4 apply, model calibration was shown to be essentially independent of the value for the saturation coefficient, K. By setting K equal to its estimated mid-range value of 7.5 mg/l, final calibration was accomplished by altering the value of YK in order to obtain the best fit for the measured data. Based upon a least sum of squared error calculation utilizing the measured BOD results, a calibrated value for YK in Equation 7 was determined to be equal to 0.0004 cm/sec.

The final results of model calibration had a total sum of squared error for the 72 simulated values equal to 112,036, roughly 47% of which resulted, again, from poor simulation of of Run 11. As with Kornegay's model, this was attributed to a pronounced response by Schroeder's model to high influent flow rate values. Nevertheless, overall first stage simulation results were essentially good, demonstrating the model's capability to predict single reactor removal efficiency over a wide range of organic loadings.

The model does, however, tend to over-estimate organic removal during the third and fourth stages of series treatment applications. This is particularly evident for runs in which the unit flow rate dropped below 0.50 gal/min; unfortunately, the theoretical formulation of this model does not permit the incorporation of a treatability factor, thereby eliminating any means to attempt mathematical correction.

To assess the model's sensitivity to variations in flow rate and influent waste strength, Figures 8 and 9 were developed. Again, single stage reactor response was selected to best illustrate model sensitivity. The linear relationship depicted in Figure 8 was generated at a constant flow rate of 0.40 gal/min using the calibrated parameter values and indicates that predicted removal efficiency is first order with respect to influent concentration.

Figure 9 indicates simulation results generated by an increase in organic loading caused by an increase in feed flow, and suggests that an upper limit on removal efficiency was approached at a hydraulic loading of 2.5 gal/day/ft^2 and that this limit may be increased by increasing reactor retention time.

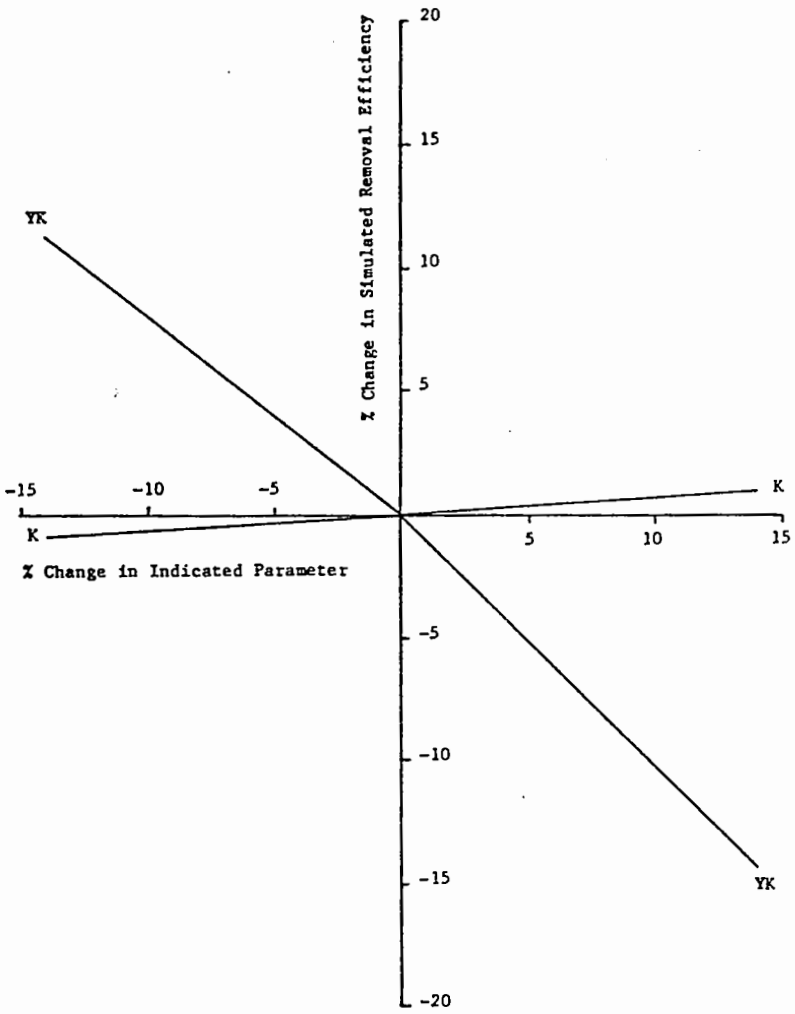


FIGURE 7: Schroeder Model-Biological Parameter Sensitivity

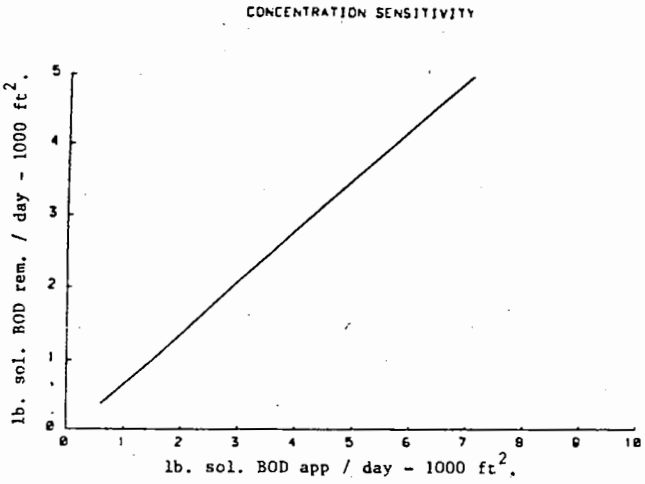


FIGURE 8: Schroeder Model-Single Stage Reactor Response at Constant Flow Rate

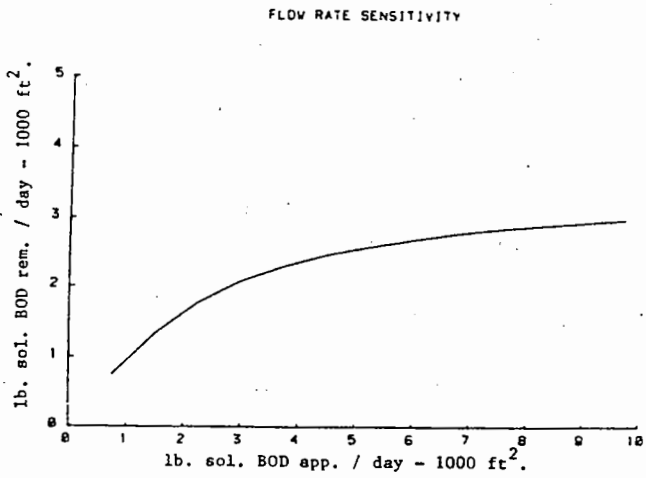


FIGURE 9: Schroeder Model-Single Stage Reactor Response at Constant Waste Strength

Grieves Model

Grieves has indicated that initial estimates for the biological model parameters of Equation 4 can be assumed to fall within the following ranges (7):

Parameter	Range	Units
K_L	0.003 - 1.00	cm/sec
μ	0.02 - 0.54	1/hr
Y	0.26 - 0.64	---
K_c	4 - 10	mg/1
X	8 - 20	mg/ml
n	1 - 15	---
Δz	50 - 200	μm

However, the calculated feasible range in values for the unknown parameter, P_1, varies by several orders of magnitude. The selection of individual parameter values would thus prove to be meaningless, especially since no attempt was made to test for these parameters. Therefore, based upon the measured results and physical characteristics of the unit, the calibration procedure was initially directed towards establishing numerically feasible ranges for both unknown model parameters, P_1 and F_f.

The term F_f was evaluated based upon theoretical and experimental analyses conducted by Zeevalink et. al. (16), Binjanta et. al. (17) and Levich (18). In these investigations, relationships were developed between disc rotational velocity and liquid film thickness. Given a disc peripheral velocity of 1 ft/sec, an appropriate range in values for the term F_f was found to extend from 6.0 to 8.0 cm^3/sec/disc face.

Through substitution of the mid-range value of F_f into Equation 4, individual P_1 values were calculated utilizing a trial and error approach, obtaining rapid convergence for this function through parameter modification via the Newton-Rapson method. An appropriate range of P_1 was found to extend from 0.25×10^{-4} to 1.75×10^{-4} cm/sec.

Model sensitivity runs revealed that simulation results were essentially independent of the value for the rotational flow rate constant, F_f, throughout its recommended range. Therefore, further calibration involved the evaluation of the biological model parameter P_1 or, more appropriately, a value for P_1' for staged treatment applications.

This was accomplished by adopting a value of 0.40 m/hr (0.0111 cm/sec) for the liquid film coefficient, K_L, a value

assumed, and successfully used, by Grieves for the simulation of a 10-stage RBC pilot plant treating municipal wastewater (7). Values for the biological coefficient, K_1, and the treatability factor, TF, which best fit the measured data were then selected. Finally, based upon a least sum of squared error calculation, calibrated parameter values were determined.

Substitution of the values for K_L and K_1 reveals a P_1 value for simulation of the overall unit (without the use of a treatability factor) equal to 1.29×10^{-4} cm/sec. This is roughly equivalent to the mean first and second stage calculated P_1 value of 1.33×10^{-4} cm/sec and resulted in a total sum of squared error equal to 69,031 for the 18 runs. However, 46% of this error resulted, once again, from poor simulation of Run 11. The P_1 values used during simulation with a treatability factor equaling 0.75 for stages 1 through 4 were 1.53×10^{-4}, 1.15×10^{-4}, 0.88×10^{-4}, and 0.65×10^{-4}. As with Kornegay's steady state model, use of a treatability factor improved simulation results in runs having an influent flow rate less than 0.60 gal/day-ft^2. Overall simulation results, however, were only found to improve approximately 1%.

Figures 10 and 11 illustrate the model's steady state response to alterations in influent feed characteristics. Because of the simplifying assumption that first order microbial kinetics govern treatment, simulated substrate removal is independent of feed strength, as indicated by the linear relationship depicted in Figure 10. Although experimental results have confirmed this hypothesis at low influent waste strength, the validity of this assumption can not be justified if feed strength were to increase above approximately 750 mg/l BOD.

As can be seen from Figure 11, model response is highly dependent upon influent flow rate. However, unlike the results obtained from the sensitivity analysis performed on Schroeder's model, an upper limit on BOD removal due to influent flow rate is not implied.

DESIGN CONSIDERATIONS/CONCLUSIONS

Although each of the steady state models evaluated in this analysis was found to provide an adequate fit of the measured data, caution must be exercised in the use of either of these models for design purposes. Inherent in their respective derivations are several simplifying assumptions regarding process performance which, if violated or neglected, can significantly affect "predicted" system response. Major factors which must be considered, particularly with respect to calibrated biologi-

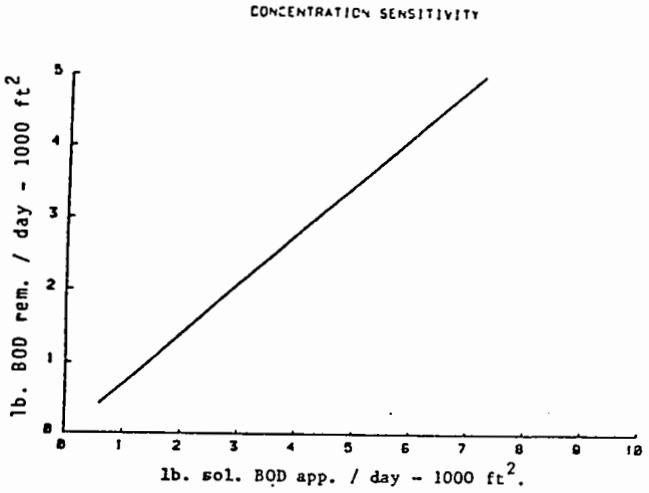


FIGURE 10: Grieves Model-Single Stage Reactor Response at Constant Flow Rate

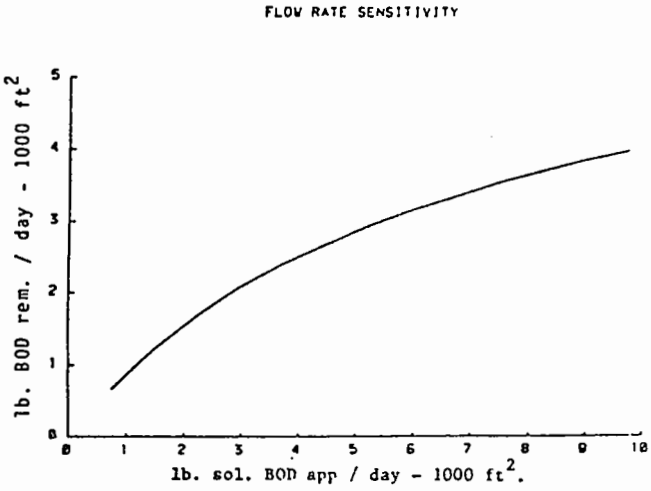


FIGURE 11: Grieves Model-Single Stage Reactor Response at Constant Waste Strength

cal constants and coefficients, are:

1) The impact that dynamic loading of the system will have upon system response.
2) The impact that scale-up will have. Preliminary research indicates that a 15 to 25% scale-up factor is appropriate depending on the physical characteristics of the pilot scale unit (19).
3) Media configuration/reactor "shor circuiting" has not been modeled explicitly.
4) Oxygen limitation has not been considered in any model.
5) Temperature effects, which can alter microbial reaction rates, diffusivity coefficients and dissolved oxygen concentrations have not been considered.
6) In reference to Schroeder's steady state model, specification of a given surface area also fixes the surface area to volume ratio of the reactor.

The best overall simulation results were obtained utiling the Grieves model for this data set. However, Kornegay's model was found to have the simplest calibration methodology. All of the models were found to exhibit a pronounced response to flow rate variations. On a single stage basis, removal efficiency was negatively impacted above a hydraulic loading of 1.5 gal/ft^2/day.

REFERENCES

1. Antonie, R.L., Welch, F.M., "Preliminary Results of a Novel Biological Process for Treating Dairy Wastes", Proceedings of the 24th Industrial Waste Conference, Purdue University, LaFayette, Indiana, 1969.

2. Joost, R.H., "Systematation in Using the Rotating Biological Surface Wastewater Treatment Process", Proceedings of the 24th Industrial Waste Conference, Purdue University, Lafayette, Indiana, 1969.

3. Weng, C.M., Molof, A.H., "Nitrification in the Biological Fixed Film Rotating Disc System", Water Pollution Control Federation Journal, Vol. 45, 1974.

4. Boyle, W.C., Berthouex, P.M., "Biological Wastewater Treatment Model Building Fits and Misfits", Biotechnology and Bioengineering, Vol. 16, No. 6, September, 1974.

5. McKinney, Ross E., Microbiology for Sanitary Engineers, McGraw-Hill Book Company, Inc., 1977.

6. Kornegay, B.H., "Modeling and Simulation of Fixed Film Biological Reactors for Carbonaceous Waste Treatment", in Mathematical Modeling for Water Pollution Control Processes, edited by Keinath, T.M. and Wanielista,M., Ann Arbor Science Publishers, Inc., 1975.

7. Grieves, C.G., "Dynamic and Steady State Models for the Rotating Biological Disc Reactor", Ph.D. Thesis, Clemson University, South Carolina, 1972.

8. Williamson, K., and McCarty, P.L., "A Model of Substrate Utilization by Bacterial Films", Water Pollution Control Federation Journal, Vol. 48, No. 1, January 1976.

9. Sanders, W.M., "Oxygen Utilization by Slime Organisms in Continuous Culture", International Journal of Air and Water Pollution, Vol. 10, 1966.

10. Tomlinson, T.G., and Snaddon, D.M.H., "Biological Oxidation of Sewage by Films of Microorganisms", International Journal of Air and Water Pollution, Vol. 10, 1966.

11. Atkinson, B., and Davies, I.J., "The Overall Rate of Substrate Uptake (Reaction) by Microbial Films", Transactions of the Institute of Chemical Engineers, Vol. 52, No. 3, July 1974.

12. Famularo, J., Mueller, J.A., and Mulligan, T., "Application of Mass Transfer to Biological Contactors", Water Pollution Control Federation Journal, Vol. 50, No. 4, April 1978.

13. Brook, T.D., Biology of Microorganisms, Prentice Hall, Inc., Englewood Cliffs, New Jersey, 1979.

14. Scroeder, E.D., Water and Wastewater Treatment: Chapter 9- Biological Film Flow Processes, McGraw-Hill Book Company, Inc., 1977.

15. Blanc, F.C., O'Shaughnessy, J.C., and LaRosa, A.P., "Treatment of Racetrack Wastewater Using Rotating Biological Contactors", New England Water Pollution Control Association Journal, Vol. 11, No. 2, October 1977.

16. Zeevalkink, J.A., Kelderman, P., and Boelhouwer, C., "Liquid Film Thickness in a Rotating Disc Gas - Liquid Contactor", Water Research, Vol. 12, No. 8, 1978.

17. Bintanja, H.H.J., Brunsmann, J.J. and Boelhouwer, G., "The Use of Oxygen in a Rotating Disc Process", Water Research, Vol. 10, No. 6, 1976.

18. Levich, V.G., Physiochemical Hydrodynamics, Prentice Hall Inc., Englewood Cliffs, New Jersey, 1968.

19. Wilson, R.W., Murphy, K.L., and Stephenson, J.P., "Scaleup in Rotating Biological Contactor Design", Water Pollution Control Federation Journal, Vol. 52, No. 3, March 1980.

MATHEMATICAL MODELING FOR ASSESSING DEVELOPMENT AND SLOUGHING OF FIXED FILMS AND THEIR EFFECTS ON WASTE STABILIZATION

Ju-Chang Huang. Department of Civil Engineering, University of Missouri-Rolla, Rolla, Missouri.

Shoou-Yuh Chang. Department of Civil Engineering, University of Missouri-Rolla, Rolla, Missouri.

Yow-Chyun Liu. Department of Civil Engineering, University of Missouri-Rolla, Molla, Missouri.

INTRODUCTION

One of the major parameters governing the performance of any biological treatment system is the food to microorganisms (F/M) ratio. In an aerobic biological unit, the rate of organic stabilization is directly proportional to the quantity of aerobic microorganisms (M) present when the substrate concentration is not limiting. Therefore, to optimize the use of any aerobic biological treatment unit, efforts must be made to keep the aerobic microbial concentration as high as possible. This is particularly important when the substrate concentration is high. However, in most biological treatment systems, the upper limit of aerobic microbial concentration is normally regulated by the oxygen availability. At a given oxygen level, only a certain limit of aerobic microorganisms are maintainable, and the rate of organic oxidation depends on this limit. If the microbial concentration is maintained beyond this limit, a portion of the microbial mass would be of either the facultative or anaerobic type, which can often develop odorous conditions in a treatment system. For example, in a suspended-growth system like con-

ventional activated sludge process, the level of mixed liquor suspended solids (MLSS) is generally kept within 4,000 mg/L to insure that molecular oxygen will penetrate into the center of biological flocs. However, when aeration is provided with pure oxygen (patented as "Unox Process"), the maximum MLSS concentration can be increased to as high as 8,000 mg/L because of the increased oxygen penetration into biological flocs. With the increased biomass in the Unox Process, its organic stabilization rate is greatly increased and its requirement of hydraulic retention can thereby be reduced. This would, of course, result in a substantial saving in the construction of the aeration tank system.

The development and maintenance of the biomass in an attached-growth (or fixed biofilms) system is considerably more complicated in comparison to the suspended-growth unit. This is because the biofilm development on a surface exposed to waste flow is the net result of physical transport and biological growth rate processes. The processes which contribute to the overall biofilm accumulation are: 1) diffusion of substrate into the biofilm; 2) diffusion of molecular oxygen into the biofilm; 3) substrate oxidation and growth of the attached microorganisms; and 4) sloughing of the biofilm. Among these processes, it is reasonable to assume that for a given substrate compound, the rate of substrate diffusion depends upon its concentration gradient in the biofilm layer. Similarly, oxygen diffusion rate also depends upon its concentration gradient. In an actual biofilm treatment unit (such as rotating biological contactor or RBC, trickling filter, and aerobic fluidized bed), under a steady-state condition the rate of substrate oxidation may be limited by either the substrate penetration or oxygen diffusion depending on the relative availabilities of these two substances. In a biological treatment system exposed to air, the maximum concentration of dissolved oxygen in wastewater seldomly exceeds 4 or 5 mg/L while the substrate concentration may be as high as hundreds or even thousands mg/L. Under such a situation, diffusion of molecular oxygen into the biofilm is normally the rate-limiting step in the waste stabilization process. For example, in a model-scale fixed film system, it has been found that for a glucose substrate with a concentration of 88 mg/L or more, the rate of organic oxidation is generally limited by the oxygen diffusion rather than by the substrate penetration (1,2). This type of oxygen limitation in the fixed film systems has also been observed by other researchers (3,4,5). Thus, all of these seem to suggest that in a biofilm treatment

system, if the influent BOD_5 is well above 100 mg/L, a significant portion, or even the majority, of the fixed-film growth will function under an anaerobic condition. When this occurs, the total oxidative capability in such a system cannot be measured by its total biomass since the anaerobic biomass does not possess the same level of biological activities as aerobic bacteria. Therefore, in order to optimize the utilization of each supporting surface area in a fixed-film system, every effort should be made to increase the "aerobic" portion of the biomass. This, of course, can be accomplished by increasing the oxygen availability in the treatment system. In fact, in a recent study using pure oxygen in the RBC operation, Huang and Bates (6) found that the use of pure oxygen was able to phenomenally increase the aerobic biomass accumulation on each unit disk surface area. Unfortunately, that study only demonstrated the qualitative evidence of the increased aerobic biofilm development; the quantitative relationship between the oxygen flux and the aerobic fixed-film accumulation was not established.

Another important parameter complicating the dynamic behavior of the fixed film development is the sloughing of biomass. Although it is known that sloughing is caused by the hydraulic shear at the biofilm layer, it is not clear as to the general frequency and exact location (or interface) that the sloughing would normally take place. It is speculated that the biofilm sloughing is most likely to take place at the aerobic-anaerobic interlayer, where the production of acidic metabolites by anaerobes is likely to weaken the binding strength of polysaccharides in the biofilm establishment.

From the above discussion, it is clear that most of the fixed film biological treatment system being used today (such as RBC, trickling filters and aerobic fluidized beds, etc.) have not been optimized to utilize their valuable surface areas to support exclusively aerobic biomass due to a lack of oxygen. Because of the oxygen limitation, several investigators (7-11) have found that the rate of substrate removal cannot be further increased once the effective thickness of the biofilm reaches a certain level. Undoubtedly, if the entire layer of biofilm is made of aerobic bacteria, the rate of substrate removal should continue to increase with the biofilm thickness as long as the substrate concentration is not limiting. On the other hand, if the biofilm is also composed of anaerobes, then the rate of substrate oxidation may not linearly increase with the thickness

of the biomass.

At the present time, our understanding of the dynamic behavior of the biofilm development and sloughing is quite meager. The relationship between the thickness of aerobic fixed film and the available substrate/oxygen concentrations has not been quantitatively established. This paper presents a rational modelling approach for developing equations which may be used to predict the development and sloughing of fixed films under defined conditions. Also, the specific experimental tests which are required to generate pertinent modelling parameters are discussed in detail.

In order to fulfill the modelling requirement, specific experimental tests have been designed to generate the following data:

1) to quantitatively relate the substrate removal rate with both the aerobic and anaerobic biofilm development under some specially-designed operating conditions; 2) to assess the impact of substrate and oxygen concentrations on the development of biofilm thickness and its impact on waste stabilization rate; and 3) to estimate the attenuation of dissolved oxygen and substrate concentrations across the biofilm layer and then to identify the interface at which biofilm sloughings are most likely to occur.

MODELING APPROACH

In order to develop fixed-film biological growths in well defined conditions, several annular reactors need to be fabricated. Each reactor will consist of a stationary outer cylinder and a rotating inner impeller, as shown in Figure 1.

The annular reactor has the advantages of providing a constant shear throughout the stationary supporting surface as well as allowing direct insertions of oxygen probes and sampling capillaries during testings. Therefore, this type of reactor will allow generation of experimental data for the development of a model to correlate the substrate removal rate with biofilm buildup. This system will also provide data to establish the attenuation of substrate and dissolved oxygen (DO) through the biofilm layer at various substrate and DO availabilities and then to identify the interface at which biofilm sloughings are most likely to occur. A glucose substrate with adequate minerals and phosphate buffer will be used as the feed. The glucose concentration may be adjusted to any level in different phases of the experiment to suit the modeling need.

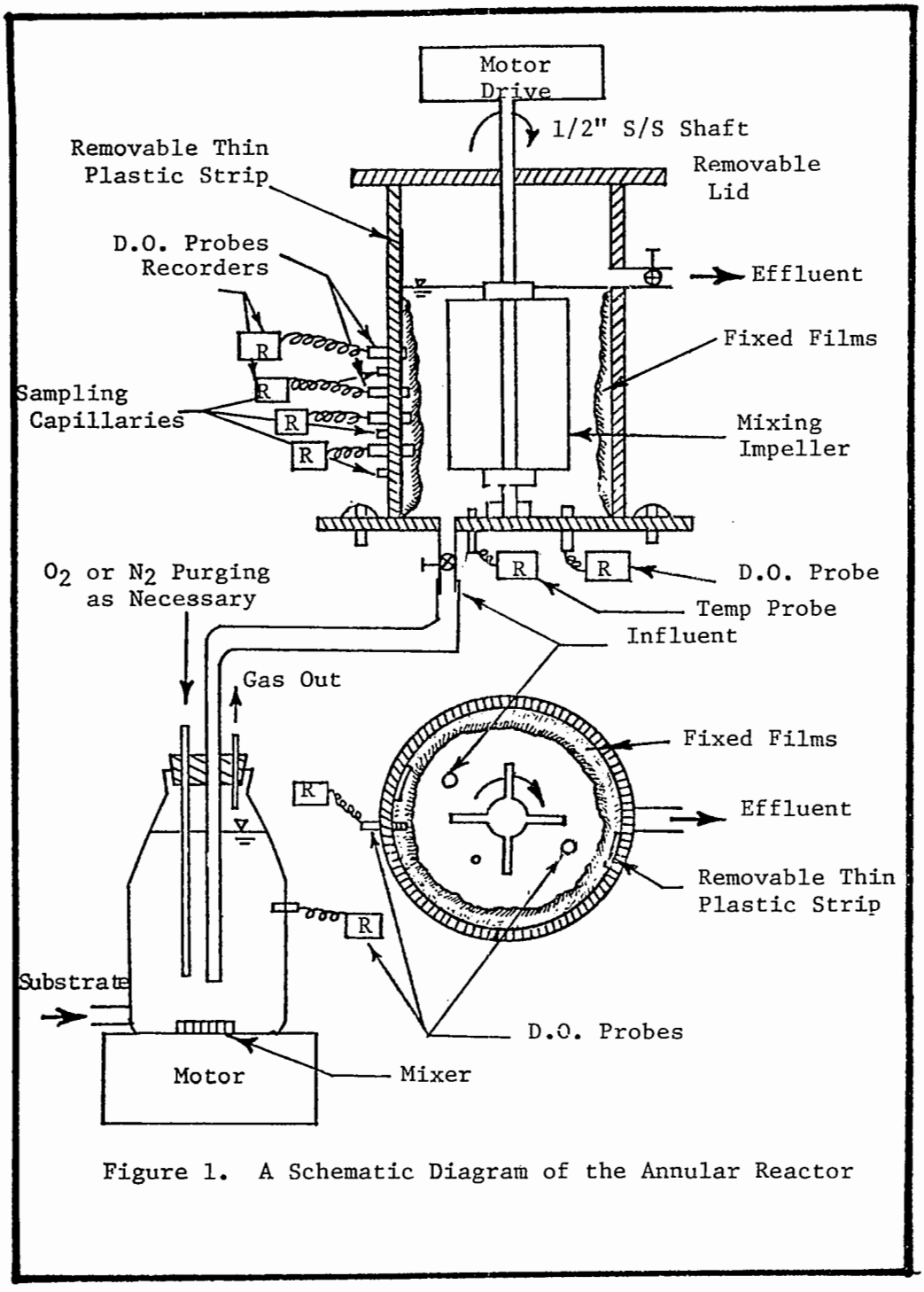


Figure 1. A Schematic Diagram of the Annular Reactor

In each testing, the substrate will be added to the annular reactor at a sufficient flow rate to intentionally maintain a hydraulic detention time of no more than 15 min so that the growth of suspended biomass can be neglected in the mathematical modeling. Before the substrate is added into the reactor, pure oxygen at various flow rates will be injected into it and briefly mixed to maintain desirable dissolved oxygen (DO) levels. The DO concentrations in both the mixer unit and the annular reactor will be monitored and recorded continuously throughout each test. The speed of the impeller rotation inside the annular reactor will be properly regulated, but in no case shall the peripheral velocity ever exceed 1 ft/sec, which is the upper limit being used in most full-scale RBC applications. Because of the short hydraulic detention inside the reactor, biomass production would be limited mainly to the attached biomass. Hence the variation in suspended solids with time can be directly attributed to the process of biofilm sloughings.

The experiment will be initiated by inoculating a small amount of sewage microorganisms and operating the reactor in a batch mode until some surface slimes start to develop. This technique will speed up the initial establishment of the primary slime layer (12,13,14) in the reactor. After the initial primary layer has developed, the reactor will be switched to the continuous-flow operation with the feed of a synthetic substrate. At this stage, the continuous development of biofilms and the associated substrate stabilization rate will be monitored as frequently as necessary. The mass balance equation for the substrate in the system is:

$$V \frac{dS}{dt} = Q\ (S_o - S) - \frac{M_a + M_x}{Y_a} - \frac{M_s}{Y_s} \quad \ldots\ldots\ldots \text{(Eq. 1)}$$

where V = liquid volume in the annular reactor (L^3)
S = substrate concentration in the reactor (ML^{-3})
t = time elapsed (t)
Q = feed rate (L^3/t)
S_o = influent substrate concentration (ML^{-3})
M_a = attached biomass growth rate (M/t)
M_x = sloughings of attached biomass in the reactor (M/t)
Y_a = attached biomass yield coefficient
M_s = suspended biomass growth rate (M/t)
Y_s = suspended biomass yield coefficient

Because of a short detention time (no more than 15 min) employed in this study, the removal of substrate due to sus-

pended biomass growth can be neglected in comparison to the substrate consumption for the attached biomass growth. Thus, Eq. 1 may be rewritten as:

$$V \frac{dS}{dt} = Q\ (S_o - S) - \frac{M_a}{Y_a} \quad \text{(Eq. 2)}$$

The mass balance of the biomass in the reactor, on the other hand, can be expressed as follows:

$$V \frac{dX}{dt} = Q\ (X_o - X) + M_x \quad \text{(Eq. 3)}$$

where X_o is the influent suspended biomass concentration and X is the biomass concentration in the reactor.

The growth rate of attached biomass can be expressed as:

$$M_a = AP \frac{dTh}{dt} \quad \text{(Eq. 4)}$$

where A = reactor surface area of the attached biofilm
P = biofilm volumetric density, and
Th= attached biofilm thickness

Since the influent suspended biomass concentration is zero, the rate of sloughing which results in the production of suspended biomass can be estimated from Eq. 3:

$$M_x = V \frac{dX}{dt} + QX \quad \text{(Eq. 5)}$$

After substituting Eqs. 4 and 5 for the terms of M_a and M_x in Eq. 2, the following equation can be obtained:

$$AP \frac{dTh}{dt} = Q(S_o - S) - V \frac{dS}{dt}\ Y_a - V \frac{dX}{dt} - QX \quad \text{(Eq. 6)}$$

After Y_a has been determined, Eq. 6 can be used to correlate the substrate removal rate with the biofilm development.

As the reactor is operated longer and longer, the biofilm layer inside the reactor will become more and more established. As the biofilm thickness becomes greater, sloughings will start to occur and suspended solids concentration in the reactor will increase. At this stage, the last two terms in Eq. 6 cannot be neglected any more. The thickness of the attached biomass will become a function of the sloughing rate. Thus the effluent suspended solids concentration (X) and $\frac{dX}{dt}$ inside the reactor over a short defined test interval must be determined to calculate the rate of change of biofilm thickness. The calculated value will then be checked against the actual measurement from the inserted thin plastic

strip during the course of the experimental study.

When steady state conditions of Eqs. 2,3 and 6 are reached, the thickness of the attached biomass, the substrate concentration and the biomass concentration in the reactor will be constant. Equation 6 becomes:

$$Q\ (S_o - S) - \frac{QX}{Y_a} = 0 \quad \text{(Eq. 7)}$$

The substrate removal rate for a constant biofilm thickness can then be calculated as:

$$Q\ (S_o - S) = \frac{QX}{Y_a} \quad \text{(Eq. 8)}$$

Note that QX is actually the amount of biomass that has been sloughed off in the reactor and can be expressed as the product of specific "yield" rate and the overall attached biomass (μ.APTh). Thus the substrate removal rate can be related to the thickness of the attached biomass as follows:

$$Q\ (S_o - S) = \left(\frac{\mu AP}{Y_a}\right).Th \quad \text{(Eq. 9)}$$

The specific yield rate, μ, is a function of the substrate concentration, oxygen concentration as well as other environmental factors. A model similar to the Monod equation and Michaelis-Menten relationship may be used:

$$\mu = \mu_{max} \frac{S}{K_s + S} \quad \text{(Eq.10)}$$

where μ_{max} = maximum "yield" rate
K_s = Monod half velocity concentration and
S = limiting substrate concentration.

After μ_{max} and K_s have been experimentally determined, the substrate removal rate for a given Th can be calculated. The calculated value will then be compared to the actual measurement in the test.

It is expected that the active biofilm thickness is dependent on both the oxygen and the substrate concentrations in the reactor. Various oxygen and substrate concentrations will be employed in the test to evaluate the dependence of the biofilm thickness on these two parameters and then to establish the quantitative relationship between the substrate removal and the biofilm thickness.

It must be noted that the theoretical considerations represented by the aforementioned equations will hold true only if the biofilm establishment in the reactor is either com-

pletely aerobic or anaerobic. A mixed aerobic-anaerobic biofilm system will complicate the calculations since their biomass yield coefficients and biofilm densities are not the same.

After the biofilm is well developed and the relationship between the substrate removal rate and the biofilm development has been established, the concentration in the influent feed will be progressively increased to effect the buildup of a thicker biofilm until it reaches a critical point at which DO concentration becomes a limiting factor. At this point, a complete aerobic condition will not prevail throughout the biofilm layer. Thus, some dark-color anaerobic biomass will develop at the biofilm's underlayer and sloughing will occur at a much greater rate. At this stage of operation, five DO microelectrodes as described by Whalen, et al. (15) and an equal number of capillary sampling tubes will be inserted into different depths of the biofilm layer from the reactor's cylindrical wall. The positions of insertion will be close together along the removable thin plastic strip so that at any particular moment, the monitored DO and substrate concentration profiles can be related to the biofilm thickness. During each separate testing, the DO and substrate concentrations in the influent feed will remain the same, while in the bulk solution the DO concentration will be monitored continuously and the substrate concentration determined as frequently as necessary. The DO monitorings will be continuously recorded throughout the test period to evaluate an expected "sigmoidal" pattern of the DO variations due to periodic sloughings of biofilms.

The concentration profile of the substrate can be obtained by establishing a mass balance equation for a differential thickness in the attached biofilm (1,2,16). A simple experimental first-order decay equation may be assumed for the limiting substrate, as follows:

$$S_{Th} = S_i 10^{-k_s Th} \quad \text{(Eq. 11)}$$

$$DO_{Th} = DO_i 10^{-k_o Th} \quad \text{(Eq. 12)}$$

where DO_i and S_i are the DO and substrate concentrations at the biofilm surface; k_o and k_s are the attenuation rate constants for DO and substrate concentrations across the biofilm; and Th is the thickness of biofilm at the point of measurement. If the substrate is not limiting, the zero order decay equation will be used:

$$S_{Th} = S_i - k_s' Th \quad \ldots\ldots\ldots\ldots\ldots\ldots \text{(Eq. 13)}$$

$$DO_{Th} = DO_i - k_o' Th \quad \ldots\ldots\ldots\ldots\ldots\ldots \text{(Eq. 14)}$$

where k_s' and k_o' are the decay rate for the substrate and oxygen, respectively. By keeping the rotating impeller at a reasonably high speed (so that the Reynolds number is in the turbulent range), the values of DO_i and S_i will be close to those existing in the bulk solution. Through an adequate number of repeated determinations, the experimental data should be able to allow for estimations of k_o and k_s. Also attention will be given to correlate the interface of sloughing with the DO and substrate profiles to establish the most likely location that the sloughings would normally take place. From the established k_o and k_s values, the attached biomass accumulation in any waste treatment system may be predicted from the available DO and substrate concentrations using Eqs. 11 through 14. The validity of such a prediction will be verified in the testing by systematically changing the DO and substrate concentrations in each study.

SUMMARY

The purpose of this paper is to present a logical approach to develop mathematical models for assessing the fixed-film buildup and sloughings in a biological waste treatment process and their resultant impacts on the rate of waste stabilization. Careful experimental testings are now being conducted at the University of Missouri-Rolla to generate pertinent parameters associated with the modeling. Besides, these testings will also be used to verify the validity of the proposed models. It is hoped that with a better understanding of the fixed-film system, future designs of RBC and aerobic fluidized-bed biological reactor can be optimized by eliminating the oxygen availability as the most common rate-limiting factor. This would result in a significant reduction of the reaction time requirement, thus achieving a corresponding capital saving associated with the tankage construction.

REFERENCES

1. Williamson, K., and McCarty, P.L., "A Model of Substrate Utilization by Bacterial Films." Journal of Water Pollution Control Fed., 48, 9 (1976).

2. Williamson, K., and McCarty, P.L., "Verification Studies of the Biobilm Model for Bacterial Substrate Utilization." Jour. Water Poll. Control Fed., 48, 281 (1976).
3. Mehta, D.S., Davis, H.H., and Kingsburg, R.P., "Oxygen Theory in Biological Treatment Plant Design," Jour. of the San. Eng. Div., ASCE, Vol. 98, SA3, P. 471 (1972).
4. Owen, D.T. and Williamson, K.J., "Oxygen Limitation in Heterotrophic Biofilms," Proceedings of the 31st Purdue Industrial Waste Conference, Purdue University, West Lafayette, IN. (1976).
5. Tropey, W., et al., "Effects of Exposing Slimes on Rotating Discs to Atmosphere Enriched with Oxygen," Proceedings of the Sixth International Conference on Advances in Water Pollution Research, P. 405 (1973).
6. Huang, J.C. and Bates, V.T., "Comparative Performance of Rotating Biological Contactors Using Air and Pure Oxygen," Jour. Water Poll. Control Fed. Vol. 52, No. 11, pp. 2686-2703 (November 1980).
7. Kornegay, B.H. and Andrews, J.F., "Characteristics and Kinetics of Biological Fixed Film Reactors," J. Water Pollution Control Fed. Vol. 40, R460 (1968).
8. Maier, W.J. et al., "Simulation of the Trickling Filter Process," Jour. of the San. Eng. Div., ASCE, Vol 93, No. SA4 (August 1967).
9. Tomlinson, T.G. and Snaddon, D.H., "Biological Oxidation of Sewage by Films of Micro-Organisms," Air and Water Poll. Int'l. Jour., Vol. 10, 865 (1966).
10. Hoehn, R.C. and Ray, A., "Effects of Thickness on Bacterial Film," Jour. Water Poll. Control Fed., Vol. 45, 2302 (1973).
11. Sanders, W.M., "Oxygen Utilization by Slime Organisms in Continuous Cultures," Intl. Jour. Air & Water Poll., Vol. 10, P. 253 (1966).
12. Baier, R.E., Shafin, E.G. and Zisman, W.A., "Adhesion: Mechanisms that Assist or Impede It," Science, 162, 2, 1360 (1968).
13. Mariappan, M., "Generation of Sulfide in Filled Pipes," Ph.D. Dissertation, Dept. of Civil Engr., Univ. of MO-Rolla, Rolla, MO (1976).
14. Characklis, W.G., "Biofilm Development and Destruction," Final Report, Electric Power Research Institute RP 902-1, Palo Alto, CA (1979).
15. Whalen, W.J., Bungay, H.R. and Sanders, W.M., "Microelectrode Determination of Oxygen Profiles in Microbial Slime Systems," Environ. Sci. & Tech., Vol. 3, 12

(December 1969).
16. LaMotta, E.J., "Internal Diffusion and Reaction in Biological Films," Env. Sci. & Tech., 10, 765 (August 1976).

EVALUATION OF RBC SCALE-UP

Yeun C. Wu, Department of Civil Engineering, University of Pittsburgh, Pittsburgh, Pa.

Ed. D. Smith, Environmenatl Division, U.S. Army Construction Engineering Research Lab., Champaign, IL.

Chiu Y. Chen, Department of Environmental Engineering, National Chung Hsin University, Taichung, Taiwan

Roy Miller, Environmental Health Branch, U.S. Army Environmental Hygiene Agency

INTRODUCTION

Development and application of Wu's model for the prediction of soluble BOD removal in rotating biological contactor (RBC) wastewater treatment systems have been discussed in detail elsewhere (1, 2). The model is capable of both precisely estimating the treatment efficiency of RBC systems and successfully determining the size of the treatment plant if the design conditions such as influent soluble BOD concentration, wastewater temperature, number of RBC stages, and % BOD removal requirement are known. Therefore, the model is very useful for predicting performance and can be easily applied for engineering design purposes.

Presently, little is known about the applicability of pilot plant data for full-scale design. As a result, there is an essential need to investigate RBC scale-up under various operating conditions. This study was primarily designed to determine the influence of wastewater temperature on process scale-up. Wu's model is capable of performing this important task.

RBC MODEL

Wu's model was developed on the basis of full-scale RBC data reported by many researchers (3-10). The model is given as follows:

$$F = \frac{14.2 \quad q^{0.5579}}{\text{Exp.}^{0.32N} \; L_o^{0.6837} \; T^{0.2477}} \qquad \text{--------(1)}$$

in which

F = fraction of influent soluble BOD remaining in the effluent, %

q = surface hydraulic loading, gpd/ft^2

N = number of RBC stages

L_o = influent soluble BOD concentration, mg/l

T = wastewater temperature, ^oC

Eq. 1 describes the relationship between % BOD removal/remaining (F) as function of process variables including q, L_o, N, and T. For instance, the effect of hydraulic loading, q, on F under varying influent soluble BOD concentrations, L_o, and number of RBC stages, N, at temperature T = 25°C is shown in Figure 1. It can be seen in Figure 1 that the F value always decreases as q, L_o, and N increase. The influence of stage number, N on F under differing conditions for q and L_o at T = 25°C is illustrated in Figure 2. Obviously Figure 2 shows that the F value decreases profoundly as a result of either decreasing q or increasing both L_o and N. However, F changed only slightly after N was greater than 6. This result becomes very obvious when L_o is high and q is low. The relationship between T and F under varying L_o, q, and N is depicted in Figure 3. It is apparent from Figure 3 that for all conditions investigated here, F appears to become independent of T after T > 15°C. In addition, Figure 3 also shows that the influence of T on F is less significant when both L_o and N are high and q is low.

The reliability and accuracy of this model has been extensively studied by using more than eighty data sets obtained from the operation of six full-scale RBC plants (2). The maximum error which results from the use of Wu's model was found to be $\pm$ 4.64% in terms of the efficiency of BOD removal.

DEVELOPMENT OF RBC SCALE-UP FACTOR

Past experience indicates that most operating full-scale RBC treatment plants were designed according to criteria generated from small-scale pilot plant studies. It is unknown whether the pilot scale data are adequate for the process engineer to physically size the full-scale plant. An early work of Famularo et al. predicted a 10% reduction in organic removal in a 4 stage RBC system if the disc size increaesd from 2 m to 6 m (11). Further, Murphy and Wilson have recently demonstrated that the removal of COD is approximately 15% lower for a 2 m RBC than for 0.5 m RBC. They propose that an additional 10% increase be made in scaling up from a 2 m RBC to 3.5 m RBC at 17°C (12). However, the effect of temperature on RBC scale-up was not reported by Famuaro et al. or Murphy and Wilson.

According to Murphy and Wilson, the inverse relationship between disc size and substrate removal efficiency could be explained by a combined physical and biological effect. They have speculated that as the RBC disc diameter is increased, the liquid film on the biomass is exposed to the atmosphere for longer times resulting in greater substrate depletions and lower substrate concentrations in the liquid layer. Under conditions of low substrate concentrations, when substrate availability and diffusion is limiting, total removal efficiency declines as disc size increases. Another possibility which may produce the aforementioned result is the operation of the small-scale pilot unit at a higher rotational speed. Therefore, the rate of oxygen transfer from gas phase to liquid phase in RBC system under the identical hydraulic/organic loading favors the small unit because of its high rotational speed.

It is evident from the discussion above that an investigation of RBC scale-up is necessary for engineering design, even though some difficulties are encountered due to a lack of field data and an available mathematical model. Since this predictive model enables one to correlate the BOD removal efficiency with the process controlling variables successfully, the scale-up factor can be determined if both pilot-scale and full-scale plant data are obtained. Sixty-four data sets including influent soluble BOD concentrations, hydraulic loading, wastewater temperature, % BOD removal, and number of RBC stages produced from seven full-scale RBC plants, along with sixty-three data sets developed

from the study of five small-scale RBC units were employed for the present investigation (13-24). Actual equations involved in the development of the scale-up factor are indicated as follows:

$$K_1 = \frac{F_1}{F_r} = \frac{F_1}{\dfrac{14.2 \quad q^{0.5579}}{\exp.^{0.32N} L_o^{0.6837} T_*^{0.2477}}} \qquad \text{-----(2)}$$

and

$$K_2 = \frac{F_2}{F_r} = \frac{\dfrac{14.2 \quad q^{0.5579}}{\exp.^{0.32N} L_o^{0.6837} T^{0.2477}}}{\dfrac{14.2 \quad q^{0.5579}}{\exp.^{0.32N} L_o^{0.6837} T_*^{0.2477}}}$$

$$= \left(\frac{T_*}{T}\right)^{0.2477} \qquad \text{-----(3)}$$

L_o, q, N, and T in Eqs. 2 and 3 are the system operating conditions for either a small-scale or a full-scale RBC plant. F_1 represents the measured % BOD remaining and F_2 is the predicted % BOD remaining obtained from the model calculation at the same conditions as F_1. However, F_r in Eq. 3 is different from F_2 because it is calculated at a referenced temperature T* instead of T. The ratios of F_1 to F_r and F_2 to F_r are designated as K_1 and K_2, respectively. K_2 is theoretically equal to K_1, if the results of % BOD remaining for both field measurement and model prediction are identical.

The effect of T* on the relationship between K_1 or K_2 and T is shown in Figure 4. It is important to point out that the theoretical curve (K_2 vs T) always passes through a point where K_2 is equal to 1 and T is the same as T*. K_1 is also a function of T*, that is, K_1 increases as a result of increasing T*. But it was decreased as the wastewater temperature T was increased, according to Figure 4.

The operational curves (K_1 vs T) as shown in Figure 4 were constructed using the full-scale RBC plant data.

Although the K_1 values are randomly dispersed, the operational curves obtained from data analyses utilizing a non-linear least square method, closely approximate the theoretical curves for different operating conditions.

Further development of the relationship between K_1, K_2 and T for pilot-scale system (disc size < 6 ft) at $T_* = 20^oC$ was made. The results are illustrated in detail in Figure 5. A comparison of performance of the pilot-scale RBC with the full-scale RBC under the same operating conditions is made with Figure 4-(C) and Figure 5. The comparison revealed the operational curve in the former system to be far below the theoretical curve. However, the reverse is found in the latter system. This result is expected because the presently employed model was developed based on the full-scale plant data.

From the above discussion, it is known that the direct application of pilot plant data for full-scale design is not acceptable. In all cases studied, the K_1 value at any particular temperature, T is always higher in the full-scale system than in the pilot scale system if the referenced temperature T_* is the same. This phenomenon indicates that the full-scale RBC plant is less effective if the system is designed in accordance with the data obtained from a treatability study of a small-scale pilot plant. As a result of this observation, the following investigation was aimed to develop the scale-up factor (SUF).

Both the operational curves as shown in Figure 4-(C) and Figure 5 are the lines of best fit, calculated by non-linear least squares regression with a 95% confidence limit. From these analyses it is found that at the referenced temperature $T_* = 20^oC$, the operational curves for full-scale RBC plants and pilot-scale RBC plants can be described by the following two equations:

$$(K_1)_{Full} = 1.5535 - 0.041666T + 0.00075233\ T^2 \qquad ------(4)$$

and

$$(K_1)_{Small} = 1.4919 - 0.036106T + 0.0005692\ T^2 \qquad ------(5)$$

At any given temperature T, the ratio of $(K_1)_{Full}$ to $(K_1)_{Small}$ is the process scale-up factor, according to Eq. 2. The relationshp between SUF and temperature T is

depicted in Figure 6. It is clearly shown in Figure 6 that the SUF value varies significantly as a function of temperature. The scale-up factor increases from 1.067 to 1.227 as the temperature increases from 3°C to 25°C. However, a decrease in SUF was found when the temperature exceeded 25°C.

It is important to point out that within the temperature range investigated the maximum scale-up (22.7%) occurs at T = 25°C and the minimum scale-up (6.7%) takes place at T = 3°C. Inhibitory effects due to high temperature begin to show when T exceeds 25°C. The relationship between SUF and T is described as

$$SUF = 1.0097 + 0.016206T - 0.00032842\ T^2 \qquad -----(6)$$

and is illustrated in Figure 6.

The following example is given to demonstrate the method for incorporating the SUF into the full-scale plant design:

The experimental data obtained from a small-scale pilot plant study are (25):

% Soluble BOD Removal Required	= 82% or F = 0.18
Hydraulic Loading in gpd/ft^2 (q)	= 1.50
Stage Number (N)	= 4
Influent Soluble BOD Concentration (L_o)	= 50 mg/l
Wastewater Temperature (T)	= 13.4°C
Design Flow Rate	= 2 MGD

Based on the design criteria specified above, the SUF was calculated by using Eq. 6. The result is 1.20256. As mentioned earlier, the SUF is mathematically defined as:

$$SUF = \frac{(K_1)_{Full}}{(K_1)_{Small}} = \frac{(F_1)_{Full}}{(F_1)_{Small}}$$

$$1.1670 = \frac{\dfrac{14.2 \quad q^{0.5579}}{\exp.^{0.32N} \; L_o^{0.6837} \; T^{0.2477}}}{0.17}$$

For the experimental data:

$$1.1670 = \frac{\dfrac{14.2 \times q^{0.5579}}{\exp.^{0.32 \times 4} (50)^{0.6837} (13.4)^{0.2477}}}{0.17} \qquad \text{-----(7)}$$

By solving Eq. 7, the q value is found to be equal to 1.348 gpd/ft^2, that is less than 1.50 gpd/ft^2 resulting from the pilot plant study. The total disc surface required is 1,483,680 ft^2 (2,000,000/1.348) instead of 1,333,333 ft^2 (2,000,000/1.5). Additionally, it is important to recognize that the effluent quality of the full-scale RBC plant will be slightly less due to the change in hydraulic loading. The resulting effluent quality is estimated as follows:

$$F = \frac{14.2 \; (1.348)^{0.5579}}{\exp.^{0.32 \times 4} (50)^{0.6837} (13.4)^{0.2477}}$$

$$= 0.210$$

Therefore, the soluble BOD remaining in the full-scale RBC plant effluent is 50 mg/l x 0.210 = 10.5 mg/l instead of 50 mg/l x 0.18 = 9.0 mg/l.

Additional calculations show the difference in hydaulic loading with and without the considering scale-up factor and are listed in Table I. The table clearly indicates that the reduction of hydraulic loading is greater when both T and q used for the operation of pilot plant are higher.

Table I

Comparison of Hydraulic Loading Calculated With and Without the Inclusion of SUF**

Operating Parameters				Hydraulic Loading		
L_o(mg/l)	N	T (°C)	F	without SUF	with SUF	(5)-(6)
(1)	(2)	(3)	(4)	(5)	(6)	(7)
45	4	15.1	0.18	2.0	1.54	0.460
64	4	16.8	0.20	3.0	2.51	0.490
60	4	20.6	0.23	4.0	3.31	0.690
53	4	23.9	0.16	2.0	1.58	0.420
49	4	24.9	0.21	3.0	2.39	0.610
57	4	17.4	0.18	2.0	1.82	0.180
73	4	24.4	0.15	3.0	2.13	0.870

**Data from Ref. (25)

CONCLUSIONS

According to this investigation, when a full-scale RBC plant design is based on the essential controlling variables of influent soluble BOD concentration, wastewater temperature, number of disc stages, surface hydraulic loading, and % BOD removal requirement, the preliminary design criteria developed from pilot plant study cannot be directly employed for design. A scale-up factor should be used.

This factor was successfully determined by the model proposed by Wu et al. Its relation to wastewater temperature was mathematically formulated by conducting non-linear least squares regression analysis on both full-scale and pilot-scale data previously reported by other investigators. It is apparent that the process scale-up increases from

1.067 at T = 3°C to 1.227 at T = 25°C. However, a decrease in the scale-up factor was found when the temperature exceeded 25°C.

This study shows the effect of process scale-up on the selection of hydraulic loading for full-scale design is significant when the wastewater temperature and hydraulic loading determined during the pilot plant study are high.

It is necessary to mention that the results of this study are valid only for the treatment of municipal wastewater by mechanical drive RBC and bio-oxidation of carbonaceous organic material in the RBC system occurs under oxygen sufficient conditions.

REFERENCES

1. Wu, Y. C., et al., "Modelling of Rotating Biological Contactor Systems," *Biotechnology and Bioengineering*, Vol. 12, p. 2055, 1980.
2. Wu, Y. C., et al., "Design of Rotating Biological Contactor Systems," In Press, *Journal of Environmental Engineering Division, ASCE*, November, 1981.
3. Antoinie, R. L., and Koehler, F. J., "Application of Rotating Disc Process to Municipal Wastewater Treatment," EPA Project No. 17050 DAM, Autotrol Corporation, Milwaukee, Wisconsin, 1971.
4. Sack, W. A., "Evaluation of the Biodisc Treatment Process for Summer Camp Application," EPA Project No. 17010 EBM, University of West Virginia, Morgantown, West Virginia, 1973.
5. Clark, et al., "Performance of a Rotating Biological Contactor Under Varying Wastewater Flow," *Journal of Water Pollution Control Federation*, Vol. 50, p. 896, 1978.
6. Hao, O., et al., "Rotating Biological Reactors Removal Nutrient - Part 1," *Water and Sewage Works*, Vol. 122, p. 10, 1975.
7. Antonie, R. L., "Fixed Biological Surfaces," *Wastewater Treatment*, CRC Press, Cleveland, Ohio, 1976.
8. Antonie, R. L., Kluge, D. L., and Mielke, J. H., "Evaluation of a Rotating Disc Wastewater Treatment Plant," *Journal of Water Pollution Control Federation*, Vol. 46, p. 498, 1974.
9. Malhortra, S. K., and Williams, T. C., "Performance of a Biodisc Plant in a Northern Michigan Community," In Proceedings of the 48th Annual Conference of the Water Pollution Control Federation, Miami, Florida, 1975.
10. Borchardt, J. A., "Biological Wastewater Treatment Using Rotating Discs," *Journal of Biological Waste Treatment*, Wiley Interscience, New York, p. 131, 1971.
11. Famularo, J., et al., "Application of Mass Transfer to Rotating Biological Contactors," Presented at the 49th Annual Conference of the Water Pollution Control Federation, Minneapolis, Minnesota, 1976.
12. Murphy, K. L., and Wilson, R. W., "Pilot Plant Studies of Rotating Biological Contactor Treating Municipal Wastewater," Report SCAT-2, Environmental Protection Service, Environmental Canada, July 1980.

13. Hitdlebauch, J. A., et al., "Full-Scale Rotating Biological Contactor for Secondary Treatment and Nitrification," in Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, February 1980.
14. Dupont, R. R., and McKinney, R. E., "Data Evaluation of a Municipal RBC Installation, Kirksville, Missouri," in Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1980.
15. Sullivan, R. A., et al., "Upgrading Existing Waste Treatment Facilities Utilizing the Bio-surf," in Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1980.
16. Andeson, E. D., "Performance of Full-Scale RBC Plant at Ft. Bragg, North Carolina," Personnel Communication, the Office of the Chief of Engineers, U.S. Department of the Army.
17. United States Army Environmental Hygiene Agency, "Phase 1 - Water Quality Engineering Special Study No. 32-24-0116-79 Sewage Treatment Plant Evaluation - Summer Conditions for Fort Knox, Kentucky," published by the Department of the Army, 1978.
18. Chow, C. S., et al., "Comparison of Full-Scale RBC Performance with Design Criteria," in Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennyslvania, 1980.
19. Smith, E. D., et al., "Recarbonation of Wastewater Using the Rotating Biological Contactor," in Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1980.
20. Khan, A. N., et al., "Rotating Biological Contactor for the Treatment of Wastewater in India," in Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, 1980.
21. Orwin, L. W., and Sieenthal, C. D., "Hydraulic and Organic Forcing of a Pilot Plant Scale RBC Unit," in Proceedings of the First National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, Pennsylvania, February 1980.

22. Torpey, W. N., et al., "Rotating Biological Disc Wastewater Treatment Process - Pilot Plant Evaluation," EPA Project No. 17010 EBM, Environmental Protection Agency, Washington, D.C., 1974.
23. Williams, et al., "The Gladstone, Michigan Experience: Performance of a 1.0 MGD RBC Plant in a Northern Michigan Community," in Proceedings of the First National Symposium/Workshop on Rotating Biological Technology, Champion, Pennsylvania, February 1980.
24. Griffith, G. T., et al., "Rotating Disc Sewage Treatment System for Suburban Developments and High Density Resorts of Hawaii," Water Resources Research Center, Technical Memorandum Report No. 56, University of Hawaii, Honolulu, Hawaii, 1978.
25. Miller, R. D., et al., "Rotating Biological Contactor Process for Secondary Treatment and Nitrification Following a Trickling Filter," Technical Report No. 7905, U.S. Army Bioengineering Research and Development Laboratory, Fort Detrick, Frederick, Maryland, 1979.

NOTATIONS

F = fraction of influent soluble BOD remaining in the effluent, %

F_1 = measured value of F, %

F_2 = predicted value of F, %

F_r = predicted value of F at given referenced temperature, %

L_o = influent soluble BOD concentration, mg/l

K_1 = ratio of F_1 to F_r

K_2 = ratio of F_2 to F_r

N = number of RBC stages

q = hydraulic loading rate, gpd/ft^2

SUF = scale-up factor

T = wastewater temperature, ^oC

T_* = referenced temperature, ^oC

PART V: SMALL-SCALE/ON-SITE SYSTEMS

SMALL WASTEWATER TREATMENT SYSTEMS USING SOIL PURIFICATION METHOD

Masaaki Niimi, Director, Soil Purification Center, Ltd.
Ueki Bldg., 2-41-8 Kabuki-cho, Shinjuku-ku, Tokyo 160, Japan

INTRODUCTION

Since night soils were used as fertilizer in agricultural land until 1950's, wastewater treatment in rural areas has generally been neglected until recently. Wide use of chemical fertilizers in recent years, however, prompted the necessity of rural wastewater treatment in Japan since night soils are no longer used in agricultural land.

Under these circumstances, Japan Ministry of Agriculture started a program in 1977 construct small system wastewater treatment facilities in rural areas, and adopted to promote soil purification systems as one of the most suitable methods of treatment.

The purpose of the paper is to describe unique features of the soil purification systems developed in Japan and to discuss construction, operation, and maintenance of the following systems:

Enhancement of Treatment by Use of Soil Cover

In this process, soils are not used as a mere construction materials, but they are effectively used as a media for supporting microbial life. In actual installations, treatment facilities are constructed underground covered by soil layer. Treatment efficiencies are observed to be greatly increased by the use of soil cover in these instances, and the ground surface can be used for lawn area or other uses for esthetic enjoyment.

Underground Trench Soil Purificaton System
By installing an impermenable sheet under the trench, capillary action of soils in horizontal directions is enhanced, thus preventing groundwater pollution due to enhanced soil purification in aerobic soil zone.

There have been already approximately 25,000 installations of our system in Japan. These systems utilize ecosystem of soilsphere and are suitable for small system wastewater treatment in rural areas. These facilities are low cost and low maintenance wastewater treatment systems and would not require extensive pipeline networks such as in a large-scale central wastewater treatment plant.

1 Oriental Tradition of Recycling Human Waste to Farmland

In the Eastern countries including Japan, human excrement has been utilized for agricultural production for as long as several thousand years, and this practice still survives even at present, though less commonly.

In my paper entitled "Do Joker Process"[1] of last September, the auther quoted the words of two Europeans who had referred to this Eastern wisdom" to our shame". One was Victor Hugo, in "Les Miserables" published in 1862 and the other was Dr. H. Maroh, a German who had visited Tokyo, then called Edo, around the same period.

For your reference, Victor Hugo and Dr. H. Maroh said as follows respectively:

"Paris Pours twenty-four million francs a year into the water. That is no metaphor. She does so by day and by night, thoughtlessly and to no purpose. She does so through her entrails, that is to say, her sewers. Twenty-five millions is the most modest of the approximate figures arrived at by statistical science.

After many experiments science today knows that the most fruitful and efficacious of all manures is human excrement. The Chinese, be it said to our shame, knew it before us."

How often do we hear our farmers talk about this manure being preferable to that manure on account of its fertilising action being 'more lasting;' yet with all our wise provision for the future, how far are we now behind the Japanese, who seem to look always to the next harvest only! As they manure for each fresh crop, and the term 'fallow' in our acceptation is entirely unknown to them, they are forced to distribute their yearly production of manure equally over the entire area of their land, which can be accomplished only by sowing

in drills or furrows, and by top-dressing."

Also, F. H. King stated in Chapter 9 of his "Farmers of Forty Centuries on Permanent Agriculture in China, Korea and Japan" (1911) as follows:

One of the most remarkable agricultural practices adopted by any civilized people is the centuries-long and well high universal conservation and utilization of all human waste in China, Korea and Japan, turning it to marvelous account in the maintenance of soil fertility and in the production of food.

The same book quoted the words by fur then, Dr. Arthur Stanley, Health officer of the city of Shanghai, in his annual report for 1899, as follows:

"(.....) while the ultracivilized Western elaborates destructors for burning garbages at a financial loss and turns sewage into the sea, the Chinaman uses both for manure. He wastes nothing while the sacred duty of agriculture is uppermost in his mind. And in reality recent bacterial work has shown that faecal matter and house refuse are best destroyed by returning them to clean soil, where natural purification takes place.

The question of destroying garbage can, I think, under present conditions in Shanghai, be answered in a decided negative. While to adopt the water-carriage system for sewage and turn it into the river, whence the water supply is derived, would be an act of sanitary suicide. It is best, therefore, to make use of what is good in Chinese hygienee, which demands respect, being as it is, the product of an evolution extending from more than a thousand years before the Christian era".

To my regret, this excellent Eastern wisdom has been utterly forgotten in present day Japan which has undergone ultracivilized 'modernization'.

The words of Socrates - "A bad law is also a law." - still survive in Japan too, even today when about 2,400 years have passed since his time. Ultracivilized modernized Japan has not only forgotten this excellent wisdom but also, on the contrary, has enacted a law to prohibit it. As a result, she has extended the life of his famous words into the 20th century.

Such being the case, I would like to tell you first of all that our proposed system for purifing or utilizing rainwater and waste water/sludge by exploiting the power of the soil has been developed under various strict restrictions imposed by this "bad law". ([3]).

2 Japanese Laws and Regulations Negating Her Good Traditions, and the Development of a New Soil Treatment System Under These Restrictions

Even under the new Japanese law enforced from the 1st of last June, our position, consistently advocated for years, that human waste and bath/kitchen waste water from smaller numbers of persons be treated of jointly was not accepted. The only one restriction relaxed by the law is that joint treatment for 51 or more persons is authorized instead of the previous 100 persons. Consequently, if law-asiding citizens wish to make onsite treatment for a small number (less than 50) of persons, we are obliged to make equipment in accordance with this bad law. In other words, human waste must first be treated independently, and then, the treated product must be re-treated together with domestic miscellaneous waste water from kitchens and baths using additional equipment. This situation is also quite different from that of foreign countries where joint treatment for a small number of people is authorized.

Nevertheless, while maintaining such a strict law for treating human waste, no law for domestic miscellaneous waste water is maintained in present day Japan, where rivers, lakes and seas are abandoned to rapid pollution, to such as extent that parts of them are being called "dead".[4] Since I believe that this miserable situation is already known to many of you experts in waste water treatment, and also that they are not the direct main subject of this paper, I would like to refrain from going into further details.

However, under the above-mentioned situation in Japan, our system mentioned below has hitherto been practiced as follows:

For the Soil-Cover type, ①, any process (e.g., Activated Sludge process) or any equipment (e.q., either aerobic or unaerobic) may be used underneath its cover soil. Whereas, for the Underground Trench type, ②, this process has been used mostly for treating domestic miscellaneous waste water and for the Tertiary Treatment since it is restricted by law.

Thence, those who emphatically supported type ② were cities, towns and villages which had resisted the bad law of the Central Government (the State). Currently, the number of such cities, towns and villages exceeds 60, and their fine results are discussed at the National Diet.[5] Finally, the Ministry of Agriculture, Forestry and Fisheries, which

hitherto had not been at all concerned with waste water treatment has adopted this process for sewerage in rural areas. At present, this process is used for as much as 90% of the rural area sewerage.

In the following, I would like to explain this system by showing you figures.

(By the way, this system is called "Dojō-Jōka" in Japanese, while the magazine in English published by us called the "Do Joker System" is a pun on the Japanese words. So, please allow me to use the term "Do Joker System" in this paper too.)

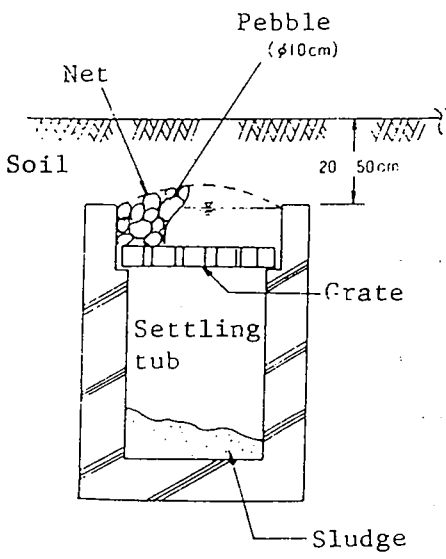


Figure 1.

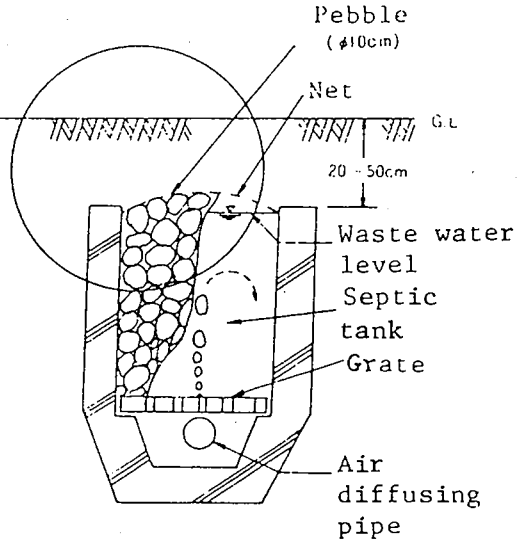


Figure 2.

Note 1. The cover soil shall be of aggregate structure and contain much organic substance.

Note 2. The net mesh shall be fine enough to support the soil but coarse enough to allow soil organisms to pass through easily, and shall be installed as convexly, as possible.

Note 3. As fillers, such natural products as river pebble, volcanic pebble, and crushed stone as well as even plastic waste may be utilized. Their grain size should be 3~10cm.

Note 4. The larger the balance between HWL and LWL, the less excess sludge is produced. If the water-covered portion is made larger as shown in Fig. 4, nitrogen will effectively be removed.

Note 5. In Fig. 1, the filling rate of the filtering material shall be determined according to the nature of the waste water. The bigger the filling rate, the better the decomposing rate of organic substances is but the harder the removal of the excess sludge.

Note 6. The equipment shown in Fig. 2 may provide tertiary treatment by changing the air diffusion method for the 2nd and tertiary treatment. The air diffusing pipe is placed either above or under the grate, but in either case it should diffuse big air bubbles to prevent clogging.

3 Do Joker System as An aerobic Fixed-Film Biological Process

The equipment shown on Fig. 1 is being utilized as a settlement tank, septic tank, sludge condensation tub, sludge storage tank, and a pumping tank, and characteristically generates no scum on the waste water surface. If larger pebbles of 7 ∿ 10cm are selected for filling, the sludge filling the gaps between the pebbles can easily be scooped up by lowering the water level. Usually, the pebble layer is as thick as approx. 50cm.

4 Do Joker System as Aerobic Fixed-Film Biological Process

The equipment shown in Fig. 2 is being utilized as secondary treatment equipment, tertiary treatment equpment, denitrodizing equipment, and purification equipment for river water or other slightly polluted water, and is characteristically of slim structure with the pebble layer as deep as 150 ∿ 300cm and as wide as approx. 100cm. (The slim structure is possible thanks to the non-generation of scum.) A typical example is the rural sewerage of Wadayama Town where it is installed under roads. The grain diameters of pebbles are 3 ∿ 7cm being a bit smaller than anaerobic filtering materials. Pebbles of cbout these sizes are selected because the peeling of the biotic film is easily solved by sending large quantities of air into the diffusing pipe. The clogging problem is usually met by using pipes

diffusing big air bubbles, but, in some cases, by combining an airlift pump or other oxygen supplying method other than the air diffusing pipe system.

For raw water whose BOD is 30ppm or less, the forced oxygen supply system is omitted.

We have consecutively succeeded in the last two years in hatching and breeding salmon fry by purifying polluted river water (approx. 50 ppm BOD) in Metropolitan Tokyo using only the equipment shown on Fig. 2. The hatching/breeding is scheduled to be continued for another five years.

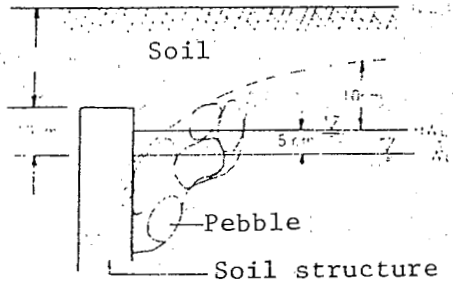


Figure 3. Soil Structure[8]

Table 1. Jidayubori Park River Water Analysis Table

Station: A = Raw Water, B = First Settling Tank, C = Discharge

Date	Station	Trans-parency	PH	BOD	COD	SS	DO	Coliform bacillius
11.Nov.	A	30<	1.5	7	9	9	8.4	940
	B	30<	1.4	13	9	2	7.5	1,100
	C	30<	7.1	9	7	9	8.8	58
18.Oct.	A	4	7.3	6	12	200	5.0	1,000
	B	10	7.4	5	11	52	5.9	1,000
	C	30<	7.4	1	6	8	10.9	0
20.Jan.	A	2.5	7.6	23	51	74.6	7.1	560
	B	12.0	7.5	13	13	40	7.3	520
	C	30<	7.7	3	5	1	11.5	14
12.Feb.	A	12.5	7.1	30	20	62	7.2	2,480
	B	16.0	7.3	25	18	29	7.8	2,000
	C	30<	7.1	3	7	1	9.9	10
19.Mar.	A	30<	7.6	12	23	230	6.3	3,200
	B	5	7.5	13	18	54	6.1	3,300
	C	7	7.6	3	7	0	10.1	6
8.Apr.	A	30<	7.7	17.4	13.6	8	8.8	-
	B	30<	6.8	29.5	14.5	8	1.4	-
	C	30<	7.2	34	7.2	-	8.7	-
7.May	A	30<	7.5	12.0	10.2	8	8.4	-
	B	30<	7.2	9.8	9.2	9	7.4	-
	C	30<	7.3	3.3	5.8	4	9.2	-
9.Jun.	A	30<	7.5	9.7	10.2	8	7.6	-
	B	30<	7.0	8.0	10.4	11	5.0	-
	C	30<	7.5	3.8	7.6	18	9.0	-
6.Aug.	A	30<	8.3	8.2	10.4	9	7.0	21,000
	B	30<	7.6	7.2	10.0	11	5.3	7,700
	C	30<	7.7	3.3	6.8	2	5.6	350
10.Jul.	A	30<	7.9	5.5	11	3	7.7	17,000
	B	30<	7.1	4.5	11	6	8.7	67,000
	C	30<	7.5	2.9	6	-	8.6	130

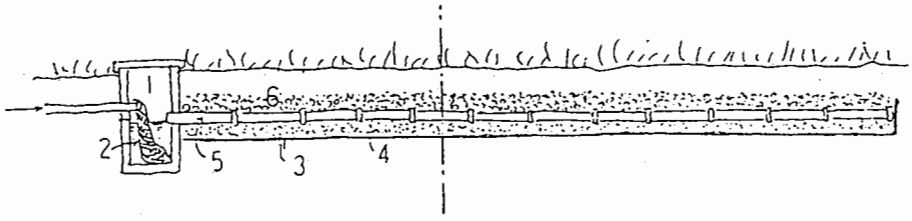


Figure 4-(a) (9)

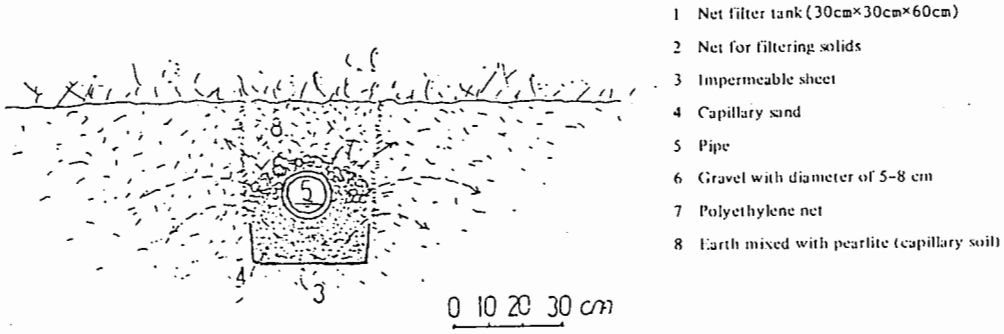


Figure 4-(b) (9)

5 Utilization of Fluctuation in Water Level within Filtering Material of Fixed Film Biological Process

Fig. 3 is an enlarged portion depicting the relationship between the covering soil and the contact filtering material of fixed biotic film under it which is the biggest feature of the Do Joker System.

Now let me explain its features.

The first feature is that the cover-soil layer on the equipment is 20 ∿ 50cm thick and is of continuous structure covering both the inside and otuside of the tank. This is based on an idea for enabling earthworms or other advanced

soil creatures to be most skillfully utilized in the waste water treatment system. Even creatures living in the soil space outside the equipment can be utilized for decomposing the sludge. (See Chapter 1.)

The second feature is that the boundary between the soil and the pebbles is of convex structure. Both ends are sex at positions lower than the wall top. This structure allows the polluted water in the tank to move either into the surface soil or into the soil layer outside the equipment by capilliary action. (This structure will be explained in the next chapter.) Thanks to this structure, we utilize the natural principle that soil organisms and microbes suited to the polluted water to be treated, propagate themselves rapidly in the soil. One example is the utilization of hemolytic bacteria in the soil when treating bloody waste water.

The third feature is the structure which causes fluctuation in the water level within the pebble layer.

Fig. 3 shows a fluctuation of 5cm in the water level illustrating that the water level drops due to capilliary action during the night when the equipment is not in use. However, if designed according to the long canal system, there would be a difference in water levels at the inlet and the outlet due to the resistance of the pebble layer. It would then be possible to have a fluctuation of approx. 10 ∿ 20cm in the water level occur at least twice a day by designing accordingly.

Further, where the water is supplied intermittently by a pump under the trench system shown on Fig. 4, if the pump is installed in the equipment shown on Fig. 2, the LWL can be lowered indefinitely. The boundary face between the soil and the pebbles is made concave for the additional purpose of allowing the septic gas to go through more easily, and of utilizing plant roots (especially root hair) as the carbon source during denitrodization, and of not having the rainwater flow into the tank.

6 Underground Trench Soil Purification System

At a first glance Fig. 4 may be thought to be not much different from the unarmored trench system, but actually it is different. The initial idea of laying impermeable sheet on the bottom of the unarmored trench in order not to have the pollutant permete by gravity is based on the following point.

Organic substances are most effectively decomposed when the three biotas of vegetable roots (GL to - 100cm), soil microbes (GL to - 50cm) and soil creatures (GL to - 30cm) are participating comprehensively. The fear of groundwater pollution cannot be removed by the conventional trench system under which the polluted water and sludge are allowed to enter into the soil locating it more deeply.

Furthermore, since there are many cracks, aqueducts, and big gaps in the soil, the gravity permeatation method by which the pollutant passed through only the big gaps is not appropriate due to the fear of probable ground water pollution. Only a purifying method making use of capilliary action which ensures that absolutely no pollutant goes through big gaps can remove the fear of ground water pollution. Because, being quite different from the gravitation method in which water permeats under positive pressure and saturation, the capilliary action enables the polluted water to pass, at a certain planned permeating speed, through the soil less than 50cm from ground level where the biotic activities are active under negative pressure and unsaturated conditions.

Also, if compared to the sprinkling system which cannot treat much water per unit area, 1m of the trench can treat 100ℓ per day - about five times - thanks to the difficult-to-clog portion of the trench wall near ground level. A detailed scientific explanation of this phenomenon, however, has yet to be clarified. One thing I have never found over the past 20 years of our study is that even polluted water of high BOD density (as high as 1,000 ppm or more) does causes unexpected clogging. I feel this is attributable to action of earthworms or other large-sized soil creatures, and if combined with the re-use for lawns, etc. of the domestic waste water and rainwater, this would become the most practical water treatment/storage system.

The surface of the filtering material shown on Fig. 4 is a fixed film biological process under both anaerobic and aerobic conditions which, if the system is home-sized, has successive water level fluctuations 5 ∿ 6 times per day. Though it has not been fully clarified what role this plays in denitrodization, I presume that, according to the line meter test using the primary treated water (actual results were 95% or more COD, SS and 65% T-N[10]), a method for heightening the removal rate of T-N may be a carbon source supplying system only. Since I obtained a 95% removal rate by adding methyl alcohol, it is a problem in Jpaan[11] as to

capilliary moisturing trenches are laid out at 3 ∿ 4m intervals, no sprinkling will be necessary for the lawn as has been attested in California.

With respect to the relationship between the soil thickness and plants, the utilization of the results of a vegetation study on artificial ground will suffice. Even tall trees may be planted in soil as deep as 80cm. The growing speed of crops on this structure gave a test result of 2 ∿ 3 times of ordinary soil when waste water from a pig farm had been used.( ) No humidity hazard for crops is found whenever the waste water level is GL - 60cm according to the study results.

The environmental pollution preventive function of this structure is acknowledged by all those who have seen its actual results as far as the non-proliferation of odor, bubbles, human pests and other readily identifiable effects are concerned. However, the function acknowledged as most practical is the non-proliferation measures taken against pathogenic bacteria (viruses) and NOx which are an invisible environmental pollution problem hard to detect by the senses.

While the conventional septic activated sludge process system needs bubble-preventing devices, this system needs no such thing. Further, diffusion into the air of fine droplets caused by exploding bubbles on the water surface is simply and completely solved by aereating the soil.

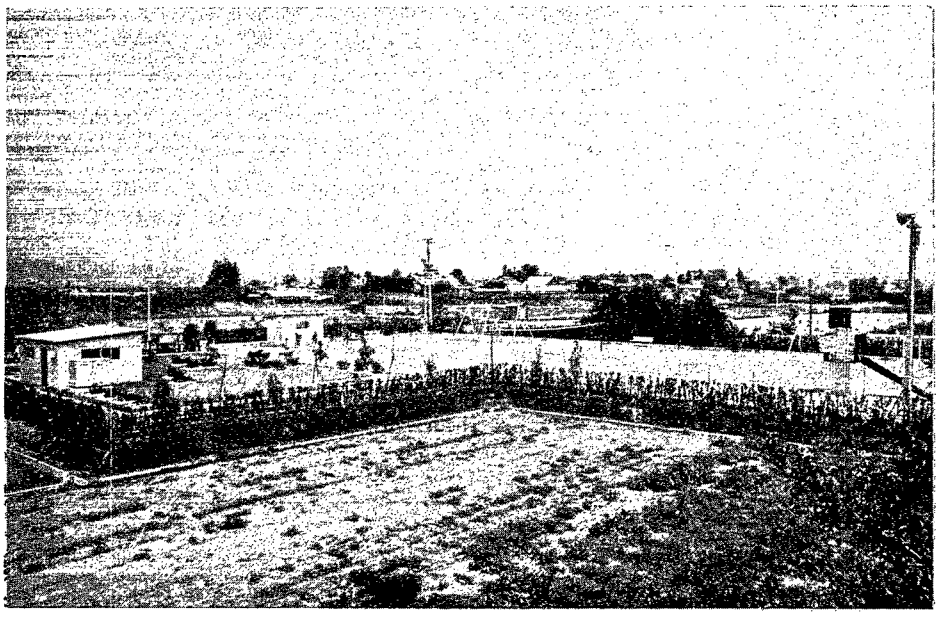
Furthermore, the over-jeneration of NO_2 and the fear of NO_2 diffusion into the air with consequent adverse affects or the human body, both of which are the biggest demerits of the F.F.B.P., can be solved through absorption by the soil and oxidization into NO_3. The biggest reason why the Do Joker System is used in 90% or more of the rural sewerage systems in Japan is the completeness of the environmental pollution preventative measures as such. Their completeness is attested by actual examples of its use under a busstop waitingroom, the lawn of an outdoor eating place, a road, a flower bed in front of a railroad station, and in the middle of a housing complex. Its deodorising system needs no excessive power, activating carbon, acid, alkali or heating. Odors from sludge treatment equipment, the covered chamber of the rotating biological contactor process, and the filtering bed of the sprinkling water could be solved by the Do Joker System under which the air is pressurized into gaps in the pebble layer at a pressure as low as that of a ventilation fan. The design speed is $300m^3$ per m^2 of ordinary soil. Before this amount is increased to $1,000m^3$, the soil

how to increase the removal rate using natural soap instead of synthetic cleansers.

7 Environmental Measures and Environmental Pollution Preventive Function

The point where the Do Joker System a differs from all other waste water treatment technologies is that environmental measures can be reasonably combined with it, and at the same time, it can so completely prevent environmental pollution that no maintenance costs are caused.

Firstly, with respect to environmental measures, the surface of the equipment may be utilized. The point in common between Figs. 1 through 4 is that a soil layer as thick as several tens of can covers the equipment. Therefore, if care is taken to grow plants over it, the facility itself will become a green area necessiating no buffer green zone around it usually, a thin soil layer needs sprinkling

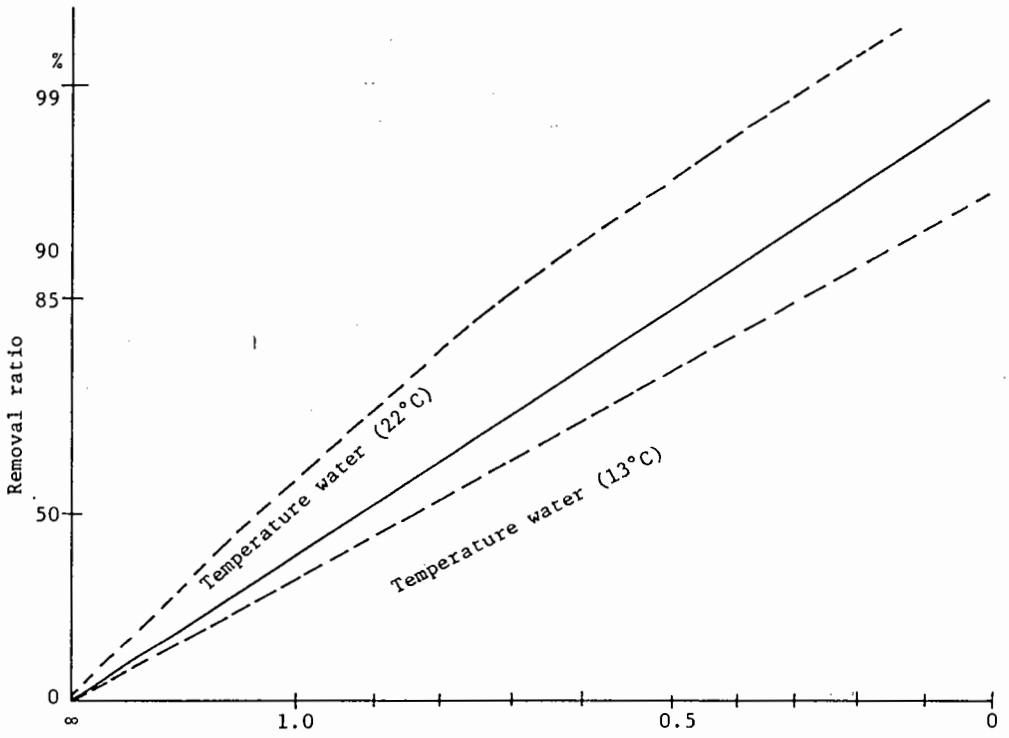


with wates for plants to grow. This structure, however, needs no sprinkling at all because the waste water level is several tens of cm below and its surface is designed so as to have water supplied by capilliary action. Usually, the capiliary water extends as far as approx. 200cm. Then, if

nature shall be determined by experiment. Anyway it is necessary to satisfy such contradicting functions as aerability, water preservation and absorptive capacity. Since they vary greatly according to the soil, construction is currently performed by determining these factors through individual experiments. Care not to make the soil too dry, and the technique for mixing perlite, comport, etc. are important.

8 The Production of Excess Sludge

A feature of the trickling filter method is that it produces less excess sludge than the activated sludge method, the contact aeration method, or the rotating biological contractor method. If, in addition to this normal feature, the F.F.B.P. filter medium in the soil and under the water surface has a continuous structure as in the Do Joker System (Figs. 1 to 4), the question is, what biota will be formed? This is not yet understood in detail. It is said that in the F.F.B.P. large-sized Metazoa, which are not found in activated sludge, live in the membrane to form a wide variety of biological groups. The formation of various biological groups can be easily hypothesized because a net-covered soil layer of 20 to 50cm thick is over a gravel layer, which is on the fixed biomembrane in water and about 20cm above the surface, offering soil organisms an area for living. Only a report of rise and fall of soil organisms within the installations was presented by a Japanese researcher of earthworms, Yoshio Nakamura, at the Darwin Centenary Symposium on Earthwoarm Ecology in cumdria U.K. last year.[12] But the report shows a profile of soil animal ecology different from that commonly thought, and helps us to understand an aspect of the complicated ecology. When there is a shortage of food, earthwoarms pass through the gravel layer connected with the soil, reach the water, take in activated sludge as food there and return to the soil. Such an earthworm habit would be suited to the configuration of the fixed biomembrane. Installations with the biomembrane are regarded as those utilizing most effectively the soil animals under natural ecosystem. Though the production of excess sludge is not estimated to account for the proportion to the amount of eliminated BOD, the amount of excess sludge produced in 132m-long Do Joker System of rural sewage buried under the farm road of Wadayama with a population of 250 persons over 36 months was $18m^3$ (moisture contents: 98%).[13] In a cold area,



Note: 1. A solid line is drawn based on the results of experiments using 10cm-diameter gravel and 2.2cm-diameter gravel the line is modified slightly downward.
2. The results of this construction method are about 1,000 of domestic waste water. BOD volume load fell mainly between 0.3 and 1.0kg/m^3/day.

Fig. 5 BOD, BOD volume load and removal ratio of soil aerobic (10cm-diameter gravel)[15]

Haguro town, Yamagata Prefecture, excess sludge has not been taken out yet as long as May, 1979 after installing.

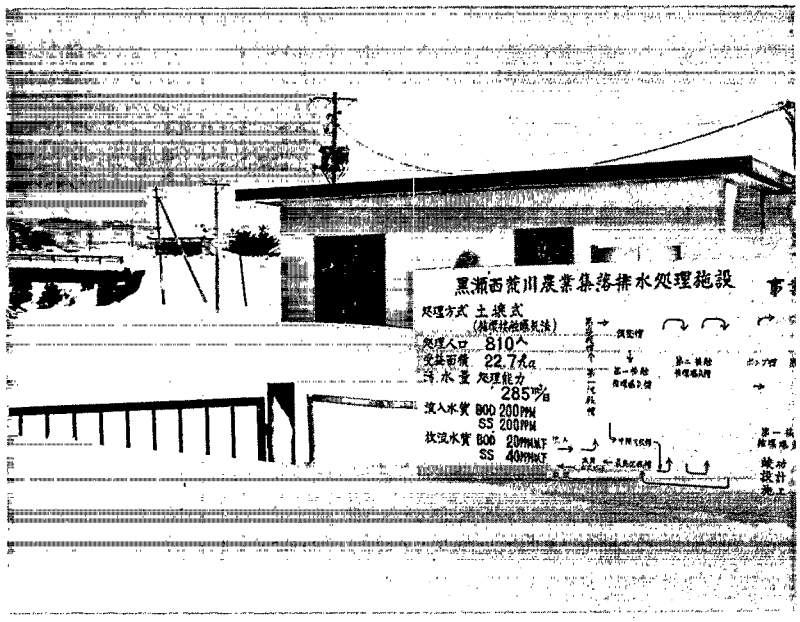


Table 2 [14]

Sampling date (1980)

Water temperature	17, Feb.	4, Jun.	19, Aug.	7, Oct.	6, Nov.
Transparency	30 or more	30 or more	30 or more	30 or more	30 or more
PH	7.2	7.0	7.2	7.4	7.4
DO	9.78	8.05	8.73	10.41	10.88
BOD	3.91	7.53	5.5	5.40	0.52
COD	8.7	21.88	12.2	11.6	8.7
T-N	(0.58)	(0.75)	(0.62)	(0.69)	(0.51)
(organ)	12.18	17.54	12.93	10.25	7.69
T-P	2.51	4.52	3.54	3.51	3.04
SS	1.0	7.5	10.0	4.0	3.6
+	+	+	+	+	+	+

Quanlity of discharged water from Haguro agricultural village sewer system

Note: 1. (organ) represents (Kj-N minus NH_4-N)

The long and narrow waterway of installations allows the filtration ability of the F.F.B.P. to work fully. The reduction of the amount of suspended solid (SS), rapid settleing speed, high rate of conversion to inorganic substances, promoted possibility of denitrification have been attained by further expanding the features of the F.F.B.P., depending on the soil.

9 Selection of Filter Mediums for Use in the Fixed Biomembrane Method

Industrially produced filter mediums (manufactured plastic products) are mainly used in Japan, and natural gravel is now used only at the author's laboratory and the River Bureau of the Ministry of Construction which is studying low level water.

Gravel includes river gravel, crushed stones, rapilli and slag. Rapilli has the greatest efficiency among these because of its high porousness and rough surface. The point of using natural gravel is to employ less industrial products from the standpoint of saving oil and energy, it is not an attitude of using no industrial products at all. Drawbacks in using natural gravel are malodor generated from sludge accumulating in the pores of gravel, and clogging. The problem of malodor has been solved completely by employing the soil-covering structure described above, but clogging of the gravel layer still remains to be solved. Therefore, the author's studies have concentrated on clogging, and the following two results have been obtained:

(a) Selection standards for gravel size of filter mediums

The author's experiences show that 7 to 10cm diameter gravel should be used for primary treatment and anaerobic filter mediums receiving high levels of sludge, and 3 to 7cm diameter gravel for secondary and tertiary treatment, and aerobic filter mediums receiving waste water containing no toilet paper so that clogging materials can be easily removed by increasing flow rate, bubbling, and lowering the water level. For those receiving river water and turbid water containing inorganic materials, it is thought to be simpler to replace the clogged gravel completely rather than to wash it. When this construction method is used in river construction, the diameter of gravel may be 1.0 to 2.0cm or more.

(b) Greater accumulations of sludge mean a wider variety of biota

The scheme for the biomembrane of a trickling filter is often used to explain the self-purifying function of the F.F.B.P. An anaerobic biomembrane is near the surface of the filter medium and an aerobic biomembrane is on the former membrane. The aerobic biomembrane in this combination eliminates malodor, making it practical for use. However, when gravel is used for the submerged biofilm, anaerobic sludge accumulates in places where the aerobic biomembrane should be formed and breaks the aerobic biomembrane to generate malodor, so that the method using gravel is absolutely unpractical. This fact is coincident with historical fact, in which the trickling filter method using gravel was replaced by the activated sludge process because the trickling filter clogged with floating sludge producing malodor. When techniques to solve the malodor problem are not available, the development of manufactured filters on which sludge hardly accumulates will play a leading part.

In contrast to the above method, the Do Joker System completely solves the malodor and sanitary pest problems and it is concluded that gravel on which sludge which might increase the variety of biota accumulates should be used to facilitae purification of waste water, self-disintegration and denitrification.

10 Advanced Treatment and Countermeasures to Trihalomethanes

The aerobic fixed biomembrane method is used for secondary treatment. But, when the long waterway system shown in Fig. 2 is employed, less BOD volume load results in the increase in removal ratio of BOD, SS, fat and fatty oil, and ABS, so that the system can be used for tertiary treatment. A removal rate of 95% can be attained by making BOD load 0.5kg/m^3/day.

The system in Fig. 2 was used for purification of polluted river water with BOD of 50ppm according to the results of experiments at Nogawa, Tokyo (1,500m^3/day). Three handred thousand young salmon were successfully hatched and reared with treated water, and it was shown that the system was very effective in removing SS, fat and fatty oil, ABS and colon bacilli as well as BOD.

The system produces treated water with very high transparency and no need for chlorination, so that it is highly regarded as a countermeasure to trihalomethanes along with the capillary saturation trench system.

In my edited book entitled "Do Joker System -- Lectures" and published in December 1980, I used the expression: "death riding on a pale horse" in conjunction with the formation of trihalomethanes by chlorine sterilization. This is because the Greek word Khloros for "pale" in the "pale horse" which, in the Book of Revelations of St. John "was allowed to kill people" is the origin of the word chlorine. So I quoted two books entitled "Pale Horse' written by Ropsin and "Look at a Pale Horse" by Hiroyuki Itsuki these are wellknown novels in Japan, (in Japan "Pale Horse, Pale Rider" written by Katherin Ann Porter, is not very famous) and I appeal to my readers that not ride on a Pale Horse.

11 Flow Sheet of Ideal DO JOKER SYSTEM

Necessary elements of an ideal treatment system are simplicity, low construction costs, easy maintenance requiring no special techniques, production of a small amount of excess sludge, perfact environmental protection, perfect pollution prevention, good treated water, and also complete treatment within a site.

The following two systems have been developed to meet the above demands:

(a) a combination of the soil settling filtration tank (Fig. 1) and capillary see page trench system (Fig. 4) uses alternately two 2m-long trenches per head, requires an area of 4 to 8m^2 per head, but needs no aeration power,

(b) a combination of the soil settling fitration tank (Fig. 1) and soil contact aeration tank requires no flow regulating tank, sludge accumulation tank or final settling tank, and needs an area of 0.5 to 1.0m^2 per head. The past long-period records of the systems, (a)=table 3 and (b)=table 4 will be shown in the following tables. (16) (17)

The records are shown in Tables 3 and 4. Table 3 shows the records at Shin-Matsuda, and Table 4 at Okutama. (These records will be rearranged.)

In the development of Do Joker System, the obtained data was far from an estimate because the anaerobic fixed biomembrane method using gravel was employed for the settling

filtration tank. The facility is the same as a part of the installation shown in Table 2, and very distinctive results were obtained, so.....

Table 3. The records of water treated by soil anaerobic fixed biomembrane method (settling filtration tank) and capillary saturation trench system (ppm) (Life Research Report, No. 11, page 81)

Water sampling \ 1978 year / Analysis items		Oct.'78	Nov.	Dec.	Jan.'79	Feb.	Mar.	Apr.	May	Jun.	Jul.	Aug.	Sep.
S1	Temperature of water	17.0	12.0	12.4	9.8	10.2	12.5	17.2	16.1	18.0	24.0	23.2	19.6
	pH (ppm)	6.0	6.6	6.5	6.1	6.8	7.0	6.4	5.3	6.5	6.4	6.1	6.9
	COD (ppm)	760	180	270	310	70	89	210	4,300	170	190	200	240
	BOD (ppm)	1,100	340	580	620	170	290	410	8,300	400	280	590	490
	SS (ppm)	3,100	920	440	1,200	73	93	1,000	19,000	730	530	290	650
	DO (ppm)		2.4	3.3	3.0	7.5	2.7	2.5	0	0	5.2	0.052	1.5
	No. of colon bacilli (No./mℓ)	170×10^3	130×10^2	120×10^3	300×10^3	94×10^2	52×10^3	75×10^3	250×10^3	100×10^3	140×10^3	230×10^3	100×10^4
	No. of common bacteria (No./mℓ)	55×10^4	89×10^4	220×10^4	99×10^4	210×10^3	45×10^4	300×10^3	97×10^4	46×10^4	150×10^3	150×10^4	110×10^4
	Chlorine ion (ppm)					21		0.76		38		70	
	Ammoniacal nitrogen (ppm)	23		48		3.4		4.2		26		14	
	Total nitrogen (ppm)	320		93		17		59		52		27	
	Nitrite nitrogen (ppm)	0.060		0.094		0.039		0.28		0.0063		0.021	
	Nitrate nitrogen (ppm)	6.0		1.5		0.70		0.93		0.1		0.16	
	phosphoric phosphorus (ppm)	21		7.8		1.8		3.3		5.4		2.7	
	ABC							2.2		2.8		2.0	1.6
	Total phosphorus (ppm)	37		24		8.3		36		26		6.1	

Table 3 (Continued)

Water sampling	Analysis items \ 1978 year	Oct.'78	Nov.	Dec.	Jan.'79	Feb.	Mar.	Apr.	May	Jun.	Jul.	Aug.	Sep.
So	Temperature of water	18.2	14.5	14.5	12.7	13.0	14.5	18.5	15.0	18.7	23.5	23.5	20.3
	pH (ppm)	6.5	6.7	6.6	6.7	6.8	6.6	6.5	6.4	6.7	6.9	6.8	6.9
	COD (ppm)	92	100	73	39	55	82	97	55	39	58	53	68
	BOD (ppm)	230	280	190	100	120	120	270	130	40	130	120	86
	SS (ppm)	44	32	63	25<	55	47	78	29	25<	33	25<	41
	DO (ppm)		0	0	0.74	0	0	0.84	0	0	0.6	0	0
	No. of colon bacilli (No./mℓ)	77×10^3	71×10^3	46×10^3	140×10	89×10^2	140×10^2	47×10^4	92×10^3	77×10^3	110×10^3	52×10^3	75×10^3
	No. of common bacteria (No./mℓ)	110×10^3	83×10^4	300×10^3	54×10^3	300×10^2	120×10^3	140×10^4	150×10^3	210×10^3	35×10^4	110×10^3	200×10^3
	Chlorine ion (ppm)					63		61		42		57	
	Ammoniacal nitrogen (ppm)	29		41		52		46		27		38	
	Total nitrogen (ppm)	38		49		61		58		32		44	
	Nitrite nitrogen (ppm)	0.017		0.013		0.019		0.028		0.025		0.011	
	Nitrate nitrogen (ppm)	0.10		0.38		<0.1		0.13		<0.1		0.21	
	Phsophoric phosphorus (ppm)	5.7		6.3		7.2		4.8		4.2		5.0	
	ABC					3.1		2.6		3.1		3.9	
	Total phosphorus (ppm)	21		18		29		30		16		6.5	

Table 3 (Continued)

Water sampling	Analysis items \ 1978 year	Oct.'78	Nov.	Dec.	Jan.'79	Feb.	Mar.	Apr.	May	Jun.	Jul.	Aug.	Sep.
W_2	Temerature of water	19.9	17.4	15.4	11.9	10.4	11,5	13,5	17,6	18,5	20.2	22.2	18.7
	pH (ppm)	6.7	6.6	6.6	6,9	7,0	6,6	6,4	6,4	6,6	6,7	6,2	6.4
	COD (ppm)	4.5	4.4	3.4	2.9	3.6	3,1	3,8	2,7	2,4	3,0	4.3	11
	BOD (ppm)	2<	2<	2<	2<	2<	2<	2<	2<	2<	2<	2<	2<
	SS (ppm)	25<	25<	25<	25<	25<	25<	25<	25<	25<	25<	25<	25<
	DO (ppm)		7.0	5.1	6.9	8.9	7.9	8.9	7.0	6.5	6.8	5.4	4.4
	No. of colon bacilli (No./mℓ)	7	11	1	13	0	6	0	51×10	0	2	120	270
	No. of common bacteria (No./mℓ)	140	113	76	250	20	80	51	80×10	2	110	32×10	67×10
	Chlorine ion (ppm)					60		47		35		23	
	Ammoniacal nitrogen (ppm)	<0.13		0.20		<0.13		<0.13		<0.13		<0.13	
	Total nitrogen (ppm)	70		62		30		41		26		40	
	Nitrite nitrogen (ppm)	0.037		0.005		0.005		0.005		0.0054		<0.005	
	Nitrate nitrogen (ppm)	68		61		30		41		26		40	
	Phsophoric phosphorus (ppm)	0.029		<0.004		0.014		0.028		0.016		0.014	
	ABC					<0.08		0.49		0.19		0.28	
	Total phosphorus (ppm)	0.45		0.10		0.30		6.0		0.23		0.21	

Note:
1. Domestic waste water from 4 farmhouses, Actual population 21 persons.
2. Measured by the Ministry of Agriculture, Forestry and Fisheries (as described in the second chapter, combination measurement for less than 50 persons is usually prohibited, but this governmental experiment was conducted under the responsibility of the Ministry of Agriculture, Forestry and Fisheries.
3. Excess sludge was removed every three years without using any power.
4. A two-meter-long trench per head was clogged three years after installation. An additional 2m-trench is planned for alternate use, but at present recovery of the existing trench is being investigated.
5. A feature of the soil settling filtration tank is to use rubber sheets and beer transporting containers filled with gravel for the inner contract filter medium.

Table 4. BOD in water treated by soil aerobic F.F.B.P. and capillary saturation trench system (ppm)

Date	BOD (mg/ℓ)			Total phosphorus (mg/ℓ)			Total nitrogen (mg/ℓ)		
	Inlet of circulated contact aeration tank	Outlet of circulated contact aeration tank	Soil leachate	Inlet of circulated contact aeration tank	Outlet of circulated contact aeration tank	Soil leachate	Inlet of circulated contact aeration tank	Outlet of circulated contact aeration tank	Soil leachate
1978									
Aug.19	402	9.4	0.8	3.7	2.6	0.2	35	32	17
Sep.10	180	0.7	0.4	6.9	3.4	0.1	60	30	22
Oct.16	510	1.1	0.4	2.5	2.1	0.0	13	24	22
Nov.18	288	1.7	0.7	2.1	1.6	0.0	12	24	18
Dec.10	164	0.6	0.0	2.6	-	0.0	79	-	15
1979									
Jan. 8	46	1.2	0.8	5.0	-	0.0	68	-	17
Feb.13	35	6.0	1.2	5.7	-	0.0	59	-	15
Mar.11	32	1.3	0.6	4.3	-	0.0	55	-	12
Apr. 9	309	4.5	1.6	2.0	0.4	0.0	10	4	11
May 14	68	2.2	2.1	5.5	-	0.0	80	-	3
Jun. 4	135	2.8	0.8	5.8	-	0.0	124	-	9
Jul.14	98	2.8	1.0	4.2	2.9	0.0	64	46	18
Aug. 6	-	-	0.2	-	-	0.0	-	-	-
Sep.10	322	7.4	0.4	4.6	3.5	0.2	43	30	14

Note: 1. Domestic waste water from the museum (stand, dining room, public lavatory), estimated 300 persons.
2. Measured by the Waterworks Bureau of Tokyo. (according to the Water Pollution Control Agreement)
3. At first, BOD at the outlet of the contact aeration tank was estimated as 60ppm, but only the activated sludge method was available in Japan at that time, so that volume load was calculated based on the activated sludge method. The removal ratio was unexpectedly high, these results and those shown in Table 3 led to the gravel filter medium being highly regarded.
4. This installation is a combination of soil settling tank (Fig. 1), soil contact aeration tank (Fig. 2) and capillary saturation trench system (Fig. 4), and was constructed in 1978.

12 Conclusion

Examples of application of this process in Japan include both anaerobic and aerobic F.F.B.P. as stated above, and the process is applied not only to primary and secondary treatment but also to tertiary treatment equipment.

Further, the recharge of treated water underground can be carried out by is same equipment simultaneously and without any other special equipment. Similarly, some equipment handles the use of treated water for plants besides treating sewage.

With this process, the beauty of a flowering plant which impresses people who look at it is not impaired in the least by foul smells or viruses from the equipment. So, guests enjoying their meal in a hotel garden on a summer night are unaware that their own excreta washed away from their rooms during the daytime are being treated under their feet -- only a few score centimeters from the ground surface.

This process is used for sludge treatment as well as sewage treatment. The faicility shown in Fig. 1 of this article is used as a sludge concentrating tank (this alone is a mere storage tank) and the equipment in Fig. 4 is installed as a supernatant liquid treating facility and surplus sludge is treated by the combination of the two. In the ordinary sense of storage, feeding is no longer possible when the container is full. But with equipment used for this process, this is not "the end of the world" but it is just the beginning, because it is provided with the capillary seepage trench of Fig. 4. In other words, the feeding of surplus sludge is continued everyday even when the tank is full. Naturally, the supernatant liquid that overflows everyday is equal in quantity to the sludge that is fed in. This supernatant liquid is treated by the capillary seepage trench and the concentration of sludge continues in the main tank.

In the case of some equipment in Shibukawa City, Gumma Prefecture, 477m^3 of surplus sludge was fed in and 120m^3 was applied to mulberry fields as a liquid fertilizer during the two months from May 1979. This means that the sludge that was fed in was concentrated to about 1/4 by the capillary seepage trench. The city, which handles night soil treatment for about 60,000 people, has installed four machiens since the installment of Machine No. 1 in 1979. When three more are constructed, bringing the total to seven, it will no longer have to use expensive petroleum to incinerate surplus sludge.

F.H. King, who was quoted in Chapter 1, refers in his two papers quoted by the author to Dr. Oskar Kellner's analysis conducted in Japan about a century ago on the fertilizer composition of human waste. Tadashi Niimi, developer of this process, graduated from the university at which Dr. (Kellner) had taught long after he had left. At the same university, the author's father was taught by Dr. Masuji Akiba about the capillary siphon movement of water in the soil and developed the equipment shown in Fig. 4 of this article about 15 years ago. With a chracteristically Japanese sentiment, the author cannot but see strange ties among those three persons to whom the same university was the stage.

Since the main theme at this conference is the Fixed-Film Biological Process, the author did not mention the fact that rain water is recharged underground by the Do Joker System. The successful percolation of as many as 41 tons/day of rain water by a trench of only 10m is the clearest proof that, whereas vertical percolation of water is difficult due to fill-up, water seeps in the horizontal direction without any fill-up. In his January 1982 letters to the Minister of Construction and the governor of Kanagawa Prefecture, the mayor of Zama City of that prefecture reported that he would adopt the above-mentioned Do Joker System for the combined purpose of underground nourishment by rain water, countermeasures against river flood and overflow, and prevention of land subsidence by excessive pumping, taking advantage of the rain water percolator installed in the relatively shallow ground. "Relatively shallow ground", as referred to in these letters, is a factor that is most important to the Do Joker System. (One will do well to recall that digging a deep well and injecting rain water deep undergound with the object of recharging it to the ground is a common practice everywhere in the wrold.)

To appeal the importance of this "relatively shallow ground", the author and my group call this part soilsphere or pedosphere -- rather than simply calling it soil in as much as they regard it as the "abode of living things" with the greatest biological density on the earth.

In this sphere, different forms of living animals, micro-organisms and plants mix together and, with the participation of sewage, "a perfect circulation of the forces of nature", as first quoted from Dr. Maron, comes into existence. This "links of the chain" is still a mysterious world which, regrettably, has not yet been sufficiently clarified. Out process is named the "Do Joker System" as a

pun on the Joker of the playing cards and the Japanese words "Dojō Jōka" (soil purification) referring to our process. It is hoped that this announcement by the author will serve as an opportunity for people particularly in the sectors of civil engineering and sanitary engineering to become interested in the "Links of the chain" in the pedosphere.

It is regrettable that this article could not describe details as the emphasis was placed on the wide-ranging application of our system. As regards contact materials, for instance, the article could not cover the use of empty cans instead of gravel or industrial wastes smaller than water in specific gravity or seawead-like strings moving freely in the sewage to improve existing facilities, e.g., aeration tank. We hope that detailed reports on these can be presented in the future.

Reference

1. Wagner, W., Die Chinesische Landwirtschaft, 1926
2. Niimi, M., "Do Joker Process" Do Joker System, Vol. 6, No. 10, Sep. 1981 (in English), p. 20∿21.
3. Notification No. 1292 of the Ministry of Construction.
4. "The Yomiuri Shimbun", 22, Feb., 1982.
5. Sepcial Committee on Environment of the House of Representatives, 22, Nov. 1977.
   Same Committee of the House of Councilors, 22, Mar., 1978.
   Same Committee of the House of Representatives, 13, Jun., 1978.
   Question to the President of the House of Councilors from a member, 19, Nov., 1980.
   Answer to the President of the House of Councilors from the Prime Minister, 28, Nov., 1980.
6. Niimi, T., Arimizu, T., Soil Purification of Sewage--General, Do Joker Center Ltd., Oct., 1977, p. 35.
7. ibid.
8. ibid., p. 36
9. Yahata, T., "Wastewater Treatment Through Surface Soil" Do Joker System, ibid., p. 5.
10. Amada, T. et. al., "The Application of Underground Piping Method to the Field of Water Purification Civil Engineering Journal, Public Works Research Center, Vol. 23, No. 10, Oct., 1981, 0. 21.
11. Studies by Aida, T., University of Ibaragi.
12. Nakamura, Y., "Colonization by Earthworms of Niimi Waste Water Treatment Trenches".
13. Namikawa, J., " New sewage system between men, earth and water" Quarterly Garden City, Japan Institute for Community Affairs, Vol. 3, No. 1, Jan., 1982, p. 28∿32.
14. Suzuki, N. et. al., The Report from the Yamagata Prefectural Institute of Public Health, 1981.
15. Yano, Y., "Application of Self-Purification to River Water Quality Control ", Journal of Water and Waste, The Industrial Water Institute, Vol. 24, No. 1, Jan., 1982. p. 21.

16. "Rural Life Research ", General Research Center for Rural Life, No. 11, Mar., 1980, p. 81~82.

17. Inchinohe, M., "Renovation of Weste Water Effluent by irrigation of Forest Land", Journal of Water and Waste, ibid., p. 22.

A NEW FIXED-FILM SYSTEM COVERED BY SURFACE SOILS

Tsutomu Arimizu, Forestry and Forest Products Research Institute, Ministry of Agriculture, Fishery and Forestry Japan

INTRODUCTION

The acceptance of a new fixed-film system covered by surface soils have been in progress last ten years for the treatment of a wide-range of low and high strength of biological wastewater, because in its very simple process removal rate of BOD,COD, total nitrogen and SS are extremely high with very few sludge production and few input of particular energy, and without daily operation and maintenace effort and skill, in the same area of the conventional wastewater treatment processes.

DEVELOPMENT OF THE SYSTEM

It is interesting to note that the prototype of this system with the name of Do Joker System(hereafter it will be abbreviated to DJS) came from studies of drain field which we call trench and has been very common in the United States.

In a DJS aerated surface soils are made much use of with capillary water having the mean infiltration rate of 0.65 gpd/sq.ft in the case of the eastern United States soils(1). One of our succesful experiments that has been carried out at a

public waste disposal site in Gifu city showed that wastewater there with BOD of 20,000 mg/1 was purified to 2 mg/1 by this trench. In many trenches which are working well sludge disposal has not been made more than last ten years. However, I would like to call your attention that our trench as shown in Fig.1 is quite different from darin field in some essential portions.

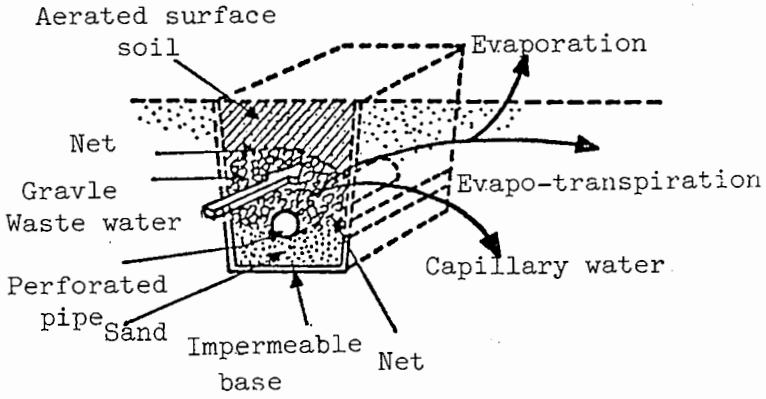


Figure 1. Typical Trench system

After the stage of trench DJS advanced to replace a septic tank as shown in Fig.2.

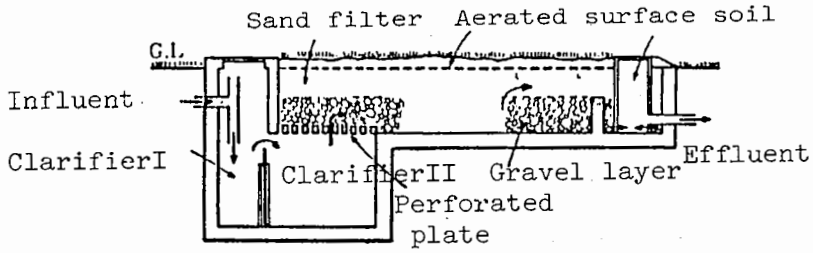


Figure 2. Small DJS

Along with them DJS was developed to provide more than tertiary wastewater treatment by adding it after activated sludge process to reduce BOD average from 20 mg/1 to one or two mg/1, or total nitrogen and SS average to one or two mg/1, eliminating phosphorus by making use of trench simultaneously that was added after the process, in order to meet the heavy requirements set to control water pollution in a number of lakes in this country.

It is since 1978 when on the basis of more than 20,000 cases of experience including trench DJS reached the stage to aim an independent and large scale facilities in wastewater treatment under a program of the Ministry of Agriculture, Fishery and Forestry to introduce sewage system to local large farmers' community.

STRUCTURE OF PRESENT DJS

This system is one of the packed-bed processes covered by surface aerated soils so that fixed-film system can develop its potential ability completely through its good contacts with gas and liquid.

Temperature

In the beds of irregular and randomly packed granular particles, heat conductivity is much large than that of liquid and the effective conductivity of it is an average with few deviation, being independent of radius of the particles. The aerated surface soils over it can contribute to not only keep in it high temperature essential for fixed-film system under extremely cold winter, minimizing thermal fluctuations and absorbing offensive odor, but also supply soil organisms having strong capacity to purify wastewater, eliminate pathogens and digest sludges, to the filter, preventing emission and airborne spread of pathogens.

When outside ambient temperature was $1^{o}C$, the effluent temperature was $10^{o}C$ in a case to be mentioned here.Everything outside was frozen, DJS could work well.

Ecological composition

Although DJS has a simple structure, it provides an attractive habitate for a more wide-range of microorganisms and for some animals than others.

The bacterial flora in a DJS consists of both Gram-negative bacilli derived from wastewater and Gram-positive from soils which are generally much more active than Gram-negative. They compete directly with fungi.

Fungi are also present in the filter beds and occasionally dominate the primary stage of the process. But population of bacteria and fungi is controlled by protozoa living in the beds which compete with nematodes and rotifers in the secondary stage of the process. These small metazoa sometimes harvest bacteria and fungi while processing the solid

materials. Mold mites are also present in it which feed on fungi in localized anaerobic portions of the process but they are preys of beetl mites and springtails. Large metazoa are present in it. They display functional roles in the recycling and communications of all types of organic debris. Adult flies of many types and species of beetles are the transporting agent of bacteria, fungi, protozoa and mites. At the tertiary stage of the process earthworms flourish as the first or second descomposers in waste materials including sludges(2).

In this way the ecological community of DJS has much more complicated and rich food chains including not only aquatic but also terrestrial microorganisms which will lead to maintain or increas overall stability(3). Considering the case of bicultural, two-stage, high-rate activated sludge process which consists of a simple food chain mainly between bacteria and protozoa, this highly developed ecological composition of DJS contribute to extremely small sludge production as well as excellent and stable effluent quality from the process(4).

Hydraulic conditions

Another typical feature of a DLS consists in hydraulic conditions.

At first, clarifier ahead of contact aeration tank or contact basin holds the maximum quantity of liquid between doses, which reaches a filter over the clarifier with bottom feed as shown in Fig.3. Then each cycle provokes a chain reaction of flow, down the filter and clarifier, smoothing out any variations in BOD loading and above all eliminating scums.

A filter of basin which comes next to the clarifier is similar to the anaerobic filter with bottom feed and is completely submerged by dosing in the waste which reaches also aerated surface soils as in the case of clarifier through synthetic net which prevents fall of soil particles into the filter as shown in Fig.4. Then immediately capillary upflow takes place in the same way with that of trench. Between dosing liquid flows down through the filter to make aerobic conditions there with reduced loading, which will prevent clogging and increas plant capacity and efficiency.

In the upflow and downflow packed-bed, liquid does not completely cover the outer surface of the porous media with biofilm and the part covered by gas contributes to reaction through absorption and desorption, producing aerobic and anaerobic conditions there. In the beds a radial variation in resistance to flow may cause appreciable maldistribution and

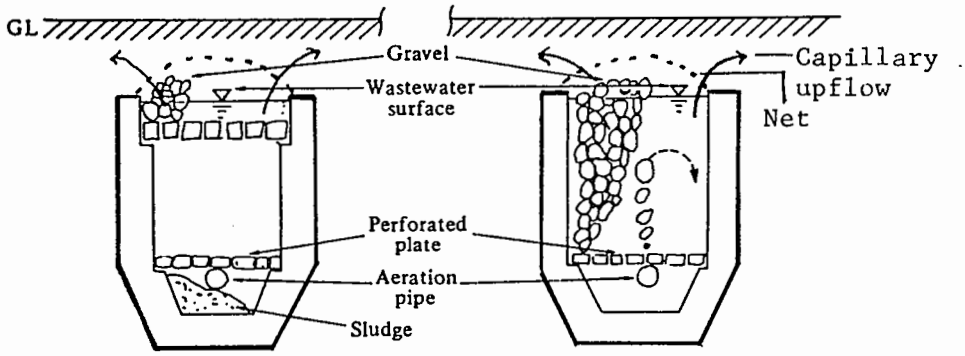


Figure 3. Clarifier

Figure 4. Contact basin

cross-flow of repetitive aerobic and anaerobic cycles in time and space, without input of particular energy. Intermittent application of liquid in a DJS is to provide alternate periods of aerobic and anaerobic conditions everywhere in the attached growth reactor all the times(5).

Filter media

Filter media used in a DJS are crushed stones, blast furnace slag, discarded cans and synthetic products which are specially manufactured for wastewater treatment.

DESCRIPTION OF RECENT DJS

The cross-section and horizontal section of a recent DJS constructed in 1979 at Haguro-cho, Yamagata prefecture, northern part of Japan, by the Ministry of Agriculture, Fishery and Forestry, are shown in Fig.5 and Fig.6,respectively.

The feed to the filter comes from a nearby community with the population of 700 and a wastewater flow of 150 m^3/day (39,600 gal/day) with a strength of 200 mg/l BOD_5 and TSS has been treated. Recirculation has not been used. They do do not use oxygen gas. The total volume of all filters are 714.5 m^3 (188,770 gal/day) with the total filter surface area of 20,403 sq.m (219,536 sq.ft). The results of chemical analysis at each filter are shown in Fig. 7 when the retention time was 60 hours in total, on August 11, 1981.

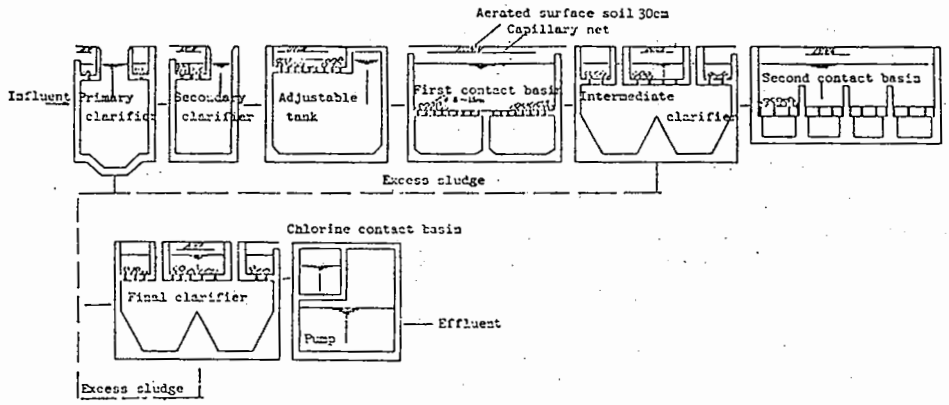


Figure 5. Cross-section of DJS at Haguro

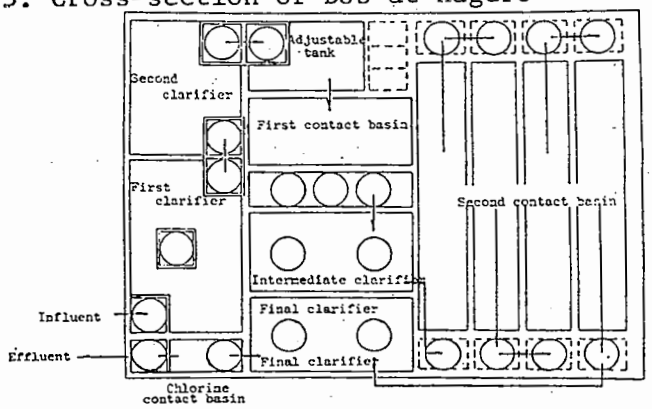


Figure 6. Horizontal section of DJS at Haguro

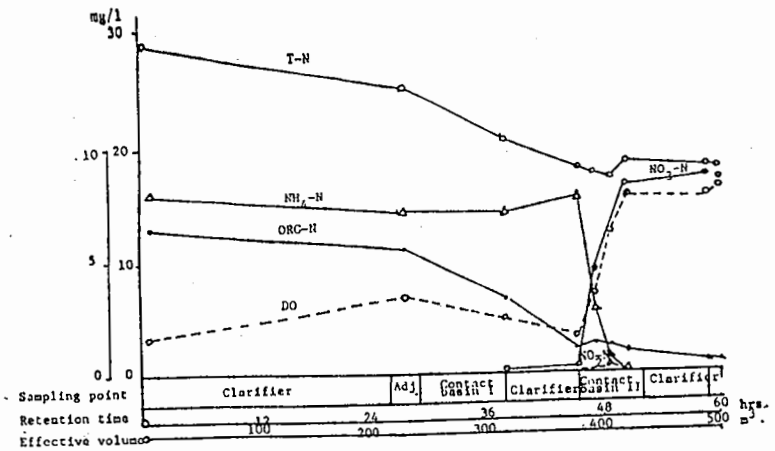


Figure 7. Results of Chemical Analysis of Flow in DJS

For the efficient nitrification organic loading up to 25 lb/BOD_5/1000-ft^3-day was employed with gravel media filter but surface loading are maintained at 0.03 gpm/ft^2. This will be one of the reasons why organic nitrogen removal efficiency was very high, reaching 96 per cent during the period of operation and sludge production was few.

At Haguro plant operation and maintenance have been made by some community people who had no skill before the plant was constructed. As far as changes of quality of treated water in Fig. 7 are concerned, excessive aeration had been made as in other areas instead of frequent changes of aerobic and anaerobic conditions. From the experience of many other DJS plants, BOD_5, TSS and total nitrogen average will be able to reach one or two mg/l in the course of time.

DESIN RELATIONSHIPS

Design relationships related to attached growth biological treatment processes can be applicable to design of DJS except for that of surface loading(6).

COSTS

Although relationships between costs and effluent quality is not clearcut, investment per capita for DJS itself was less than $1,000 at the time of construction.

CONCLUSION

A DJS is a valuable and successful experiment which has displayed the possibility of solving many problems of biological wastewater treatment by the fixed-film biological processes. If a DJS could be added to after any wastewater treatment processes in question or such processes were converted to DJS, the situations concerned will undoubtedly be much improved.

Mathmatical modeling has been under way but this model must be different much from the conventional ones in that instead of solving at the expense of creating new problems, all problems of internal and external environments must be solved towards increasing overall stability with healing process in an automatic way(3).

REFERENCES

1. Kropf,F.W., et al.,"Equilibrium Operation of Subsurface Absorption Systems," Journal of WPCF.Sept.1977,

pp.2007-2016.
2. James,A.,"An Ecological Model of Percolating Filters ," in "Mathmatical Models in Water Pollution Control ," edited by James, A.,John Wiley & Sons, N.Y.,1978 pp.303-318.
3. Goldsmith,E.,"Thermodynamics or Ecodynamics," The Ecologist, Vol.11, No.4,1981,pp.178-195.
4. Dixit,N.S.S.,"Bicultural, Two-stage, High-rate Activated Sludge Process," Illinois Institute of Technology, PhD.Dissertation, 1976.
5. Sharma,B.,et al.,"Nitrification and Nitrogen Removal " Water Research, Vol.11,1977,pp.897-925.
6. Benefield, L.D., et al,"Biological Process Design for Wastewater Treatment," Prentice-Hall,Inc.,Englewood Cliffs,N.J.,1980.

STUDY OF FIXED-FILM BIOLOGICAL CONTACTORS FOR RECREATIONAL AREA WASTEWATER TREATMENT APPLICATION

Calvin P.C. Poon. Department of Civil and Environmental Engineering, University of Rhode Island, Kingston, Rhode Island.

Edgar D. Smith. Department of the Army, Construction Engineering Research Laboratory, Champaign, Illinois.

Vicki A. Strickler. Department of Civil and Environmental Engineering, University of Rhode Island, Kingston, Rhode Island.

INTRODUCTION

A survey was conducted in 1981 to evaluate the type and performance of existing wastewater treatment facilities in U.S. Army Corps of Engineers Civil Works (CE) recreational areas (1). It was found that septic tank-leaching field or septic tank-sand filter systems for subsurface discharge are by far the most used treatment systems, followed by extended aeration and lagoon systems. Occasional high suspended solid (SS) concentration in lagoon effluents is not uncommon since dispersed growth and algal cells do not settle well. Upsets of extended aeration treatment plants are experienced by many recreational areas from time to time resulting in biochemical oxygen demand (BOD) and SS concentrations higher than the acceptable limits. This phenomenon is typical of an extended aeration process which has dispersed growth leading to poor settling in the final

clarifier.

A prime concern of applying a suspended growth biological reactor to the treatment of organic wastewater is the occurence of periodic shock of hydraulic and/or organic loads. A successful control of the suspended culture population in the reactor by sludge return requires a skillful operation. Even so, a washout of the suspended culture occurs quite often with hydraulic shock loads. The previously mentioned survey identifies the flow fluctuations at the CE recreational areas being $\frac{\text{weekend}}{\text{weekday}}$ flow ratio = 1.62 to 15.0, $\frac{\text{holiday}}{\text{weekday}}$ flow ratio = 1.93 to 27.50, and $\frac{\text{offseason day}}{\text{weekday}}$ flow ratio = 0.17 to 0.5. These flow fluctuations apparently present an operational problem to numerous extended aeration and oxidation pond treatment facilities in recreational areas. It is believed that because of the simpler operational requirement of a fixed-film biological contactor and its ability to retain its biological culture with hydraulic and/or organic loading, a rotating biological contactor (RBC) will lend itself a favorable alternative to suspended growth reactors in recreational area sewage treatment.

RBC TREATMENT KINETICS

Under a steady hydraulic and organic loading condition of 2.5 to 7.5 g soluble BOD/m^2.d (approximately 0.5 to 1.5 lb SBOD/1000 ft^2.d), a RBC is able to remove consistently from 57 to 90 percent of the soluble BOD (SBOD). Outside of this range of loading, the percentage removal is definitely lower at lower loadings (2). It is not certain that the same percentage of SBOD removal can be maintained at steady high loadings higher than 7.5 g SBOD/m^2.d, (Figure 1). It appears that with limits and with any given SBOD loading, a lower percentage of removal can be expected when the influent SBOD concentration is lower, particularly when the influent has already received some degree of biological treatment. It is noted that below the loading of 7.5 g SBOD/m^2.d, the effluent SBOD concentration is consistently below 20 mg/l, of which 33 to 70 percent is made up of soluble nitrogenous BOD.

Successful nitrification of wastewater using RBC have been demonstrated. O'Shaughnessy et al (3) show 81 to 96 percent NH_3-N removal from secondary effluent by RBC when optimal pH and alkalinity are under control. Beyond a loading of 4.0 g NH_3-N/m^2.d (0.8 lb/1000 ft^2.d) however, the percentage is decreased significantly. With approximately the same range of NH_3-N loading (0.2 to 4.0 g/m^2.d), Zenz et al (4) report 70 to 94 percent NH_3-N removal, while Reh (5) report 85 percent removal. For the nitrification of primary effluent where the process is sensitive to organic loading and the subsequent sloughing of nitrifying biofilm, Zenz et al (4) report from 20 to 95 percent NH_4-N removal within the range of 0.25 5o 2.0 g NH_4-N/m^2.d (0.05 to 0.40 lb/1000 ft^2.d). On the other hand, Poon et al (6) show 50 percent removal within the range of 0.1 to 0.65 g NH_3-N/m^2.d loadings to 83 percent removal for up to 2.8 g NH_3-N/m^2.d loading (Figure 2).

In a suspended growth complete-mix reactor a Monod or Michaelis-Menten enzyme kinetics is applicable. The kinetics assumes a hyperbolic saturation phenomenon with a graduate change in reaction rates. Biofilm kinetics however may involve three distinct regions with abrupt transition in the order of the bulk reaction from one region to the other shown in Figure 3 according to Harromoës (7). Data from Kornegay et al (8) indicate that from 0 to 65 μ biofilm thickness, glucose fully penetrates the biofilm, resulting in zero-order reaction or the rate of glucose removal increases proportionally to the film thickness for a given glucose concentration. For biofilm thickness greater than 65 μ, the reaction rate becomes constant corresponding to a partly penetrated biofilm. The same data also show that a zero-order reaction rate can be obtained for a biofilm thickness of 200 μ if the glucose concentration is 1300 mg/l. Below this concentration, only a half-order reaction rate is obtained. LaMotta's work (9) shows zero-order reaction rate obtainable with a biofilm thickness of 10 μ when the glucose concentration is 5.2 mg/l and 70 μ when the glucose concentration is increased to 200 mg/l.

In the study with steady loads, (2), the biofilm thickness is not measured. Instead, the biofilm of a unit

surface area is collected and its dry weight is measured. Knowing the moisture content and the density of the dry biofilm, the thickness of the biofilm on the media is calculated. The average biofilm thickness for the 1st, 2nd, 3rd and 4th stages of the RBC are respectively 250μ, 180μ, 120μ and 102μ. It is apparent that an influent BOD from 40 to 186 mg/l in the study do not fully penetrate the biofilm, resulting in a reaction kinetics ranging from half-order to first-order.

Assuming that the biological contactor is rotating in an ideally mixed compartment, the following mass balance equations can be written:

Half order

$$V \frac{dC_n}{dt} = QC_{n-1} - Q \cdot C_n + A(-k_{\frac{1}{2}a} C_n^{\frac{1}{2}}) = 0$$

$$\text{or} \quad Q(C_{n-1} - C_n)/A = k_{\frac{1}{2}a} C_n^{\frac{1}{2}} = r_{\frac{1}{2}a} \qquad (1)$$

First Order

$$Q(C_{n-1} - C_n)/A = k_{1a} \cdot C_n = r_{1a} \qquad (2)$$

in which Q is the flow rate; C_n is the substrate concentration at the n stage; C_{n-1} is the substrate concentration at the (n-1) stage; A is the surface area of the rotating media at stage n; $k_{\frac{1}{2}a}$ and k_{1a} are respectively the half-order and first order rate constants; and $r_{\frac{1}{2}a}$ and r_{1a} are respectively the half-order and first-order rates of substrate removal. Plotting the substrate removal rates versus $C_n^{\frac{1}{2}}$ or C_n should yield a straight line, the slope of which is the rate constant $k_{\frac{1}{2}a}$ or k_{1a} according to equations 1 and 2. Using the total BOD data (non-settled, carbonaceous BOD and nitrogenous BOD combined) such plots yield a $k_{\frac{1}{2}a}$ value of 0.93 $g^{\frac{1}{2}}/m^{\frac{1}{2}}.d$ (correlation coefficient $r = 0.63$) and a k_{1a} value of 0.08 m/d ($r = 0.64$). If SBOD data are used, the plots yield a $k_{\frac{1}{2}a}$ value of 0.75

$g^{½}/m^{½}.d$ (r = 0.68) and a k_{1a} value of 0.09 m/d (r = 0.69).

In the same study (2), NH_4Cl is introduced into the sewage to create a very high NH_3-N loading for a period of 10 days. Figure 4 shows that a relatively constant rate of removel of soluble NH_3-N is reached at 2.8 $g/m^2.d$ when the soluble NH_3-N loading is about 5.0 $g/m^2.d$ or the soluble NH_3-N concentration is about 50 mg/l. The S-shaped curve therefore suggests that the removal rate could be first-order initially, changed to half-order as the soluble NH_3-N concentration increases and finally reaching the maximum or the zero-order with very high NH_3-N concentrations. The result also suggests that where nitrification primarily takes place in the third and fourth stages of the RBC, a soluble NH_3-N concentration at or higher than 50 mg/l is able to penetrate fully a biofilm of 102 to 120μ thick. One can not take advantage of the maximum removal rate in the design, however, because the percentage of removal is lower and the effluent would be unacceptable.

SIMULATED RBC STUDY IN RECREATIONAL AREAS

A simulated study is carried out in laboratory using a 4-stage RBC 0.5m in diameter with a total area of 23.3m^2 (250 ft^2) of media. The purpose of this study is to investigate the effect of shock loads typical of recreational areas on the treatment performance of the RBC.

Three series of experiments are conducted. The first series uses a synthetic sewage of the following composition:

Glucose	100 mg/l
Bacto-peptone	55 mg/l
$FeCl_3$	0.35 mg/l
$MgSO_4.7H_2O$	62.5 mg/l
NH_4Cl	92 mg/l
K_2HPO_4	22 mg/l
KH_2PO_4	8.4 mg/l
$Na_2HPO_4.7H_2O$	37.5 mg/l

This sewage simulates that of a recreational area facility where urine is the major component of the wastewater. Facilities for short term visits (visitors center, swimming, boating, hiking, etc. but no camping) usually have wastewater of this characteristic relatively weak in BOD but high in NH_3-N compared to a typical municipal sewage. At first a steady hydraulic load of 460 liter/day (0.02 $m^3/m^2.d$ or 0.5 gpd/ft^2) is maintained over a period of time. After a steady performance is reached, load fluctuation simulating the frequency of use by visitors is applied with 16 hours of high load (0.06$m^3/m^2.d$) followed by 8 hours of normal load (0.02 $m^3/m^2.d$). This 24 hour cycle is repeated for 2 days in this series of experiments. Figure 5 shows that the RBC consistently produces a very low BOD effluent under the fluctuating load condition. BOD removal is 92 percent, comparing to that of a control experiment (a steady normal load maintained for a period of several days) of 86.4 percent. Ammonia nitrogen in the effluent as depicted by Figure 6, is high throughout the experiment despite the fact that 3.9 to 9.2 mg/l of NO_3-N repeatedly shows up in the effluent. However, there is a significant removal of the organic-N from the synthetic wastewater. The conversion of organic-N to NH_3-N adds to the already high NH_3-N concentration, resulting in very high NH_3-N loadings to the RBC unit. This may explain the phenomenon of low percentage of NH_3-N removal and strong nitrification taking place at the same time. This phenomenon is unique and reflects the special characteristics of a recreational area wastewater or a similar wastewater with high NH_3-N and organic-N concentrations. If a more complete nitrification is desirable, more stages or additional media can be added to the RBC unit to reduce the NH_3-N loading.

The other two series of experiments use a synthetic sewage similar to the one aforementioned except that glucose is increased to 300 mg/l, NH_4Cl concentration remains relatively the same and Bacto-peptone is eliminated. The sewage is stronger in BOD and contains a relatively high nitrogen concentration. The strength is equivalent to that of a typical municipal wastewater. It simulates the

characteristics of a sewage from a recreational area with camping, shower and laundry facilities. Both experimental series start with a steady normal load for a long period of time. In one series, this period is followed by 18 hours of high load (approximately 3 times the normal load) and then 6 hours of normal load. The (3Q-1Q) cycle is repeated twice in the experiment. The other series is similar except that the high load is approximately 4 times the normal load (4Q-1Q) series. As shown in Figure 7, the effluent SBOD concentration is relatively stable at or below 17 mg/l under the fluctuating load condition (4Q-1Q). Even with a short term shock (almost 3 times the high BOD load or 10-12 times the normal BOD load) that is applied to the RBC by mistake, the effluent SBOD concentration is only 24 mg/l and the unit recovers quickly once this unusually high load is eliminated.

Although most engineers use the SBOD parameter in monitoring RBC performance, the result of the 3Q-1Q series as depicted in Figure 8 indicates the importance of total BOD rather than SBOD in monitoring the effluent quality. Again the RBC is able to produce a good quality effluent under the fluctuating load condition. However, towards the end of the second high-load period the effluent BOD is increased to 28 mg/l. After a short period of recovery the effluent BOD is further increased to 36 mg/l. The increase of the effluent BOD coincides with biofilm sloughing initially from the first-stage and later on from the second-stage. When sloughing occurs in either one of the first two stages, the biological solids do not settle well even though the overflow rate of the clarifier is low at $20.4 m^3/m^2.d$ (500 gpd/ft^2). The suspended solid (SS) concentrations corresponding to the effluents with 28 mg/l and 36 mg/l total BOD are respectively 23 and 134 mg/l. It is expected that the effluent SS contributes to some effluent BOD, making the effluent SBOD values lower than the respective 28 and 36 mg/l values. The effluent quality expressed in SBOD concentration in effect would be acceptable under the fluctuating load condition. An implication of this finding is that for a small RBC treatment facility in a recreation area the period of sloughing could yield a higher total BOD and therefore a poorer effluent quality. The frequency of sloughing is not monitored in this study. Consequently it is not known how often a lower quality effluent occurs. It should be noted that this problem is

greatly minimized in larger RBC facilities because only a small fraction of the entire media would experience sloughing at any given time.

To test the kinetics of BOD removal, removal rates are plotted versus effluent concentration or (effluent concentration)$^{\frac{1}{2}}$. When the data of all three series are put together with the normal loads as one group and the high loads as another, such plots yield a $k_{\frac{1}{2}a}$ value of 0.54 $g^{\frac{1}{2}}/m^{\frac{1}{2}}.d$ (r = 0.37) and a k_{1a} value of 0.1 m/d (r = 0.55) for normal loads, but a $k_{\frac{1}{2}a}$ value of 1.56 $g^{\frac{1}{2}}/m^{\frac{1}{2}}.d$ (r = 0.60) and a k_{1a} value of 0.23 m/d (r = 0.56) for the high loads. This indicates that first-order kinetics is more applicable to the normal loads (lower effluent BOD concentration) and half-order kinetics a better fit with high loads (higher effluent BOD concentration). The finding is in conformity with the fixed-film reactor kinetics of Harremoës (7).

Ammonia nitrogen removal is low in these two series of experiments with high influent BOD concentrations. Percentage of removal is 36.2% for the (3Q-1Q) series and 30% for the (4Q-1Q) series. Only trace amount of nitrate is detected in the effluents, indicating that nitrification can not be estiablished in these high and fluctuating BOD load conditions. The NH_3-N removal is due to biofilm synthesis alone as NH_3 stripping is unlikely at the wastewater pH of 4.8 to 6.0. It should be remembered that nitrification takes place in the (3Q-1Q) series with fluctuating but low BOD load condition even though nitrification is not complete. The incomplete nitrification is partly due to insufficient media area for the high NH_3-N load and partly due to the relatively unfavorable pH (4.8-6.0) of the simulated recreational area wastewater. Because of the unsuccessful nitrification in fluctuation load conditions, investigation of nitrification kinetics is omitted from this work.

SUMMARY

Under a steady load condition, the biofilm thickness on all 4 stages of a RBC unit is indirectly measured and

calculated to be 250, 180 120 and 102 μ respectively. Zero-order reaction kinetics is not to be expected since the substrate (BOD) does not fully penetrate the biofilm. The data indicate that both first-order and half-order kinetics apply equally well for SBOD loadings within the range of 0 to 8.0 $g/m^2.d$ (0 to 1.6 $lb/1000\ ft^2.day$) using NH_3-N loadings from 0 to 5.0 $g/m^2.d$ and beyond. The data suggest that the removal rate could be first-order followed by half-order and then reaching zero-order when the NH_3-N concentrations and loadings are high. A full penetration of the biofilm 102 to 120 μ thick is possible when NH_3-N concentration is 50 mg/1 or above.

Two studies are conducted, one with a synthetic sewage relatively weak in BOD but strong in organic-N and NH_3-N, and the other strong in BOD and NH_3-N, simulating two different recreational area wastewaters. BOD removal is good under fluctuating load conditions. Three problems are identified in the application of RBC for the treatment of this special waste. One is that despite the nitrification taking place when the weak BOD wastewater is treated, a great deal more media is required if near complete nitrification is desired since the NH_3-N loading is high. Secondly, nitrification can not be established when the wastewater is strong in BOD probably due to the sloughing of nitrifiers. Thirdly, sloughing of biofilm from small RBC facilities periodically yields effluents with high total BOD even though the SBOD concentration may be acceptable.

ACKNOWLEDGEMENT

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REFERENCE

1. Poon, C.P.C., and Smith, E.D., "Rotating Biological Contactor Technology Evaluation for Civil Works Recreational Area Application," First Report to Construction Engineering Research Laboratory, Department of the Army, July, 1981.

2. Poon, C.P.C., Chin, H.K., Smith, E.D., and Milkueki, W.J., "Upgrading With Rotating Biological Contactors for BOD Removal," J. Water Pollution Control Fed., Vol. 53, No. 4, April, 1981. pp. 474-481.

3. O'Shaughnessy, J.G., et al., "Nitrification of Municipal Wastewater Using RBC," Proc. 1st National Sypm/Workshop on RBC Technology, Vol. 2, February, 1980, pp. 1194-1219.

4. Zenz, D.R., et al., "Pilot Scale Studies on the Nitrification of Primary and Secondary Effluents Using RBC at the Metropolitan Sanitary District of Greater Chicago," Proc. 1st National Sypm/Workshop on RBC Technology, Vol. 2, February, 1980, pp. 1222-1246.

5. Reh, C.W., et al., "An Approach to Design of RBC for Treatment of Municipal Wastewater," Paper presented at the ASCE National Environ. Engr. Conf., Nashville, July, 1977.

6. Poon, C.P.C., Chin, H.K., Smith, E.D., and Mikucki, W.J. "Upgrading with RBC for Ammonia Nitrogen Removal," J. Water Pollution Control Fed., Vol. 53, No. 7, July, 1981 pp. 1158-1165.

7. Harremoes, P., Biofilm Kinetics, Chap. 4, Water Pollution Microbiology, ed. R. Mitchell, Vol. 2, John Wiley & Sons.

8. Kornegay, B.H., Andrew, J.F., "Characteristics and Kinetics of Biological Film Reactors," Fed. Water Pollution Control Admin., Final Report, WP-011811, 1969.

9. LaMotta, J., "Internal Diffusion and Reaction in Biological Films," Environ. Science and Tech., Vol. 10, No. 8, August, 1976, pp. 765-769.

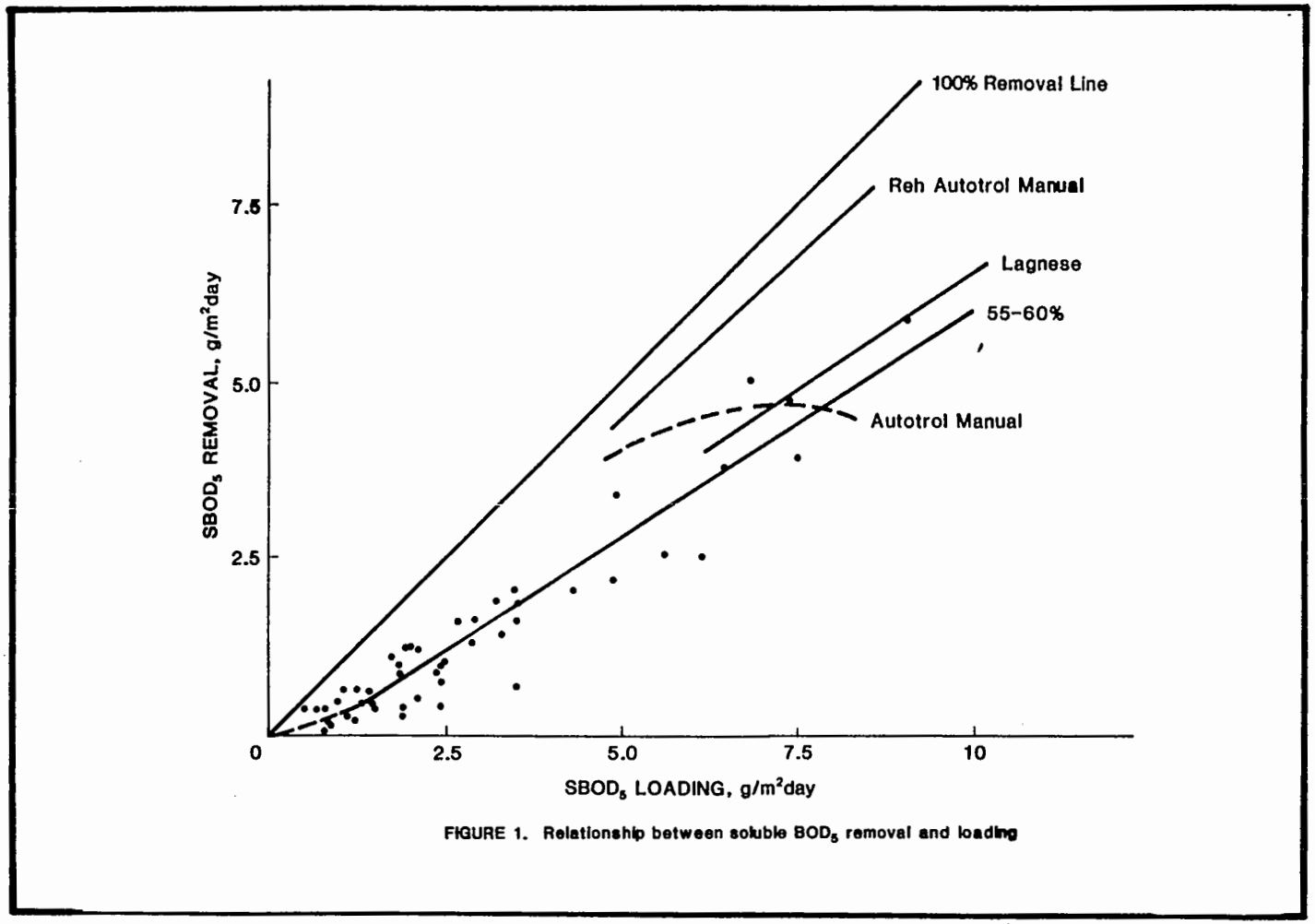

FIGURE 1. Relationship between soluble BOD_5 removal and loading

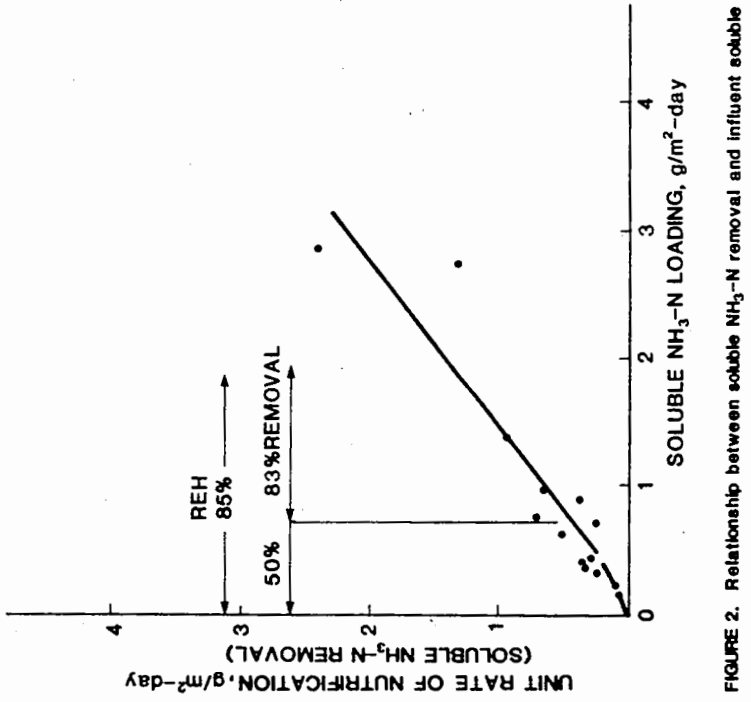

FIGURE 2. Relationship between soluble NH_3-N removal and influent soluble NH -N loading.

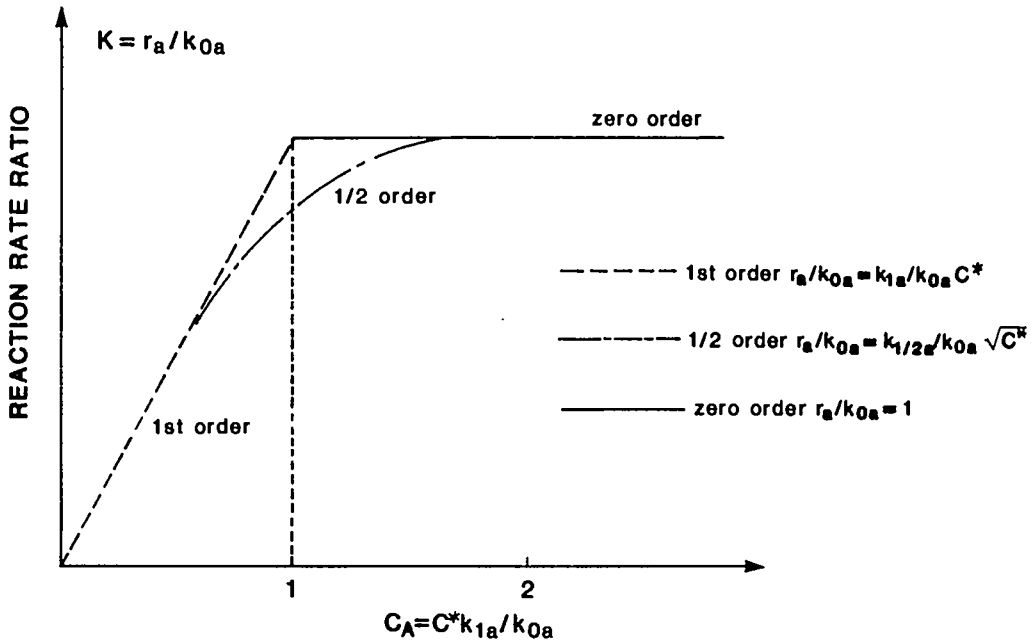

FIGURE 8. Dimensionless plot of reaction versus concentration in the bulk liquid

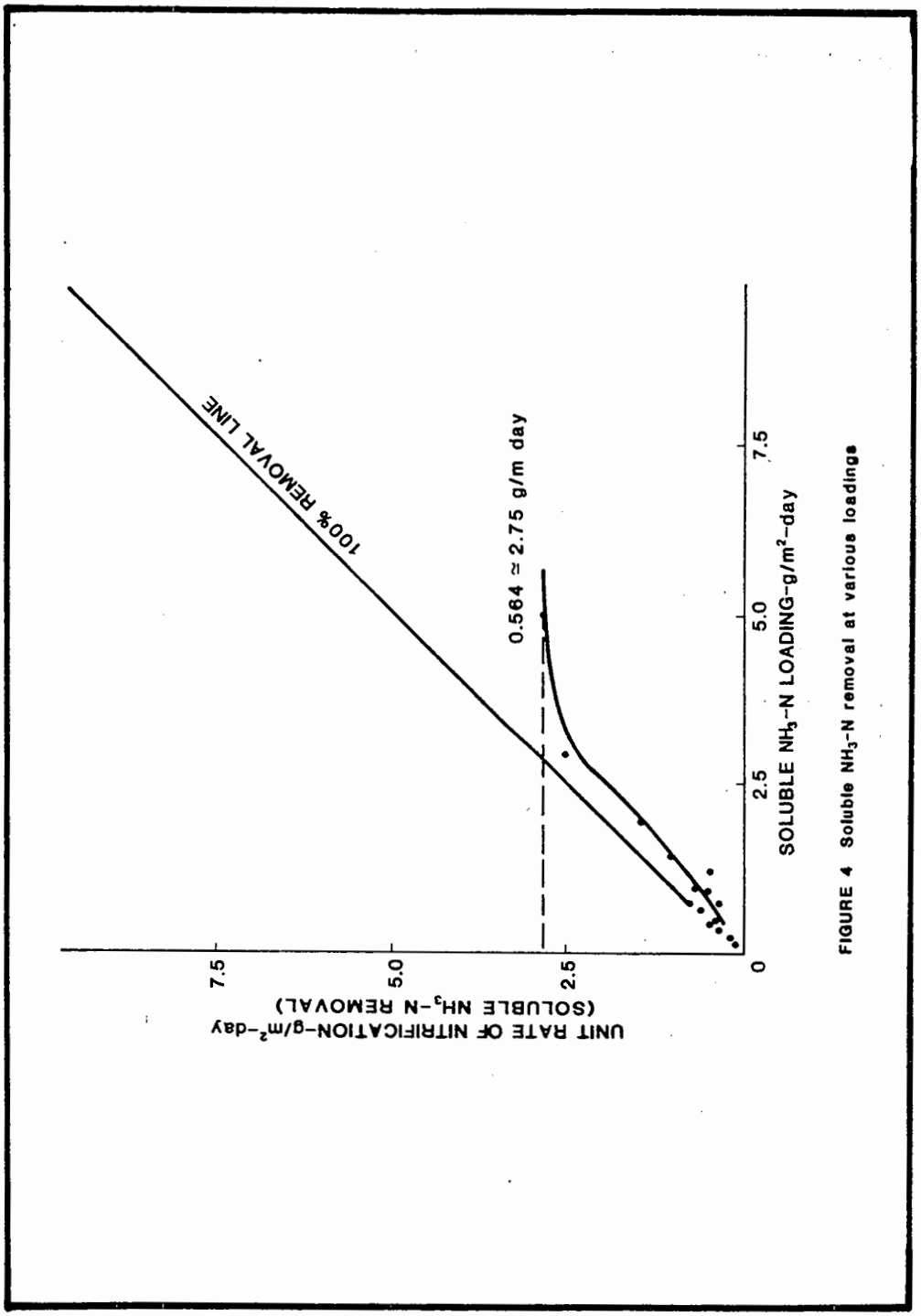

FIGURE 4 Soluble NH_3-N removal at various loadings

Normal load	Normal load	High load	Normal load	High load	Normal load
Experiment starts →	8 hrs.	16 hrs.	8 hrs.	16 hrs.	8 hrs.
Hydraulic load m^3/m^2d	0.02	0.059	0.02	0.059	0.02
BOD load, g BOD/m^2d	1.7	5.5	1.7	5.1	1.7
influent BOD concentration, mg/l	87.3	92.0	85.3	87.0	86.0

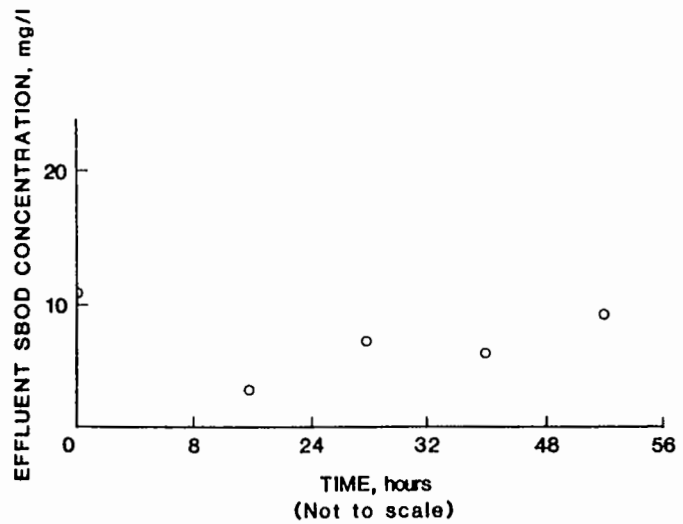

FIGURE 5. RBC effluent SBOD concentration under fluctuating load condition – low influent BOD concentration

Normal load	Normal load	High load	Normal load	High load	Normal load
Experiment starts →	8 hrs.	16 hrs.	8 hrs.	16 hrs.	8 hrs.
Hydraulic load m^3/m^2d	0.020	0.059	0.02	0.059	0.02
NH_3-N load gNH_3-N/m^2d	0.46	1.73	0.34	1.33	0.49
Influent NH_3-N mg/l concentration	23.4	29.1	17.2	22.3	24.6

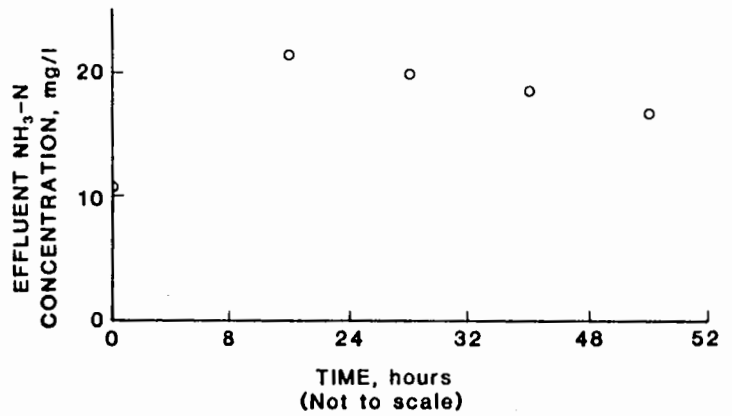

FIGURE 6. RBC effluent NH_3-N concentration under fluctuating load condition - low influent BOD concentration

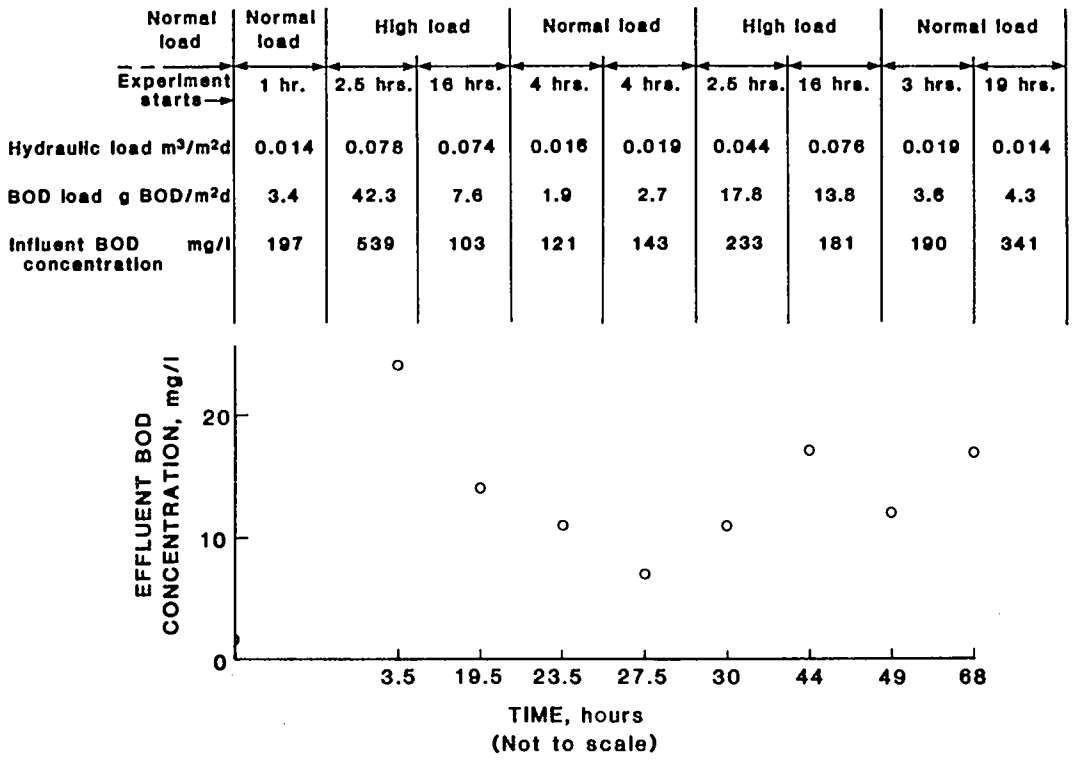

FIGURE 7. RBC effluent SBOD concentration under fluctuating load condition - high influent BOD concentration, (4Q-1Q) series

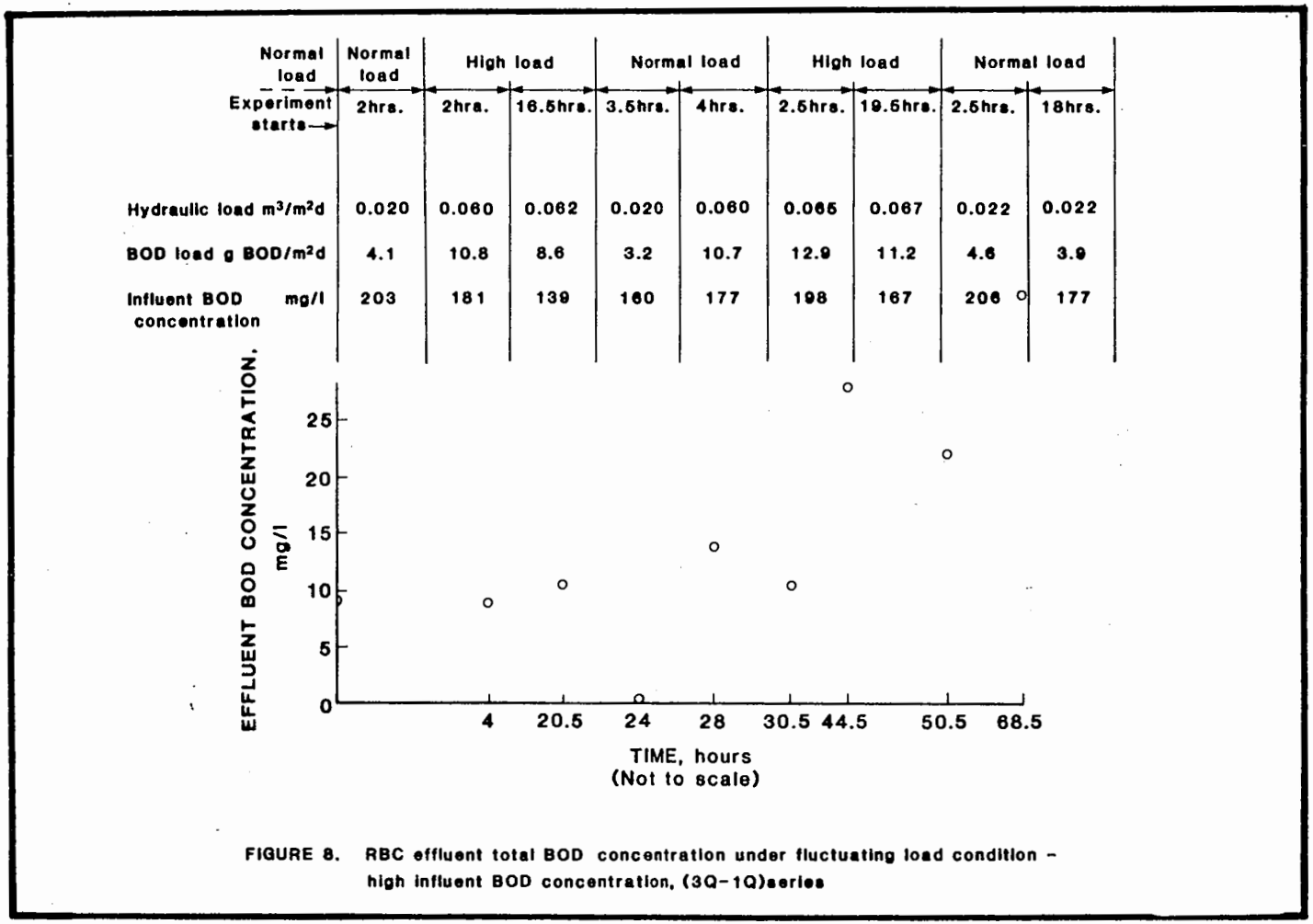

FIGURE 8. RBC effluent total BOD concentration under fluctuating load condition – high influent BOD concentration, (3Q-1Q)series

START-UP AND SHOCK LOADING CHARACTERISTICS OF A ROTATING BIOLOGICAL CONTACTOR PACKAGE PLANT

Farley F. Fry. Department of Civil Engineering, Virginia Polytechnic Institute, Blacksburg, Virginia.

Tom G. Smith. C.M.S. Rotordisk Limited, Mississauga, Ontario, Canada.

Joseph H. Sherrard. Department of Civil Engineering, Virginia Polytechnic Institute, Blacksburg, Virginia.

INTRODUCTION

Rotating biological contactors have become increasingly popular as a wastewater treatment alternative within the last several years. Because this method of treatment is relatively new, design and operation procedures are still being developed. Information that is currently lacking for use in design and operation includes a) the time needed for development of sufficient biofilm mass to insure an acceptable effluent quality, and the progress of organic removal and development of nitrification during the start-up period, and b) the shock loading response of an RBC to increases in hydraulic and organic loading.

The purpose of this investigation is to provide start-up and shock loading response data for an RBC pilot plant receiving primary effluent from a municipal wastewater treatment plant. Data reported in this study are abstracted from a more comprehensive study performed by Fry (1). Due to space limitations only a small portion of this study are reported herein.

BACKGROUND INFORMATION

A review of pertinent literature pertaining to the start-up and shock loading response of RBC's is presented in this section to provide a background for understanding the observations and analysis of results which follow.

Start-Up Characteristics of RBC's

In general, little information is available on start-up characteristics of RBC units. References which are encountered are brief, incomplete and incidental in nature because the research was not focused on start-up. With the issuance of so-called "tiered" National Pollutant Discharge Elimination System (NPDES) permits, the time required to attain steady-state conditions has become very important. For example, the assimilative capacity of a body of water receiving effluent from a wastewater treatment facility may require the operation of nitrification facilities only during the summer months. If the RBC process is used for nitrification purposes, what length of start-up time is required for the units to discharge acceptable quality effluent? Obviously start-up characteristics, especially the length of time required to discharge acceptable effluent quality, is desirable information.

Information is available on two aspects of the start-up period. One aspect is the establishment of the biomass. According to one authority the attached biofilm generally ranges from 2 to 4 mm thick one week after start-up (2). Although useful as a general guide, this statement does not consider the effects of varied organic and hydraulic loads. In another study, 9 days were required to establish a thin layer of biomass covering the entire outside of the media (3). This observation was made in a study using RBC units to upgrade trickling filter effluent.

Establishment of an observable biomass required two weeks in an RBC treatability study for phenol-formaldehyde resin wastewater (4). Prior to introducing the industrial wastewater, domestic primary effluent was fed to the RBC at the rate of 1.6 gal/ft^2/day. The phenol-formaldehyde resin waste was introduced after the biomass was established. Average organic strength of the primary effluent was not provided.

Based on the preceding studies it is apparent that a measurable or observable biofilm will result 1 or 2 weeks after start-up begins. At the current time it is not known how other factors such as wastewater characteristics and loading

rates affect the development of the biomass. It should be noted there was no distinction made between heterotrophic and autotrophic growth.

Another important aspect of the start-up period which relates to the time required to achieve steady-state operating conditions, was evaluated by the Ontario Ministry of Environment (5). During the start-up period 300 gpd of raw domestic sewage was fed to a five stage unit at the average rate of 1.05 lb BOD_5 (5-day Biochemical Oxygen Demand)/1000 ft^2/day. After three weeks of operation acceptable effluent quality (i.e. 15 mg/l BOD_5 and 15 mg/l SS) was discharged. Several more months of operation were required for a comparable growth in the last two stages. Effluent concentrations of BOD_5 and ammonia-nitrogen (NH_3-N) remained the same.

Nitrification of a high strength ammonia waste by use of RBC units was examined by Lue-Hing et al. (6). Sludge lagoon supernatant diluted by 50% with water was introduced to an RBC unit for 10 days of batch aeration. After batch aeration, the eight stage pilot plant RBC unit was continuously fed diluted supernatant with a 12 day hydraulic detention time. Following three weeks of continuous flow operation, effluent nitrate-nitrogen (NO_3-N) approximately equalled the total Kjeldahl nitrogen (TKN) removed. Typical TKN removal was approximately 600 mg/l and influent BOD averaged approximately 100 mg/l.

Trinh (7) reported the acclimation of an RBC unit (in terms of BOD_5 removal) within two weeks with a loading rate of 7.3 kg/100 m^2/day (1.5 lb/1000 ft^2/day). This investigator compared the performance of an extended aeration activated sludge package unit with an RBC package unit using domestic waste from an isolated work camp.

An RBC pilot unit required approximately three weeks to reach steady-state conditions using primary municipal effluent in a study conducted by Srinivasaraghavan et al. (8). Soluble organic loading ranged from 0.5 to 1.2 lb $SBOD_5$ (Soluble BOD_5)/1000 ft^2/day. Since nitrification did not occur in any phase of this study, steady-state operation was based on organic substrate removal.

Based on the examples cited above it appears that 2 to 3 weeks of operation are required for an RBC to attain steady-state operating conditions. This appears evident not only in terms of BOD_5 but also for nitrification when the TKN:BOD_5 ratio is very large.

Shock Loading Characteristics of RBC's

Statements concerning the excellent ability of RBC units to successfully handle shock organic and hydraulic loadings are frequently encountered. Wu et al. (9) for example, noted a chief advantage of an RBC is the ability to resist organic and hydraulic loads. These statements are generally based on one or two characteristics of RBC plants. One important characteristic is the ability to retain the attached biomass when exposed to large hydraulic shocks.

Welch observed this ability in one of the first investigations of RBC units in the United States (10). Welch focused his attention on the response of a two-stage RBC to different variables. Variables included concentrations of synthetic feed, disc speed, hydraulic residence time, intermediate settling and sludge recycling. Data on shock loading characteristics were not presented, but it was observed that the process did not experience biological upsets encountered as with the activated sludge process.

An analysis of phenol-formaldehyde resin wastewater treatment by an RBC process found effluent Chemical Oxygen Demand (COD) values to be a function of the influent COD concentration. More importantly the pilot plant RBC units functioned effectively "under varying climatic and loading conditions and exhibited excellent stability in withstanding periodic shock loadings" (4).

Trinh (7) reported that the biological slime of an RBC system weathered shock loads without sloughing and produced consistent effluent quality. However, diurnal flow variations caused a slight deterioration of effluent quality. These comments were based on a study comparing an extended aeration activated sludge process with a full-scale RBC system.

Researchers in California also reported a stable biomass (11). Municipal primary effluent was used to investigate the response of an RBC pilot plant to increases in hydraulic loading rate. Over a 15 day period, the feed rate was increased from 6 gpm (1.1 gal/ft^2/day) to 70 gpm in 5 steps. $SBOD_5$ removal remained relatively constant within the RBC while the hydraulic loading was up to 1,040% of design values and organic loading up to 370% of design values.

Later, the RBC received a two-fold hydraulic peak on the first day, a three-fold hydraulic peak on the second day, a four-fold hydraulic peak on day three and five-fold hydraulic peak on day four. These daily peaks were timed to include an increasing organic load in the municipal primary effluent. The result was a significant increase in both organic and hydraulic loading rates. With soluble organic loading increases of up to 700%, the total mass of soluble Total Organic Carbon (TOC) removed increased although admittedly effluent soluble TOC increased considerably. Of major importance was

the lack of operational difficulties encountered in contrast to occasional biological "washout" encountered in a suspended growth system.

It appears conclusive that RBC units are more resistant to the loss of biological solids than suspended growth systems. Perhaps the key words for describing this attribute are "biofilm stability."

Another important characteristic is the alledged ability of RBC units to produce a consistent and acceptable quality effluent while exposed to shock loads. A main objective of a study conducted by the Ontario Ministry of Environment in 1973 (5) was to determine the performance of a full-scale RBC system under intermittent feed conditions. During a 6 week period raw sewage was fed to the unit at the rate of 320 gph for 2 consecutive days per week. The average organic loading of 0.92 lb BOD/1000 ft^2/day for each consecutive 2 day period was previously determined to be the approximate maximum capacity of the system for the continuous feed phase. When compared to the data from the continuous feed period, little difference was found in terms of organic removal efficiency. Noticeable evaporation losses were evident in the two central stages which were isolated from the primary and secondary clarifiers.

Kinner and Bishop (12) reported similar findings while investigating saline RBC microbial populations. The RBC units were set-up at a sewage pumping station in Durham, New Hampshire and received the diurnal loading characteristic of a small town. An effluent SBOD below 30 mg/1 was consistently observed.

Srinivasaraghavan et al. (8) evaluated the effect of diurnal flow variations with no impairment of $SBOD_5$ removal efficiency. For this study primary municipal effluent was fed to a four-stage, air-driven, pilot plant RBC unit. The RBC was 10 feet long, 10.4 feet in diameter and was preceded by a 10.4 foot diameter aerated wet well. In the diurnal flow phase the organic loading rate ranged from 0.47 to 0.78 lb $SBOD_5$/1000 ft^2/day, which was typical of other phases of the study. Nitrification did not occur in any phase of the study. The diurnal flow pattern consisted of periodic four-fold hydraulic increases. Unfortunately, the time between simulated diurnal peaks was only 20 minutes. It is probable the large detention time of the wet well and RBC dampened or eliminated all effects of the 20 minute diurnal cycle.

In contrast, Dupont and McKinney (13) after studying the performance of a municipal RBC installation in Kirksville,

Missouri, found treatment efficiency was reduced as a result of variable hydraulic loadings. These workers evaluated monthly reports of the treatment plant and not the RBC unit alone. Reduced treatment efficiency was attributed to reduced contact time within the RBC units and hydraulic surges on the final clarifiers.

The results of a study by Poon et al. (3) agree with the Kirksville study. Trickling filter effluent was fed to a pilot plant RBC treatment system (including primary and secondary clarification) at the moderate hydraulic rate of 0.045 $m^3/m^2/day$ (1.1 $gal/ft^2/day$). As expected, the trickling filter effluent supplied a low $SBOD_5$ influent concentration. The RBC system was exposed to a series of hydraulic shocks ranging from 120 to 220% of the steady-state loading. Effluent $SBOD_5$ from the RBC system increased rapidly as the hydraulic shocks increased. An organic shock was simulated by coupling a high hydraulic feed rate with a moderate $SBOD_5$. Total $SBOD_5$ removal actually improved but effluent quality deteriorated significantly.

Using a laboratory scale two stage RBC unit combined with primary, intermediate and secondary clarification Antonie (14) examined treatment during intermittent flow conditions. To simulate an industrial wastewater flow cycle, synthetic wastewater was introduced only during the regular eight hour working day. Performance was generally consistent throughout the eight hour period with the exception being a delayed response period in percent COD reduction for during the first several of hours. Continued sloughing of the biofilm during the night led to a five-fold increase in mixed liquor suspended solids (MLSS).

To reduce this problem the author repeated the experiment but maintained a low wastewater flow and reduced disc revolutions per minute (RPM) overnight. Instead of a delayed response period of COD removal, the COD removal initially was greater than steady-state operation. The author noted in an actual application this could be accomplished by recycled effluent. Antonie concluded that intermittent flows could be effectively treated by the RBC process as long as a low wastewater flow was maintained between cycles.

In addition, Antonie evaluated the RBC system with the intermediate clarifier bypassed under varying flow conditions. During each day the treatment system was exposed to periods of decreasing, increasing and constant flow. The COD concentration remained constant with only the flow rate changing. Overall performance in terms of COD reduction actually improved over steady-state performance.

In the final phase of testing, Antonie investigated the

response of a 10 stage RBC unit without clarification to hydraulic surges. The 60 gallon unit was fed a synthetic waste with strength of 500 mg/l COD. Shock loads of 500 gph for 6 minutes, 750 gph for 4.5 minutes, and 1000 gph for 3 minutes were used. In all cases the severe hydraulic surge drastically impaired percent COD reduction although the total mass of COD removed increased dramatically. The unit required one hour to return to steady-state conditions. Even though effluent quality was impacted it is important to note deleterious effects on the biofilm were absent.

This study by Antonie was probably the best and most comprehensive study of shock loadings currently available. However, Antonie neglected to examine the effects of pure organic shocks and did not include nitrification in his research.

In a well conducted study, Stover and Kincannon (15) found that nitrification was more easily inhibited than COD removal. By using a synthetic waste of known composition, nitrification and carbon oxidation could be carefully monitored. The steady-state hydraulic loading was 0.5 gal/ft^2/day with respective COD and NH_3-N concentrations of 250 mg/l and 27.6 mg/l. Complete nitrification was achieved during this study. On two separate occasions the workers introduced quantitative shock loads to the RBC. The unit was exposed to two-fold and four-fold shocks. The percent COD removal remained relatively constant in all instances. In contrast, percent NH_3-N remaining increased while effluent NO_3^--N concentrations decreased. The authors attributed the depressed nitrification rate to possible intermediary metabolic by-products resulting from the increased heterotrophic growth rates.

MATERIALS AND METHODS

This research was conducted to determine the start-up characteristics of a full-scale RBC unit and to determine the response of the same unit to controlled shock loadings. In this section descriptions and details of the procedures and methods used to attain these goals are provided.

Rotating Biological Contactor

In 1978 CMS Rotordisk Limited of Mississauga, Ontario, loaned an S5 Rotordisk unit to the Department of Civil Engineering, Virginia Polytechnic Institute and State University, for research purposes. Primarily intended for small commercial establishments and single family dwellings, the S5 Rotordisk is designed to treat 600 US gallons per day. The fiberglass unit

includes the rotorzone (compartment containing the rotating disks), subjacent primary clarifier and secondary clarifier shown in Figure 1. The primary and secondary clarifiers had respective detention times of six and four hours. These detention times do not include ample space reserved for sludge storage.

Support for the biofilm is provided by 500 square feet of high density polyethylene 3/8 inch mesh divided into four stages. A 1/4 horsepower motor provided power for continuous rotation at three RPM (approximately 0.5 ft/sec tip speed). Wastewater enters the primary clarifier, flows under the rotorzone and enters the first stage through a slot located in the opposite corner. A smaller slot is provided at the bottom of the first stage to provide the recirculation of some aerated wastewater into the primary clarifier. The flow proceeds through the four stage RBC unit in a serpentine manner finally exiting to the secondary clarifier and eventually discharges with gravity flow utilized throughout the unit. A baffle in the secondary clarifier inhibits the discharge of floating solids.

The S5 Rotordisk was placed next to a primary clarifier at the Blacksburg and Virginia Polytechnic Institute Sanitation Authority Stroubles Creek Wastewater Treatment Plant, near Blacksburg, Virginia. Approximately 30,000 people in the Blacksburg vicinity are served by this treatment plant. This population includes approximately 21,000 students engaged in studies at Virginia Tech. Very few industries are within the service area so the wastewater is primarily of domestic origin.

An ECO C-15 Centrichem Pump was purchased by the Department of Civil Engineering to supply primary effluent to the package plant. Three C-clamps on the discharge hose provided effective and economical flow rate control. C.M.S. Rotordisk Ltd. supplied a pin timer manufactured by Hydro-Aerobics International, Inc. of Milford, Ohio for positive pump control. The pin timer provided 24 hour off/on pump control in 15 minute increments. Additionally, the pin timer provided the capability of turning the pump off any or all days of the week. Operation of the motor for disc rotation was independent of the pin timer.

Sample Collection Points

Throughout the entire research period grab samples were collected from the water surface at the same locations in the

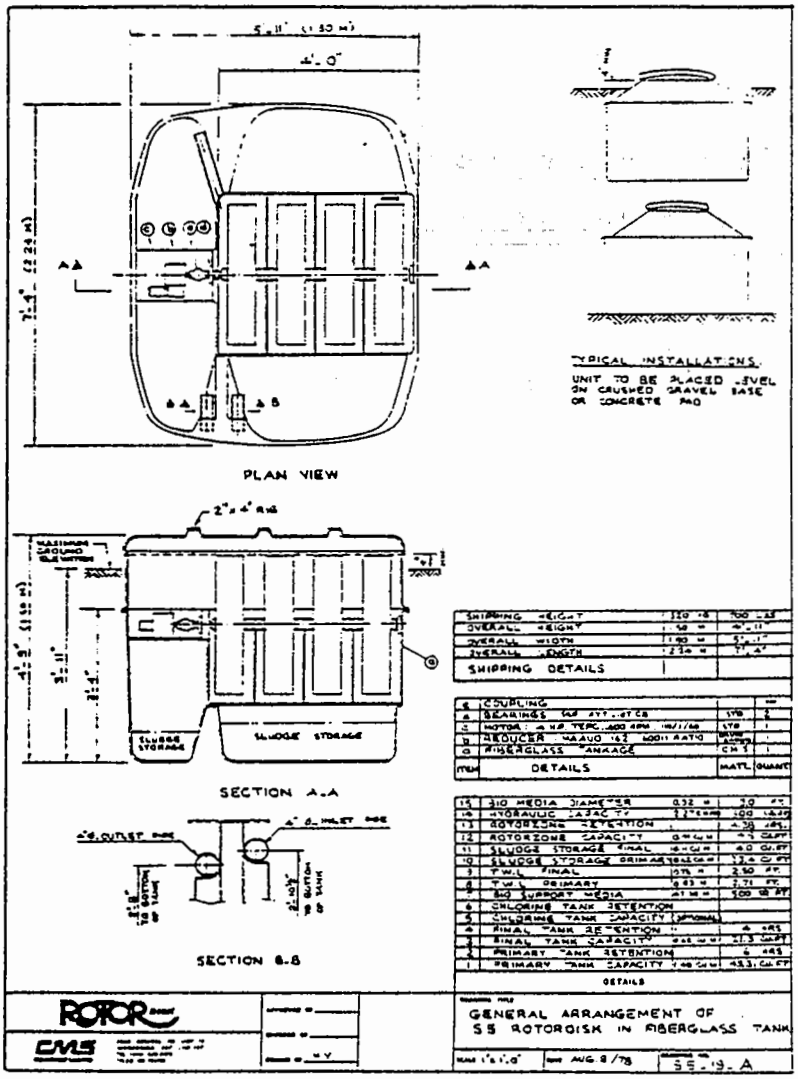

FIGURE 1 PROFILE AND PLAN VIEW OF S5 ROTORDISK

package plant. Samples were collected in the primary clarifier near the inlet pipe, in each of the four RBC unit stages and in the secondary clarifier on the discharge side of the baffle. A sample was collected from the secondary clarifier instead of the discharge because of intermittent flow. After three and one-half weeks it became apparent the sample collected in the primary clarifier of the rotordisk was not necessarily reflective of the influent composition. Therefore an additional grab sample was regularly collected from the clarifier supplying primary effluent to the Rotordisk. This sample is referred to as the influent to avoid confusion. The influent has been labeled sample collection point A, with the Rotordisk primary clarifier referred to as collection point B. Stages 1 through 4 of the rotorzone are denoted by C, D, E and F respectively. The letters G and H are used to designate the sample collected in the secondary clarifier.

Operations

Operations began on May 8, 1981 when the package plant was filled with primary effluent. A normal feedrate of 480 pgd was maintained during all phases of research. One exception was the introduction of hydraulic shock loadings. The desired flow rate was achieved by using the pin timer to alternately turn the pump on for 15 minutes and off for 30 minutes.

Start-up

The first set of samples were collected at 8:00 a.m. on May 9. Additional sample sets were collected every other day at 8:00 a.m. until the end of the start-up phase on June 22. Samples were quickly transported to laboratory facilities on the Virginia Tech campus. Suspended solids (SS) and total alkalinity determinations were immediately conducted after which all samples were acidified and cooled to $4^{o}C$ for later analysis of COD, NH_3-N, NO_3^--N, and Organic Nitrogen (Org-N). Dissolved oxygen (DO) concentrations were recorded at each sample collection point in the Rotordisk after collection of the samples.

Hydraulic Shock Loading

Three separate hydraulic shocks were conducted to evaluate the response of the unit. The initial hydraulic shock was applied on July 17, 1981. Starting at 6:00 a.m. samples were

collected at 8:00 a.m. the pump was operated continuously for eight hours. This created a three-fold increase with an eight hour duration. Each set of samples was immediately placed in an ice cooler for preservation and DO concentrations were recorded at each sample collection point. After the final set of samples were collected at 6:00 p.m., all samples were transported to the laboratory where they were preserved and cooled to 4°C following analysis of SS and alkalinity. To monitor recovery, samples were collected the following morning at 8:00 a.m. and again on June 20 at 8:00 a.m. All samples were later analyzed for COD, NH_3-N, and NO_3^--N.

The entire procedure was repeated on July 25, 1981 with the duration of the hydraulic shock extended to 10 hours. Samples were not collected at 6:00 a.m. and 10:00 a.m. Otherwise, procedures were the same as the previous hydraulic shock. In addition, monitoring of the recovery period was extended to seven days after the shock. Once again DO was recorded on site while SS, alkalinity, COD, NH_3-N, and NO_3^--N were determined at a later time.

Another eight hour hydraulic shock test was conducted on August 11, 1981 to examine reproducibility. Seven sets of samples were collected at two hour intervals from 6:00 a.m. until 6:00 p.m. on the day of the increased hydraulic loading. Samples were collected at 8:00 a.m. on August 12, 14 and 16 to monitor the return to steady-state conditions. All other collection and analytical procedures remained the same.

Organic Shocks

To examine the effects of an organic shock without an increased hydraulic loading, Kroger Incorporated (Cincinnati, Ohio) Nonfat Dry Milk was added to the primary clarifier of the Rotordisk on two separate occasions. A step feed increase was produced by thoroughly stirring the milk into the primary clarifier.

The first organic shock was conducted on August 18, 1981. Prior to the addition of two pounds of milk, samples were collected and DO recorded at 6:00 and 8:00 a.m. Normal pumping routines of 480 gpd were maintained. Samples were collected every two hours until 2:00 p.m. when it became visibly apparent the organic removal capacity was grossly exceeded. Samples were collected the following morning at 8:00 a.m. and again at 8:00 a.m. on August 21 and 23. All samples which could not be immediately transported to the laboratory were placed in an ice chest. Upon arrival at the laboratory SS and alkalinity were

evaluated. Afterwards all samples were acidified and stored at 4°C until later analysis for COD, NH_3-N, and NO_3^--N.

The entire organic shock procedure was duplicated on August 25 with two differences; 1.2 lbs. of nonfat dry milk was used, and samples were collected every two hours from 6:00 a.m. until 6:00 p.m.

ANALYTICAL PROCEDURES

Unless otherwise stated, each parameter evaluated for this investigation was determined in accordance with Standard Methods for the Examination of Water and Wastewater (16).

All dissolved oxygen concentrations were measured by means of a Yellow Springs Instrument Company, Inc. (Yellow Springs, Ohio) Model 54 Oxygen Meter.

Unfiltered chemical oxygen demand determinations were performed on all samples by use of the dichromate reflux method as prescribed in Standard Methods

The procedures found in Section 208.C of Standard Methods were utilized to measure suspended solids. All samples were filtered through 5.5 cm glass fiber filters (Grade 934 AH, Fisher Scientific Company, Clifton, New Jersey). All weight measurements were made by use of Mettler Instrument Corporation (Princeton, New Jersey) balance Model AC100, Model H 10 or Model H 18.

Total alkalinity determinations were performed on all samples by titration to a pH of 4.5. A Fisher Scientific Company Accumet Model 120 or Corning Glass Works (Corning, New York) Model 7 pH meter was used to measure the pH.

Unfiltered ammonia-nitrogen and organic-nitrogen concentrations were determined in accordance with Standard Methods. After distillation and digestion, the acidimetric method was used to determine NH_3-N and Org-N concentrations.

Unfiltered nitrate-nitrogen determinations were made in accordance with the Brucine method presented in Standard Methods. A Bausch & Lomb Incorporated (Rochester, New York) Spectronic 100 was used to measure absorbance.

COMPUTER GRAPHICS

Data from the start-up, hydraulic shock and organic shock phases is presented in three-dimensional graphs. The three variables presented are sampling location, time and parameter concentration. All three-dimensional graphs were drawn by use of the Surface II Graphics System (17). Each graph was

plotted by the perspective block diagram mode of the Surface II program. To enhance the view of the diagram, 30^{o} was selected as the angle of the observation point above the horizon. To reduce distortion from convergence of lines, the distance from the center of the block diagrams to the point of observation was assigned the value of 10,000 grid units. As a result the perspective block diagrams appear as conventional three-dimensional plots. Finally, the diagrams were placed at a 25^{o} azimuth to aid the viewer. An azimuth of -155^{o} was utilized, when it was desirable to view a diagram from the reverse side.

Difficulties were encountered when plotting the data from the organic and hydraulic shock exercises. Variables contained in the Surface II program could not be adjusted to accommodate the transition from two hour sampling intervals to 48 hour sampling intervals. Using SAS (18) multiple linear regression, intermediate sampling values were generated to eliminate this problem.

RESULTS AND DISCUSSION

The goals of this research were to determine the start-up characteristics of a full-scale RBC and to examine the response of the same unit to controlled shock loadings. The RBC package plant contained a primary clarifier, four stages of discs, and a secondary clarifier. Samples were collected from seven locations ranging from the influent, the Rotordisk primary clarifier and through each stage of the rotorzone into the secondary clarifier. The influent and primary clarifier are respectively referred to as sample collection points A and B while the four stages of discs are labeled C through F. A final collection point in the secondary clarifier is designated by the letters G and H.

Samples were analyzed for DO, COD, SS, nitrogen forms, and alkalinity. After careful consideration three-dimensional plots were selected to illustrate the trends revealed by the data. In this section the three-dimensional plots will be analyzed and discussed. Due to space limitations only eight representative three-dimensional graphs will be illustrated. Emphasis will be placed on observable trends rather than quantities consumed or generated. A trend analysis such as this will be more applicable to other RBC treatment systems. In fact, the chief advantage of three-dimensional graphs is the easy perception of surface trends. For the purposes of this study this feature compensates for the difficulty encountered when trying to read precise values from the graphs.

The three-dimensional graphs for the hydraulic and organic shock loading experimets utilize supplemental multiple linear regression data. These supplemental or intermediate values were merely intended to reflect general trends of the actual data without replacing any actual data.

START-UP

A slight bacterial growth was observed 24 hours after start-up. Two days later an increased biofilm thickness was observed, but a COD reduction trend did not begin until the fourth day. Chemical oxygen demand profiles slowly changed until 20 days passed. During the next 10 days effluent COD values were uniform and the reduction of COD concentrations occurred primarily in the first two stages of the rotorzone.

At the same time, changes in the appearance of the biofilm occurred. Initially growth on the discs was brown to grey-brown in color. This remained true for approximately three weeks of operation. By this time growth was greater on the first two stages while the last two stages began acquiring a reddish brown appearance. In addition, growth in the first two stages became filamentous.

An examination of Figure 2 reveals an average influent DO concentration of 0.5 mg/l rising to a peak of approximately 7.0 mg/l in stages 3 and 4 of the rotorzone. The DO concentration decreased to an average of 4.5 mg/l in the secondary clarifier. The graph reveals that peak DO values began decreasing on the tenth day and decreased for 18 days. At this point respective DO concentrations in stage 4 and the secondary clarifier are 3.2 and 2.3 mg/l. During the first 9 days this trend resulted from increasing COD reduction whereas the last half coincided with the start of nitrification.

As can be seen from Figures 3 through 5, nitrification began slowly after 18 days with vigorous activity recorded six days later. Influent NH_3-N concentrations consistently ranged from 17 to 22 mg/l until the end of the start-up period. Concentrations in the effluent started decreasing rapidly after 20 days and remained close to zero after 36 days. On day 18, NO_3^--N was detected in the secondary clarifier and fourth stage of the rotorzone (Figure 4). Two days later concentrations had increased to over 12 mg/l and were detected in the third stage of the rotorzone. After 30 days effluent NO_3^--N concentrations slowly decreased while NO_3^--N appeared in stage 2.

Influent total alkalinity concentrations shown in Figure 5 averaged 165 mg/l as calcium carbonate ($CaCO_3$) with a range from 151 to 180 mg/l as $CaCO_3$. After 28 days effluent

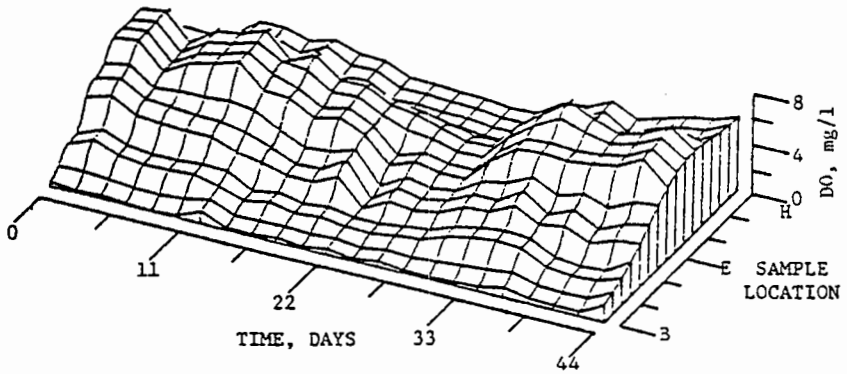

FIGURE 2 DISSOLVED OXYGEN CONCENTRATIONS FOR THE START-UP PERIOD

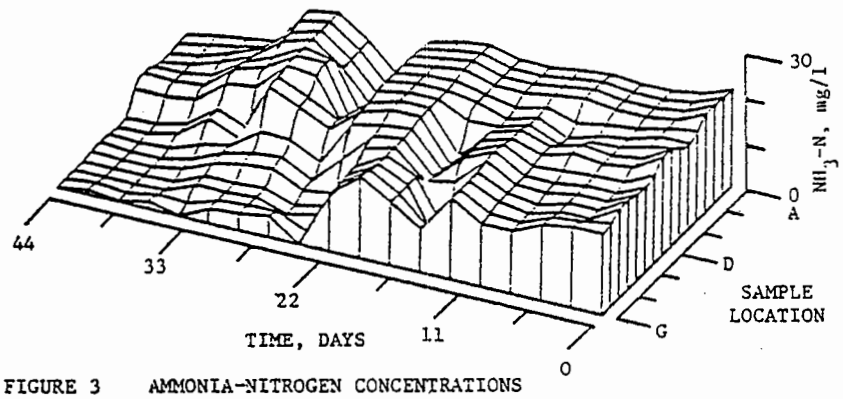

FIGURE 3 AMMONIA-NITROGEN CONCENTRATIONS FOR THE START-UP PERIOD

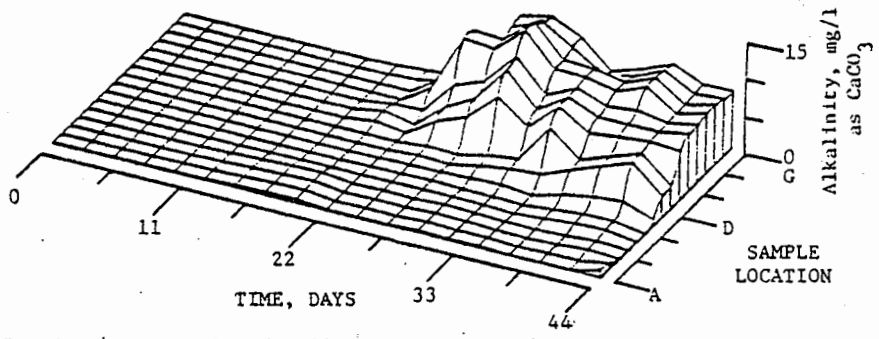

FIGURE 4 NITRATE-NITROGEN CONCENTRATIONS FOR THE START-UP PERIOD

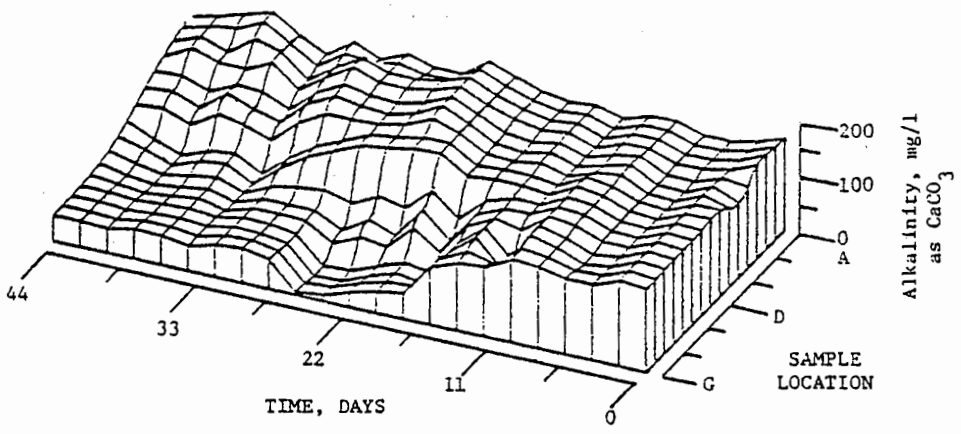

FIGURE 5 ALKALINITY CONCENTRATIONS FOR THE START-UP PERIOD

alkalinity concentrations averaged 105 mg/l as $CaCO_3$ with approximately 60 mg/l consumed. Nitrification was confined to the last two stages where a distinctive red-brown growth was present. Sufficient DO concentrations were present for nitrification throughout the start-up period.

On June 11, 33 days after testing began, final examinations were concluded at Virginia Tech. An exodus of most of the 21,000 students changed the character of the wastewater. With the exception of SS and alkalinity, influent concentrations were reduced. As a result, peak DO levels began increasing and sloughing occurred in the second stage, accompanied by a gradual change of the biofilm to a red-brown color. Afterwards COD reduction occurred only in the first stage while nitrification migrated forward to the second stage. The graphs also indicate lower effluent COD, NH_3-N, and NO_3^--N concentrations while effluent alkalinity values increased. When samples were collected on the last day, sloughing was negligible and the color change appeared complete. By this time, the second stage had changed to a red-brown color and the first stage changed from a grey-brown to a grey-white color. It is suspected that septic conditions in the sludge zone of the primary clarifier stimulated the growth of sulphur organisms in the first stage.

Suspended solids and Org-N concentration trends did not change during the start-up period. As a general rule SS and Org-N concentrations decreased as the wastewater passed through the treatment system.

In summary, it may be stated that the attached biofilm was present within 24 hours of start-up and steady-state operations in terms of COD removal and nitrification were definitely achieved in 44 days. This statement is tempered by the sudden change of wastewater characteristics during the start-up period. Consistent effluent COD concentrations were observed after 20 days while nitrification appeared to approach steady-state conditions at 30 days. The latter observation can not be confirmed because of the change in wastewater characteristics. These statements are all based, on observation at a lightly-loaded RBC package plant and may not be true under other conditions.

HYDRAULIC SHOCK LOADINGS

This phase consisted of three separate hydraulic shock loadings. On each occasion the hydraulic flow rate was tripled for a specific period of time. As expected the response characteristics were similar on each occasion.

First 8-Hour Hydraulic Shock Loading

A three-fold step increase in hydraulic loading was imposed on the RBC for 8 hours. A diurnal pattern was exhibited by the influent COD and NH_3-N concentrations. Therefore the RBC unit was exposed to increased concentrations coupled with a tripled flow rate. Dissolved oxygen concentrations decreased until two hours after the shock loading ended. This can be attributed to an increased biological activity and/or decreased hydraulic detention time.

An interesting progression of high COD concentrations was observed in the first 8-hour hydraulic shock. The high COD concentrations first appeared in the primary clarifier and stage one of the rotorzone. Two hours later the peak appeared in the second stage, disappeared at the eight hour sample, but reappeared in the third stage only to disappear at 12 hours. Because floating and rising sludge had been previously observed in the start-up phase, it might appear that the increased flow rate disturbed the primary clarifier. However, SS concentrations do not corroborate this hypothesis. Therefore it appears logical that the biofilm could not quickly accommodate the large influx of soluble organic matter. In addition to the peak values, slowly increasing COD concentrations may be observed in the first and second stages. Interestingly, effluent COD concentrations did not increase during the hydraulic shock loading.

As in the start-up period SS concentrations tended to decrease as the wastewater flowed through the RBC package plant. Generally, SS concentrations in the unit did increase towards the end of the hydraulic shock loading. Visual observations of increased turbidity confirmed this pattern.

In contrast to COD removal, nitrification was greatly inhibited. Effluent NH_3-N and alkalinity concentrations increased immediately while NO_3^--N formation decreased. Nitrification began to recover within 24 hours and was fully recovered within 74 hours.

Ten-Hour Hydraulic Shock Loadings

Trends of the 10-hour hydraulic shock are very similar to those of the first shock. Influent COD concentrations were typical of a diurnal pattern whereas NH_3-N and alkalinity concentrations were relatively constant. Dissolved oxygen concentrations decreased during the shock but began increasing as the hydraulic shock ended.

Chemical oxygen demand concentrations started to slowly increase throughout the unit six hours after the shock began.

The increases are first noticeable in the second stage and soon appeared in the remainder of the treatment unit. Effluent COD concentrations had only begun to increase when the intensive sampling period ended. Chemical oxygen demand removal returned to normal within 24 hours.

Suspended solids fluctuated considerably even when the unit was exposed to normal flow conditions before imposing the shock load. During the increased hydraulic loading, SS fluctuations intensified throughout the entire treatment system.

Nitrification was quickly and significantly impacted by the 10-hour hydraulic shock loading. At the end of the shock, alkalinity consumption, NO_3^--N formation and NH_3-N were barely noticeable. In this case nitrification recovered quickly and appeared normal after 24 hours.

Second 8-Hour Hydraulic Shock Loading

The second 8-hour hydraulic shock resembled the first 8-hour hydrualic shock. The influent COD concentrations exhibited a diurnal pattern. As the increased hydraulic loading began, DO concentrations decreased until the shock ended, thereafter DO concentrations increased. DO concentrations increased immediately when flow rates returned to normal. Chemical oxygen demand concentrations are presented in Figure 6. Influent COD concentrations fluctuated from 110 to 154 mg/l with an average value of 134 mg/l. Effluent values consistently less than 25 mg/l with only one exceptions occurring between 8 and 21 hours. Once again the last three stages were not capable of removing excess COD. In contrast to the first 8-hour hydraulic shock, effluent COD concentrations increased significantly. At the same time SS concentrations were also larger than for the first 8-hour hydraulic shock. This indicates the larger COD concentrations were probably caused by increased turbidity from the primary clarifier. The sudden decrease of SS and COD concentrations after the shock ended affirms this belief. As with the other hydraulic shock loadings, increased turbidity was visually observed but was not accompanied by biofilm sloughing.

Total alkalinity concentrations are presented in Figure 7. Influent alkalinity concentrations averaged 148 mg/l as $CaCO_3$ with a narrow range of 142 to 157 mg/l as $CaCO_3$. Under normal

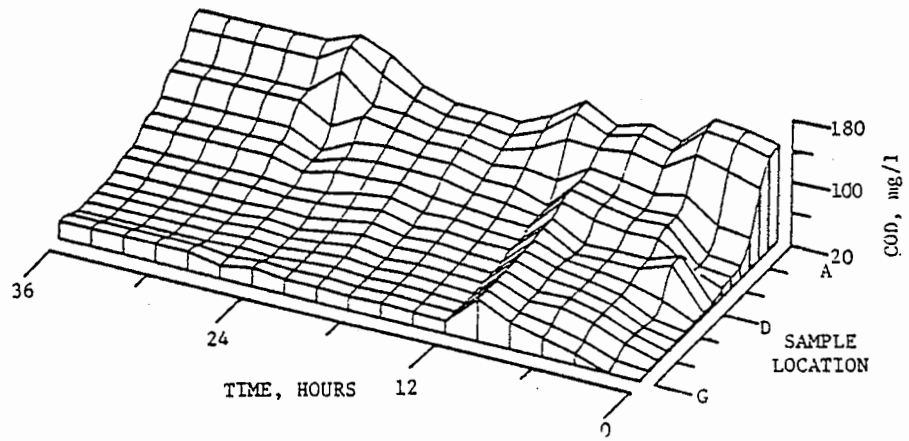

FIGURE 6 CHEMICAL OXYGEN DEMAND CONCENTRATIONS FOR THE SECOND 8-HOUR HYDRAULIC SHOCK LOADING

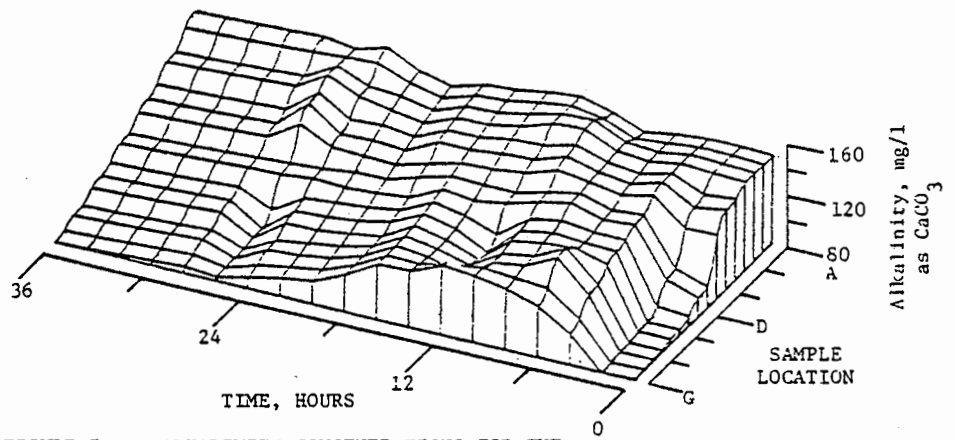

FIGURE 7 ALKALINITY CONCENTRATIONS FOR THE SECOND 8-HOUR HYDRAULIC SHOCK LOADING

conditions alkalinity consumption averaged 65 mg/l as $CaCO_3$ leaving an average effluent concentration of 83 mg/l as $CaCO_3$. As expected, nitrification was significantly inhibited by the second 8-hour hydraulic shock loading. Immediately after the shock began NH_3-H concentrations increased dramatically in the latter stages of the RBC while NO_3^--N concentrations decreased significantly. In addition, alkalinity consumption decreased to 20 mg/l as shown in Figure 7. Recovery began as soon as the flow rate returned to normal with complete recovery within 24 hours. The recovery rate compares favorably with the 10-hour shock loading response but is faster than the original 8-hour hydrualic shock.

Summary

In general, COD removal was only moderately affected by the hydraulic shocks. As expected COD removal was affected least by the shorter 8-hour shock loadings. Under normal operating conditions the last three stages of the rotorzone removed very little COD. These stages were incapable of removing larger COD concentrations for short term hydraulic shock loadings.

Nitrification was more inhibited than COD removal. In each instance nitrification was quickly inhibited to a significant degree. Recovery was slower than that of COD removal with a minimum of 24 hours required.

Peak DO concentrations declined during each hydraulic shock loading. The decrease began when the flow rate increased and recovered when the flow rate returned to normal. Sufficient DO concentrations were available for COD removal and nitrification at all times.

Biomass stability was excellent throughout the hydraulic shock loadings. Unusual or excessive sloughing did not occur as evidenced by SS concentrations.

Data from the two 8-hour hydraulic shock loadings does not indicate excellent reproducibility. In general the response of the Rotordisk to the first 8-hour shock loading was less severe. Nitrification and COD removal inhibition were greater in the second test but restoration of full nitrification was quicker than the first 8-hour experiment. This phenomenon can not be adequately explained.

ORGANIC SHOCK LOADINGS

The organic shock loading phase consisted of two large step increases in organic loading. A normal flow rate of

480 gpd was maintained during each test.

First Organic Shock Loading

During the first organic shock, influent COD remained fairly consistent. DO concentrations decreased slightly in the first 12 hours of sampling whereas the decreases encountered with the hydraulic shocks were much larger.

Chemical oxygen demand increased tremendously when powdered milk was added to the primary clarifier. Six hours later the large influent COD values were only slightly smaller and remained constant throughout the rotorzone. A normal COD concentration profile was encountered when samples were collected at 26 hours.

The addition of powdered milk to the primary clarifier did not cause a general increase in SS concentrations. Fluctuations and peak concentrations were found as usual, but with greater magnitude. A recognizable SS concentration trend could not be found.

Nitrification was seriously inhibited by the organic shock loading. When compared to the hydraulic shock loadings, the response differed in two significant ways. First the inhibition of nitrification occurred gradually. Alkalinity consumption and effluent NH_3-N concentrations increased slowly while NO_3^--N production decreased at a slightly more rapid rate. Secondly, when the 26 hour sample t was collected, nitrification was barely evident. In comparison, nitrification began to revive when the hydraulic shocks ended, and with the exception of the first 8-hour shock, were completely recovered in 24 hours. Full recovery from the powdered milk occurred within 72 hours.

Second Organic Shock Loading

Trends for the second organic shock closely resembled those of the first shock. As envisioned, the effects were less severe because less powdered milk was added to the primary clarifier. Dissolved oxygen concentrations decreased slightly for the first 4 hours but soon recovered.

Changes in COD concentrations are presented in Figure 8. Influent COD values fluctuated widely from 108 to 174 mg/l, but appear small when compared to the high value of 480 mg/l recorded in the primary clarifier. The addition of powdered milk to the primary clarifier caused high COD concentrations

throughout the unit. When the intensive sampling period ended at 12 hours, COD concentrations were decreasing throughout the unit except for the secondary clarifier. Secondary clarifier effluent concentrations increased from 45 to 159 mg/l, but declined to near steady-state conditions within 24 hours.

In contrast to the hydraulic shock loadings a general increase in SS concentrations was not encountered. A recognizable pattern was not found, except for the general decrease from the influent to effluent.

As shown in Figure 9, effluent NO_3^--N had an initial average of 10 mg/l and slowly decreased to approximately 5.0 mg/l. This demonstrates that nitrification was reduced by the second organic shock loading. In a manner similar to the first organic shock, nitrification was slowly inhibited but recovery was more rapid. Although nitrification did not completely recover within 24 hours, it recovered quicker than for the first organic shock. Recovery was completed within 72 hours as indicated by the raw data.

The second organic shock was the only shock loading which affected the stability of the biofilm. Six days after the shock was applied severe sloughing was observed in the second stage and the first stage had changed from a grey-white color to grey-brown color. This color change is the exact opposite of what occurred in the latter part of the start-up period. Since this study did not include biofilm examinations, it is impossible to conclusively state what caused the color change.

Summary

Chemical oxygen demand concentrations increased dramatically throughout the unit during the organic shock loading testing. The last three stages of the rotorzone were incapable of reducing COD concentrations. This observation was also found during the hydraulic shock loading studies.

Nitrification was inhibited by both organic shock loads. The inhibition of nitrification occurred gradually and recovered slowly in comparison to the response of nitrification to the hydraulic shock loads.

Dissolved oxygen concentrations declined only slightly during the organic shock loadings. This directly contrasts the sharp declines encountered in the hydraulic shock loadings. It is speculated that this can be attributed to a change in the organic constituents of the wastewater, i.e., the biofilm was not acclimated to the change in organic composition.

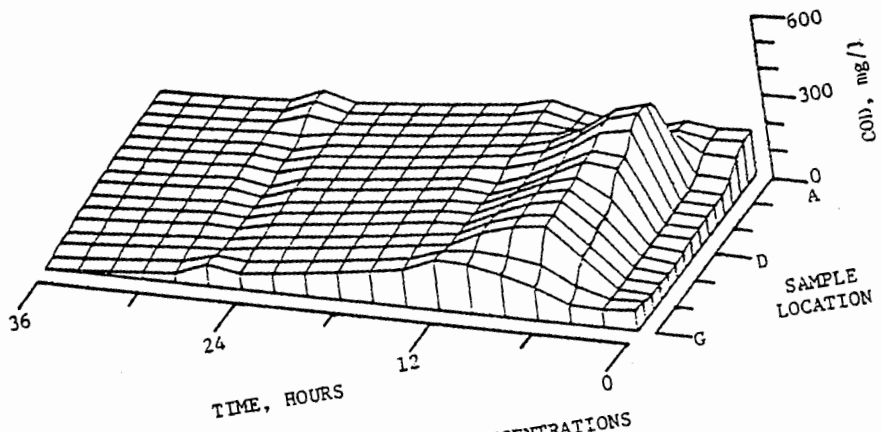

FIGURE 8 CHEMICAL OXYGEN DEMAND CONCENTRATIONS FOR THE SECOND ORGANIC SHOCK LOADING

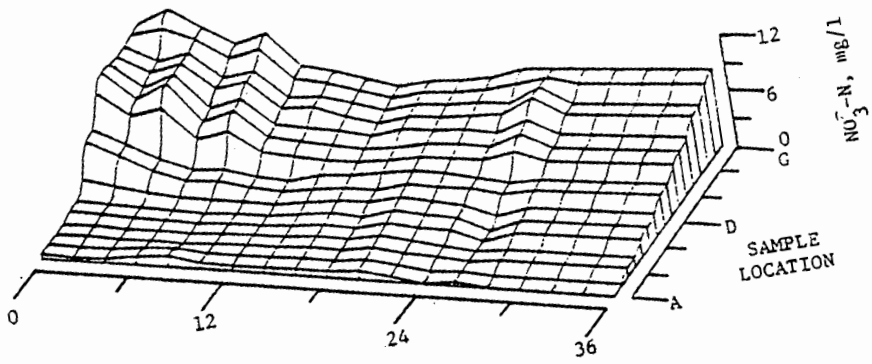

FIGURE 9 NITRATE-NITROGEN CONCENTRATIONS FOR THE SECOND ORGANIC SHOCK LOADING

Biofilm stability was affected by the second organic shock loading. The sloughing was observed six days after the package plant was exposed to the increased organic loading. This occurred approximately 72 hours after the data indicated the treatment system had returned to normal steady-state operations. A satisfactory explanation cannot be offered for this phenomenon.

CONCLUSIONS

Start-up characteristics of a full-scale RBC unit and the response of the same unit to controlled shock loadings were examined in this study. Based on an analysis of the results obtained the following conclusions are drawn.

1. Start-Up
   a) Growth of the biofilm began within 24 hours.
   b) The autotrophic biofilm was easily identified by a distinct non-filamentous red-brown color.
   c) Twenty days were required to achieve steady-state conditions in terms of COD removal.
   d) Approximately 30 days were needed for nitrification to approach steady-state conditions. This statement is based on observable trends before wastewater characteristics suddenly changed in early June.

2. Controlled Shock Loadings
   a) Hydraulic shock loads depressed DO concentrations due to decreased hydraulic detention times and/or increased biological activity.
   b) The response to hydraulic shock loadings were not very reproducible.
   c) A DO depression did not occur for the organic shock loads. Most likely this can be attributed to a change in the organic constituents of the wastewater.
   d) Nitrification was more easily inhibited by shock loads than was COD removal.
   e) Nitrification was inhibited immediately and recovered more quickly from hydraulic shock loadings when compared to organic shock loadings.
   f) The attached biofilm was not adversely affected by the shock loads.

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REFERENCES

1. Fry, F. F., "Start-Up and Shock Loading Characteristics of a Rotating Biological Contactor Package Plant." Thesis submitted in partial fulfillment of requirements of Master of Science Degree, Virginia Polytechnic Institute and State University (1982).

2. Antonie, R. L., "Fixed Biological Surfaces-Wastewater Treatment." CRC Press, Cleveland, Ohio (1976).

3. Poon, C. P. C., Chin, H. K., Smith, E. D., and Mikucki, W. J., "Upgrading Trickling Filter Effluents with a RBC System." Proceedings: First National Symposium/ Workshop on Rotating Biological Contactor Technology, (1980).

4. Bracewell, L. W., Jenkins, D., and Cameron, W., "Treatment of Phenol-Formaldehyde Resin Wastewater Using Rotating Biological Contactors." Proceedings: First National Symposium/Workshop on Rotating Biological Contactor Technology, (1980).

5. Ahlberg, N. R., and Kwong, T. S., "Process Evaluation of a Rotating Biological Contactor for Municipal Wastewater Treatment." Research Paper No. W2041, Wastewater Treatment Section, Pollution Control Planning Branch, Ministry of Environment, Ontario, Canada (November 1974).

6. Lue-Hing, C., Obayashi, A. W., Zenz, D. R., Washington, B., and Sawyer, B. M., "Nitrification of a High Ammonia Content Sludge Supernatant by Use of Rotating Discs." Proc. 29th Ind. Waste Conf., Purdue University, 245 (1974).

7. Trinh, D. T., "Exploration Camp Wastewater Characterization and Treatment Plant Assessment." Report No. EPS 4-WP-81-1, Environmental Protection Service, Environment Canada, Ottawa, Canada (1981).

8. Srinivasaraghavan, R., Reh, C. W., and Lilegren, S., "Performance Evaluation of Air Driven RBC Process for Municipal Waste Treatment." Proceedings: First National Symposium/Workshop on Rotating Biological Contactor Technology, (1980).

9. Wu, Y. C., Smith, E. D., and Gratz, J., "Prediction of RBC Plant Performance for Municipal Wastewater Treatment." Proceedings: First National Symposium/Workshop on Rotating Biological Contactor Technology, 887, (1980).

10. Welch, F. M., "Preliminary Results of a New Approach in the Aerobic Biological Treatment of Highly Concentrated Wastes." Proc. 23rd Purdue Ind. Waste Conf., 428, (1968).

11. Orwin, L. W., and Siebenthal, C. D., "Hydraulic and Organic Forcing of a Pilot Scale RBC Unit." Proceedings: First National Symposium/Workshop on Rotating Biological Contactor Technology, 119, (1980).

12. Kinner, N. E., and Bishop, P. L., "High Salinity Wastewater Treatment Using Rotating Biological Contactors." Proceedings: First National Symposium/Workshop on Rotating Biological Contactor Technology, 259, (1980).

13. Dupont, R. R., and McKinney, R. E., "Data Evaluation of a Municipal RBC Installation, Kirksville, Missouri." Proceedings: First National Symposium/Workshop on Rotating Biological Contactor Technology, 205, (1980).

14. Antonie, R. L., "Response of the Bio-Disc Process to Fluctuating Wastewater Flows." Proceedings of the 25th Purdue Ind. Waste Conf., 425, (1970).

15. Stover, E. L., and Kincannon, D. F., "One Step Nitrification-Carbon Removal." Water & Sewage Works, 122, 66, (June 1975).

16. Standard Methods for the Examination of Water and Wastewater, 14th Edition, Washington, D.C., American Public Health Association (1976).

17. Sampson, R. J., "Surface II Graphics System." Kansas Geological Survey, Lawrence, KS (1978).

18. Barr, A. J., Goodnight, J. H., Sall, J. P., Blair, W. H., and Chilko, D. M., SAS User's Guide, SAS Institute Inc., Raleigh, NC, (1979).

UPGRADING WITH SUBMERGED BIOLOGICAL FILTERS

Orval Q. Matteson. Mid-South Distributor, Jacksonville, Alabama.

It may sound like Utopia, but it is now possible to materially upgrade any aerobic wastewater treatment system by just adding three types of very simple devices to the secondary aeration and settling tanks, with the work done in-house with off-the-shelf materials. It makes no difference how big the systems are or of what type. And, the cost of doing this, related to a gpd basis, is low for a very small system and nominal for a big system. Further, if the primary treatment component is then eliminated, except for non-organic grit removal, the total overall treatment will be even better than it was and much less expensive. Digester loads will be decreased, and in many cases the digesters can be eliminated, as can tertiary treatment.

This observation is not based on just theory or on the results of bench-type experiments. A working system incorporating these techniques and devices has been operating in Jacksonville, Alabama, for over eight years, with rotifers clearly visible in its clarifier, consistently producing an effluent of 10 BOD_5/SS mg/l, + or -, even under periods of forced extreme overloads. Although this is an extended aeration unit housed in a 1000 gallon tank serving a single home, the techniques and devices employed are equally adaptable to any type of aerobic treatment unit or system regardless of its design or size. Also, as the gpd rate increases,

the cost-benefits ratio increases in geometric progression.

In the first forty-some years after the activated sludge process was developed in England in 1913, any improvements made came from operators. Just so did my concept and its application come from an operator, me, and not from a design engineer or a research laboratory. I started on this route trying to solve special problems I found in the early 70's while handling what was then the most effective package extended aeration secondary treatment unit on the market. However, the manufacturer finally stopped making it because of operational problems: it frequently clogged up.

Problems with wastewater treatment systems, even in the best designed and best operating situation, basically come from widely fluctuating growth patterns of the organisms, which cause oscillation and continuous imbalance.

However, as few systems are designed to meet the needs of or to take advantage of the natural capabilities of the wastewater treatment organisms, most systems do not fall into the category of "best designed." Also, as operators usually understand so little of the biological/biochemical aspects of their treatment systems, of whatever type, few systems can be classified as "best operated."

Let's see what such established experts as Ross McKinney and W.W. Eckenfelder have had to say on this matter.

Ross E. McKinney, in the 1962 edition of his book, Microbiology for Sanitary Engineers, said, "Fundamental microbiology offers the means for the sanitary engineer to base the design of biological waste treatment systems. It is important for the engineer to realize that all microbial systems operate on the same general biochemical principles and that the differences between the various biological systems lie in the environment imposed by the mechanical aspects of the system" (1).

W. Wesley Eckenfelder, Jr., at one of the sessions he presented at Vanderbilt University in 1971, said, "Wastewater treatment systems should be designed so that the bugs would be very happy and thus eat, reproduce and die at a great rate" (2).

McKinney also wrote that the activated sludge process is the most versatile of the biological treatment processes; that activated sludge is simplicity personified; that no other treatment process has more advantages or disadvantages; that the chief disadvantage with it lies in the lack of understanding of the basic process by both design engineers and plant operators; that the design of any biological waste

treatment system can be made properly only if the designer has a thorough understanding of the microbiology and the biochemistry of the process; that engineers never consider microbiology in the design of waste treatment systems and that the sanitary bacteriologist is not interested in the design of treatment systems; and that activated sludge is a pure biological process and yet biology never entered into its design or operation until the past few years (3).

McKinney's comments in 1962 seemed to indicate that he then expected that we today should find great improvement in the design of systems. To his observations I say, amen; to his prediction, I would have to say, wishful thinking. Think about it; how many municipal or package systems would you say were designed by a sanitary biologist? What percentage of the classes offered for BS degrees in engineering or biochemistry or in schools for plant operators, and how many of the questions on operators' tests, are on the biology or biochemistry of waste treatment?

Apparently McKinney did not think his predictions had come through by 1972 for the paper he presented at a conference in Atlanta is filled with comments such as, "The recommended design criteria for activated sludge systems employed by the various state regulatory agencies clearly demonstrates the lack of concern for the biological factors affecting activated sludge." Or, the statement, "At best the design engineer gives lip service to the fact that activated sludge is a biotreatment process" (4). He did comment that by then young engineers were getting some instruction in the "why" as well as the "how." He said that most research scientists and university professors have attempted to make the biological process more, rather than less, complicated; that there is going to have to be a drastic change in the philosophy and attitude of everyone involved. He contended that operators have failed to apply basic biological concepts to understanding how their biosystems should be operated, and that the design engineers have been of no help.

I liked what he said about it being necessary to make the system so simple that everyone can understand it. Particularly I liked his comment to the effect that if you have a system designed around sound biological principles the microbes will run their part with little operator attention, and that such attention, mainly directed to control of the MLSS through balanced sludge wasting, can be designed to operate automatically.

Incidentally, these comments of McKinney and Eckenfelder

all happen to be made in relation to the activated sludge process. However, I am confident that they would agree that they apply to all types of aerobic/anerobic systems/processes as well.

McKinney's closing comment takes the prize: "If we are to make real progress in solving the current water pollution problems, we are going to have to recognize that we must design and construct as simple systems as possible to minimize problems."

Well, statements like those of McKinney and Eckenfelder, and others, and their implications, and a review of all types of conventional systems (including the ones I had been handing) are what brought me to develop these new applications of old techniques and devices. I set out to apply the common-sense knowledge gained by hands-on experience to marry what mechanics, biochemistry and microbiology I acquired to meet the ultimate demands of wastewater treatment. I tried out my ideas until I got the results I wanted, then I got patent rights, and I now suggest that these techniques and devices be used to upgrade every aerobic system, of whatever type or size or state of operation or development.

The application is new, or I would not have been able to patent it. The possibilities these techniques and devices offer are news to the wastewater treatment business or everyone would have already used them. They are now available to everyone as license to use the features of the patent described herein can be obtained for an extremely small fee, to match with minimal costs related to the on-the-job construction and installation of the devices.

The Environmental Protection Agency puts out a big loose leaf publication titled, Process Design Manual for Upgrading Wastewater Treatment Plants (5). It emphasizes that only the most effective design and operation of treatment facilities, with the latest techniques, will meet the future water quality objectives, and that it is essential that this new technology be incorporated into the contemporary design of waste treatment facilities to achieve maximum benefits. I am pleased to note that although it seems to contradict the title, they do recognize that the term "upgrading" should also apply to systems on the drawing board or in the manufacturing process.

Unfortunately, but, as I expressed earlier, to be expected, the emphasis in the Foreword and throughout its content is on "engineering." There is very little included on ways to make the "bugs" happier that is not directly

related to a pump, pipe, or tank, and even when the structural or mechanical feature is directed to making the "bugs" happier the "whys" of it are not provided. Microbiology and biochemistry seem to have no place in the upgrading process.

Also emphasis is placed on removing pollutants by the use of chemicals that, although effective for the purpose, also always have a collaterally adverse impact on the system, rather than in the possible use of cultured varieties of specific microorganisms products such as LS-1471, BPS-2020, or GSHGD-1, or of enzymes such as Septictrine, or when applicable, of the use of the non-toxic algaecide, Cutrine-Plus.

A most disturbing feature is that, even though it is a loose-leaf manual, and thus easy to update, there seems to have been no advances in upgrading techniques of note since 1974, as the EPA manual I received in January 1982 seems to be the same as the one I received in 1974.

In categorizing the reasons for upgrading, the manual lists meeting more stringent treatment requirements, and increased hydraulic or organic loads, and to overcome improper plant design or operations. All are quite valid needs. But, they seem to have ignored more basic reasons for upgrading which are common to all systems, even those working just fine: to just improve system performance, perhaps to get "more bang" for the "total bucks" invested; or, even though all is working fine, to upgrade to improve the system's capability to meet potential shock loads, or to simplify procedures, or to reduce capital or O & M costs.

When I talk about upgrading I'm addressing any or all needs, for systems in trouble and for those working OK, for those in place or those yet to be.

Incidentally, did you ever examine the wastewater treatment patents in the Search Room of the Patent Office in Arlington, Virginia? It is an interesting experience that I recommend to any student of the science of wastewater treatment. It is a must, I think, for anyone who is thinking about trying to get an idea patented.

Even if all you do is look at the pictures, one thing that will strike you is the complexity of so many of the devices. You keep expecting to see Rube Goldberg's name on them. But then, the devices on operating treatment systems are complicated--a lot of those patent ideas got incorporated into systems on the market. You might think the objective is to establish a cult that believes that if it is not complicated it can't do the job; with a creed that simple ideas or things won't work.

I am here to attack and dispel that belief. My creed is to keep it simple and inexpensive. So, what I have to offer here to meet the objectives of McKinney and Eckenfelder, et alii, and I am sure yours as well, are inexpensive simple devices, simply applied. That is, they are simple in the sense of being easy to do and understand and operate. They are simple to construct and install and simple to care for. They are also very inexpensive to construct and install, have no moving parts, and require no O & M efforts or costs.

If your objectives are just to prevent things getting into the secondary clarifer, or to upgrade the microbiology and biochemistry aspects of treatment in activated sludge, trickling filter or oxidation ditch systems far more than is possible in conventional systems, or to increase the organic or hydraulic capacities, or to eliminate some or all of the features of primary treatment except non-organic grit removal, or to reduce final clarifier or digester or trickling filter or oxidation ditch loads, or to reduce digester problems, or even to eliminate the digester phase, then you place a series of "permeable retaining members" in each aeration tank.

If your objectives are to control velocities and currents coming into the clarifier element and to accelerate settling and to concentrate sludge both as to content and location far more than is possible in conventional systems, then you install a "permeable deflection member" in the clarifier or settling tank.

If your objectives are to produce a highly clarified effluent with very low BOD/SS, maintain a state of quiescence prior to discharge and at the same time to be able to keep up a return sludge/skimmer rate far greater than ever possible in conventional systems, and to also maintain DO levels in the clarifier capable of supporting animals such as protozoa and rotifers, then you place a "permeable restraining member" in the clarifier or settling chamber.

Of course the use of the retaining members in the aeration tanks will materially add to the effectiveness of the deflection and restraining members in the clarifier, and vice versa.

Why, or how, do my devices produce these results? Let us for the moment refresh our thinking on the fundamental microbiology and biochemistry of wastewater treatment.

Metcalf & Eddy state: "By controlling the environment of the microorganisms, the decomposition of wastes is speeded

up. Regardless of the type of waste, the biological treatment process consists of controlling the environment required for optimum growth of the microorganisms involved...Effective environmental control in biological waste treatment is based on an understanding of the basic principles governing the growth of microorganisms" (6).

Without getting too technical, let us review what they are.

Bacteria (single-cell plants) grow in a pattern of competition in mixtures of species, each organism and each species competing with the others. The prime factor is competition for food, with the dominate strains surviving. Which are dominate depends upon the type of nutrients available, the DO, temperature and pH. In aerobic systems with a proper balance of nutrients the bacteria species which survive are those that can oxidize the organic matter completely to carbon dioxide and water. Both aerobic and faculative bacteria will be found in aerobic treatment systems; the faculative use the free oxygen as long as it is available. Bacteria absorb the nutrients, produce and use enzymes to speed up the processes, metabolize the organic and inorganic compounds and produce energy and protoplasm, thus producing new bacteria (usually by splitting into two cells). For this as well as for mobility and to just stay alive they require oxygen. If the oxygen is free (available in water) they produce energy faster and more efficiently, thus absorbing food faster than if they have to make oxygen out of the wastes. If lots of nutrients are available, then available dissolved oxygen is the principal limiting factor to organic loading--increase the available oxygen and you increase the eating rate. Two of the most critical nutrients for growth are nitrogen and phosphorous, another is carbon; nitrogen deficient nutrients stimulate filamentous fungi over bacteria, which prevents good settling.

Not all organisms are beneficial. If the DO goes down below 0.5 mg/l the faculative bacteria (those which can use free oxygen or produce it from the wastes, and which always take the free if it is available) start to metabolize anerobically. At this stage filamentous microorganisms (strict aerobes) can still use the low rate free oxygen and they start to predominate; they also dominate when the critical nutrients of nitrogen and phosphorus are deficient, or at low pH. These organisms keep the floc from compacting. Filamentous microorganisms also tend to predominate over long

periods when waste food is absent because they can use the cell wall material of dead bacteria for food, which naturally the bacteria can not use.

It is easier to provide food than oxygen, for water does not take up oxygen easily; turbulence and contact time are needed. The idea is to have small bubbles, which have more surface area for contact than do large ones, bounced around and broken up, so that the water area which surrounds the bubble and thus is oxygenated will move aside and allow unsaturaged water (oxygen deficient) to contact the bubble. So, you need turbulence. You also need contact time. If you get the turbulence by increasing air pressure, or velocity, you lose transfer efficiency as the contact time is reduced; if you have too little pressure the bubbles can be too small for effective transfer or too slow to be able to break the liquid film which is resistant to the passage of oxygen. Time of contact is usually controlled by velocity and vertical depth (distance traveled).

The metabolism or growth pattern of bacteria, individually or as a mass, involve these phases.

The Lag phase is the time required for them to become acclimated to a new environment, which could extend for hours or days. A surge of food in the morning after the drop during the night, or re-entering the aeration tank from the clarifier usually produces the Lag phase. We want to cut this Lag time away down by assuring no low-food periods and reducing the holding time in the clarifier.

The Log Growth phase is a period of constant growth, individual and mass, when there is always more food than bacteria, and the only thing that holds them back is their individual capacities to eat and reproduce, and available oxygen. We want to promote this phase by providing balanced nutrients all the time and accelerating the oxygen uptake of the water. These bacteria will handle organic and hydraulic shocks. However, these bacteria are too active to floc as they do not stick together and thus do not settle well.

The Stationary or Declining Growth phase is a time when food and bacteria balance out to a level population matching growth and death, and then start to have death rates exceed growth. We have to have this phase, but want to shorten this part of the cycle as it is less productive than Log Growth.

The Endogenous or Log Death phase is when food gets progressively scarcer and the organisms metabolize their own protoplasm without replacement, keeping the mass constant to the food, which is mainly nutrients from dead cells. Bacteria are

not cannibals, they don't eat each other, they are scavengers. As their energy level drops these bacteria floc. We have to isolate this phase to get concentrated sludge. We also need to create and hold to this condition in isolation near the effluent point. (Self-metabolism is constantly occurring to some degree in each phase.) These organisms are very susceptible to hydraulic/organic shock.

The relationship between the plants and the animals is the secret of success in biological systems of any sort.

Animals, such as worms, snails and crustaceans, eat the waste and start it on its way to faster oxidation. The bacteria and other plants such as fungi and slimes eat the animals' wastes and materials coming in with the influent. Animals such as the various types of protozoa and rotifers eat the bacteria; they must have DO equal to or higher than that for aerobic bacteria. As bacteria populations develop the protozoa appear to eat the bacteria, and some organic matter, and some eat each other. Some types predominate in the Log Growth environment and others take charge through the Declining Growth and Log Death phases, depending on the numbers of bacteria and their activity (energy level), and the energy levels required by the different types of protozoa. The rotifer, a multicell animal, is a strict aerobe and thus requires a higher DO level than the others. Rotifers eat the bacteria as well as any small organic particles, such as the residue from bacteria cells which bacteria cannot process. You will have rotifers only if the water has low organic content, so if you have rotifers you have a highly efficient aerobic biological process. Some rotifers are macroscopic and can be seen without magnification.

All these animals preying on the bacteria keep the bacteria population in balance. As the bacteria population gets too low then the animals start to die off in proportion to their available food. The animals are never able to eat all the plants or other animals nor ever die off completely so the cycle is never stopped.

Treatment is never complete in an aerobic unit unless bacteria and animals are in proper balance.

The challenge is to operate the system so that it always has food available to the bacteria to control a smooth growth pattern preventing imbalance, and yet to also have a semi-starvation condition to achieve flocking, and then also to maintain an aerobic effluent staging area practically void of organic materials in order to assure a low BOD/SS effluent.

Further, for high quality treatment it is necessary to

metabolize materials and compounds that are slow to oxidize, inorganics, and those which will change from an inorganic to an organic state with time. In conventional systems most of these substances go out with the effluent, some not registering in the BOD test but producing an eventual DO demand on the receiving waters. To just meet typical effluent standards it is necessary to hold the bacteria (activated sludge) in the system for several days (6-15) striving for sufficient sludge age (mean cell residence time, or MCRT). For conventional systems this means MCRT to meet prescribed standards. However, high quality treatment requires sufficient MCRT to completely metabolize all materials and compounds (20 days, 90-95% oxidation). Nitrification requires at least 6-10 days MCRT as the nitrifying bacteria have a very slow growth rate.

Also, the objective is to hold hard-to-oxidize solids such as grease, hair, seeds, or rubber in the aeration tank. This is not generally accomplished in conventional systems. These materials either pass out with the effluent or settle out in the aeration or secondary clarifier tanks, where they produce rising and bulking sludges and keep the bacteria out of the Log Death phase.

Secondary clarifiers have to be capable of controlling velocity and turbulence to permit settling and clarification, to produce concentrated sludge, to be able to remove the sludge fast enough to keep it from turning anerobic, to keep the bacteria in condition to minimize the shock when they return to the aeration tank, and to support protozos and rotifers. In current designs these objectives can seldom be achieved: sludge is not concentrated; it becomes anerobic; Lag periods can last several days; it is impossible to achieve anything like quiescence in the effluent holding area; DO levels are low.

When all these most complicated challenges are met we have, except for pH and temperature control, succeeded in meeting the objectives of McKinney, et alii: we have a system based on the concept of controlling the environment of the plants and animals. But, in order for everyone to be able to have a system with such an environment we need to change the engineering of conventional systems, whether in operation or on the drawing board.

It can be done now, using my devices.

Opinions I will offer on the effectiveness of these devices and the effect they have on wastewater treatment will seem to some to be iconoclastic. That is good because we need to jolt many of our sanitary engineers and biochemists

(remember, no new pages to the EPA manual on upgrading systems since 1974). Also, analysis of the possibilities presented and the mechanics, microbiology and biochemistry involved in changes of this nature can be a fertile field for laboratory and operational evaluations. Further, the possibilities which are created for new designs for total systems and their various elements, as well as conversions of existing systems, by the use of these devices may be a boon to engineers as well. We may have opened a veritable treasure house of possibilities for the waste treatment industry!

Let us consider what these devices do, and how they create a system based on biological/biochemical principles, and what changes they can effect, and why.

Let's follow the sewage through the treatment system.

Conventional systems use racks, mechanical or otherwise, mechanical screens and grinders of various types in primary treatment to prevent materials from getting any further because they will upset the procedure, clog pumps, pipes and equipment, and cause delays and generate the need for costly repairs or replacements. Primary clarifiers, and sometimes skimming and preaeration tanks separate solids and liquids. Capital, energy, and O & M costs for these items are high.

It is not necessary to have all these machines, tanks, pipes and pumps to accomplish the purposes for which they are used.

My "permeable retaining devices," installed in the aeration tank do a better job and cost practically nothing. But, even when used in conjunction with all the primary system's apparatus, they still meet the challenge of providing the means to materially upgrade any treatment system's operation at practically no cost.

The "permeable retaining devices" installed in the aeration tank operate, to put it simply, as do nets or filters. Framed, they are installed so as to extend from side to side and from the bottom to above the water level. They are made of any material which is impervious to the wastewater, e.g., treated metal, plastic, synthetic fibers. The mesh in the network of materials may be formed by any means, such as punching, weaving, braiding, molding, into whatever size or shape is desired. The retaining element can be in any configuration, such as that of a fish net or a furnace air filter.

Several devices are installed per tank; the more used the more effective the results--mainly because each device functions both as an habitat for organisms and to create

water turbulence. The buildup of the materials caught and the slimes and activated sludge will permit particularly the first device to entrap materials much smaller than their mesh. Also, even a final mesh that will entrap seeds will, because of the overall area of the device, have a total void space considerably greater than the maximum square inches of all the influent and return sludge/skimmer pipes combined. Mesh sizes are decreased progressively along the direction of flow; the largest mesh size depending on the nature of the solids expected to be received.

Spacing between devices is not too material; however, biological and oxygen transfer benefits can be increased by placing two or more devices very close together to permit the development of a biological labyrinth between them.

Even for the most sophisticated treatment plant, simple devices will meet the need, for example ones made with treated angle iron frames and fish netting, braced across the face to support the netting against the pressure of the flowing water with more angle irons, and with the frames supported in place by angle irons fastened to each side wall. The use of off-the-shelf materials and on-the-job construction (and a very low license fee for a permit to use the devices) results in a very unimposing total outlay of capital funds. (Materials only, one retaining device 14' x 26' abt. $140.00.)

If the use of my retaining devices is carried to its potential all primary treatment can be eliminated except for non-organic grit removal. The retaining devices catch everything that is in motion at the place in the tank where you want it caught, depending on its size, shape and composition from hogs' heads to grease, sand, seed, hair and some live and dead organisms. These caches of solids and compounds are held in place until decayed and the bacteria, and some animals, have stabilized the organic matter and inorganic compounds. Things not at all biodegradable, even plastic bottles, stay there with the retaining devices, they and the devices constantly bathed in oxygen, with each acting to accelerate oxygen transfer by decreasing liquid circulating velocities, bubble size and the thickness of the liquid film. Collectively they create turbulent habitats for plants and animals the likes of which should enrapture a student of the science of wastewater treatment: fungi, slimes, bacteria, worms, snails, roaches, protozoa, rotifers. All gather eating and reproducing, challenging man's ability to make water take up oxygen.

Satisfy the organisms' nutrient and oxygen supplies and

control temperature and pH and they can have their Utopia. My retaining devices make it possible to fully utilize all the wastewater's nutrients. They help to maximize oxygen transfer (as does another device in my patent which is not a biological filter).

Things which are quickly biodegradeable as well as those requiring very extended oxidation are held up in transit either physically or by ad-or-absorption until changes in size or composition permit them to flow on to their next retaining station. Benefits increase when the devices are used in systems processing water with fiber contents.

The physical effects of this retention are evident to all, but the microbiology and the biochemistry effects are more dynamic--the "bugs" are happier.

Because of the around-the-clock availability of nutrients the primary bacteria in those areas of the aeration tank eat and grow/reproduce faster and continually; the food to organism ratio (F:M) is high. Because of the rapid recirculation, lag periods for these strains of bacteria are controlled. Therefore, the protozoa increase in numbers and activity. In turn, the rotifers increase and prosper, which in turn results in high quality effluent. Because habitats are provided, other plants and animals can flourish, accelerating decomposition.

All diets are fortified by the now plentiful elements and nutrients in the things being served to them that heretofore have been hauled away to be burned, drowned or buried--all of which is expensive, a nuisance, detrimental to the environment, and whose fate rightly should be oxidation in the system, not in a fire or the ocean or a landfill. Food is always available where materials are retained; in those areas the primary bacteria which thrive in a high food ratio environment can approach Eckenfelder's goal: eating and reproducing at a great rate. More technically, these bacteria can stay in a Log Growth or increasing rate phase; here they produce the maximum in the removal of organic and inorganic matter per unit of organisms.

With the restricted flow of solids and even microscopic materials as the wastewater flow continues on through the series of compartments created by the progressively smaller mesh retaining devices, the plants' food ratio drops, producing decreased energy levels. This promotes progressive stages of bacteria life, on through the Stationary phase of declining growth and increasing death into the Endogenous or Log Death phase where growth is exceeded by death, which is

the state so necessary for good floc formation and settling. During these phases, secondary bacteria strains which best utilize the products of dead bacteria are predominate. Although they have a longer Lag phase upon recirculation than do the primary species, their rapid recirculation will reduce their normal time of recovery, just as it does for all others.

Because of these biochemical reactions many things occur.

Nitrogen and phosphorus removals are heightened because of the increased availability of nutrients and energy and increased MCRT. The more balanced nutrients available in all the waste assure that bacteria which oxidize organic matter completely to carbon dioxide and water will predominate and survive over bacteria with incomplete metabolic patterns.

The relationship between fluctuating influent flow rates and BOD variations loses its importance in both design and operations, as bacteria at the influent end will always be in the Log phase. Preaeration is eliminated, as is the need for rough trickling filters or add-on RBC's. Hydraulic loading can be increased, or volume requirements can be decreased.

Sludge age, F:M, mean cell residence time, MLSS, all critical factors are positive. Thus, waste stabilization is intensified, formation of a strong floc is enhanced, nitrification can occur in the aeration tank and denitrification during the Log Death phase, and foaming is minimized. The five-day and past-five-day BOD/COD and SS usually passed on to receiving waters are substantially reduced as materials not oxidized in conventional systems are retained until oxidized or reduced to true inorganics.

All of the primary treatment phase can be eliminated except for non-organic grit removal. Some equipment and facilities can be converted to other uses, e.g., primary clarifiers can be converted to aeration tanks or to post-aeration to handle the biological solids sloughed off of trickling filters thus eliminating expensive microscreens or settling tanks.

Oxidation ditches, trickling filters and lagoons reap the same physical and biochemical benefits from using these devices as does the conventional activated sludge system. RBCs are not fouled, nor is the media in trickling filters, nor are aeration devices. By changing the primary clarifier to an aeration chamber with my devices, roughing filters can be converted to conventional filters for increased gpd capacity or for recirculation to increase quality, or conventional filters can be upgraded. Tighter media can be used to increase quality.

Eliminating primary treatment not only greatly increases the efficiency of secondary treatment when my devices are used, it also eliminates problems with digesters including those caused by grease and cellulose. Digesting times are reduced; shock loads are eliminated, as are volatile materials. Hair, which raises havoc with digesters, especially in towns like Auburn, Alabama, where the university generates most of the wastewater, does not get into the digester. Sludge going to the digester is decreased in bulk, can be pumped easily, pump and pipe sizes are reduced; the digester's sludge does not contain fertile seeds.

All in all, many generally accepted contentions are now no longer valid, e.g., the statement in the EPA manual, "...since clarification is the most economical way to remove suspended and collodial pollutants, every effort should be made to improve primary clarification process before additional primary or secondary facilities are considered" (7). Nor is Metcalf & Eddy's statement, "Primary sedimentation is most efficient in removing coarse solids" (8). Neither is the conclusion reached by Hoyland and Harwood: "It appears that the most cost effective treatment is without primary sedimentation; however, the increased costs of maintenance resulting from the accumulation of rags and coarse solids would not warrant secondary treatment only" (9).

Now let's get on to the secondary clarifier, or settling chamber.

The EPA's manual on upgrading states that "...Of all the process design variables which can effect overall plant performance, those which are selected for secondary clarification are the most critical" (10).

Quality secondary clarification is measured by how well it produces concentrated sludge, how clear is its effluent, and how high is the effluent DO. To accomplish these objectives requires: weak bacteria in the final stages of the Endogenous phase; the means to concentrate settling and remove the sludge without affecting the effluent; the elimination of turbulences in the pre-effluent holding area; the means to practically eliminate organic residue from the effluent; and a high DO in all parts of the clarifier.

Concentrated settling and turbulence control starts with my "permeable deflection member." This is a porous baffle with a fine mesh, constructed somewhat like the retaining device. Tanks without a weir entry should if possible be converted as the weir not only lessens the weight and force of the incoming liquids as compared to a pipe feed, it also aug-

ments aeration. The deflector device is mounted against the weir, under the clarifier liquid level, inclined upward and outward into the clarifier, sized to fit the flow volume Its depth is such that it catches the flow before it starts to divert from the vertical to the horizontal. Its width is such that the deflected flow over the top of its lip will be out of the main thrust of the downward flow, causing the two flows to partially intersect, thus producing resistance and counter currents which retard the deflected upward flow toward the deflector lip. The mesh construction of the deflector permits some of the incoming flow to pass through it at a reduced rate and in a dispersed condition, carrying with it some of the activated sludge. These actions at the deflector result in a slowed down horizontal flow into the clarifier causing the sludge to settle just off the deflector and out to the porous restraining device (see next par.). Both zone and compression settling result in a concentrated sludge in a controlled area. The returned sludge is removed from here, as is sludge for discharge; skimming capability is provided. (See device in my patent.)

Adjacent to the permeable deflector, but outside any horizontal currents it has created, I install a "permeable restraining device" similar in its physical characteristics to the aeration tank's retaining device, but with a very fine mesh; this separates the clarifier into two compartments. Water only enters the pre-effluent holding compartment to replace that lost via its return sludge/skimmer devices and as effluent. As this flow is relatively low per square inch of the restraining devices very little sludge passes the resistance of the mesh into the pre-effluent compartment.

The end result is a pre-effluent discharge compartment with very few solids which free-settle within a continuous state of almost complete quiescence, and in which rotifers are clearly visible.

To install these devices in some types of clarifiers will require modifications. However, the same principles will be applicable.

As a measure of the effectiveness of these devices, I operate with a return sludge/skimmer rate over 600% more than the designed gpd influent rate, with about 60% of the flow from the first compartment. Even when for testing purposes this flow was combined for eight continuous hours with a forced influent flow 80 times the designed gpd rate, the effluent compartment not only retained its quiescent state, but the effluent's typical high SS quality was maintained.

The DO quality of both compartments stays about the same as that in the aeration compartment because of the rapid return rate and low demand in the pre-effluent compartment. That extends the system's purification capacity and provides for the final stages of the Endogenous phase, where the lack of energy sources forces bacteria to metabolize their own protoplasm without replacement and the protozoa and rotifers consume the active bacteria and residue.

In Bloodgood's publication is the statement, "Sedimentation is influenced by the laws of physics" with the inference that laws of bacteriology and chemistry do not apply. I disagree, they all apply. I also disagree with the statement which followed, "In treatment of sewage by sedimentation there are not feasible ways of modifying the process to improve it." I do agree with his later comment, "Perhaps, in time, a settling tank can be designed that will eliminate all turbulence caused by entering sewage" (11). I have one.

Yes, my observations are iconoclastic, but they do open up fertile fields for students of the science of wastewater treatment.

High rate recirculation and the concentrated sludge generated by these devices produce profound results in the aeration tank. For example: rapidly recirculated organisms do not require many hours, or generations, to acclimate to the food in the aeration tank; incoming wastewater is rapidly mixed making food available throughout the retaining areas; organic, hydraulic or toxic shocks are dispersed; rapid horizontal water movement, and flow through the retaining devices deflect and disperse rising air streams, extending their contact time; mixture of the concentrated return sludge with incoming wastes tends to overcome foaming, partially because of the high ash content of the sludge.

Further, widely fluctuating growth patterns are leveled out as the constant oscillation and continuous imbalance of organisms common to conventional systems are practically eliminated. As MCRT is extended, exposure to DO throughout the system allows the bacteria to get a little further into the Endogenous phase with each cycle; as a bacteria's age increases the reduced activity permits the slime layer around the bacteria to be retained--not sheared off, promoting floculation and thus clarity. Rising and bulking sludges are limited regardless of fluctuations of incoming wastewater. Also plug flow type systems develop conditions similar to those of complete mix, Biosorption, complete oxidation, multistaging, and extended aeration systems. As wastes are retain-

ed and recirculation is rapid, the retention periods for waste and bacteria are no longer a factor of tank volume.

The use of chemicals to augment waste stabilization, to promote settling, to remove compounds or to disinfect, with all their costly up-front and after-use effects is at least reduced.

Because of the actions caused by all three types of devices, the aeration tank and clarifier will not be as susceptible to washout as in conventional systems. Thus, there should be a decrease in bypassing, which, though universally practiced, is the curse of pollution prevention objectives.

Particular application of these devices is evident for systems designed to prepare wastewater for reuse as grey water, irrigation, fish farming, ground water recharge, industrial purposes, etc. With their use aerobic digestion will become more attractive; the cost effectiveness of pure oxygen systems will improve. In some cases it will be feasible to eliminate digesters and dispose of the activated sludge directly to the land, or into surface or ocean waters, or to process it for use as fertilizer or soil conditioning/ reclamation, or for animal food.

Benefits far in excess of the cost will be realized through upgrading even if my devices are just used in either the aeration or clarifier tanks, or in both, without any further changes to existing systems or to conventional designs. However, maximum benefits on a progressive scale can be realized if the primary treatment system, less inorganic grit removal, is eliminated and as changes are made to other elements of the total system.

Yes, additional air delivery capabilities will be required if the primary system is eliminated, but perhaps not if it is retained as the devices generate more efficient use of the conventional air supply. However, any such costs will be more than offset by the savings realized and upgrading accomplished.

Extraordinary savings would be realized for both existing systems and for new construction if the adoption of the concepts created by the use of these devices would result in eliminating the primary treatment phase of aerobic wastewater treatment. That would be compounded if the digester phase or tertiary phase could also be discontinued. Such prospects far fetched as they may seem at first consideration, are most exciting!

The degree of benefits in any case increases in direct relationship to the extent the devices are used, progressively

as conventional facilities and equipment are eliminated and in any case exponentially as the size of the plant increases (the more gpd, the more "bang" for the "buck").

Yes, I believe a new horizon of opportunities for elevating the science and the art of wastewater treatment has arrived, made possible by these simple mechanical means of maximizing the potential of environmental control in biological-biochemical wastewater treatment. This concept, answers, to a considerable degree, the challenge McKinney made in Atlanta in 1972 to design systems around biological principles so that the microbes could run their part of the plant--perhaps someone will develop the automatic sludge-wasting control he recommended which will match the conditions created in this type of system (12).

Remember, to materially upgrade your treatment system now, you don't have to go so far as to eliminate the primary system, or be concerned about technical matters such as the organisms' Lag or Endogenous phases. You can keep everything but water and activated sludge from leaving your aeration tank, or slow down the rush of water into your clarifier, and get all the benefits of almost floc-free water in the outlet end of your clarifier now, by using these permeable devices I have described. You will be happier (and so will the "bugs").

ADDENDUM:

That aerobic package unit I mentioned earlier has a nylon filter bag hung in a round fiberglass tank, and the water had to go through the bag to get out of the tank, only frequently the bag clogged. I made a few changes in it to make the "bugs" happier and to make it work better and installed it as a tertiary system behind this secondary system I patented. For over eight years it has been passing water about as fast as it comes in, even during that eight hour period of forced flow. The effluent is as clear as drinking water; the laboratory never could find any trace of BOD/SS, but they always rated it at 1 BOD/SS mg/l, + or -, because, they said, there just had to be some.

References:

1. McKinney, R. E., "Microbiology for Sanitary Engineers," *McGraw-Hill Book Co., Inc.,* 1962, p. 194.
2. Eckenfelder, W. W. Jr., Seminar, "Process Design in Water Quality Engineering," held at Vanderbilt University, Nov. 1-5, 1971.
3. McKinney, R. E., ibid., pp. 213, 233.
4. McKinney, R. E., "The Value and Use of Mathematical Models for Activated Sludge Systems," presented at the October 5, 1972, International Conference sponsored by The International Association on Water Pollution Research and The Georgia Institute of Technology, a loose leaf manual w/cps of the several presentations, presented to participants, pp. 1, 2, 3, 7, 11, 12, 23, 24.
5. "Process Design Manual for Upgrading Existing Wastewater Treatment Plants," *U. S. Environmental Protection Agency,* October, 1974, Foreword.
6. Metcalf & Eddy, Inc., "Wastewater Engineering," *McGraw-Hill Book Co., Inc.,* 1972, p. 386.
7. "Process Design Manual for Upgrading Existing Wastewater Treatment Plants," ibid., p. 6-1.
8. Metcalf & Eddy, Inc., ibid., p. 481.
9. Hoyland, G., and Harwood, N. J., "Design of Biological Filtration Plants," *Journal Water Pollution Control Federation,* June, 1981, pp. 694, 699.
10. "Process Design Manual for Upgrading Existing Wastewater Treatment Plants," ibid., p. 6-2.
11. Bloodgood, D. E., "Sewage Treatment Practices," *A Scranton Gillette Publication*, undated, pp. 39, 41.
12. McKinney, R. E., ibid., No. 4, p. 23.

PART VI: MUNICIPAL WASTEWATER TREATMENT-CASE HISTORIES

RBC FOR BOD AND AMMONIA NITROGEN REMOVALS AT PRINCETON WASTEWATER TREATMENT PLANT

Shundar Lin. Water Quality Section, Illinois State Water Survey, Peoria, Illinois.

Ralph L. Evans. Water Quality Section, Illinois State Water Survey, Peoria, Illinois.

Warren Dawson. Princeton Wastewater Treatment Plant, Princeton, Illinois.

INTRODUCTION

The rotating biological contactor (RBC) process has become an attractive alternative for treating wastewater. Currently there are over 300 RBC installations in the United States. Nevertheless there have been only a few long term on-line assessments (1-8) of the process. And despite reported advantages of its simplicity, small space and low energy requirements, stability, and capability for carbonaceous and nitrogenous removals there remains skepticism of its utility as a reliable wastewater treatment process in Illinois. Design criteria are still questionable; and frequent mechanical and structural failures remain unresolved.

The installation of an RBC system at Princeton, Illinois in 1979 provided an opportunity to evaluate the overall efficiency of an RBC process, in terms of BOD_5 and ammonia removals, as well as some of the changes that occur in the chemical and biological components of the wastewater

during its passage in series from one contactor to another. Observations and sampling were performed over a 12-month period at intervals, generally, of two times per week.

A comprehensive evaluation of all the data and observations assembled during the course of the study is not within the scope of this paper. Rather, this presentation is limited to selective data and observations. It will provide some insight on the influence each contactor has on the waste stream. It will also offer a basis for a more rational design of RBC systems than is the current practice. A more complete report, including all data gathered, is in preparation.

STUDY PLANT

The City of Princeton is a community of about 7,000 persons located 120 miles (193 km) southwest of Chicago. Its climate is characterized by cold, snowy winters, and hot humid summers. Four major industries contribute about 15 percent of the waste flow. The BOD_5 of the industrial waste flows is similar to domestic sewage with suspended solids concentrations considerably lower. Information regarding ammonia concentrations is not available.

For about 30 years an activated sludge process served the waste treatment needs of the city. And before that a trickling filter installation sufficed. In 1979 an RBC system went on line to replace activated sludge process. A sand filter arrangement was provided as a "finishing" process for the final effluent. The waste water treatment facilities are designed to meet effluent requirements of 10 mg/l BOD_5 (total), 12 mg/l suspended solids (SS) and 1.5 mg/l ammonia nitrogen (NH_3-N).

RBC PROCESS AND DESIGN

A layout of the waste treatment facilities is shown in Figure 1. During dry weather flow raw sewage is pumped to the primary clarifiers. Settled sewage flows by gravity to two trains of RBC units operating in parallel thence to secondary clarifiers where the clarified effluent flows to sand filters. During high flow periods a portion of the settled sewage (44 percent) passes through the trickling filter unit then through intermediate clarifiers. From there

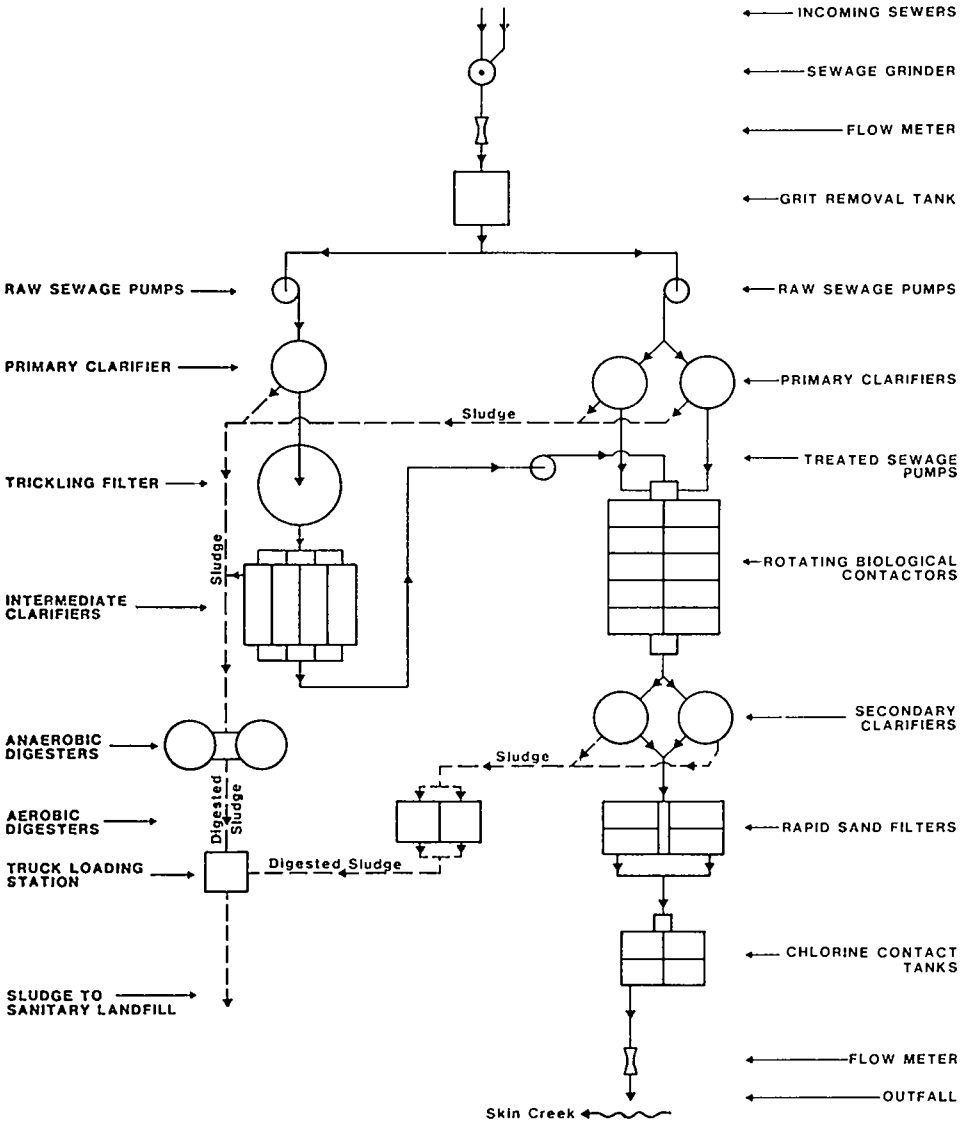

Figure 1. Schematic flow diagram of Princeton Wastewater Treatment Plant

by pumpage the filters clarified effluent is combined with settled sewage from the primary units at the head end of the RBC units.

In each train of the RBC units there are five Bio-Surf contactors manufactured by Autotrol Corporation. Each of the 10 shafts supports media 12 feet in diameter and 25 feet in length consisting of corrugated polyethylene. And each shaft is mechanical driven by a 7.5 hp motor at a design rotation speed of 1.6 rpm. Media submergence is about 40 percent. Wastewater flow is perpendicular to shaft rotation. All units are protected from weather by fiber-glass housings.

Each of the first two stages in each train provides 100,000 square feet (9290 m^2) of media (standard). Each of the three remaining stages in each train provides 150,000 square feet (13935 m^2) of media (high density). Thus a total of 1,300,000 square feet (120,770 m^2) of fixed film media are provided for the process. The tanks housing the contactors are flat-bottomed with a trapezoidal section at each end. Each contactor is separated by an underflow-overflow baffle thus providing a complete mixed reactor type of arrangement in series. These details are shown in Figures 2 and 3.

The pertinent design features for the RBC system are as follows:

Design flow:	1.63 mgd	(6170 m^3/d)
Peak flow:	4.58 mgd	(17,300 m^3/d)
Hydraulic loading:	1.25 gpd/s.f.	(51 $l/m^2/d$)
BOD_5 loading:	1.12 lbs/d/1000 s.f.	(46 $g/m^2/d$)
Ammonia loading:	0.1 lbs/d/1000 s.f.	(4 $g/m^2/d$)
Detention:	2 hours	

In terms of concentrations in the settled sewage applied to the RBC system the anticipated average BOD_5 and ammonia, for design purposes, was 110 mg/l and 13 mg/l, respectively.

METHODS AND PROCEDURES

Observations and sample collections were performed on the south train of the RBC system from January 6, 1981 to January 14, 1982. Samples of wastewater were collected as 24-hour composites generally during the periods Monday 1000 to Tuesday 1000 and Wednesday 1000 to Thursday 1000 with ISCO (Model 1392) samplers. Seven stations were monitored.

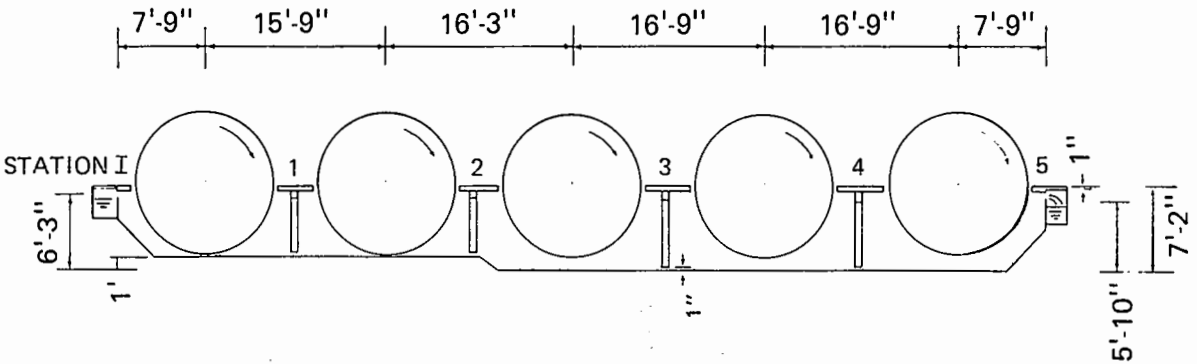

Figure 2. Profile of an RBC Train

to doubling the load that would be normally applied to the 5-unit RBC system. For the purpose of this report reference to the single train operation is designated as the shock period.

RESULTS AND DISCUSSION

The results shown in Figure 4 are typical of the observations recorded during the tests to determine the degree of short circuiting. As shown, there is no basic difference within the water column during passage from one unit to another in terms of temperature, DO, pH, and alkalinity. The tank of each unit behaves as a completely mixed reactor. Similar observations have been reported by others (1,10,11,12).

The observed diurnal changes in the waste stream for temperature, DO, pH, and alkalinity are shown in Figure 5. These are probably typical of warm weather conditions. Temperatures differed about $3^{o}C$. Dissolved oxygen fluctuated 1.5-2.5 mg/l. The diurnal ranges for alkalinity and pH were narrow.

A summary of the seasonal quality of the influent to the RBC system is included in Table I. Also included is the influent quality during the shock period. Wastewater temperatures ranged from 7.5^{o} to $20.9^{o}C$. Dissolved oxygen concentrations were frequently below 1 mg/l except during spring months when snow melt and infiltration contributed to the flow. Total (T) and soluble (S) components of ammonia and Kjeldahl nitrogen, and BOD_5 were quite variable. Ammonia concentrations were generally higher during winter months. The several-fold variations depicted in Table I suggest that inflexible modelling procedures may not be productive.

Dissolved Oxygen and Suspended Solids

As mentioned earlier DO measurements were not made over a 24-hour period. Rather they were instantaneous measurements performed during the time of sampling equipment set-up and the gathering of sample containers. Nevertheless sufficient data was recorded to estimate the pattern of DO changes in the waste stream during its passage through the 5-unit RBC system. A typical pattern is shown in both Figures 5 and 6. Generally there was little change in DO concentrations during passage through the first three units. The fourth unit generally increased the dissolved oxygen concen-

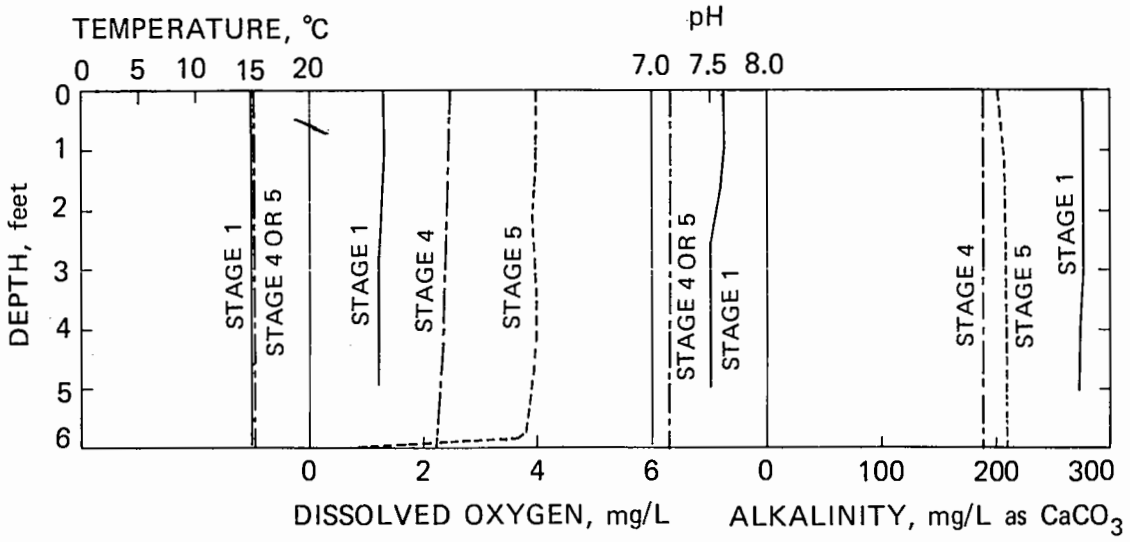

Figure 4. Temperature, dissolved oxygen, pH, and alkalinity profiles in RBC units

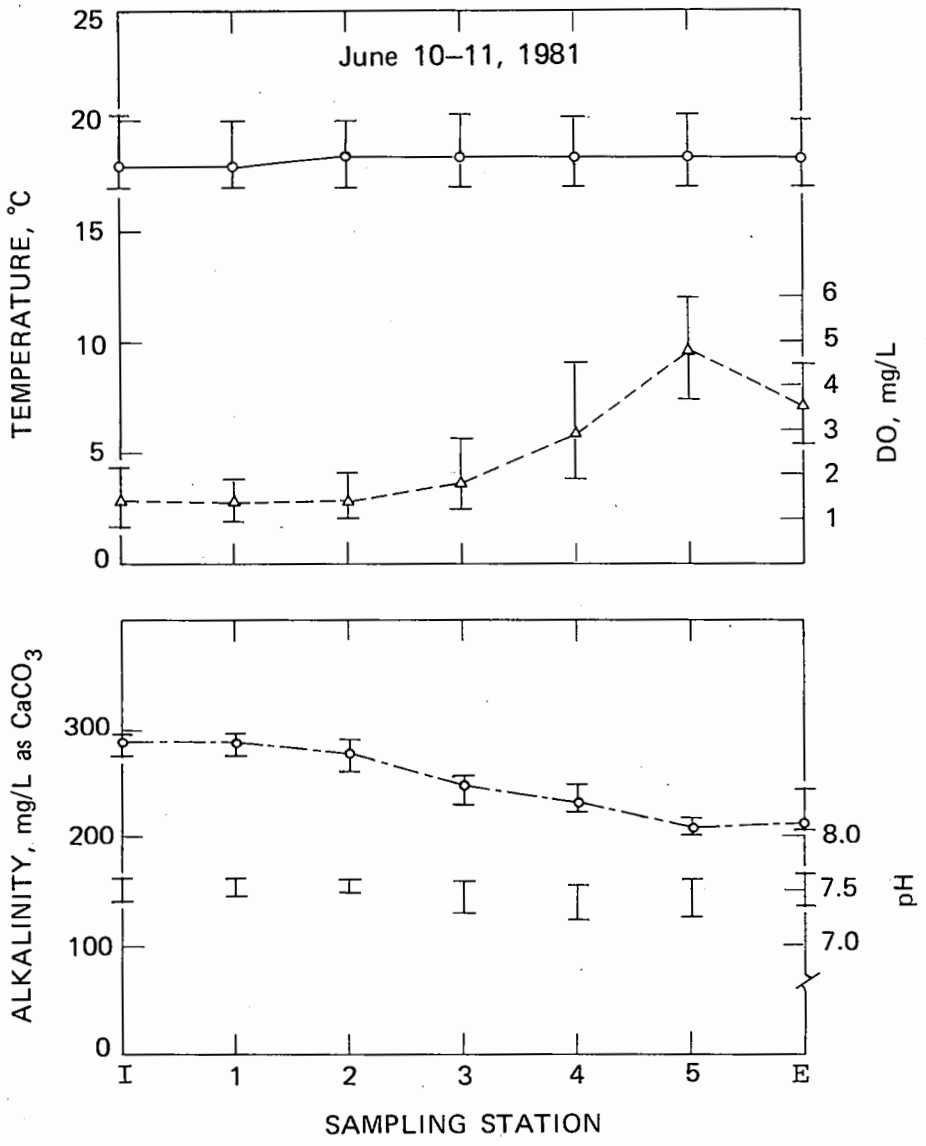

Figure 5. Means and ranges of temperature, DO, pH, and alkalinity for the 24 hourly collection on June 10 and 11, 1981

Table I. Water Quality of RBC Influent at Princeton

	Winter 1/6-2/18 11/9-12/28	Spring 2/23-5/20	Summer 5/25/-9/8	Fall 9/14-11/4	Shock Period 7/8-7/22
Temperature, ^{o}C	7.5-16.2	9.2-15.0	15.5-20.8	17.0-20.9	19.9-21.1
DO, mg/l	0.8-4.5	2.70-7.88	0.72-5.25	0.05-3.91	0.23-2.64
pH	7.38-7.95	7.65-7.92	7.52-7.81	7.00-7.86	7.52-7.72
Alkalinity, mg/l as $CaCO_3$	223-323	245-291	234-299	253-310	265-289
TNH_3-N, mg/l	8.33-20.84	2.26-16.01	5.23-11.99	4.41-16.05	10.00-15.08
SNH_3-N, mg/l	8.07-19.96	1.65-15.86	4.06-11.17	4.41-15.17	9.58-14.93
NO_3-N, mg/l	0.07-3.74	0.16-4.84	0.10-4.65	0.01-2.15	0.13-2.16
NO_2-N, mg/l	0.04-0.44	0.12-0.48	0.05-0.62	0.01-0.27	0.01-0.41
TKN, mg/l	12.82-32.24	5.82-22.05	6.82-19.52	12.29-25.05	12.58-24.40
SKN, mg/l	9.11-22.96	3.88-17.46	5.23-14.11	8.23-18.56	10.64-17.64
Solids, mg/l					
Dissolved	400-540	444-528	424-544	440-540	472-504
Suspended	52-160	32-214	30-228	50-132	52-184
Volatile Susp.	48-116	24-150	26-144	46-108	46-120
Settleable	0-2.4	0.2-5.0	0.1-7.5	0.02-1.60	0.4-1.20
$TBOD_5$, mg/l	47.6-126.7	25.5-107.8	24.9-74.6	28.8-96.5	32.2-93.3
$SBOD_5$, mg/l	12.7-56.8	6.1-32.7	3.8-28.6	12.8-44.0	8.6-37.4
Flow, mgd	0.45-0.87	0.54-2.04	0.71-1.51	0.56-1.01	1.0-1.12

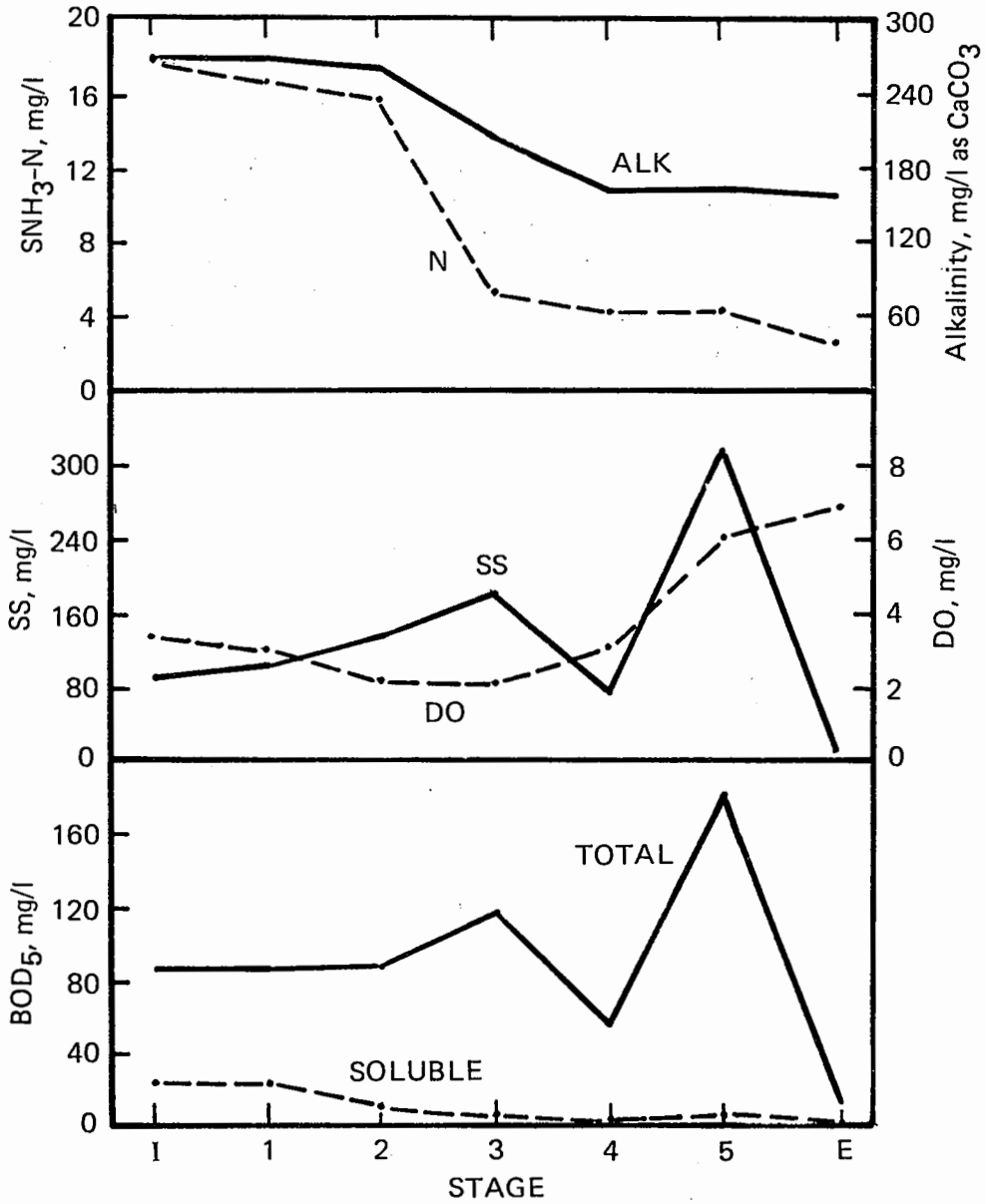

Figure 6. BOD_5, DO, SS, SNH_3-N, and alkalinity profiles, February 2, 1981

tration, on the average, about 0.5 mg/l; and the fifth unit increased the DO concentration, on the average, about 2.0 mg/l. Similar DO profiles through RBC systems have been reported by others (1,3,13,14).

The concentration of suspended solids through the RBC system at Princeton appears to be a function of the bottom configuration of the reaction basins. As shown in Figure 2 there is a 12-inch "drop" in the bottom for stages 3, 4, and 5. Presumably this change in elevation has been incorporated to provide addition detention time for the nitrification phase. As shown in Figure 6, a typical pattern of suspended solids concentration during passage through the system involves a substantial increase occurring at stage 5. On the average the increase is from about 90 mg/l at the influent of the system to about 260 mg/l at the effluent of stage 5. The average concentration in the clarified effluent was 17 mg/l. Apparently the solids synthesized within the RBC system tend to settle in the basins rather than being swept or scoured by the rotation of the RBC media with the exception of the unit 5. The solids in the stage 5 effluent were scoured and cut in fine size. This build-up of solids to an equilibrium has a profound effect upon measurements for $TBOD_5$ and may be the basic reason why $TBOD_5$ is not considered a reliable measurement to be applied to the RBC process at Princeton. The same findings have been expressed by others (2,14,15). As shown in Figure 6, $TBOD_5$ and suspended solids concentrations maintain similar patterns during passage through the system.

$\underline{BOD_5}$

The design loadings and observed operational loadings, for normal and shock periods, are set forth in Table II. During normal operations that is, with sewage flows about equally distributed between the two RBC trains, all observed loadings were generally within the range of design loadings except BOD_5 loading which was about one half of the design. Under these conditions the average influent $TBOD_5$ was about 63 mg/l and the $SBOD_5$ was about 22 mg/l. The respective effluents after clarification was 14 mg/l and 3 mg/l. The overall percent removal for $SBOD_5$ was about 85 at normal operations.

During the shock period the average hydraulic loading exceeded design by 30 percent but the BOD_5 loading was not exceeded. The average influent concentrations of $TBOD_5$ and $SBOD_5$ were similar to that observed for normal operation i.e. 72 mg/l and 25 mg/l, respectively. Clarified effluent aver-

Table II. Loadings on RBC Process

Parameter	Design	Observed	
		Normal Operations	Shock Period
Flow, mgd			
Mean	0.82	0.76	1.06
Range	2.29 (peak)	0.45-2.04	1.00-1.12
Detention, hours			
Mean	2.0	2.16	1.56
Range		0.80-3.60	1.45-1.64
Hydraulic (gals/d/sf)			
Mean	1.26	1.17	1.62
Range		0.69-3.14	1.53-1.72
$SBOD_5$ (lbs/d/1000sf)			
Mean	1.12 (T)	0.21 (0.62T)	0.34 (0.98T)
Range		0.07-0.37	0.11-0.54
SNH_3-N (lbs/d/1000sf)			
Mean	0.1 (T)	0.09 (0.11T)	0.16 (0.17T)
Range		0.04-0.14	0.13-0.22

aged 18 mg/l $TBOD_5$ and 4 mg/l $SBOD_5$. Solely on the basis of $SBOD_5$ the RBC system performed well despite excessive hydraulic loadings and a corresponding diminishment in detention time.

The changes that occur in BOD_5 concentrations during passage through the RBC process under normal operations are summarized in Table III. The review of the data for $TBOD_5$ suggest that the most efficient unit for $TBOD_5$ removal is the clarifier. This is consistent with the earlier findings that $TBOD_5$ is significantly influenced by suspended solids concentrations.

A review of the data for $SBOD_5$ indicates that the first three stages of the process are the principal "reducers". A 73 percent reduction of the influent $SBOD_5$ is achieved at these three stages. The last two stages contribute to an additional 12 percent reduction. This pattern of $SBOD_5$ reduction, as shown in Figure 6, has been reported by others

Table III. Statistical Summary of BOD_5 Data for Normal Operations

	Mean mg/l	Maximum mg/l	Minimum mg/l	Number of samples	Standard deviation
Total BOD_5					
RBC influent	63	127	16	86	21
Stage 1	55	147	20	86	22
Stage 2	56	150	16	86	21
Stage 3	59	119	13	84	26
Stage 4	37	72	13	84	13
Stage 5	90	209	11	84	45
2^o effluent	14	43	6	86	7
Soluble BOD_5					
RBC influent	22	57	3	86	11
Stage 1	14	33	3	86	7
Stage 2	9	18	4	86	3
Stage 3	6	14	3	86	2
Stage 4	4	8	1	85	1
Stage 5	3	6	1	86	1
2^o effluent	3	6	1	86	1
$SBOD_5/TBOD_5$					
RBC influent	0.34	0.55	0.11	84	0.11
Stage 1	0.24	0.48	0.11	84	0.09
Stage 2	0.18	0.43	0.06	84	0.06
Stage 3	0.11	0.48	0.04	82	0.06
Stage 4	0.11	0.41	0.03	82	0.05
Stage 5	0.04	0.13	0.02	82	0.02
2^o effluent	0.21	0.50	0.08	84	0.02

(3,13,14,16,17,18).

Also shown in Table III are the values of $SBOD_5/TBOD_5$ during passage through the system. As would be expected of the data thus far reviewed the fraction of $SBOD_5$ to the whole diminishes from an average of 34 percent in the influent to 4 percent in the stage 5. This is simply a case where the $SBOD_5$ diminishes with time and the $TBOD_5$ does not materially change or increases.

Nitrogen

The SNH_3-N/TNH_3-N ranged from 0.81 to 1.00 with an average of 0.97 for all sampling locations. The extremely low values of about 0.7 observed during April 15 to April 22 due to high

flow were excluded. For all practical purposes the total and soluble NH_3 are the same.

The removal of nitrogen through a wastewater treatment systems is achieved by biological assimilation and nitrification. For the purpose of this study a reduction of soluble organic nitrogen is evidence of biological assimilation; and increases in nitrite and nitrate nitrogen is evidence of nitrification. The transformation of various nitrogen forms through the RBC system is shown in Figure 7. The total mean nitrogen (ammonia, organic, nitrate, and nitrite) applied to the system during normal operation is about 18 mg/l. Theoretically this concentration should remain constant as the waste stream passes through the stages. Any substantial reduction should occur at the clarifier where the insoluble organic nitrogen fraction will be removed by sedimentation.

Figure 7 depicts these expectations except at stage 5 where the accumulation of suspended solids distort the anticipated results. More important however are the changes in concentration occurring for the various forms of nitrogen. Nitrite concentrations, represented in Figure 7 by a solid bar, remain fairly constant with a variation of 0.21-0.38 mg/l. On the other hand there are reductions in ammonia nitrogen and organic nitrogen (except stage 5) but substantially increases in nitrate nitrogen during passage through the system.

Hydraulic and ammonia-nitrogen loadings for normal operations and the shock period are shown in Table II. During normal operations the mean loadings were within the design limits. The pertinent data regarding nitrogen removal within the RBC system during normal operations are included in Table IV. The mean soluble ammonia nitrogen concentration in the influent was about 10.5 mg/l. This represents a loading of 0.09 lbs/d/1000 square feet of media. The mean ammonia nitrogen concentration in the clarified effluent was 1.52 mg/l.

On many occasions there was a slight increase in ammonia nitrogen as the waste stream passed stages 1 and 2. This slight increase was due to the production of ammonia nitrogen by the hydrolysis of organic nitrogen. Similar observations have been reported by others (13,14,18,19). Nevertheless, as an examination of Tble IV will show, the mean values for nitrogen during passage through the RBC system depict a downward trend. With reference to Table IV, ammonia nitrogen reduction is accomplished mainly at stages 3 and 4. Although some reduction occurs at stages 1, 2 and 5. In fact stages 3 and 4, alone, account for about 73 percent of total reduction. This is confirmed by the significant increases in nitrite and nitrate nitrogen, also depicted in Figure 7, at stages 3 and 4. And

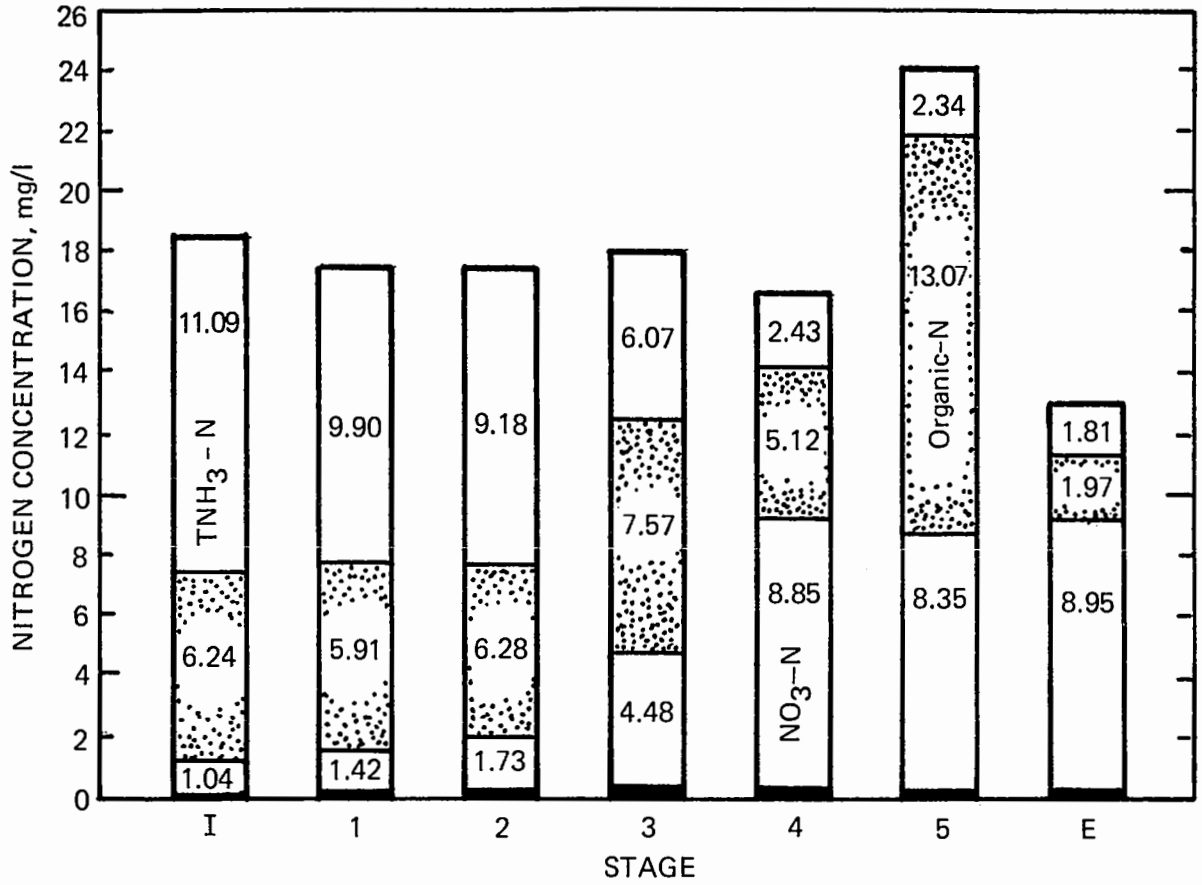

Figure 7. Changes of nitrogen forms in RBC system

Table IV. Summary of Nitrogen Related Parameters Within RBC System* During Normal Operations

	RBC Influent	Stage 1	2	3	4	5	Secondary effluent
Soluble ammonia nitrogen							
Mean, mg/l	10.49	9.34	8.58	5.49	2.04	1.58	1.52
Maximum	19.96	17.08	18.23	20.29	14.52	6.44	3.97
Minimum	1.65	2.65	1.88	0.87	0.17	0.21	0.21
Standard deviation	3.95	3.61	3.31	2.92	1.99	1.02	0.87
Soluble organic-N							
Mean, mg/l	1.99	1.52	1.22	1.21	1.01	0.92	0.91
Soluble nitrate + nitrite nitrogen							
Mean, mg/l	1.25	1.69	2.00	4.86	9.16	8.72	9.27
Alkalinity							
Mean, mg/l as $CaCO_3$	276	272	269	249	221	219	215
Maximum	323	320	310	294	263	262	264
Minimum	223	222	218	183	151	155	155
S. D.	20	20	19	22	24	26	27
pH, median	7.69	7.71	7.68	7.60	7.59	7.56	7.71

* 87 observations per station

this suggests that nitrification is the principal mode for ammonia reduction.

Stover and Kincannon (20) reported an ammonia reduction of 82 percent at the first stage of a 6-stage RBC system. This was not the case at Princeton. Several investigations have suggested that heterotrophic bacteria, that are responsible for most $SBOD_5$ removal, and autotrophic bacteria, that are responsible for most ammonia nitrogen removal are not too compatible on the same medium. In fact Antoni (18), Banerji (13), Autotrol (15), and Miller et al. (14) have suggested that significant nitrification did not begin until wastewater organic content in the stage of an RBC system reduced to 30 mg/l $TBOD_5$, 20 mg/l $TBOD_5$, 15 mg/l $SBOD_5$, and 10 mg/l $SBOD_5$, respectively. At Princeton significant nitrification commenced at stage 3. As shown in Table III the mean $SBOD_5$ applied to that stage was 9 mg/l. If $SBOD_5$ is indeed a regulator regarding ammonia removal than the findings of Miller et al. (14) apply to Princeton during normal operations.

The amount of ammonia oxidation also can be estimated by the alkalinity analysis. It is generally accepted that the ratio of alkalinity reduction to ammonia removal is about 7:1 (21). From the data summarized in Table IV, the ratios at stages 3 and 4 were 7.0 and 6.5 respectively - a reasonable expectation. The overall ratio, from influent to clarified effluent, was 6.8

Median values of pH, ranging from 7.56 to 7.69 per stage were within the optimum suggested by Chou (6) for the nitrification of ammonia nitrogen. At Princeton, typical high influent alkalinity concentrations of 275 mg/l as $CaCO_3$ did not allow a significant pH depression due to the alkalinity consumption of the nitrification reaction.

During the shock period the hydraulic loading of the RBC system exceeded design about 30 percent. The ammonia loading was 0.16 lbs/d/1000 square feet of media - exceeding the design loading of 0.10 lbs/d/1000 square feet by 60 percent. As shown in Table V the influent ammonia nitrogen concentration was 12.2 mg/l. The effluent concentration at stage 5 was 6.42 mg/l. This is an overall reduction of about 47 percent compared to a reduction of about 85 percent during normal operations.

As in the case of shock loading operations significant ammonia removal did not occur except at stages 3 and 4 (see Table IV). For that soluble ammonia nitrogen removed in the RBC system, 29 percent was accomplished at stage 3 and 77 percent was accomplished at stage 4. This is consistent with increases in nitrate nitrogen at these two stages and corre-

Table V. Mean Values of Nitrogen Related Parameters During Shock Period

	RBC influent	Stage 1	2	3	4	5	Secondary effluent
Soluble Ammonia-N	12.22	12.54	12.15	10.47	6.03	6.42	7.28
Soluble Organic-N	2.04	2.04	1.23	1.21	1.09	2.29	1.67
Nitrate + Nitrite-N	0.64	0.30	0.32	0.92	4.90	4.33	3.88
Alkalinity (as $CaCO_3$)	280	281	281	274	240	255	250

Unit: mg/l

sponding reductions in alkalinity all as shown in Table V.

The mean concentrations of $SBOD_5$ applied to each stage of the system during the shock period are as follows:

Stage:	1	2	3	4	5
Applied, mg/l:	25	20	16	14	7

This summary of applied $SBOD_5$ concentrations suggests that 10 mg/l $SBOD_5$ may not be the critical limiting factor for the occurrence of nitrification on the medium of a contactor. Under conditions where the mean $SBOD_5$ applied was 16 mg/l and 14 mg/l (stages 3 and 4) the ammonia removed was 29 percent and 77 percent respectively of the RBC overall removal. And despite an applied concentration of 7 mg/l BOD_5 at stage 5 ammonia removal was ineffectual.

The average DO concentration in the system during normal operations was 2.8 mg/l with concentrations varying from 2.2 mg/l to 4.4 in the system. The average during the shock period was about 1.0 mg/l with concentrations varying in the system from 0.7 mg/l to 1.7 mg/l. Dissolved oxygen was not considered a limiting factor during the shock period because the lowest values (0.7-0.8 mg/l) occurred at stages 3 and 4, the most efficient stages while the DO concentration at stage 5 was 1.7 mg/l, a most inefficient stage.

It was observed that the nitrifiers did not grow well on the media of stage 5 of the south train. The biomass was very thin and dark brown in color. Twenty to 30 percent of the media surface area was often clear (peeled off). However, the phenomena were not found in stage 5 of the north train which usually has healthy nitrifiers growth. The reason for the difference between the two corresponding units it unknown.

Rational Analysis

The length of the study, frequency of sampling, and scope of analyses at Princeton requires a more rigorous examination of the data gathered than here presented. The discussion here is limited to mean conditions for two different operational modes - one labelled "normal operation"; the other "shock period". With this in mind a closer examination of the loadings and response for each stage is now offered.

Conventional design of RBC systems insist on applying the hydraulic and mass (BOD_5 and NH_3) loadings directly to the total area of the media without any regard to the loadings on each stage of the system. Loading based on surface area of each stage should be used. At Princeton the total area of

media, the sum of all five stages, is 650,000 square feet. The design flow is 820,000 gpd. Therefore the hydraulic loading is 1.26 gals/d/sf. Similar calculations are made for the design $TBOD_5$ loading (1.12 lbs/d/1000 sf) and the design total NH_3 loading (0.10 lbs/d/1000 sf).

As shown in Tables VI and VII the conventional loading rates do not have any relevancy to the loadings applied per stage in the RBC system. In Table VI, where a conventional hydraulic loading of 1.2 gals/d/sf is designated for normal operation, the actual mean hydraulic loadings on the rotating contactors varies from 7.6 to 5.1 gals/d/sf. Similarly, where a conventional organic ($SBOD_5$) loading of 0.21 lbs/d/1000 sf is designated for normal operations the actual mean organic loadings on the units varies from 1.38 to 0.16 lbs/d/1000 sf. For the shock period actual mean hydraulic load-

Table VI. Soluble BOD_5 Loadings and Removal and Hydraulic Loadings per Stage

	Normal operation		Shock period	
	Applied	Removed	Applied	Removed
Conventional calculation				
Hydraulic loading*	1.17		1.62	
$SBOD_5$ loading+	0.21		0.34	
Actual				
Hydraulic loading*				
Stage 1	7.6		10.5	
Stage 2	7.6		10.5	
Stage 3	5.1		7.0	
Stage 4	5.1		7.0	
Stage 5	5.1		7.0	
$SBOD_5$ loading†				
Stage 1	1.38	0.52	2.19	0.44
Stage 2	0.86	0.27	1.75	0.34
Stage 3	0.39	0.14	0.93	0.14
Stage 4	0.25	0.09	0.80	0.39
Stage 5	0.16	0.02	0.41	0.02
Overall average		0.21		0.26
2^{o} effluent $SBOD_5$, mg/l	2.6		4.4	

Note: * = gals/d/sf;
† = lbs/d/1000 sf

Table VII. Soluble Ammonia Loadings and Removal and Hydraulic Loadings per Stage

	Normal operation		Shock period	
	Applied	Removed	Applied	Removed
Conventional calculation				
Hydraulic loading*	1.17		1.62	
SNH_3-N loading+	0.09		0.16	
Actual				
Hydraulic loading*				
Stage 1	7.6		10.5	
Stage 2	7.6		10.5	
Stage 3	5.1		7.0	
Stage 4	5.1		7.0	
Stage 5	5.1		7.0	
SNH_3-N loading+				
Stage 1	0.66	0.07	1.08	(-0.03)
Stage 2	0.59	0.05	1.11	0.04
Stage 3	0.36	0.13	0.72	0.10
Stage 4	0.23	0.14	0.62	0.26
Stage 5	0.09	0.02	0.36	(-0.02)
Overall average		0.08		0.07
2^{o} effluent SNH_3-N, mg/l	1.5		7.3	

Note: * = gals/d/sf
+ = lbs/d/1000 sf

ings per stage varied from 10.5 to 7.0 gals/d/sf while the mean $SBOD_5$ loading varied from about 2.2 to 0.4 lbs/d/1000 sf.

The meanings of these loadings, for design purposes, are not clear at this time. It is interesting however that the average overall removal of $SBOD_5$ for each of the two operating modes was in a narrow range of 0.20 to 0.26 lbs/d/1000 sf (Table VI). And since the concentrations of $SBOD_5$ in the effluent were limited to 2.6 to 4.4 mg/l for the two operating modes, it appears that the hydraulic and organic loadings (carbonaceous) applied to the system have not exceeded the treatment capability of the system.

Under normal operations (see Table VI), stages 1 and 2 removed most of the $SBOD_5$ on the basis of unit area. And the pounds per unit removed becomes progressively less through each succeeding stage. However during the shock period the

removal at stage 4 (0.39 lbs/d/1000 sf) was about equal to removal performance of stages 1 and 2, i.e. 0.44 and 0.34 lbs/d/1000 sf, respectively. This unexpected occurrence requires closer examination at a later date.

The SNH_3 loadings and removal per stage as shown in Table VII suggest a different pattern. At normal operations with conventional mean hydraulic and ammonia loadings of 1.2 gpd/sf and 0.09 lbs/d/1000 sf a mean effluent of 1.5 mg/l was achieved. However, during the shock period with mean conventional loadings of 1.6 gals/d/sf and 0.16 lbs/d/1000 sf an effluent of 7.3 mg/l NH_3-N was produced. This indicates that during the shock period the treatment capability of the system was exceeded for ammonia nitrogen removal.

Unlike the carbonaceous removal ($SBOD_5$) ammonia nitrogen removal progressively improved through each succeeding stage except at stage 5. Stage 5 seemed to be an idler during both operational modes.

An interesting aspect of ammonia removal was the average overall removal per unit area in the system. As shown in Table VII the overall removal was 0.08 and 0.07 lbs/d/1000 sf for the normal and shock period operations. Does this mean that the RBC process is limited? Is its treatment capability for ammonia removal, in the presence of $SBOD_5$, to be about 0.08 lbs/d/1000 sf regardless of loading? This also will require further examination of the data.

SUMMARY AND CONCLUSIONS

The RBC system at Princeton is designed as a secondary treatment for the removal of BOD_5 and ammonia nitrogen. An intensive study on each stage under normal and artificial shock operations was carried out for over one year. This paper deals with preliminary evaluation of mean values. The following conclusions can be made:

1. The Princeton RBC units gave an average BOD removal efficiency of 85 percent with the secondary effluent of 2.6 mg/l mean $SBOD_5$ under a normal mean conventional hydraulic loading of 1.2 gals/d/sf and a mean $SBOD_5$ loading of 0.21 lbs/d/1000 sf.

2. At a mean conventional hydraulic loading of 1.6 gsl/d/sf and a mean $SBOD_5$ loading of 0.34 lbs/d/1000 sf a mean effluent of 4.4 mg/l $SBOD_5$ was produced.

3. The RBC system also removed 85 percent of ammonia nitrogen under a mean conventional hydraulic loading and a mean ammonia nitrogen loading of 0.09 lbs/d/1000 sf. The mean ammonia nitrogen concentration in the secondary effluent was 1.5 mg/l.

4. At a mean conventional hydraulic loading of 1.6 gals/d/sf and a mean ammonia nitrogen loading of 0.16 lbs/d/1000 sf an unacceptable mean effluent of 7.3 mg/l of ammonia nitrogen was produced.

5. The average overall $SBOD_5$ removal under two different mean loadings varied from 0.20 to 0.26 lbs/d/1000 sf.

6. The average overall ammonia nitrogen removal under two different mean loadings varied of 0.07 to 0.08 lbs/d/1000 sf.

7. Based solely on mean loadings (hydraulic and organic), the RBC system was not stressed for $SBOD_5$ removal; however the system was stressed for ammonia removal.

8. For the loadings experienced at Princeton it appears that the RBC system is limited to ammonia nitrogen removal of about 0.08 lbs/d/1000 sf.

9. Significant ammonia nitrogen removal is limited to stages 3 and 4 of the RBC system but stages 1 and 2 support the nitrification process under normal operations.

10. Nitrification occurred significantly in the presence of 15 mg/l $SBOD_5$.

11. Nitrification progressed at mean dissolved oxygen levels as low as 0.8 mg/l.

12. There is some evidence that a closer examination of loadings applied to each stage in the system can provide a more rational approach to the design of RBC system.

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REFERENCES

1. Antonie, R. L., Kluge, D. L., and Mielke, J. H., "Evaluation of a Rotating Disk Wastewater Treatment Plant," Journal Water Pollution Control Federation, Vol. 46, No. 3, Mar., 1974, pp. 498-511.
2. Dupont, R. R., and McKinney, R. E., "Data Evaluation of a Municipal RBC Installation, Kirksville, Missouri," Proceedings of First National Symposium/Workshop on Rotating Biological Contactor Technology, held at Champion, PA, Feb., 1980, pp. 205-234.
3. Hitdlebaugh, J. A., and Miller, R. D., "Full Scale Rotating Biological Contactor for Secondary Treatment and Nitrification," Proceedings of First National Symposium/ Workshop on RBC Technology, Feb., 1980, pp. 269-294.
4. Hynek, R. J., and Iemura, H., "Nitrogen and Phosphorus Removal with Rotating Biological Contactors," Proceedings of First National Symposium/Workshop on RBC Technology, Feb., 1980, pp. 295-324.
5. Sullivan, R. A., and Hynek, R. J., "Upgrading Existing Waste Treatment Facilities Utilizing the Bio-Surf Process," Proceedings of First National Symposium/Workshop on RBC Technology, Feb., 1980, pp. 1015-1033.
6. Chou, C. C., Hynek, R. J., and Sullivan, R. A., "Comparison of Full Scale RBC Performance with Design Criteria," Proceedings of First National Symposium/Workshop on RBC Technology, Feb., 1980, pp. 1101-1125.
7. Williams, T. C., and Berner, R. J., "The Gladston, Michigan Experience Performance of a 1.0 MGD RBC Plant in a Northern Michigan Community," Proceedings of First National Symposium/Workshop on RBC Technology, Feb., 1980, pp. 1137-1146.

8. Morley, W., "Experiences at Gladstone, Michigan Utilizing Rotating Biological Contactors for BOD_5, Phosphorus, and Ammonia Control," Proceedings International Seminar on Control of Nutrients in Muncipal Wastewater Effluents, U.S.E.P.A., Sept. 1980, Vol. I, pp. 91-112.
9. APHA, AWWA, and WPCF, "Standard Methods for the Examination of Water and Wastewater," American Public Health Association, New York, NY, 1980, 1134 p.
10. Scheible, O. K., and Novak, J. J., "Upgrading Primary and Secondary Waste Treatment System with RBC," Proceedings of First National Symposium/Workshop on RBC Technology, Feb., 1980, pp. 961-996.
11. Steiner, B. C. G., "Take a New Look at the RBC Process," Water & Wastes Engineering, May 1979, Vol. 16, No. 5, pp. 41-46, 103-107.
12. Tait, S. J., and Friedman, A. A., "Anaerobic Rotating Biological Contactor for Carbonaceous Wastewaters," Journal Water Pollution Control Federation, Vol. 52, No. 8, Aug., 1980, pp. 2257-2269.
13. Banerji, S. K., "ASCE Water Pollution Management Task Committee Report on Rotation Biological Contactor for Secondary Treatment," Proceedings of First National Symposium/Workshop on RBC Technology, Feb., 1980, pp. 31-52.
14. Miller, R. D., et al., "Rotating Biological Contactor Process for Secondary Treatment and Nitrification Following a Trickling Filter," Proceedings of First National Symposium/Workshop on RBC Technology, Feb., 1980, pp. 1035-1063.
15. Autotrol Corporation, "AUTOTROL Wastewater Treatment Systems - Design Manual," Milwaukee, WI, 1978, p. C-3.
16. Stover, E. D., and Kincannon, D. F., "Evaluating Rotating Biological Contactor Performance," Water and Sewage Works, Vol. 123, No. 3, Mar., 1976, pp. 88-91.
17. Poon, C. P. C., and Mikucki, W. J., "Rotating Biological Contactors Treat Island's Saline Sewage," Water and Sewage Works, Vol. 125, No. 2, Feb. 1978, pp. 62-68.
18. Antonie, R. L., "Fixed Biological Surfaces - Wastewater Treatment," CRC Press, Inc., West Palm Beach, FL, 1975, 200 p.
19. Weng, C., and Molof, A. H., "Nitrification in Biological Fixed-film Rotating Disk System," Journal Water Pollution Control Federation, Vol. 46, No. 7, July 1974, pp. 1674-1685.
20. Stover, E. L., and Kincannon, D. F., "One-step Nitrification and Carbon Removal," Water and Sewage Works, Vol. 122, No. 6, June 1975, pp. 66-69.
21. U.S.E.P.A., "Process Design Manual for Nitrogen Control," USEPA Technology Transfer, Washington, D.C., Oct. 1975.