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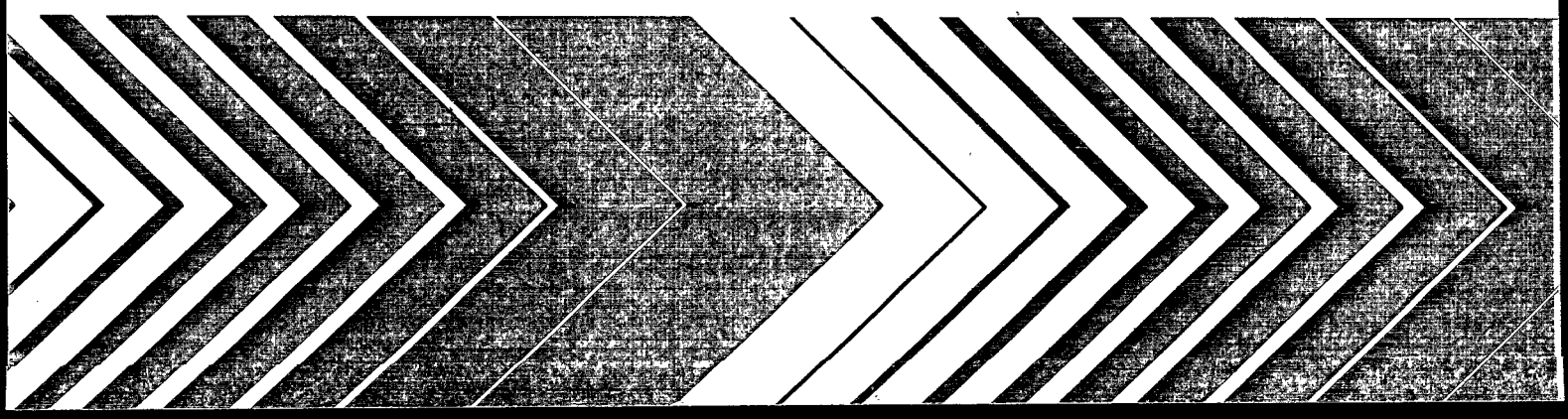
October 12-13, 1989
Cincinnati, Ohio



EPA/600/9-90/036
August 1990

PROCEEDINGS

TWELFTH UNITED STATES-JAPAN CONFERENCE ON SEWAGE TREATMENT TECHNOLOGY

OCTOBER 12-13, 1989
CINCINNATI, OHIO

RISK REDUCTION ENGINEERING LABORATORY
U.S. ENVIRONMENTAL PROTECTION AGENCY
OFFICE OF RESEARCH AND DEVELOPMENT
CINCINNATI, OHIO 45268

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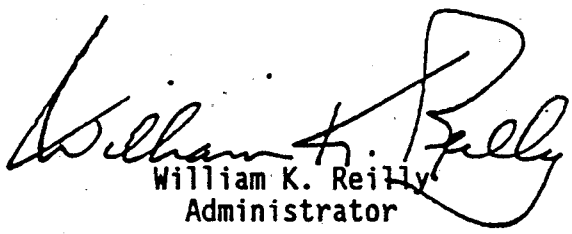
FOREWORD

The maintenance of clean water supplies and the management of municipal and industrial wastes are vital elements in the protection of the environment.

The participants in the United States-Japan cooperative project on sewage treatment technology have completed their Twelfth Conference. These conferences, held at 24-month intervals, have given the scientists and engineers of the cooperating agencies an opportunity to study and compare the latest practices and developments in the United States and Japan. These Proceedings of the Twelfth Conference comprise a useful body of knowledge on sewage treatment, which will be available not only to Japan and the United States but also to all nations of the world who desire it.



William K. Reilly
Administrator

Washington, D.C.

TABLE OF CONTENTS

Foreword iii

List of Japanese Delegates and Presentation Topics vi

List of United States Delegates and Presentation Topics viii

Joint Communique 1

Japanese Papers 3

United States Papers 5

JAPANESE DELEGATION

Delegates	Presentation Topics
Dr. Ken Murakami (Head of Delegation) Director, Regional Sewerage Division Sewerage & Sewage Purification Dept. City Bureau Ministry of Construction	New Movement in Sewerage Construction Projects in Japan
Mr. Taigo Matsui Director, Water Quality Control Dept. Public Works Research Institute Ministry of Construction	Renovation of an Extended Aeration Treatment Plant for Biological Nutrient Removal Process
Mr. Shunsoku Kyosai Research Coordinator for Wastewater Treatment Water Quality Control Dept. Public Works Research Institute Ministry of Construction	Innovative Wastewater Treatment Processes - Interim Results of Biofocus WT Project
Mr. Eiichi Nakamura Chief, Water Quality Section Water Quality Control Dept. Public Works Research Institute Ministry of Construction	Toxicity Monitoring Method by Biosensor
Dr. Kazuhiro Tanaka Director, Research & Technology Development Division Japan Sewage Works Agency	Innovative Sludge Handling System Using Pelletization for Simultaneous Thickening and Conditioning
Mr. Masanobu Aoki Deputy Director Research & Technology Development Division Japan Sewage Works Agency	Development of a High Efficiency Nitrogen Removal System Using Immobilized Nitrifiers in Synthetic Resin
Mr. Kazuo Hoya Director, Engineering Division Regional Sewerage Center Sewerage Bureau Tokyo Metropolitan Government	Fundamental Study on Application of High Gradient Magnetic Separation to Municipal Wastewater Treatment System

JAPANESE DELEGATION (continued)

Mr. Senji Kaneko Chief, Research & Development Division Sewage Works Bureau Yokohama City	Operation of Centralized Sludge Treatment Plant
Mr. Hiroshi Oniki Director, Planning Department Sewage Works Bureau Fukuoka City	Development of Sludge Treatment Technology in Biological Phosphorus Removal Process
Dr. Takeshi Kubo (Advisor) Counselor, Japan Sewage Works Agency	

UNITED STATES DELEGATION

Delegates	Presentation Topics
Mr. John J. Convery (Head of Delegation) Deputy Director Risk Reduction Engineering Laboratory U.S. Environmental Protection Agency	
Mr. Dolloff F. Bishop Chief, Treatment Assessment Branch Water & Hazardous Waste Treatment Research Division Risk Reduction Engineering Laboratory U.S. Environmental Protection Agency	
Dr. William C. Boyle Dept. of Civil & Environmental Engineering University of Wisconsin-Madison	Performance of Fine Pore Aeration Systems in Process Water
Mr. William F. Brandes Permits Division Office of Water Enforcement & Permits U.S. Environmental Protection Agency	Current Status and Future Direction of Organic Toxics Control in the Water Environment - Discharge Regulations and Stream Standards
Mr. Richard C. Brenner Environmental Engineer Treatment Assessment Branch Water & Hazardous Waste Treatment Research Division Risk Reduction Engineering Laboratory U.S. Environmental Protection Agency	
Mr. Donald S. Brown Environmental Engineer Municipal Wastewater Branch Water & Hazardous Waste Treatment Research Division Risk Reduction Engineering Laboratory U.S. Environmental Protection Agency	In-Vessel Composting of Municipal Wastewater Sludge
Mr. Randall J. F. Bruins Environmental Scientist Environmental Criteria & Assessment Office U.S. Environmental Protection Agency	The Development of Risk Assessment Methodologies for Use in Regulation of Sewage Sludge Disposal

UNITED STATES DELEGATION (continued)

Dr. Carl A. Brunner
Acting Chief, Hazardous Waste Treatment Branch
Water & Hazardous Waste Treatment Research Division
Risk Reduction Engineering Laboratory
U.S. Environmental Protection Agency

Dr. Arthur J. Condren James M. Montgomery Consulting Engineers, Inc.	A Preliminary Assessment of High Biomass Systems
Dr. Richard A. Dobbs Research Chemist Treatment Assessment Branch Water & Hazardous Waste Treatment Research Division Risk Reduction Engineering Laboratory U.S. Environmental Protection Agency	Partitioning of Toxic Organic Compounds Between Wastewater and Wastewater Solids
Dr. Rahesh Govind Dept. of Chemical and Nuclear Engineering University of Cincinnati	Prediction of Biodegradation Kinetics of Toxic Organic Compounds
Dr. James A. Heidman Sanitary Engineer Municipal Wastewater Branch Water & Hazardous Waste Treatment Research Division Risk Reduction Engineering Laboratory U.S. Environmental Protection Agency	
Mr. James F. Kreissl Acting Chief, Municipal Wastewater Branch Water & Hazardous Waste Treatment Research Division Risk Reduction Engineering Laboratory U.S. Environmental Protection Agency	Experience with Intrachannel Clarifiers in the U.S.A.
Dr. Vincent P. Olivieri Department of Geography and Environmental Engineering The Johns Hopkins University	Levels and Removal of Selected Indicator and Pathogenic Microorganisms During Conventional Anaerobic Sludge Digestion
Dr. O. Karl Scheible Hydro-Qual, Inc.	UV Disinfection

UNITED STATES DELEGATION (continued)

Mr. Henry H. Tabak
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Research Division
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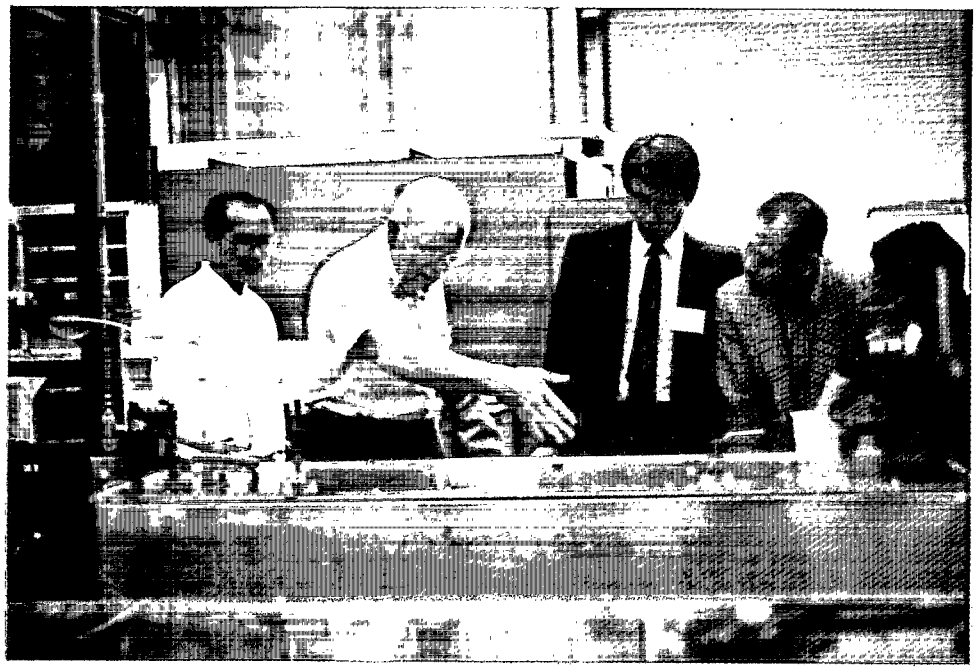
Biodegradability and
Biodegradation Kinetics
Assessment for Toxic
Organic Compounds Using
Oxygen Uptake Measurements

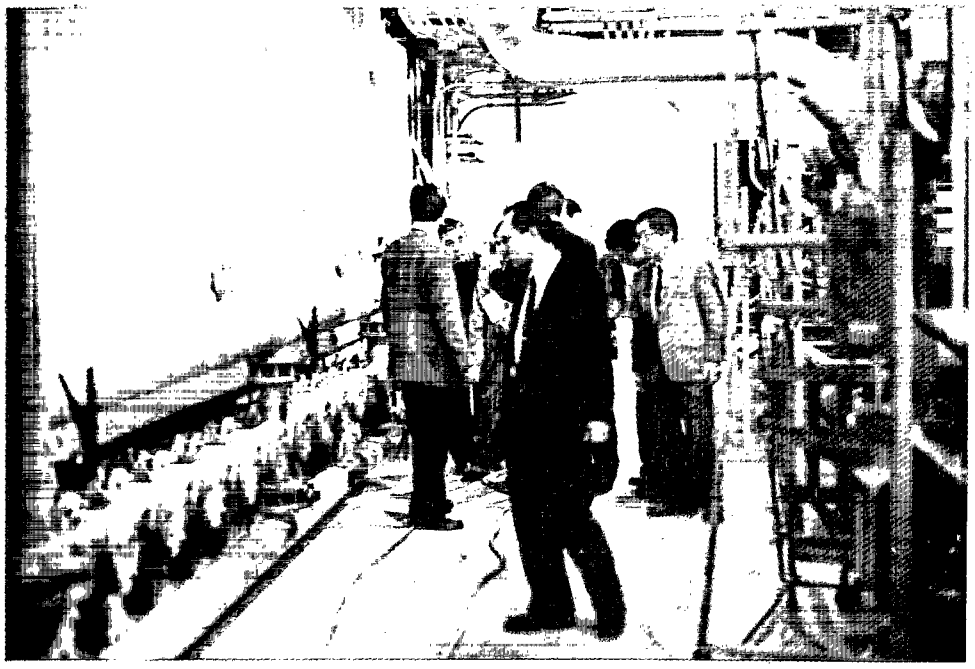
Dr. Albert D. Venosa
Microbiologist
Municipal Wastewater Branch
Water & Hazardous Waste Treatment Research Division
Risk Reduction Engineering Laboratory
U.S. Environmental Protection Agency

UNITED STATES AND JAPANESE DELEGATES TO THE
TWELFTH CONFERENCE

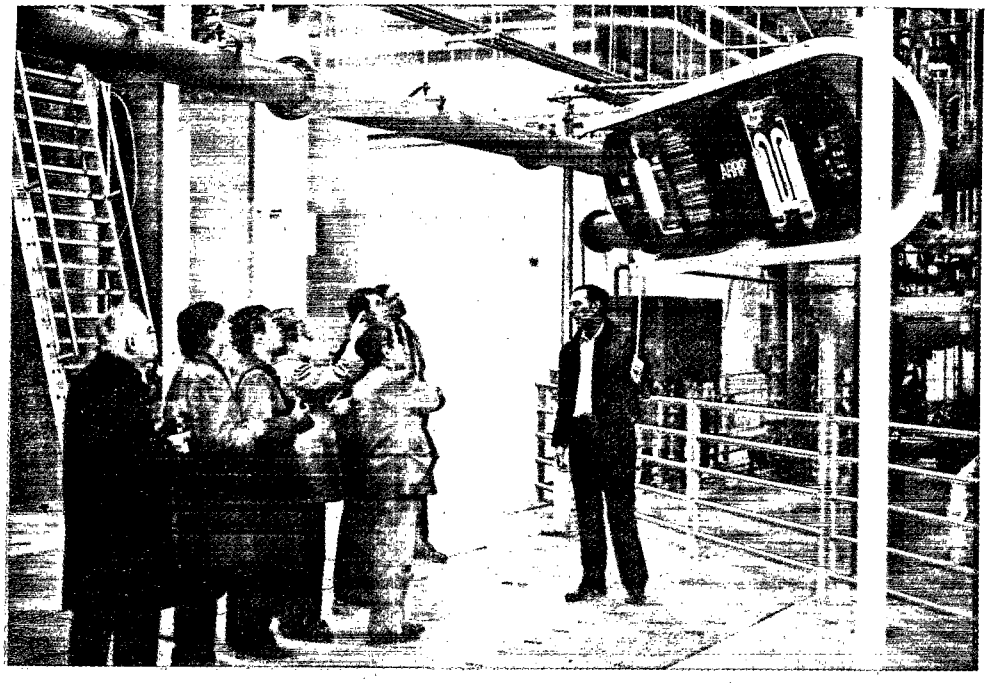
DR. KEN MURAKAMI, HEAD OF THE JAPANESE DELEGATION, AND
MR. JOHN CONVERY, HEAD OF THE U.S. DELEGATION, MAKING
OPENING REMARKS AT THE CONFERENCE



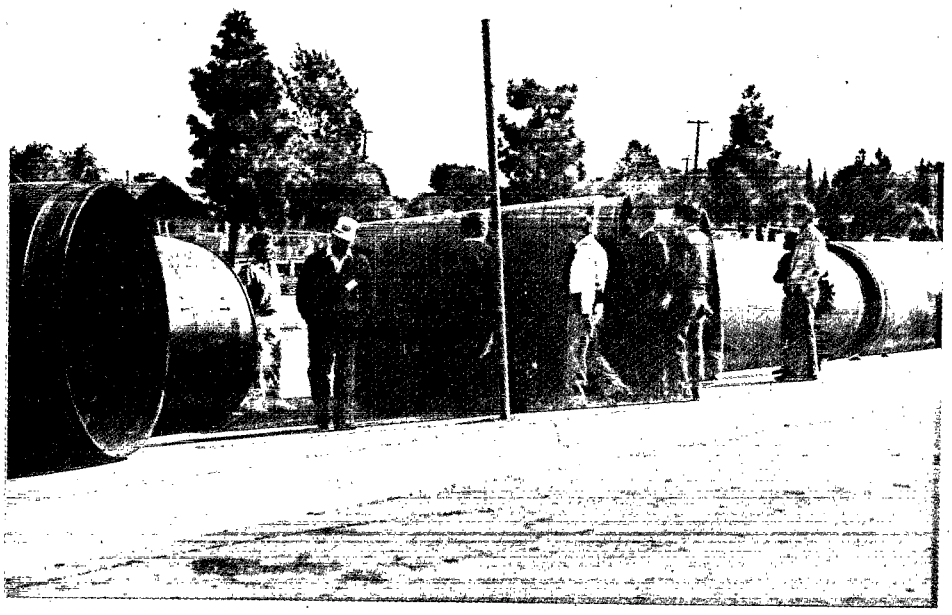
DELEGATES OBSERVING PILOT ROTATING BIOLOGICAL CONTACTOR STUDIES AT THE USEPA TEST AND EVALUATION FACILITY, CINCINNATI, OHIO



DELEGATES OBSERVING MECHANICAL COMPOSTING OF SLUDGE AT HAMILTON, OHIO



WASTEWATER REUSE PLANT OPERATION AT
DENVER, COLORADO, BEING DESCRIBED TO DELEGATES



DELEGATES OBSERVING SEWER RELINING AT LOS ANGELES COUNTY, CALIFORNIA

DELEGATES OBSERVING THE COLLECTION AND EVALUATION OF MARINE LIFE NEAR SEWAGE OUTFALLS, LOS ANGELES COUNTY, CALIFORNIA

JOINT COMMUNIQUE

Twelfth United States/Japan Conference on Sewage Treatment Technology

Friday, October 13, 1989

1. The Twelfth United States/Japan Conference on Sewage Treatment Technology was held in Cincinnati, Ohio, October 12 and 13, 1989.

2. The US Delegation headed by Mr. John J. Convery, Deputy Director, Risk Reduction Engineering Laboratory, U.S. Environmental Protection Agency (USEPA), Cincinnati, Ohio, was composed of twelve representatives from the USEPA, two representatives from consulting engineering companies, and three representatives from the universities.

3. Dr. Ken Murakami, Director, Regional Sewerage Division, Sewerage and Sewage Purification Department, Ministry of Construction was the Head of the Japanese Delegation, which consisted of four National Government representatives, three Japan Sewage Works Agency representatives, and three representatives from the Tokyo Metropolitan Government, Yokohama City and Fukuoka City.

4. During the Conference, papers related to toxics control in municipal wastewater, and sludge treatment and disposal technologies were presented by both sides. Additional presentations were made by the Japanese side concerning sewage construction projects and innovative wastewater treatment processes including nutrient removal. The US side, too, presented additional papers related to aeration methods, pathogens and disinfection, and innovative treatment. Data and findings were useful to the development of sewage treatment technology for each country.

5. Field visits in Chicago, IL, Denver, CO, and Los Angeles, CA are planned to inspect sewers and wastewater treatment facilities.

6. A proceedings of the Twelfth Conference shall be printed in English and Japanese.

7. Regarding the operation of the Sewage Treatment Technology Project under the United States/Japan Environmental Cooperation Agreement (hereinafter referred to as the Project), both sides agreed that the Project be amended following this conference to have a workshop for technology exchanges and visits in either country by sending one or two representatives to the host country. Each side would host the workshop at least once every two years alternatively.

8. Both sides agreed that engineer exchanges and the cooperative research program be continued as before.

9. Summaries of the research projects concerning sewage treatment technology of each country shall be exchanged every year.

10. A progress report shall be prepared each year by both the US and Japan to summarize briefly the activities of the Project. The progress reports shall be exchanged at the time of the workshops discussed in paragraph 7.

11. The Co-chairmen of the Project shall be the Branch Chief responsible for municipal wastewater research, Water and Hazardous Waste Treatment Division, Risk Reduction Engineering Laboratory, USEPA for the US side and Director, Sewerage and Sewage Purification Department, Ministry of Construction for the Japanese side.

12. It was proposed by the Japanese side that the forthcoming workshop based on paragraph 7 of this Joint Communique shall be held in Tsukuba around May 1990.

JAPANESE PAPERS

INNOVATIVE WASTEWATER TREATMENT PROCESSES - INTERIM RESULTS OF BIOFOCUS WT PROJECT 7
Shunsoku Kyosai, Research Coordinator for Wastewater Treatment, Yutaka Suzuki, Masahiro Takahashi, Katsumi Moriyama, Masashi Ogoshi, Toshiyuki Nakajima, Takashi Masuda, Suuichi Ochi, and Hiroaki Tanaka, Water Quality Control Department, Public Works Research Institute, Ministry of Construction

FUNDAMENTAL STUDY ON APPLICATION OF HIGH GRADIENT MAGNETIC SEPARATION TO MUNICIPAL WASTEWATER TREATMENT SYSTEM 59
Kazuo Hoya, Director, Engineering Division, Regional Sewerage Center, Yasuo Tanaka, Ph.D., Researcher of Research and Technological Development Section, Planning Division, and Khouichi Ishida, Asst. Head of Research and Technology Development Section, Planning Division, Sewerage Bureau, Tokyo Metropolitan Government

NEW MOVEMENT IN SEWERAGE CONSTRUCTION PROJECTS IN JAPAN 101
Dr. Ken Murakami, Director, Regional Sewerage Division, Sewerage and Sewage Purification Department, City Bureau, Ministry of Construction

INNOVATIVE SLUDGE HANDLING SYSTEM USING SLUDGE PELLETIZATION FOR SIMULTANEOUS THICKENING AND CONDITIONING 119
Dr. Kazuhiro Tanaka, Director, Shuzo Koike, Research Engineer, Research and Technology Development Division, Japan Sewage Works Agency, and Naohiro Taniguchi, Director, General Planning Section of General Planning Division, City Planning Bureau, Tokyo Metropolitan Government

OPERATION OF CENTRALIZED SLUDGE TREATMENT PLANT 157
Senji Kaneko, Chief, Research and Development Division, Sewage Works Bureau, Yokohama City

DEVELOPMENT OF SLUDGE TREATMENT TECHNOLOGY IN BIOLOGICAL PHOSPHORUS REMOVAL PROCESS 215
Hiroshi Oniki, Director, Planning Department, Sewage Works Bureau, Fukuoka City

TOXICITY MONITORING METHOD BY BIOSENSOR 239
Eiichi Nakamura, Chief, and Hiroaki Tanaka, Senior Research Engineer, Water Quality Section, Public Works Research Institute, Ministry of Construction

RENOVATION OF AN EXTENDED AERATION TREATMENT PLANT FOR BIOLOGICAL NUTRIENT REMOVAL PROCESS 253
Taigo Matsui, Director, K. Sato, K. Moriyama, Water Quality Control Department, Public Works Research Institute, Ministry of Construction, M. Imamichi and Y. Yarada, Sewage Works Division, Hamamatsu City

DEVELOPMENT OF A HIGH EFFICIENCY NITROGEN REMOVAL SYSTEM USING IMMOBILIZED NITRIFIERS IN SYNTHETIC RESIN 265
Masanobu Aoki, Deputy Director, Minoru Tada, Deputy Director, Takashi Kimata, Shouji Harada, and Yuhko Fujii, Research and Technology Development Division, Japan Sewage Works Agency

UNITED STATES PAPERS

CURRENT STATUS AND FUTURE DIRECTION OF ORGANIC TOXICS CONTROL IN THE WATER ENVIRONMENT - DISCHARGE REGULATIONS AND STREAM STANDARDS 295
William F. Brandes, and Catherine Crane, Permits Division, Office of Water Enforcement and Permits, U.S. EPA, Washington, DC

PERFORMANCE OF FINE PORE AERATION SYSTEMS IN PROCESS WATER 303
Dr. William C. Boyle, Professor, Department of Civil and Environmental Engineering, University of Wisconsin, Madison, Wisconsin; Richard C. Brenner, Environmental Engineer, Water and Hazardous Waste Treatment Research Division, Risk Reduction Engineering Laboratory, U.S. EPA, Cincinnati, Ohio; James J. Marx, Consulting Engineer, Donohue and Associates, Milwaukee, Wisconsin; and Thomas C. Rooney, Consulting Engineer, Waukesha, Wisconsin

BIODEGRADABILITY AND BIODEGRADATION KINETICS ASSESSMENT FOR TOXIC ORGANIC COMPOUNDS USING OXYGEN UPTAKE MEASUREMENTS 325
Henry H. Tabak, Research Chemist, Water and Hazardous Waste Treatment Research Division, Risk Reduction Engineering Laboratory, U.S. EPA, Cincinnati, Ohio; Senjay Desai, Chemical Engineer, and Dr. Rakesh Govind, Professor, Department of Chemical and Nuclear Engineering, University of Cincinnati, Cincinnati, Ohio

PREDICTION OF BIODEGRADATION KINETICS OF TOXIC ORGANIC COMPOUNDS 383
Sanjay Desai, Chemical Engineer, Dr. Chao Gao, Visiting Chinese Scholar, Dr. Rakesh Govind, Professor, Department of Chemistry and Nuclear Engineering, University of Cincinnati, Cincinnati, Ohio; and Henry H. Tabak, Research Chemist, Water and Hazardous Waste Treatment Research Division, Risk Reduction Engineering Laboratory, U.S. EPA, Cincinnati, Ohio

PARTITIONING OF TOXIC ORGANIC COMPOUNDS BETWEEN WASTEWATER AND WASTEWATER SOLIDS 401
Dr. Richard A. Dobbs, Research Chemist, Water and Hazardous Waste Treatment Research Division, Risk Reduction Engineering Laboratory, U.S. EPA, Cincinnati, Ohio

UNITED STATES PAPERS (continued)

LEVELS AND REMOVAL OF SELECTED INDICATOR AND PATHOGENIC MICROORGANISMS DURING CONVENTIONAL ANAEROBIC SLUDGE DIGESTION 425

Dr. Vincent P. Olivieri, Professor, Lynne Cox, and Mohammed Sarai, Department of Geography and Environmental Engineering, The Johns Hopkins University, Baltimore, Maryland; Dr. Jan L. Sykora, Professor, and Patrick Gavagahn, University of Pittsburgh, Pittsburgh, Pennsylvania

IN-VESSEL COMPOSTING OF MUNICIPAL WASTEWATER SLUDGE 441
Donald S. Brown, Environmental Engineer, Water and Hazardous Waste Treatment Research Division, Risk Reduction Engineering Laboratory, U.S. EPA, Cincinnati, Ohio; John Johnston, Associate Professor, California State University-Fresno, Fresno, California; and Leslie Bayer, Eastern Research Group, Arlington, Massachusetts

THE DEVELOPMENT OF RISK ASSESSMENT METHODOLOGIES FOR USE IN REGULATION OF SEWAGE SLUDGE DISPOSAL 475
Randall J. F. Bruins, Physical Scientist, Dr. Norman E. Kowal, Research Medical Officer, Cynthia Sonich-Mullin, Environmental Health Scientist, and Steven D. Lutkenhoff, Physiologist, Environmental Criteria and Assessment Office, U.S. EPA, Cincinnati, Ohio

UV DISINFECTION 485
Dr. O. Karl Scheible, HydroQual, Inc., Mahwah, New Jersey

A PRELIMINARY ASSESSMENT OF HIGH BIOMASS SYSTEMS 515
Dr. Arthur J. Condren, James M. Montgomery Consulting Engineers, Inc., Pasadena, California; Dr. Bjorn Rusten, Aquateam Norwegian Water Technology Centre AS, Oslo, Norway; and Dr. James A. Heidman, Sanitary Engineer, Water and Hazardous Waste Treatment Research Division, Risk Reduction Engineering Laboratory, U.S. EPA, Cincinnati, Ohio

EXPERIENCE WITH INTRACHANNEL CLARIFIERS IN THE U.S.A.....531
James F. Kreissl, Environmental Engineer, Water and Hazardous Waste Treatment Research Division, Risk Reduction Engineering Laboratory, U.S. EPA, Cincinnati, Ohio; Dr. A. T. Wallace, Professor, College of Engineering, University of Idaho, Moscow, Idaho; and Jon H. Bender, Environmental Engineer, Technical Support Division, Office of Drinking Water, Office of Water, U.S. EPA, Cincinnati, Ohio

INNOVATIVE WASTEWATER TREATMENT PROCESSES

- INTERIM RESULTS OF BIOFOCUS WT PROJECT -

by

Shunsoku Kyosai, Yutaka Suzuki, Masahiro Takahashi,
Katsumi Moriyama, Masashi Ogoshi, Toshiyuki Nakajima,
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October 1989
Cincinnati, Ohio

1. PREFACE

The Ministry of Construction of the Government of Japan started a five-year research project, "The Development of New Wastewater Treatment Systems Employing Biotechnology", which is commonly called Biofocus WT, in fiscal year 1985 as one of the comprehensive research and development projects of the Ministry.

By broadly applying biotechnology, which is a technology that utilizes the life maintenance functions of biological organisms, to wastewater treatment, Biofocus WT intends to solve the present and future problems in wastewater treatment listed below.

a. lower energy requirement and lower O & M costs
b. smaller land requirement for wastewater treatment facilities
c. better effluent quality
d. recovery of valuable resources from wastewater treatment
e. easier operation & maintenance of wastewater treatment facilities
f. discovery of useful microorganisms for wastewater treatment

The Public Works Research Institute and the Building Research Institute of the Ministry of Construction are the main organizations carrying out Biofocus WT. The former is mainly in charge of matters related to publicly owned wastewater treatment plants (POWTP), while the latter oversees matters related to household wastewater treatment tanks.

Table 1.1 shows the subjects of research and development in Biofocus WT related to POWTP. This paper presents the interim results of Biofocus WT being carried out by the Public Works Research Institute.

Table 1.1 Subjects of Research and Development in Biofocus WT

a. Study on Microorganism Bank for Wastewater Treatment
b. Study on Application of Genetic Engineering to Microorganisms for Wastewater Treatment
c. Study on Immobilization Methods of Microorganisms for Wastewater Treatment
d. Development of Bioreactors for Wastewater Treatment
e. Development of Bioreactors for Sludge Treatment
f. Development of Solid-Liquid Separators for Raw Wastewater
g. Development of Biosensors for Water Quality Measurement
h. Development of New Wastewater Treatment Systems

2. STUDY ON MICROORGANISM BANK FOR WASTEWATER TREATMENT

2-1 OUTLINE OF MICROORGANISM BANK

Recently, environmntal microbiology has achieved great progress. Though there is much microbiological information useful for wastewater systems, most sanitary engineers can not fully utilize such information. There may be several reasons for this situation: (1) The results of microbiology research are too difficult for sanitary engineers to utilize. (2) The esults are too specialized in a certain field. (3) Sanitary engineers can not obtain the information easily.

A microorganism bank for wastewater treatment has been investigated in order to apply the results of microbiology research to wastewater engineering. There are several key points for establishing such a bank. A technical operation strategy is one of the most important issues as well as a financial operation strategy. Development of the technical operation strategy would be the result of this research.

The bank deals with the microorganisms which are principally related to wastewater systems. A unique feature of the bank is that it consists of two parts, a data bank and a preservation bank. The data bank accumulates data sheets in which microbiological and ecological data on microorganisms related to wastewater systems are summarized. The preservation bank is an ordinary one. It will not collect every strain listed in the data bank, but only those which are not kept at other places. The preservation places where the strains are kept are listed in the data bank.

2-2 DATA BANK

Most microbiological data on any individual strain are obtained from pure cultures. However, most microbiological phenomena in biological wastewater treatment processes are observed in mixed cultures of bacteria, protozoa and other microorganisms. The results of the pure cultures are usually different from the phenomena in the mixed cultures. This understanding impedes the utilization of microbiological information for wastewater treatment systems. The data bank would reconcile this situation by providing the ecological properties of each strain. It also employs special classification of microorganisms. All strains are classified by their relationship to wastewater systems. A tentative classification is shown in Table 2.1.

Table 2.2 is the data sheet from the data bank. It contains ecological information as well as ordinal microbiological data. Data concerning wastewater treatment systems are also summarized.

Table 2.3 shows the list of bacteria species which have been registered in the bank. These species were collected from papers, or were obtained directly from researchers.

Table 2.1 Tentative Classification of Microorganisms in Relation to Wastewater Treatment

WASTEWATER TREATMENT
Removal of Biodegradable Organics
Removal of Refractory Organics
Removal of Synthetic Hazardous Organics
Nitrification
Denitrification
Phosphorus Removal
Iron, Manganate Removal
Removal of Trace Heavy Metals
Removal of Soluble Inorganic Materials
Oxidation of Sulfur Compounds
Reduction of Sulfur Compounds
Sedimentation of Suspended Solids
Flotation of Suspended Solids
Liquefaction of Suspended Solids
Odor Control
Color Removal
Epidemic Risk Control
SLUDGE TREATMENT AND DISPOSAL
Liquefaction of Sludge
Volatilization of Sludge
Heavy Metal Removal from Sludge
Composting
Resource Recovery from Sludge

Table 2.3 List of Strains

Strains	Characteristics
Nitrosomonas europaea	Nitrification
Nitrobacter sp.	Nitrification
Thiothrix sp.	Sulfur Oxidation, Bulking
Beggiatoa sp.	Sulfur Oxidation, Bulking
Thiosheara pantotropha GB17	Sulfur Oxidation
Rhodopseudomonas sp.	Removal of Organic Material
Rhodospirillum rubrum	Removal of Organic Material
Flavobacterium sp., or Pseudomonas sp.	Removal of Synthetic Hazardous Organic Material
Alcaligenes eutrophus H850	Removal of Synthetic Hazardous Organic Material
Acintobacter sp. GJ70	Removal of Synthetic Hazardous Organic Material
Rhodococcus sp. TTD-1	Removal of Synthetic Hazardous Organic Material, Odor Control, Sulfur Oxidation
Acintobacter calocaeticus SH-1	Removal of Synthetic Hazardous Organic Material
Pseudomonas sp. K172	Removal of Synthetic Hazardous Organic Material, Denitrification
Pseudomonas sp. S100	Removal of Synthetic Hazardous Organic Material, Denitrification

Table 2.2 Data Sheet of Data Bank

1. CODE
 (1) Bank Code
 (2) Scientific Classification: 1) Family 2) Species 3) Common Name
2. RELATION TO WASTEWATER TREATMENT
 (1) Functional Name
 (2) Relationship to Wastewater Treatment
3. DATA SHEET WRITER
 (1) Name
 (2) Affiliation
 (3) Date of Register
4. RESEARCHER
 (1) Name
 (2) Affiliation
5. PRESERVATION
 (1) Organization
 (2) Code Number in the Organization
6. ORIGIN OF STRAIN
 (1) Purification from Natural Environment
  a) Purification Data: Source of Strain, Date of Purification, Researchers, Affiliation
  b) Identification Data: Researchers, Affiliation
 (2) Genetic Engineering
  a) Applied Genetic Engineering Method
  b) Conditions: Host, Vector, Other
  c) Characteristics of Strain
 (3) Other
  a) Obtained from Other Researcher
  b) Characteristics of Strain
7. CHARACTERISTICS OF STRAIN
 (1) Physiological Characteristics: Growth Rate, Substrate, Products, Genetic Data, so forth
 (2) Ecological Characteristics: Environmental Condition for Growth, Competition, Symbiosis, so forth
8. NOTES FOR EXPERIMENT
 Epidemic Notes, Hazardous Products, Patents, so forth
9. CONDITION FOR PURIFICATION
 (1) Cumulative Cultivation: Temperature, Media, Method
 (2) Purification: Temperature, Media, Method
10. PRESERVATION OF VIABLE CELL: Temperature, Media, Subculturing Method
11. PRESERVATION AND REACTIVATION OF FREEZING CELL
 (1) Preservation: Temperature, Media, Period of Preservation
 (2) Reactivation: Temperature, Media, Reactivation of Specified Nature
12. RELATED LITERATURE

3. STUDY ON APPLICATION OF GENETIC ENGINEERING TO MICROORGANISMS FOR WASTEWATER TREATMENT

3-1 ECOLOGICAL AND EVOLUTIONARY STABILITY OF ENGINEERED MICROORGANISMS

3-1-1 INTRODUCTION

There are various technical problems in the application of engineered microorganisms to wastewater treatment systems. From the view point of stability of useful genetic traits or genetic information of engineered microorganisms, the problems can be divided into three as to stability at each level of the biological system.

Firstly, the useful genes must be stably retained in the cell. This stability includes synchronous replications of plasmids with replications of the host cell. This stability is called physiological stability here.

Secondly, a sufficient population of the cells with physiologically stable genetic information is required to survive in a microbial community within a reactor under ecological interactions such as interspecific competition, predator-prey interactions and so forth. This stability is called ecological stability of genetic information here.

Thirdly, various mutants may arise from an ecologically stable population with useful traits after a long period. Then, intraspecific competition between the cells of original genotype with useful traits and mutants without useful traits would occur. If the fitness of the original phenotypic cells are lower than that of mutants, the population of the original genotype would decrease and disappear from the reactor. Thus, engineered organisms are required to have higher fitness in a given reactor than any other mutants which might arise from themselves. This stability, called evolutionary stability of the genotype of a microbial organism here, may be a kind of ecological stability, because evolutionarily stable bacteria with useful traits are able to exist under intraspecific competition with the mutants. The notion of evolutionary stability, however, can be useful to differentiate the ecological stability of a population in evolutionary (i.e., long) time scale where the effects of mutants can not be ignored from that in ecological (i.e., short) time scale where there is no significant effect of mutants on the population dynamics.

In this study, from the view point of the ecological and evolutionary stability of bacterial populations, a basic approach to the application of genetically engineered microorganisms to wastewater treatment technology was done using long term experimental cultures of enteric bacteria.

3-1-2 MATERIALS AND METHODS

Bacteria used : Escherichia vulneris JCM1688, Escherichia blattae JCM1650, Enterobacter cloacae JCM1232, Enterobacter aerogenes JCM1235, Citrobacter freundii IFO12681, Citrobacter freundii IFO13539, and Citrobacter freundii IFO13545. Long term cultures of bacteria were conducted by a serial transfer culture illustrated in Figure 3.1 and changes in ecological traits related to their fitness in the culture were examined as to specific growth rate, cell dispersiveness, biochemical characteristics, etc.

3-1-3 RESULTS

The experimental results are shown in Figures 3.2 and 3.3, and summarized as follows:

(1) The fitness of bacterial populations, which is expressed by the specific growth rate in the culture, increased with the number of transfers. This is considered to be caused by the appearance and increase of mutants with higher fitness in the culture.
(2) The fitness of a bacterial population eventually came to a plateau level.
(3) The plateau values of the fitness of bacteria were not changed by the treatment of a chemical mutagen with NTG.
(4) The plateau values were different in species (see Figure 3.2), and in case of C. freundii the values varied in strains.
(5) Genetically identical bacterial strains which were derived from the same clone have the same value for the plateau level.
(6) From the experimental results of mixed cultures consisting of two kinds of bacteria, both of which had reached each plateau level, it was revealed that the population with the lower plateau level decreased and became extinct in the mixed cultures, while that with the higher plateau level remained (see Figure 3.3).

3-1-4 DISCUSSION

It was revealed that ecological traits of bacteria can genetically change through intraspecific competition. In addition, the changes in their growth rates were limited to certain values which were depending on species. From the view point of the application of engineered microorganisms to wastewater treatment, it is important to use the bacterial species or strains with higher plateau levels for treatment, whether they are selected from natural microbial communities or genetically modified in laboratories.

3-2 GENETIC MANIPULATION OF ECOLOGICAL TRAITS OF MICROORGANISMS RELATED TO THEIR FITNESS

3-2-1 INTRODUCTION

Many genes with useful traits, which are integrated into particular plasmids, have been transferred into bacterial cells in order to genetically modify the bacterial ability for some industrial demands. This method, however, can not manipulate a number of genetic traits which would be related to ecological fitness of the bacteria in some way. It commonly deals only with genes related to some biochemical or physiological traits which are considered to be useful for such demands. For wastewater treatment, engineered microorganisms should grow under various kinds of ecological interactions with other species in a given reactor. Then, physiological and ecological traits of microorganisms required for the stable existence in a reactor should be open to genetic modification, if it becomes necessary.

In this study, focusing on how we can genetically manipulate such quantitative phenotypes as ecological traits which are considered to be expressed by many genes on the chromosomes, for the first step, we investigated the genetic changes of ecological traits of bacteria by genetic recombination at a genomic level.

Hfr strains of Escherichia coli K12 are known as conjugational donors of the chromosome DNA into F^- strains at high frequency. When an Hfr cell joins to a F^- cell, conjugation-induced replication of F^- factor which is integrated into the chromosome begins, and transfer of the chromosome DNA follows. Complete transfer of the chromosome requires about 100 minutes at 37°C. By breaking apart the male (Hfr) and female (F^-) conjugant unions by violent agitation at different times, we can get various kinds of recombinant strains which vary in the degree of recombination between parental strains. Using these strains, we investigated the genetic changes of ecological traits of bacteria by genetic recombination at a genomic level.

3-2-2 MATERIALS AND METHODS

E. coli K12 JE5879 (suc, man) and E. coli K12 ME7767 (ara, mal, xyl, mtl, lac, suc, str, gal) were used as an Hfr strain and an F^- strain, respectively.

Conjugation between JE5879 and ME7767 and isolation of recombinant strains between the two were carried out by the method illustrated in Figure 3.4.

Each strain was cultured in an L-broth medium. 6 ml of JE5879 and 3 ml of ME7767 suspensions at a stationary phase of their growth were mixed together in a 300 ml flask. The mixture was statically incubated for 180 minutes at 37°C.

After 30, 60, and 100 minutes of incubation, 1 ml of the bacterial suspension were taken from the flask. The samples were agitated with a violent blender for 30 seconds, and plated onto three kinds of EMB agar plates containing maltose and streptomycin, xylose and streptomycin, and galactose and streptomycin. After incubation for 1-2 days, dark-violet bacterial colonies, which indicated acid forming ability from the sugars, were randomly isolated from the plates with a loop. The ten strains isolated from the plates that were incubated for 30 minutes are here referred to as hy30-1 through hy30-10. Each isolated strain was purified two times on the same kind of EMB agar plate as the plate for the isolation. These isolates were transplanted to L-broth slant agar, and stored at 4°C. To obtain bacterial growth curves, optical densities of bacteria were monitored with a spectrophotometer in their cultures.

3-2-3 RESULTS

The changes of viable cell counts on the three kinds of EMB agar plates during the conjugation experiment are shown in Figure 3.5. The growth curves of ME7767, hy100-7, hy100-8, and hy100-9 are shown in Figure 3.6. Various types of growth curves which varied in the growth rate and maximum density at a stationary phase of the population growth were observed. While both the growth rate and the maximum density of hy100-7, for example, were higher than those of parental strains, the growth rate was lower than that of hy100-9. The growth rate of JE5879 was much lower than that of ME7767.

3-2-4 DISCUSSION

Although the ecological traits of hy-30 and hy-60 are not examined at

this time, from the result of hy-100 it is revealed that ecological traits such as growth rate, maximum density at a stationary phase can be changed by genetic recombination at a genomic DNA level.

4. STUDY ON IMMOBILIZATION METHODS OF MICROORGANISMS FOR WASTEWATER TREATMENT

The immobilization of microorganisms in bioreactors for wastewater treatment and sludge treatment is useful to increase the concentration of microorganisms in the reactor, and to maintain a large amount of microorganisms for specifically removing particular pollutants. Three immobilization methods will be discussed: binding method, entrapping method and self-immobilization method.

Microorganisms themselves attach to the surface of media in the binding method. Microorganisms are immobilized by chemicals in the entrapping method. Immobilization is made by microorganisms themselves with no other materials in the self-immobilization method.

4-1 BINDING METHOD

Suitable shape and materials of attached media depend on the features of reactors in which the media are used. The reactors can be classified into fixed bed type or fluidized bed type and aerobic or anaerobic.

4-1-1 ATTACHED MEDIA FOR FIXED BED TYPE REACTOR

(1) For Aerobic Treatment

Attached biomass was compared among eight different media under conditions of moderate aeration and synthetic substrate addition. Figure 4.1 shows the results. Though attached biomass was different between media, sufficient biomass grew after 3 months on all media tested.

Table 4.1 is an example of physical characteristics of attached media. The media with more specific surface area accumulate more biomass but require regular washing due to clogging. The media with diameter less than 1 cm remove SS, then additional SS removal is unnecessary, though washing is required. On the other hand, washing is not necessary for media with bigger diameters but solids separation from the effluent is essential.

Table 4.1 Examples of Physical Characteristics of Attached Media

Media	Shape	Void Ratio (%)	Size (cm)	Specific Surface Area (g/cm^3)	Specific Gravity (cm^2/cm^3)
Ceramics	Saddle	77	2	335	2.64
Ceramics	Ball	40	1	400	1.63
Polypropylene	Cylinder	89	5	76	0.95
Polypropylene	Ring	89	5	185	1.45

(2) For Anaerobic Treatment

A comparison test of attached media for anaerobic treatment revealed that porous media produce gas more than media with smooth surfaces, showing that microorganisms attach to the porous media more than the smooth media. As no aeration results in insufficient mixing in anaerobic treatment compared with aerobic treatment, it was necessary to select and arrange media so that

adequate influent distribution was attained.

4-1-2 ATTACHED MEDIA FOR FLUIDIZED BED TYPE REACTOR

Attached biomass was compared among different media under aerobic and anaerobic conditions. Table 4.2 shows the results.

Table 4.2 Results of Comparison Tests of Attached Media for Fluidized Bed Type Reactor

Media	Size (mm)	Specific Gravity (g/cm^3)	Aerobic or Aerobic	Attached Biomass (mg*VSS/ g*media)	Biomass in Reactor (g/l)
Silica Sand A	0.2-0.3	2.65	Aerobic	40	4.4
Silica Sand B	.074-0.15	2.65	Aerobic	1,200	12
Silica Sand C	0.09-0.25	2.63	Anaerobic	3.5	0.8
Zeolite	0.35	1.98	Anaerobic	10-30	10-20
Polypropylene	2-5	1.03	Aerobic	26	3.8
Plastic Foam	1-2	1.3	Aerobic	95-150	4.3-6.7

Attached biomass and total biomass in reactors was more for sand B than A, which has a smaller diameter under aerobic conditions. Comparison of sand B and C shows that aerobic microorganisms tend to attach to media more than anaerobic microorganisms. Zeolite had more biomass than sand in an anaerobic condition. As polypropylene pellets have the least specific gravity, they were fluidized in a reactor with the least mixing. As about 90 % of the volume of plastic foam has pores with diameters 0.03 to 1 mm, it accumulates many microorganisms.

Granular activated carbon and porous ceramics had problems with abrasiveness.

4-2 ENTRAPPING METHOD

Polyvinyl alcohol (PVA), acrylic amide (ACAM) and polyethylene glycol (PEG) were compared as chemicals to entrap microorganisms.

4-2-1 POLYVINYL ALCOHOL (PVA)

(1) PVA-Boric Acid Method

In this method, a mixture of PVA and activated sludge is dropped into saturated boric acid solution, then activated sludge is entrapped in lattices of pellets made of PVA and boric acid. Appropriate conditions of the method were as follows:

- -PVA concentration in the mixture of PVA & activated sludge: 7.5-10%
- -Activated sludge concentration: 80-90 g/l
- -Contact time of the mixture with the saturated boric acid solution: 24 hours

Though the respiration rate of the entrapped activated sludge was low at the beginning of cultivation, the rate increased rapidly to 3 - 4 times the initial value after 10 days, as shown in Figure 4.2.

(2) PVA-Refrigeration Method

The mixture of PVA and activated sludge is refrigerated and becomes a gel. The activity of microorganisms is only slightly reduced after the immobilization by this method compared with the method using boric acid, which is harmful to microorganisms.

The appropriate refrigeration temperature was minus 20°C. The strength of the gel did not increase at temperatures lower than the value. The PVA concentration of 10 % was not enough to make pellets with enough strength. Pellets of 22.5 % showed a lower respiration rate than that of 15 % as shown in Figure 4.3, thus PVA concentration of about 15 % is suitable.

4-2-2 ACRYLIC AMIDE (ACAM)

ACAM is relatively cheap and a typical polymer for entrapping microorganisms. However, it has the ability to impair the activity of microorganisms due to a harmful monomer present during the immobilization.

(1) Agar-ACAM Monomer Method

Microorganisms are first immobilized by agar, then an ACAM monomer is added to the immobilized pellets to polymerize and harden them. The more ACAM monomer added resulted in less activity of microorganisms. But less addition of monomers leads to weak gels. The proper concentration of ACAM was about 7.5 % as shown in Figure 4.4.

(2) ACAM Monomer Method

A cross-linking agent and a polymerization agent are added to a mixture of ACAM monomer and activated sludge to make gels. Figure 4.5 shows the test results. Proper concentration of ACAM monomer was about 15 % in this method. Variation of concentration of a cross-linking agent had no effect on the activity of microorganisms.

4-2-3 POLYETHYLENE GLYCOL (PEG)

An immobilization method by PEG was developed as a method to immobilize nitrifiers. Activity of nitrifiers in gels by PEG was higher by 5 times than those by ACAM, melamine and urethane.

A respiration rate of gels prepared by PEG after sufficient cultivation was 1,100 mg*O_2/l*hr which was higher by 2 times than those prepared by other materials. This was mainly because the penetration rate of ammonia from water to gels was the highest in PEG gels among the chemicals tested.

4-3 SELF-IMMOBILIZATION METHOD

Immobilization is made by microorganisms themselves with no other materials in this method. Dense granules made of microorganisms are naturally formed without sludge recirculation in upflow sludge blanket type bioreactors. UASB (upflow anaerobic sludge blanket) process applied for treatment of wastewater with high organic concentration has been well known as a process utilizing this method. In Biofocus WT, four processes using self-immobilization methods have been investigated. There are two UASB processes, an AUSB (aerobic upflow sludge blanket) process and MRB (multi-stage reversing-flow bioreactor).

4-3-1 UASB PROCESSES

(1) For Municipal Wastewater Treatment

Anaerobic sludge blanket of 10 g*SS/l was obtained under conditions of temperature 10 to 25°C, linear velocity (LV) 5 to 6 cm/min and hydraulic retention time 6 to 9 hours. Biological granules have formed after 6 months acclimation. Though they are small in diameter (1 mm), they are basically the same as biological granules generally found in the UASB process treating wastewater with high organic concentration.

(2) For Treatment of Supernatant from Heat Treatment of Sludge

Anaerobic sludge blanket shown in Figure 4.6 was obtained. COD_{cr} in influent was 5,000 mg/l and temperature in the reactor was 35°C. Biological granules were 0.5 to 2 mm in diameter and had 1 to 5 cm/sec of a settling velocity. Methanothrix seemed to dominate in the granules.

4-3-2 AUSB PROCESS

Aerobic granules were developed in this process. DO in influent to the reactor is as high as 50 to 70 mg*O_2/l for keeping the reactor aerobic. This high DO concentration is achieved by contact of wastewater to pressurized pure oxygen as shown in Figure 4.7. Granules of several milimeters in diameter have been observed. The mechanism of the aerobic granulation is under investigation.

4-3-3 MRB

Multi-stage Reversing-flow Bioreactor (MRB) which utilizes new biological interaction for removal of organic substrate in wastewater has been developed in Biofocus WT.

(1) Principle of MRB

Figure 4.8 is a flowsheet of the MRB. Influent is introduced to the first downflow aeration vessel (AV). No suspended solids remain here. The next vessel is a biological reaction vessel (BV) of upflow without aeration. Suspended solids with a higher settling velocity than the upflow velocity remain here. The following vessels have the same sequences. The AV is merely for oxygen supply. In the BV, suspended solids and biomass are gently mixed by the upflow, resulting in formation of self-granulated sludge (SGS) with diameters of 2 to 10 mm. The primary effluent of a municipal wastewater treatment plant was influent to the MRB. In the initial period of the operation, anaerobic sludge was accumulated in the first BV and then the sludge was carried over from the first BV to the second and third BVs, the SGS was formed in due course as shown in Figure 4.9. Once the sludge blankets of SGS were built up in these BVs, the fourth and fifth BVs receiving partially treated wastewater became aerobic and developed an aerobic sludge blanket. The aerobic sludge blankets did not have the firm SGS, but could be kept in these BVs against 144 m/day upflow velocity. Sludge returning, liquid circulation and a final sedimentation tank were not required.

(2) Mechanism of Self-Granulation of Sludge in MRB

One of the interesting topics of the MRB is the formation of SGS. The SGS is firm enough to be picked up by tweezers, but is crushed by strong shear stress, such as when it is pumped out. Most of the SGS is covered with a thin white film as shown in Figure 4.9. The film consists of filamentous bacteria

as shown in Figure 4.10. The bacteria, having small sulfur granules in their cells and showing active gliding mobility, are identified as Beggiatoa. The inside of the SGS is black and anaerobic. The plate count method for sulfate reducing bacteria indicated that there were 10^8 N/g-SS of sulfate reducing bacteria (SRB). In the MRB, oxygen is supplied at the AV where few microorganisms are present. DO in the AV is limited to the saturation concentration for wastewater at a given temperature. On the other hand, a significant amount of microorganisms accumulate in the BV, and the DO supplied in the AV is consumed rapidly if organic substrate in the wastewater is high. Consequently, near-anaerobic conditions allowing the growth of the SRB are developed in the BV. From the above-mentioned observations, the mechanism of self-granulation of sludge in the BVs is considered as follows. Figure 4.11 explains the interaction among the SRB, Beggiatoa as sulfide oxidizing bacteria and other anaerobic bacteria. Organic substrate diffusing into the SGS is hydrolyzed to organic acids by anaerobic bacteria, then utilized by the SRB resulting in sulfate reduction.

Sulfide produced from this reaction diffuses into bulk liquid through the SGS surface. Though the oxygen supply is limited in the BV, there is the chance of the microorganisms contacting oxygen on the SGS surface. Because an oxygen consumption rate of bacteria oxidizing sulfide is much higher than that of organic substrate oxidizing bacteria, most the oxygen is utilized by the sulfide oxidizing bacteria which are microaerophilic. Beggiatoa, which is one of the most common bacteria existing in such an environment, increases on the SGS surface in a thin film.

5. DEVELOPMENT OF BIOREACTORS FOR WASTEWATER TREATMENT

Many municipal sewage systems are being designed or constructed in rural areas in Japan. Even in rural areas, the land for wastewater treatment plants is limited in many cases. In addition, small municipalities tend not to be able to employ enough skilled technicians.

Many local governments will have to reconstruct or to improve their old wastewater treatment plants in the future. An increase in treatment capacity, and/or upgrading removal efficiency of nutrients or refractory organic materials might be required in such cases.

Development of bioreactors for wastewater treatment aims to answer these needs mainly by introducing immobilization of microorganisms. There are three targets of the development, in addition to develop bioreactors with features of easier O & M, phosphorus removal or refractory organic removal.

(1) To reduce the retention time in reactors to 1/2 to 1/10 compared with the conventional processes.
(2) To reduce the energy requirement and/or sludge production to 1/2 compared with conventional processes.
(3) To reduce the retention time in reactors for biological nitrogen removal to 6 to 12 hours so that existing plants can be modified to remove nitrogen.

The first target would be accomplished mainly by the development of aerobic bioreactors. Anaerobic bioreactors are for the second target. The third target would be concluded by immobilization of microorganisms related to nitrogen removal.

In total, 23 bioreactors are under development. All of the bioreactors have been developed as joint researches between the Public Works Research Institute and private companies. Twenty-three pilot plants are operated at present.

5-1 AEROBIC BIOREACTOR

Thirteen aerobic bioreactors have been developed. They are divided into three groups: fixed bed type, fluidized bed type and sludge blanket type.

5-1-1 FIXED BED TYPE

Figure 5.1 shows an example of fixed bed type aerobic bioreactors which have been developed. The aerobic biofilter has several experimental and full scale trials in Japan and abroad. Influent is introduced to the top of the filter. Air is dispersed from the bottom of the filter. Wastewater is treated by the biofilm on the surface of filter media. Suspended solids are removed by a filtration mechanism.

Suspended solids in influent is better to be low in order not to cause clogging in the filter. The filter is washed by backwashing. The operational conditions and the performance are summarized in Table 5.1.

Table 5.1 Operational Conditions and Performance of Fixed Bed Aerobic Bioreactor

CONDITIONS						
Linear Velocity, m/day		31		50		70
Hydraulic Retention Time, hr		1.65		1.01		0.7
BOD Volumetric Load, kg-BOD/m^3/day		1.1		1.5		2.3
PERFORMANCE						
		Inf.	Eff.	Inf.	Eff.	Inf.	Eff.
SS	mg/l	42	4	39	4	35	4
BOD	mg/l	72	6	62	9	69	11
T-K-N	mg/l	28.8	4.1	27.1	5.7	24.1	11.1
T-P	mg/l	2.8	1.9	2.2	1.5	2.1	1.5

5-1-2 FLUIDIZED BED TYPE

Figure 5.2 shows an example of fluidized bed type aerobic bioreactors which have been developed. Air diffusers are installed in the reactor. Fine sand, fine slag, granulated activated carbon, plastic media, and chemical media which are made of polyacryl amide and poly-ethylenglycol for immobilizing microorganisms are used as the fluidizing media in these developments.

Media separation from liquid is one of the engineering problems to be solved. When effluent from bioreactors has little SS, final solid-liquid separation might be accomplished not by sedimentation but by filtration because settlability of the SS is poor. How to fluidize the media is another hurdle. Lighter and finer media are better for fluidizing, but such media are difficult to separate from liquid.

Table 5.2 Operational Conditions and Performance of Fluidizing Aerobic Bioreactor

RUN		1	2	3
CONDITIONS				
Hydraulic Retention Time of Reactor, hr		2.0	2.0	4.0
Linear Velocity of PF, m/day		80	80	53
Linear Velocity of SF, m/day		161	161	115
PERFORMANCE				
T-BOD mg/l	Influent	138	91.9	132
	PF Effluent	20.6	16.0	17
	DF Effluent	9.4	11.1	9.1
S-BOD mg/l	Influent	44.3	47.0	44.7
	PF Effluent	10.9	9.5	8.5
	DF Effluent	8.0	8.0	6.3
SS g/l	Influent	109	40.1	102
	PF Effluent	10.1	7.5	6.7
	DF Effluent	3.2	2.6	2.2

PF: Plastic Media Filter
DF: Dual Media Filter

A plastic medium of 3 to 5 mm in diameter and about 1.0 in density has been employed in a development. It can be separated by a screen. The operational conditions and the performance of the plastic media bioreactor followed by a two stage filtration process are summarized in Table 5.2.

5-1-3 SLUDGE BLANKET TYPE

A flow diagram of an aerobic sludge blanket reactor developed in this project is shown in Figure 5.3. Influent flows into a pressurized oxygen tank. Effluent from the tank with 50 to 70 mg/l of DO is introduced to the bottom of the upflow reactor. At the beginning of the operation, activated sludge was seeded in the reactor. Within several weeks, aerobic biological granules formed in the reactor. The operational conditions and the performance are summarized in Table 5.3. Though the mechanism of self-granulation has not been clear, sludge that settles easily has been maintained for one year.

Table 5.3 Operational Conditions and Performance of Sludge Blanket Type Aerobic Bioreactor

CONDITIONS			
Hydraulic Retention Time, hr			1.72
Recirculation Ratio of Effluent			1.57
Up Flow Velocity in Reactor, m/day			125
Influent DO mg/l			63
Effluent DO mg/l			4.8
PERFORMANCE			
		Influent	Effluent
T-BOD	mg/l	112.6	19.9
S-BOD	mg/l	42.4	10.2
SS	mg/l	91.3	9.6
T-N	mg/l	30.9	24.6
T-P	mg/l	2.7	1.8

5-2 ANAEROBIC BIOREACTOR

Six anaerobic bioreactors are under development. They are also divided into 3 groups: fixed bed type, fluidized bed type and sludge blanket type. Anaerobic wastewater treatment requires the post-treatment due to the following two reasons. Effluent from anaerobic treatment can hardly satisfy the BOD standard of 20 mg/l for secondary treatment. It also contains sulfide which causes corrosion and odor problems.

Most post-treatment processes under development are aerobic biological treatments which require additional energy and facilities. A pilot plant has installed a modified trickling filter for post-treatment. The filter media are cubic sponges which catch up fine suspended solids. This system can eliminate further SS removal facilities and aeration devices for post-treatment.

5-2-1 FIXED BED TYPE

Non-woven fabrics, saddle-shaped ceramics and plastic braids are used as

media for fixed beds.

Figure 5.4 shows an example of a fixed bed type anaerobic bioreactor under development. Concentrated anaerobic sludge can shorten the hydraulic retention time to less than 8 hours for primary effluent. There is no recirculation of effluent.

5-2-2 FLUIDIZED BED TYPE

One fluidizing anaerobic bioreactor with fine zeolite of 0.5mm in diameter as media has been developed. The flow diagram is shown in Figure 5.5. Recirculation of effluent is required to fluidize the media and the recirculation ratio depends on the fluidized velocity of the media. The retention time is less than 3 hours, but this reactor requires energy for recirculation.

5-2-3 SLUDGE BLANKET TYPE

An upflow anaerobic sludge blanket (UASB) reactor shown in Figure 5.6 has been applied for domestic wastewater treatment. Pellets of anaerobic microorganisms formed 6 months after seeding digested sludge to the reactor. The effluent from the reactor has been good as the influent to the modified trickling filter mentioned before. The operational conditions and the performance are summarized in Table 5.4. The system has neither liquid recirculation, nor aeration devices. If there is enough water head difference available between influent and effluent in a plant, this system can be an energy saving system.

Table 5.4 Operational Conditions and Performance of Sludge Blanket Type Anaerobic Bioreactor

CONDITIONS			
Hydraulic Retention Time of Reactor, hr			8
Hydraulic Retention Time of MTF, hr			2
Water Temperature, C°			13-16
PERFORMANCE			
	Inf.	Eff.(UASB)	Eff.(MTF)
BOD mg/l	71-111	24-40	7-17
SS mg/l	45-99	8-15	2-7
Sulfide as S mg/l	-	6-8	0

MTF: Modified Trickling Filter for post-treatment

5-3 MULTI-STAGE REVERSING-FLOW BIOREACTOR (MRB)

The Multi-stage Reversing-flow Bioreactor (MRB) which utilizes new biological interaction for organic substrate removal in wastewater has been developed. As shown in Figure 4.8, MRB has several stages, one stage consists of a downflow aeration vessel (AV) and an upflow biological reaction vessel (BV). Successful accumulation of self-granulated sludge (SGS) can be achieved in BV. As described in 4-3-3, the biological interaction between sulfate reducing bacteria and sulfide oxidizing bacteria (Beggiatoa) is considered to contribute to form SGS with 2 to 10 mm in diameter. SGS can be maintained in BV with linear velocity of 144 m/day. A prototype plant study was conducted to demonstrate the performance of the MRB.

The design parameters, the operational conditions and the performance of the prototype MRB are summarized in Tables 5.5 and 5.6

Table 5.5 Design Parameters of Prototype MRB

Diameter:	AV	10 cm
	Lower part of BV	10 cm
	Upper part of BV	18 cm
Depth	Average	220 cm
Total volume		213 l
Hydraulic retention time		3-4.5 hour
Up-flow velocity at lower part of BV		144-216 m/day

Table 5.6 Operational Conditions of Prototype MRB

Influent		Settled municipal wastewater
BOD volumetric load	for total volume	0.42 kg/m^3.day
	for BV volume	0.73 kg/m^3.day
Hydraulic retention time		4.5 hour
Dissolved oxygen concentration in AV		5-9 mg/l

Table 5.7 Typical performance and Sludge Concentration in MRB

No. of BV		No.1	No.2	No.3	No.4	No.5
Total Solids (mg/l)		21,000	14,600	8,900	5,900	14,000
		Influent	Effluent			
BOD	(mg/l)	80	14			
D-BOD	(mg/l)	50	9			
SS	(mg/l)	30	4			

Though the influent BOD was sightly low, the effluent quality was as good as that of a conventional activated sludge process. The influent was municipal primary effluent. The total hydraulic retention time was 4.5 hours including the final solid-liquid separation in the last BV. The MRB has several advantages over conventional biological wastewater treatment, such as lower energy requirements, less sludge production, and no necessity for a final settling tank.

5-4 BIOREACTORS FOR NITROGEN REMOVAL

This study has been conducted to develop biological nitrogen removal processes utilizing immobilization methods of microorganisms related to biological nitrogen removal in pilot plants. The immobilization methods used are a binding method and an entrapping method.

5-4-1 BIOREACTOR FOR NITROGEN REMOVAL WITH BINDING METHOD

(1) Experimental Conditions of Pilot Plant Study

The flowsheet of the pilot plant is shown in Figure 5.7. , The size of the cross section of the reactor is 700 mm by 700 mm and 2,800 mm in height.

The volume is 1.37 m^3. Sands with diameters of 0.2 mm to 0.3mm constitutes 8 % of the reacting volume in the reactor. The specific surface area of the sand is 1.5 m^2/g. Sand was selected as attaching-media after a comparison test among clistobalite, sand and two kinds of light weight aggregate.

Influent flow was 11.5 m^3/day and the retention time in the reacting volume in the reactor was 2.9 hours. BOD, total nitrogen and ammonia nitrogen in influent were 110 to 160 mg/l, 40 to 60 mg/l and 30 to 40 mg/l, respectively. The water temperature was 28 to 29°C.

The operation mode of the reactor was intermittent aeration with continuous inflow. Both aeration time and anaeration time were 15 minutes. The volumetric loadings of BOD and total nitrogen in the reactor were 0.9 to 1.3 kg/m^3/day and 0.3 to 0.5 kg/m^3/day, respectively. The DO level was set to 2 to 3 mg/l during aeration period.

(2) Results of Pilot Plant Study and Discussion

Total nitrogen in the effluent was 8 to 15 mg/l as shown in Figure 5.8. The fact that ammonia nitrogen and nitrate nitrogen in the effluent were about 2 mg/l and 8 to 10 mg/l, respectively, shows that it is necessary to enhance denitrification for upgrading nitrogen removal.

MLVSS in the reactor was 7,000 to 9,000 mg/l. About 90 % of the MLVSS was attached to sands, which demonstrates the effectiveness of sand to increase microorganisms in the reactor.

The following are necessary to study in order to improve this biological nitrogen removal process.

a. optimum combination of aeration time and anaeration time
b. DO level during aeration time
c. air flow rate to fluidize sand and to supply DO
d. solid-liquid separation of effluent from the reactor

5-4-2 BIOREACTOR FOR NITROGEN REMOVAL WITH ENTRAPPING METHOD

(1) Experimental Conditions of Pilot Plant Study

Figure 5.9 is the flowsheet of the pilot plant. The volume of the nitrification tank is 4 m^3 and that of the denitrification tank is 6 m^3. 7.5 % of the nitrification tank volume is filled by pellets in which activated sludge taken from a biological nitrification-denitrification plant are entrapped. Polyethylene glycol was selected as a suitable chemical to make the pellets with activated sludge containing nitrifiers based on a selection test among 5 chemicals. As it was revealed that chemical entrapping of denitrifiers for denitrification was not effective, microorganisms including denitrifiers were suspended in the reactor. Experimental conditions are summarized in Table 5.8.

(2) Results of Pilot Plant Study and Discussion

Figure 5.10 is the performance of nitrogen removal in the pilot plant. BOD, TN and SS in effluent were 10 mg/l, 7 mg/l and 7 mg/l, respectively in RUN 1 in which the total retention time in the reactor was 8 hours. This performance was satisfactory, considering the water temperature was between 10 to 12°C. The total retention time was reduced to 6 hours in the following RUN 2 in which BOD and TN in the effluent were less than 20 mg/l and 10 mg/l.

Table 5.8 Experimental Conditions

	RUN-1	RUN-2
Inflow (m^3/day)	31.5	42
Retention Time in Biological Reactor (hr)	8	6
Retention Time in Nitrification Tank (hr)	4.8	3.6
Retention Time in Denitrification Tank (hr)	3.2	2.4
Percent of Pellets Filled into N. Tank by Volume (%)	7.5	7.5
Air Supplied (Nm^3/day)	200	230
Recirculation Ratio of Mixed Liquor (%)	300	300
Sludge Return Ratio (%)	40	40

This process can be applied to existing activated sludge plants for upgrading TN removal. The following should be studied in order to make this biological nitrogen removal process feasible.

a. life of pellets
b. a production method of huge amounts of pellets for full-scale plant application
c. cost of pellets, including disposal

6. DEVELOPMENT OF BIOREACTORS FOR SLUDGE TREATMENT

6-1 ANAEROBIC SLUDGE DIGESTION

There are two objectives in this development. One is to shorten the period of sludge digestion to less than 10 days, and the other is to increase the percentage of solid decomposition for recovering more digested gas by 1.5 times that of the conventional process. Two techniques have been investigated in order to attain the objectives. They are liquefaction of feed sludge and immobilization of bacteria related to anaerobic sludge digestion by the binding method.

It is well known that liquefaction of feed sludge, especially excess biological sludge, is the most limiting factor in the digestion. As attempts were unsuccessful to find bacteria and enzymes which accelerate the liquefaction of sludge, the direction has been changed to use physical-chemical methods. The following three methods have been investigated:

(1) Crushing excess sludge by a mill with NaOH of 0.05 N addition --- An example of the experiment is shown in Table 6.1.
(2) Heat treatment of excess sludge with conditions of 60°C for 1 hour
(3) Heat treatment of excess sludge with conditions of 90°C for 1 hour and protease addition to the heat treated sludge flowing to an acidification process

Table 6.1 Liquefaction of Excess Sludge by Crushing with NaOH

Items (mg/l)	---- CRUSHING ---		----- CRUSHING WITH NAOH ----		
	Control	Crushed	Control	NaOH Addition	Crushed with NaOH
TS	24,100	23,900	24,800	25,000	24,800
VS	19,700	18,800	19,500	19,300	18,400
VSS	18,400	12,500	18,100	16,400	8,100
pH	6.0	6.2	6.1	7.1	7.0
Sol. Sugars	199	458	128	232	554
Sol. Proteins	250	2,420	205	1,060	3,030
VFA	1,836	3,070	1,361	3,222	3,847
formic	-	114	-	28	51
acetic	580	2,082	460	1,824	3,162
propionic	425	-	346	350	-
isobutyric	84	348	250	99	-
butyric	401	-	327	239	-
isovaleric	157	526	119	169	-
valeric	189	-	109	513	168

Attached media for bacteria in sludge digestion are different from those for bacteria in wastewater treatment, because sludge is much more viscous and, of course, contains more SS than wastewater. Cement balls as attached media for a fluidized bed type reactor and ceramics as attached media for a fixed bed type reactor have been investigated.

Three pilot plants are under operation. Their flowsheets are in Figure

6.1. In addition to their digestion performance, dewaterability of the digested sludge, biodegradability of the filtrate and heat balance of each system are scheduled to be examined.

6-2 BACTERIA LEACHING OF HEAVY METALS FROM SLUDGE

The bacteria leaching technique, which is practiced in recovering metals from ores by oxidizing sulfur and lowering pH by sulfur oxidizing bacteria and iron oxidizing bacteria, was applied to sewage sludge in order to remove heavy metals from it.

6-2-1 GROWTH OF BACTERIA IN SLUDGE

Thiobacillus thiooxidans and Thiobacillus ferrooxidans were selected as sulfur oxidizing bacteria and iron oxidizing bacteria. The conditions for the two kinds of bacteria to grow in sludge were examined. Results of the study are summarized as follows:

(1) There was no difference in the growth of the bacteria in primary sludge, excess sludge, mixed sludge or digested sludge.
(2) Sterilization of sludge was not necessary for the growth of the bacteria.
(3) Though it was indicated that Thiobacillus thiooxidans and Thiobacillus ferrooxidans or bacteria similar to those existed in sludge, addition of Thiobacillus thiooxidans and Thiobacillus ferrooxidans accelerated the decrease of pH in sludge.
(4) Initial pH of sludge had no influence to the growth of the bacteria if the pH was between 4 to 7.
(5) Addition of powdered sulfur or ferric sulfate as a substrate was necessary for the growth of bacteria.

6-2-2 HEAVY METAL REMOVAL FROM SLUDGE BY BACTERIA LEACHING

Bench scale tests on the bacteria leaching of mixed sludge and digested sludge were conducted. Figure 6.2 is the procedure. The results were compared with those attained by sulfuric acid addition to sludge. Results of the study are summarized as follows.

(1) Mixed sludge and digested sludge came to pH equilibrium within 7 days after the start of the incubation. The pH values were 1.0 and 1.8 for mixed sludge and digested sludge, respectively.
(2) Leaching ratios depended on heavy metals. Cd, Zn and Mg were leached more than 80 %. Little Hg was removed from the sludge. Figure 6.3 shows changes of leaching ratios of heavy metals.
(3) There was no significant difference in the bacteria leaching and "chemical leaching" as to the leaching ratios of heavy metals.

7. DEVELOPMENT OF SOLID-LIQUID SEPARATORS FOR RAW WASTEWATER

Almost all bioreactors for wastewater treatment being developed in Biofocus WT incorporate immobilization methods in the reactors in order to increase concentration of microorganisms. A problem may exist in reactor clogging due to SS in influent for such reactors, especially fixed bed type reactors. In addition, as dissolved organics are decomposed by microorganisms faster than particulate organics, removal of the bigger particles from raw wastewater makes bioreactors more efficient.

Solid-liquid separators for raw wastewater, of which SS removal rate is higher than that of a conventional primary sedimentation tank, have been developed in this subject.

7-1 SIZE DISTRIBUTION OF SUSPENDED SOLIDS IN RAW WASTEWATER

Filtrate were prepared by sieves of 2 and 74 mm and filter papers of 8 and 1 microns. SS, BOD, COD, TN and TP in the filtrates were analyzed. Examples of the study are shown in Figures 7.1 and 7.2, and summarized as follows:

(1) Percentage of particulate BOD in raw wastewater was higher than those of particulate nitrogen and phosphorus. Then C/N and C/P ratios in influent to bioreactors become small, if the removal rate of SS in raw wastewater is high (see Figure 7.1).
(2) Size distribution of SS in raw wastewater is different during high influent loading and during low influent loading. Percentage of SS more than 8 microns is higher during high influent loading than during low influent loading (see Figure 7.1).
(3) There was a possibility that influent to small plants with shorter time of flow contained soluble BOD more than that of large plants with longer time of flow (see Figure 7.2).

7-2 BIODEGRADABILITY OF PARTICULATE ORGANICS

Biodegradability of particulate TOC was examined under aerobic and anaerobic conditions. Examples of the study are shown in Figures 7.3 and 7.4. Initial biodegradation rate of particulate TOC in particle size more than 1 micron meter was much lower than that of smaller than 1 micron meter in both aerobic and anaerobic conditions.

7-3 SOLID-LIQUID SEPARATORS FOR RAW WASTEWATER

Five solid-liquid separation processes have been developed in this subject.

(1) Upflow sedimentation process
(2) Floating-bed filtration process
(3) Filtration process with tube-shaped filtration media
(4) High rate bioflocculation process with lamella settler
(5) Two stage filtration process with upflow moving bed and downflow fixed bed

An upflow sedimentation process has been investigated using an experimental column of 280 mm in diameter and 1,500 mm in effective height and mixing paddles. Figure 7.5 is the result.

Removal rates of the processes of (2), (3) and (4) are summarized in Figure 7.6. Removal rates of SS larger than 8 microns in raw wastewater were more than 80 % in these three processes. On the other hand, SS smaller than 8 microns were difficult to remove by these processes.

8. DEVELOPMENT OF BIOSENSORS FOR WATER QUALITY MEASUREMENT

Biosensors, which use biomaterials such as enzymes and microorganisms as detectors to measure objective matter, have been developed as for BOD, ammonia-nitrogen and organic acids.

8-1 PRINCIPLE OF WATER QUALITY MEASUREMENT BY BIOSENSOR

Aerobic microorganisms are applied to these biosensors. Microorganisms are immobilized in a thin biofilm which is attached to an oxygen electrode. When the microbial electrode is immersed into a substrate solution, the substrate begins to diffuse into the biofilm. If microorganisms in the biofilm metabolize the substrate, they consume DO in the biofilm resulting in decline of oxygen electrode current. In a few minutes the oxygen consumption rate becomes equivalent to the oxygen diffusion rate towards the biofilm. Consequently the current of the electrode becomes steady (peak current).

When the microbial electrode is removed from the substrate solution and put into a solution without substrate, oxygen consumption rate decreases the level of endogenous metabolism (base current). An oxygen consumption rate, which is expressed as the difference between the peak current and base current, is in proportion to substrate concentration within a certain range.

The flow chart of a system is shown in Figure 8.1. The microbial electrode attached to the flow cell sends the current to a microcomputer (CPU) by way of an amplifier and an analog digital (A/D) transfer. A standard solution or a sample is injected into the flow cell for calibration or measurement. DO in the flow cell is almost completely eaturated. The flow cell is set in an isothermal water bath to maintain a constant temperature.

8-2 BOD SENSOR

8-2-1 SELECTION OF BIOFILM FOR BOD SENSOR

Microorganisms and immobilization methods to make biofilm for BOD sensor were experimentally compared among the following candidates.

Microorganisms:
- Sacharomyces cerevisiae(yeast)
- Sacharomyces uvarum(yeast)
- Schizosacharomyces pombe(yeast)
- Trichosporon cutuneum(yeast)
- Candida lypolylica(yeast)
- Candida tropicalis(yeast)
- soil microorganisms
- wastewater microorganisms
- activated sludge

Immobilization methods:
- entrapping method by glutaraldehyde treatment
- entrapping method by PVA-boric acid
- immobilization by acetylcellulose dialysis membrane absorption method

Three combinations of microorganisms and an immobilization method were selected for further BOD sensor development based on performance of the difference between the peak current and the base current, the transient time to

the peak current and the recovery time to the base current. They were;
a) Trichosporon cutuneum + gelatin-glutaraldehyde film,
b) Trichosporon cutuneum + acetylcellulose film, and
c) soil microorganisms + PVA film

8-2-2 PERFORMANCE OF BOD BIOSENSOR

The three selected biofilms were evaluated. Glucose-glutamic acid (GGA) solution and synthetic wastewater were used as samples. The results are shown in Figure 8.2. Each biofilm had a linear relationship between BOD concentration and the current difference. The most sensible biofilm was of soil microorganisms immobilized by PVA in both GGA solution and synthetic wastewater. The current difference of a GGA solution shows larger than the artificial sewage in each biofilm.

A biofilm of soil microorganisms immobilized by PVA was used for further study in which BOD in several kinds of raw wastewater, secondary effluent and polluted urban river water was measured by the BOD sensor and by a manual method. Calibration curves were prepared with GGA solution and synthetic wastewater. Figure 8.3 shows the relationship between the sensor BOD and manual BOD_5.

In the case of raw wastewater, the sensor BOD showed enough output current. The sensor BOD calibrated by the synthetic wastewater was more agreeable with the manual BOD on the whole than that by GGA.

In cases of secondary effluent and river water, however, the current difference was so small that the sensor BOD showed much lower than the manual BOD. The difference between the two calibration methods seemed very small. Samples which have been biologically treated seem difficult to detect by this BOD sensor in a short period. We have been trying to improve the detection ability for such refractory organic solution.

8-3 AMMONIA SENSOR

8-3-1 PERFORMANCE OF AMMONIA BIOSENSOR

Nitrosomonas europaea was immobilized between two porous acetylcellulose membranes to make a biofilm for an ammonia biosensor. The relationship between ammonia nitrogen concentration below 10 mg/l and a current difference was linear as shown in Figure 8.4. According to the biomass increase, the upper sensible concentration became lower, but the resolution became better. It takes 7 to 8 minutes for the sensoring system to reach the peak current after sample injection. It takes 7 to 12 minutes to return to the base current after stopping the sample injection.

Environmental samples such as river water and effluent from wastewater treatment plants were measured by the sensor system in batch bases. The sensor ammonia agreed well with the manual ammonia. The sensor system was tested as a monitor for effluent from a wastewater treatment plant. The monitoring of ammonia in the effluent was conducted every 15 minutes. A stable agreement between sensor ammonia and manual ammonia was observed during more than 3 weeks. Ammonia nitrogen standard solution with a glucose glutamic acid solution (100mg/l) was sometimes injected into the sensor system to check the interference of other microorganisms. But only the response to the ammo-

nia nitrogen solution was recognized. Therefore, the interference of other microorganisms seemed negligible for this sensor.

8-3-2 APPLICATION OF AMMONIA BIOSENSOR TO TOXIC SENSOR

As a nitrification bacterium is said to be very sensitive to toxic substances, application of the developed ammonia biosensor to a toxic sensor has been expected. Ortho-chlorophenol (OCP) was selected as a toxic material. It was solved in the ammonia nitrogen standard solution (1 mg/l). The response of sensor output was observed when the concentration of OCP was added. The results are shown in Figure 8.5. When the OCP concentration was below 0.5 mg/l, any decrease of the current difference was not observed. The inhibitory effect of OCP to the ammonia sensor was recognized over 0.5 mg/l. The current difference of the electrode decreased with increasing OCP concentration. When the OCP concentration was over 4 mg/l, the current difference became almost zero. After the OCP solution was removed, the current difference recovered to the initial current difference.

This experiment shows that this ammonia sensor has the ability to detect toxic materials. Additional experiments have been carried out to study the detective concentration of other chlorinated organics.

8-4 ORGANIC ACID BIOSENSOR

A yeast of Trichosporon brassicae is used as a detector. The yeast is immobilized between two porous acetylcellulose membrane filters. The biofilm was sandwiched between an oxygen electrode and a gas permeable teflon membrane filter. The yeast utilizes gasified organic acid and consumes oxygen in the biofilm when the microbial electrode is immersed into an organic acid solution.

A linear relationship was observed between an organic acid concentration below 600 mg/l and current difference as shown in Figure 8.6. The current difference decreased with the ascent of pH from 3 in the sample and fell to zero at a pH of 7 or more due to ionization of acetic acid. It shows pH in a sample must be kept lower than 3.

The sensitivities of other organic acid samples were tested. The sensor could detect formic acid, n-butyric acid, propionic acid as well as acetic acid. However lactic acid, tartaric acid, succinic acid could not be detected. That shows it is difficult to convert the sensor value to the actual acetic acid equivalent value in solutions containing a mixture of organic acids.

A field test in an actual digestion tank has been carried out for the check of maintenance .

9. DEVELOPMENT OF NEW WASTEWATER TREATMENT SYSTEMS

Bioreactors for wastewater and sludge treatment, liquid-solid separators and biosensors are being developed in addition to the basic studies on the microorganism bank and genetic engineering in Biofocus WT, as mentioned above. Systematizing the individual development to propose new wastewater treatment systems is being investigated. Table 9.1 summarizes new wastewater treatment systems to be proposed based on results of pilot plant studies. Table 9.2 is the list of pilot plants in Biofocus WT.

Table 9.1 New Wastewater Treatment Systems to be Proposed in Biofocus W.T.

a. Energy-saving Type Wastewater Treatment System
b. Area-saving Type Wastewater Treatment System
c. Nitrogen Removal Type Wastewater Treatment System
d. Simultaneous Nitrogen & Phosphorus Removal Type Wastewater Treatment System
e. Wastewater Treatment System for Small Flow

A self-sustaining energy system has been proposed as an ultimate system of an energy-saving type wastewater treatment system. The system is illustrated in Figure 9.1. Most SS in raw wastewater are removed by a high-efficiency liquid solid separator. Energy consumption in the secondary treatment can be greatly reduced by two means. One has influent to the secondary treatment containing not only less organics, but less particle organics which are more difficult to decompose than soluble ones. The other is adopting an energy-saving type bioreactor.

In sludge digestion, the feed sludge has a large amount of primary sludge which has higher potential for generating digestion gas than that of excess sludge. Also, an energy-creating type bioreactor for sludge digestion is employed, so that a greater amount of energy can be recovered. A key point of the system development is how energy consumption of a high-efficiency liquid solid separator can be reduced.

10. POSTSCRIPT

As Biofocus W.T. is a research project which intends to broadly apply biotechnology to biological wastewater treatment for its improvement and greater understanding, it includes various research subjects from very basic ones to practical ones, as mentioned above. Therefore, results of Biofocus W.T. will be varied, from those of identifying further research needs to practically feasible technologies which may break through present technologies.

Table 9.2 List of Pilot Plant in Biofocus WT

RESPONSIBLE ORGANIZATION	OUTLINE OF PILOT PLANT	LOCATION (average flow, d/day)	MAIN DEVELOPMENT TARGET
Aerobic Bioreactor for Organic Removal			
Asano Engineering CO., LTD.	FLUIDIZED BED REACTOR WITH ATTACHMENT MEDIA	Biofocus Yard (38)	to reduce retention time to between one second and one tenth of that of conventional activated sludge process
Ebara Infilco CO., LTD.	AEROBIC UPFLOW SLUDGE BLANKET PROCESS	Biofocus Yard (20)	
Komatsu, LTD.	ROTATING BIODISK-TYPE REACTOR WITH ENTRAPPED MICROORGANISMS	Shinomiya WTP, Kanagawa Pref.(1)	
Sanki Engineering CO., LTD.	MOVING BED REACTOR WITH ATTACHING MEDIA	Chubu WTP, Yamato Ciy (50)	
Nippon Steel Corporation	FLUIDIZED BED & FIXED BED REACTORS WITH MEDIA OF SLAG	Kougasaki WTP, Kitakyushu City (22.4, 6)	
Takenaka Komuten CO., LTD.	FLUIDIZED BED REACTOR WITH POLYACRYLAMIDE GEL	Biofocus Yard (3)	
Chiyoda Chemical Engineering & Construction CO., LTD.	FIXED BED REACTOR & FILTRATION	Biofocus Yard (35)	
Toshiba Corporation	FLUIDIZED BED REACTOR WITH ULTRA FILTRATION	Kase WTP, Kawasaki Ciy (0.7)	
Nishihara Environmental Sanitation Research Corporation	FILTRATION & FIXED BED REACTOR	Biofocus Yard (35)	
Nippon Kokan K.K.	FLUIDIZED BED REACTOR WITH PLASTICS ATTACHMENT MEDDIA & FILTRATION	Biofocus Yard (24)	
Mitsui Construction CO., LTD.	FLUIDIZED BED REACTOR WITH ACTIVATED CARBON	Biofocus Yard (40)	
Hazama-Gumi, LTD.	FIXED BED OXYGEN REACTOR WITH ATTACHMENT MEDIA	Laboratory in Hazama-Gumi, LTD. (1)	
Anaerobic Bioreactor for Organic Removal			
Ishikawajima-Harima Heavy Industries CO., LTD.	UASB WITH PERMIATING AERO-FILTER	Biofocus Yard (16)	· to reduce energy reqirment to one second of that of conventional activated sludge process · to reduce sludge generation
Kajima Corporation	ANAEROBIC & AEROBIC FIXED BED REACTORS WITH MOOL-TYPE ATTACHING MEDIA	Biofocus Yard (20)	
Kubota, LTD.	HYBRID REACTOR WITH UASB & FIXED BED FOR CONDENSED WASTEWATER TREATMENT	Sakaino WTP, Kiryu City (10)	
Kurita Water Industries LTD.	FLUIDIZED BED REACTOR WITH PORUS MEDIA & H_2S REMOVAL BY VACCUMN	Biofocus Yard (40)	
Fuji Electric Co., LTD.	ANAEROBIC & AEROBIC FIXED BED REACTOR WITH NONWOVEN FABLIC MEDIA	Biofocus Yard (30)	
Meidensha Corporation	FIXED BED REACTOR WITH PORUS CERAMICS AND AEROBIC BATCH REACTOR	Totsuka-Daiiti WTP, Yokohama City (3)	
Bioreactor for Nutrient Removal			· to reduce reten. time for N removal to 6 to 12 hours so that existing WTP can adopt
Kyowa Kako CO., LTD.	FLUIDIZED BED REACTOR WITH PORUS MEDIA	Tsurukawa WTP, Machida City (10)	
Kobe Steel, LTD.	FLUIDIZED BED REACTOR WITH SAND & INTERMITTENT AERATION	Tarumi WTP, Kobe City (10)	
Hitachi Plant Engineering & Construction CO., LTD.	FLUIDIZED BED REACTOR WITH ENTRAPPED NITRIFIERS	Kitano WTP, Hachioji City (50)	
Bioreactor for Anaerobic Sludge Digestion			· to upgrade digestion persentage by 50 % · to reduce retention time to 5 days
Tsukishima Kikai CO., LTD.	FLUIDIZED BED REACTOR WITH CEMENT MADE ATTACHMENT MEDIA	Ichinokawa WTP, Higashimatsuyama City (0.5)	
NGK Insulators, LTD.	THERMOPHILIC TWO PHASE DIGESTION WITH CERAMICS ATTACHMENT MEDIA	Ohno WTP, Osaka City (1.6, 0.8)	
Fujita Corporation	DIGESTION WITH PHISICAL-CHEMICAL SOLUBOLIZATION OF SLUDGE	Chubu WTP, Yokohama City (1)	
Solids/Liquid Separator			· to reduce SS loading flowing to bioreactor to one third of that of pri. sed. effluent
Shimizu Construction CO., LTD.	FLOATING-BED FILTRATION SYSTEM WITH MULTIPLE SUPPLY	Biofocus Yard (50)	
Dic-Degremont CO., LTD.	HIGH RATE BIOFLOCCULATION PROCESS WITH LAMELLA SETTLER	Fujimi Relay Pumping Station, Saitama Pref. (100)	
Maezawa Industries, INC.	TWO-STAGE FILTRATION WITH MOVING BED & FIXED BED	Biofocus Yard (150)	
Bioreactor for Oaganic Removal			· to reduce ret. time & energy requirment
The Sewage Treatment Plant Constructors Association of Japan	MULTI-STAGE REVERSING-FLOW BIOREACTOR	Biofocus Yard (225)	

Biofocus yard is located in Kasumigaura wastewater treatment plant, Ibaraki prefecture.

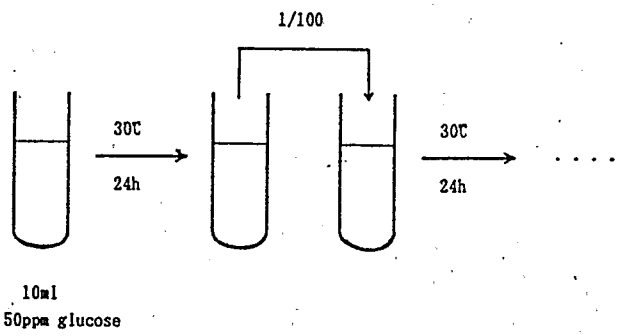


Figure 3.1 Diagram of Serial Transfer Culture

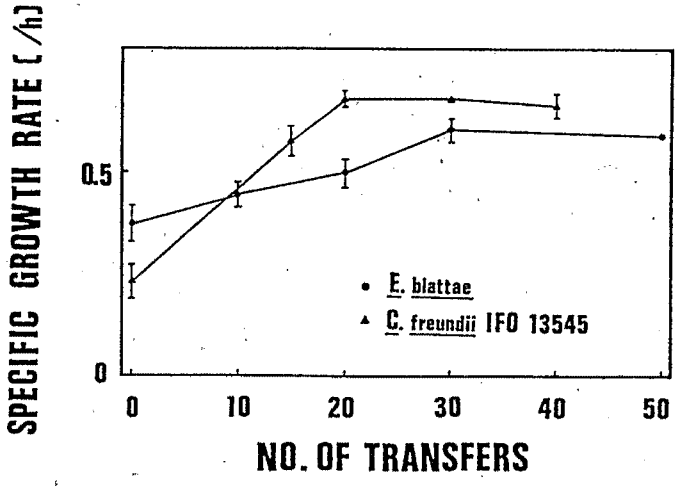


Figure 3.2 Changes of specific growth rates of E. blattae and C. freundii IFO13545 isolated from the serial transfer cultures (i.e., from their single species cultures, not a mixed culture)

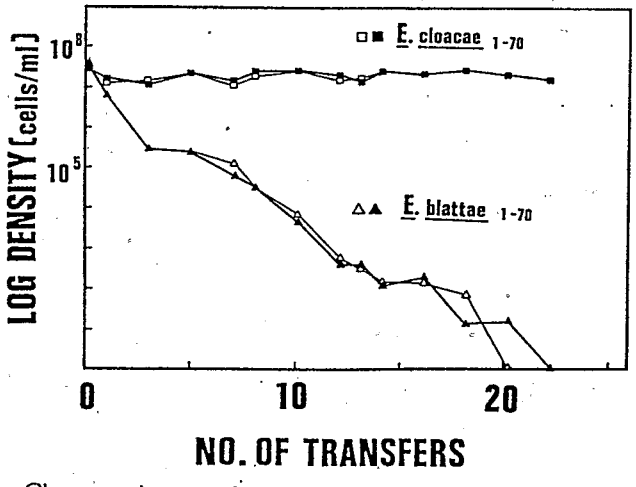


Figure 3.3 Changes in population densities of E. cloacae and E. blattae in their mixed culture. Both were isolated from the serial transfer cultures of 70 times transfer

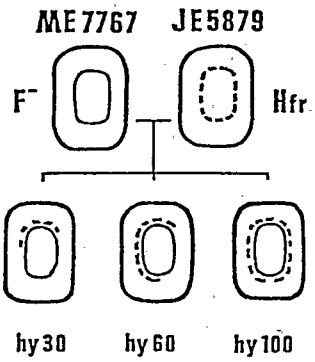


Figure 3.4 Schematic Diagram of the Relationship between Parental Strains and Recombinant Strains

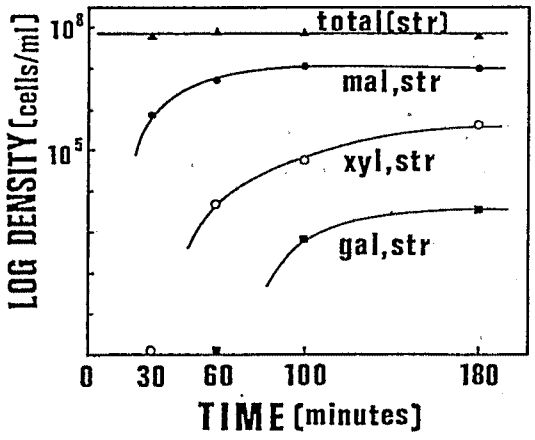


Figure 3.5 Changes in Density of Recombinant Cells in Mixed Cultures

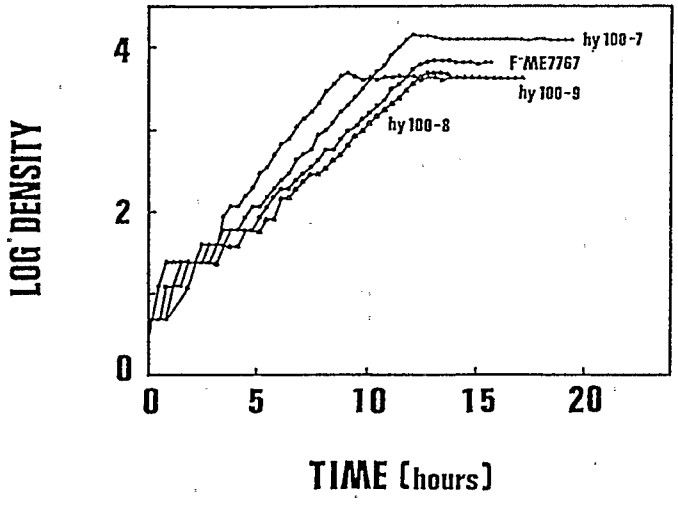
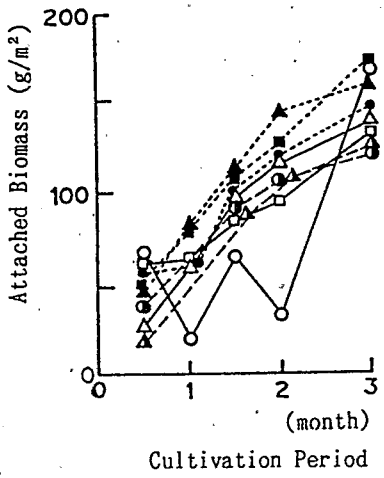


Figure 3.6 Growth Curves of ME7767, hy100-7, -8, and hy100-9



Material	Surface Characteristics
◑Stainless Steel	smooth
●Vinylchloride (a)	smooth
▲Vinylchloride (b)	roughed with fine sand paper
■Vinylchloride (c)	roughed with coarse sand paper
○Ceramics (a)	pore size 0.1～0.2mm , 1～4 mm
△Ceramics (b)	pore size 0.01～0.1mm
□Ceramics (c)	pore size 0.045mm
⚠Polyethylene	smooth

Figure 4.1 Comparison of Growth of Biomass Attached on Media

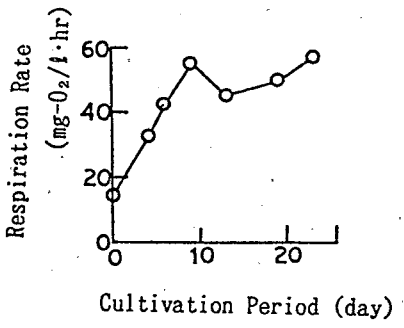


Figure 4.2 Transition of Respiration Rate of Microorganisms Immobilized by PVA-Boric Acid Method

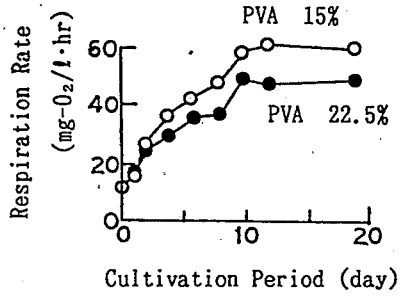


Figure 4.3 Transition of Respiration Rate of Microorganisms Immobilized by PVA-Refrigeration Method

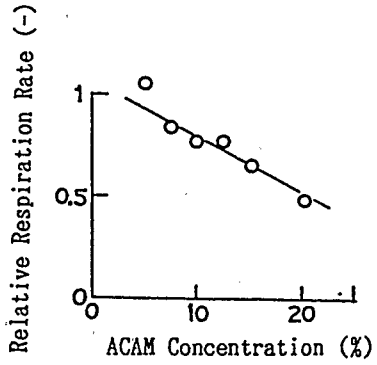


Figure 4.4 Effect of ACAM Concentration on Respiration Rate of Microorganisms Immobilized by Agar-ACAM Monomer Method

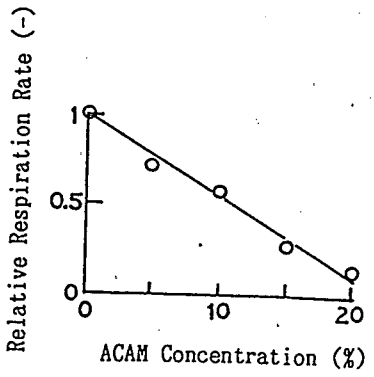


Figure 4.5 Effect of ACAM Concentration on Respiration Rate of Microorganisms Immobilized by ACAM Monomer Method

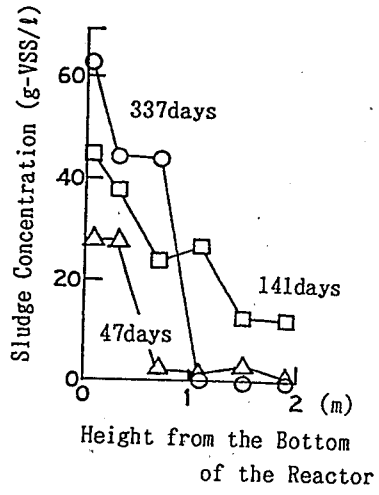


Figure 4.6 Distribution of Sludge Concentration in UASB Reactor for Treatment of Supernatant from Heat Treatment of Sludge

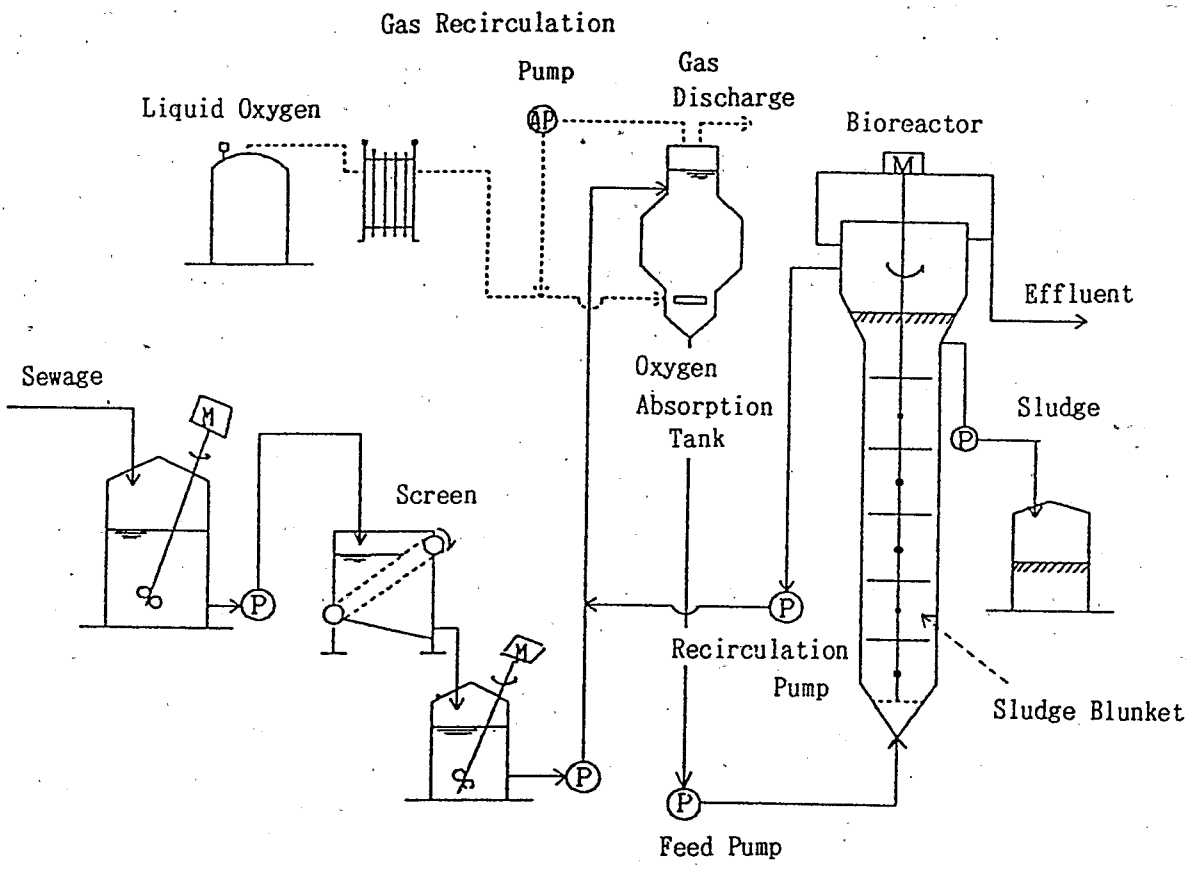


Figure 4.7 Flowsheet of AUSB Process

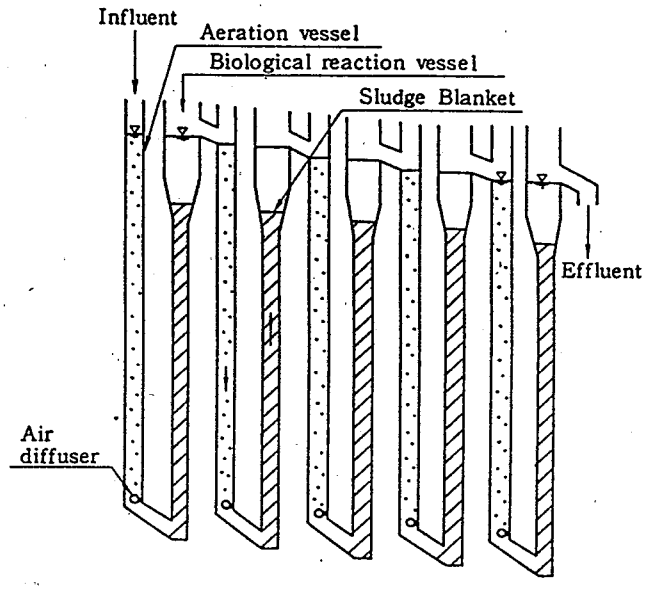


Figure 4.8 Flowsheet of MRB

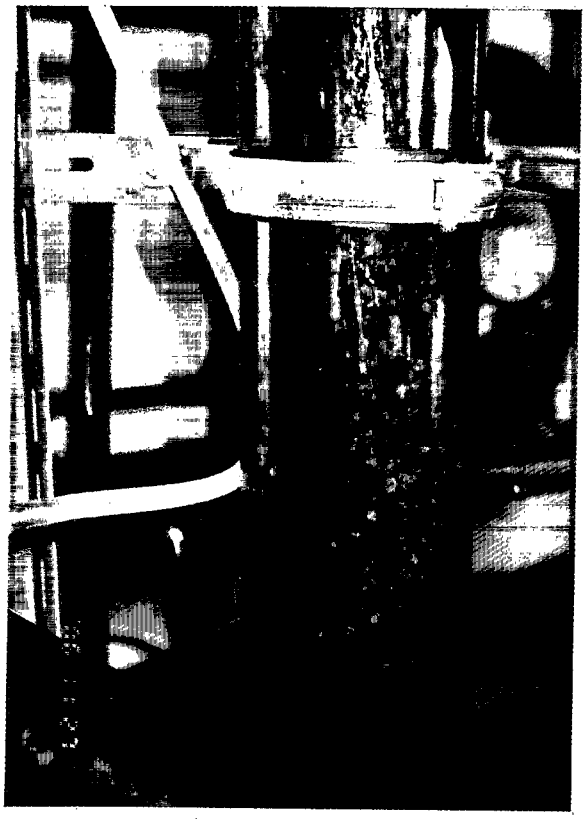


Figure 4.9 Self-Granulated Sludge (SGS)

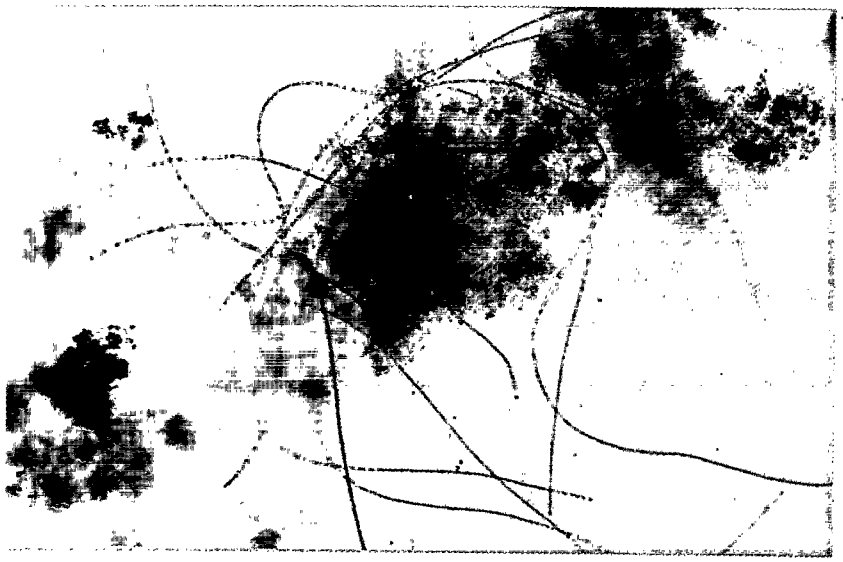


Figure 4.10 Microscopic View of Film

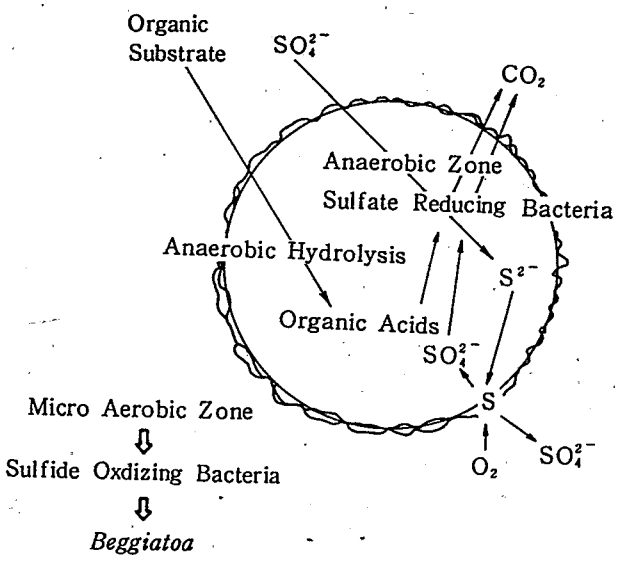


Figure 4.11 Mechanism of Self-Granulation

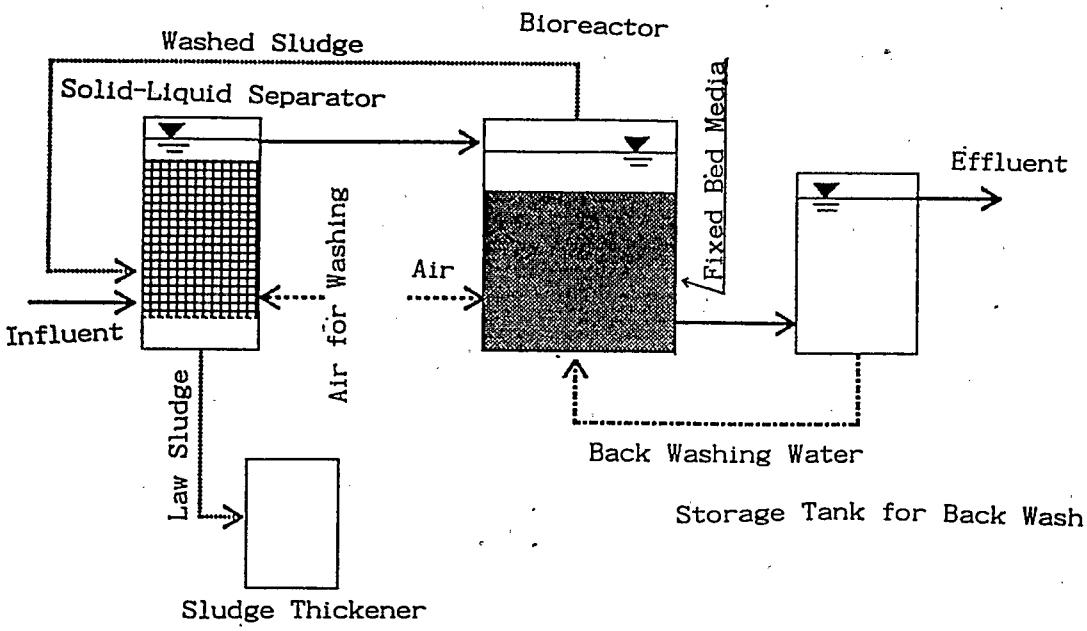


Figure 5.1 Flow Diagram of Fixed Bed Type Aerobic Bioreactor

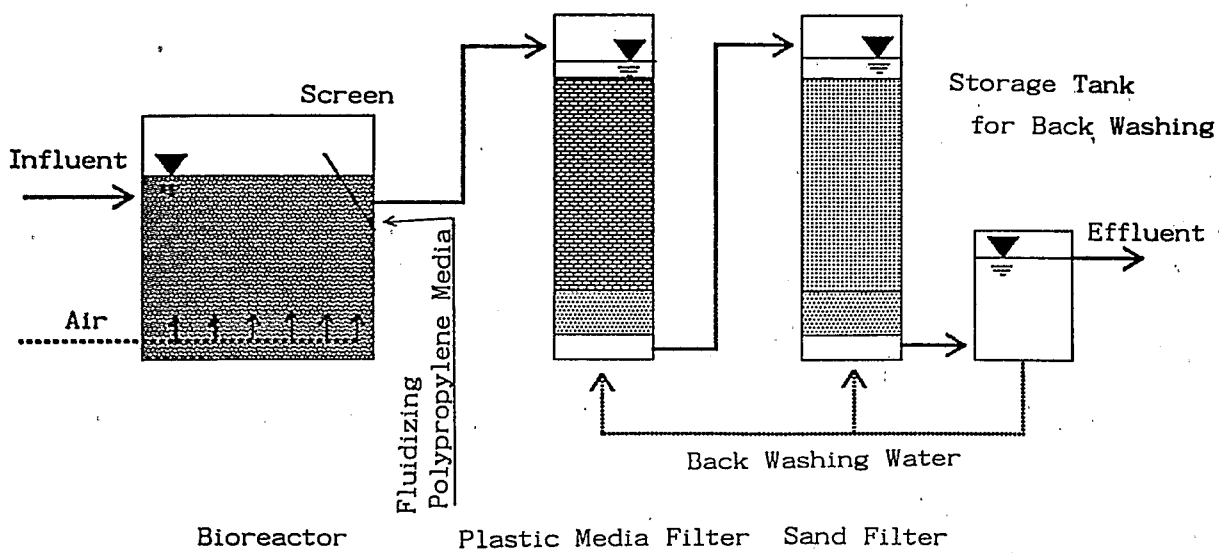


Figure 5.2 Flow Diagram of Fluidized Bed Type Aerobic Bioreactor

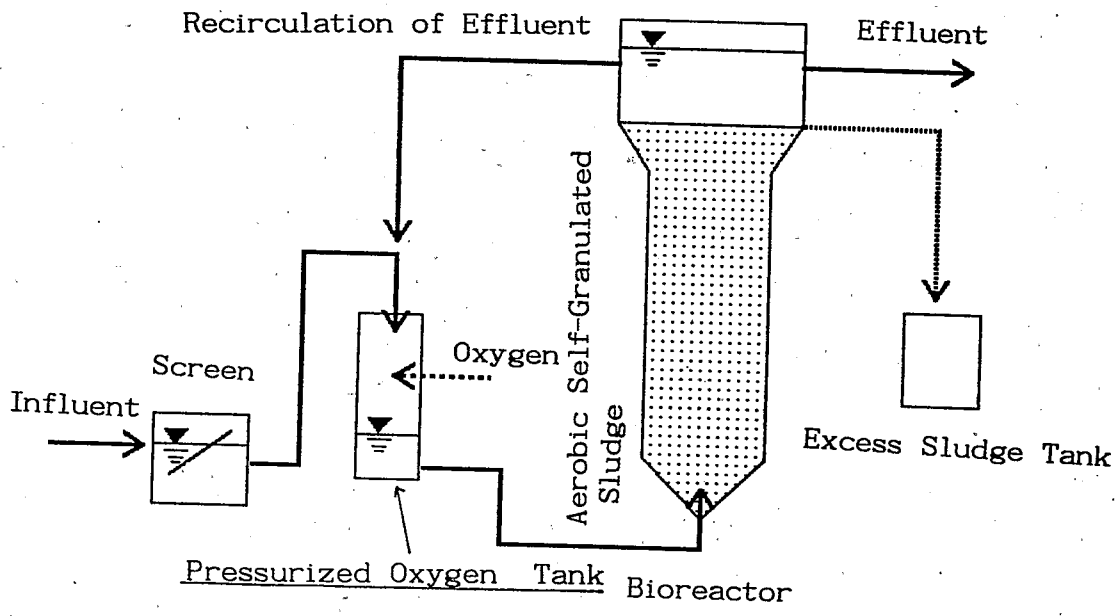


Figure 5.3 Flow Diagram of Aerobic Sludge Blanket Bioreactor

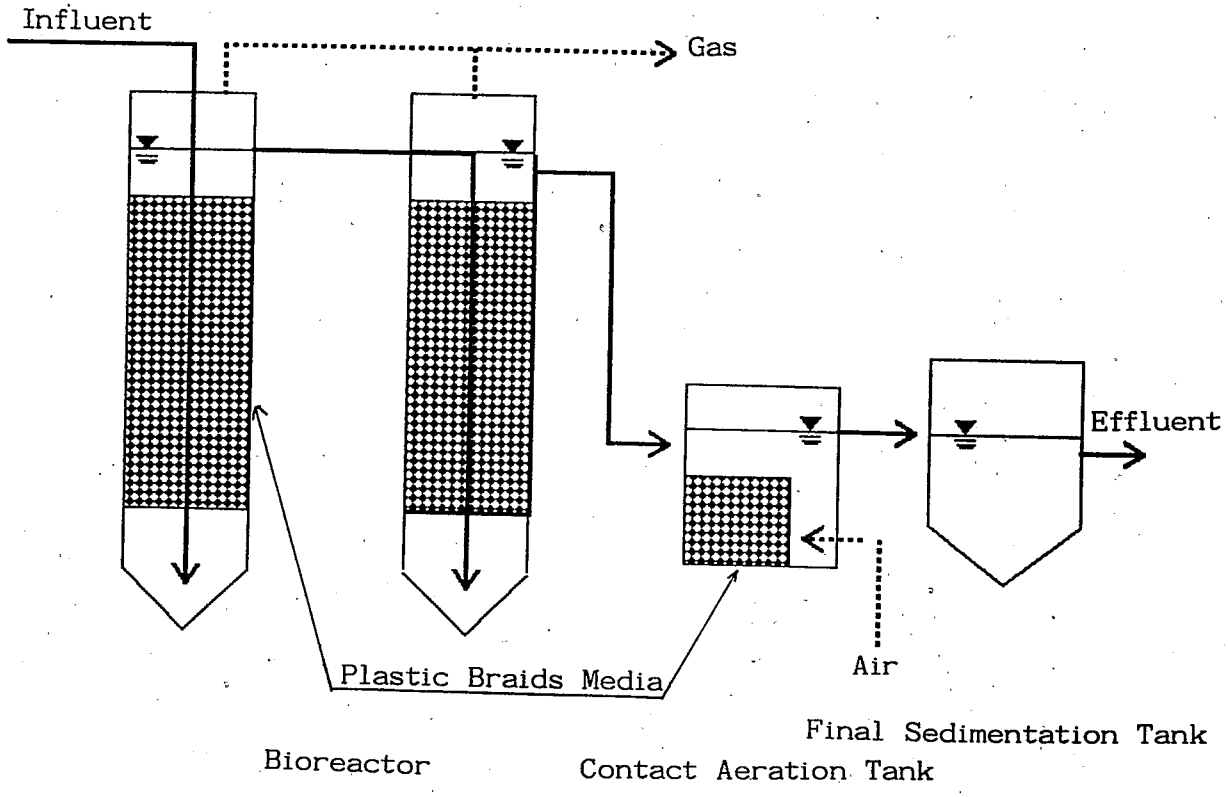


Figure 5.4 Flow Diagram of Fixed Bed Type Anaerobic Bioreactor

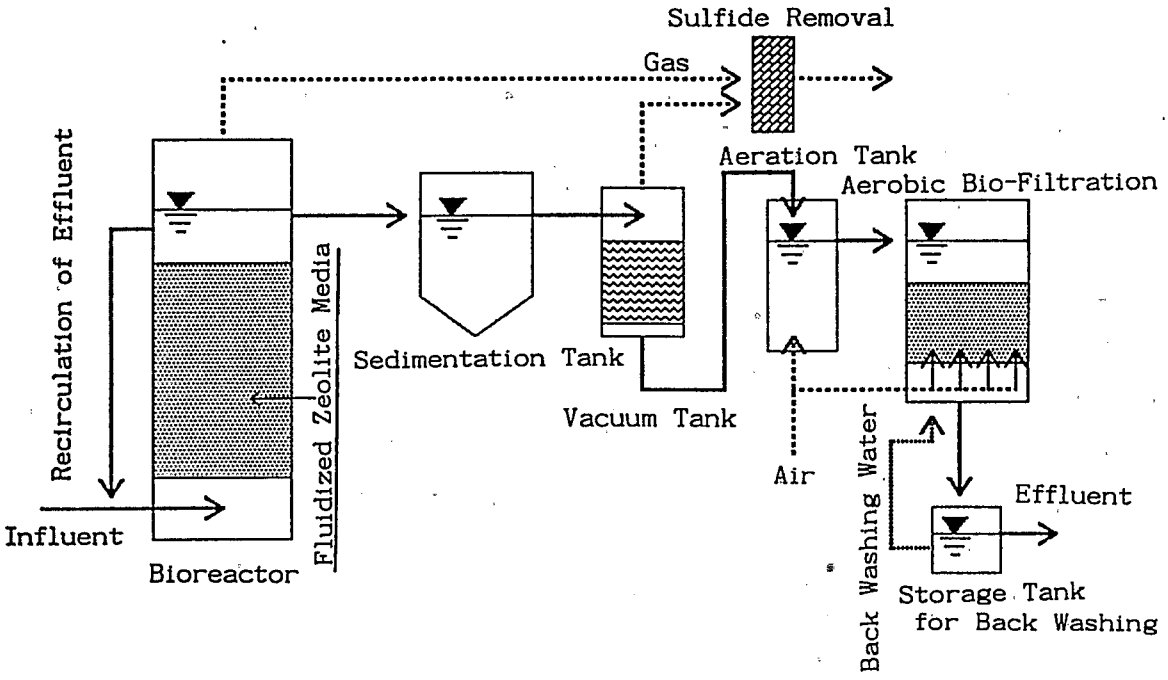


Figure 5.5 Flow Diagram of Fluidized Bed Type Aerobic Bioreactor

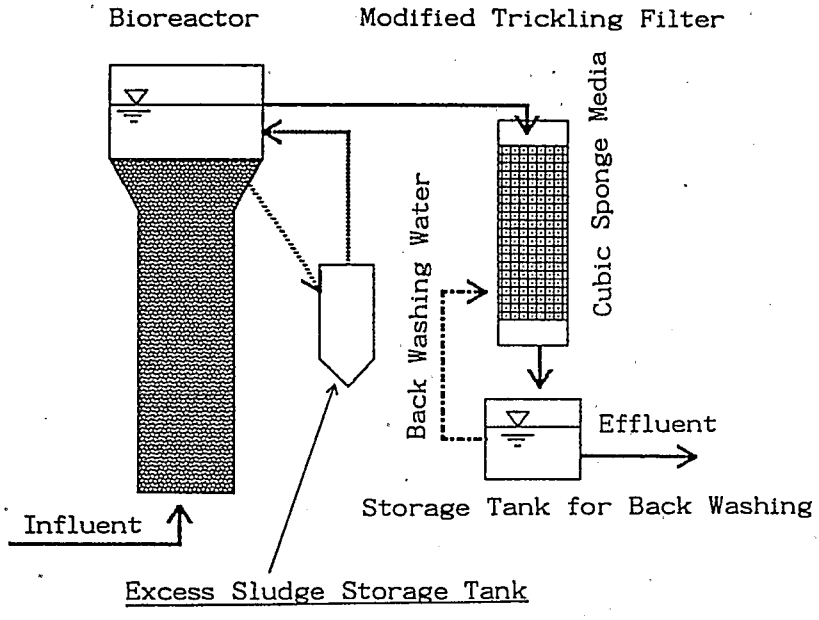


Figure 5.6 Flow Diagram of UASB with Modified Trickeling Filter

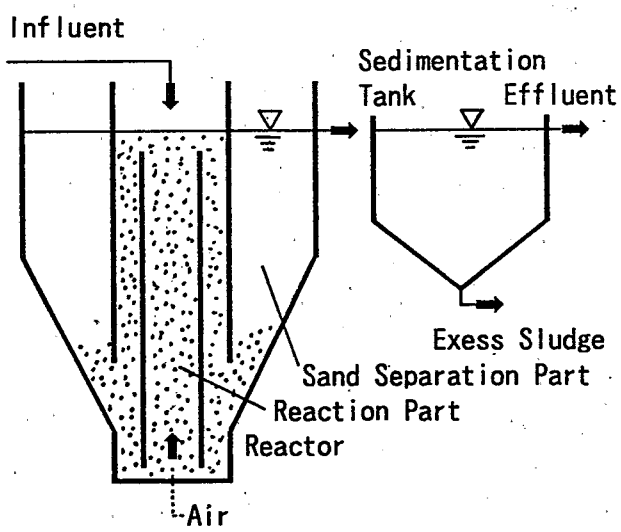


Figure 5.7 Flowsheet of Pilot Plant

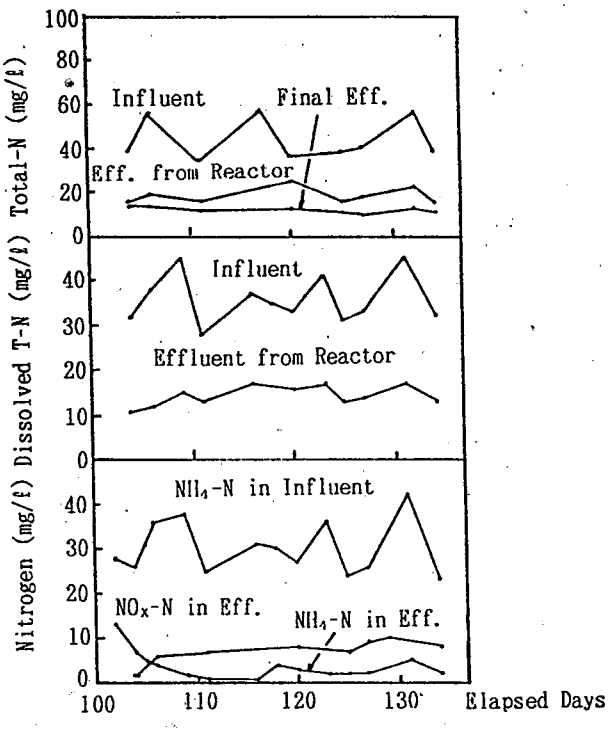


Figure 5.8 Nitrogen Removal

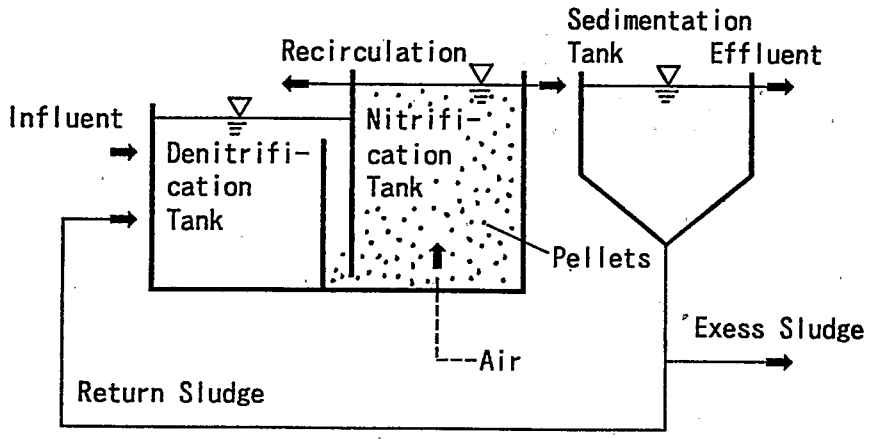


Figure 5.9 Flowsheet of Pilot Plant

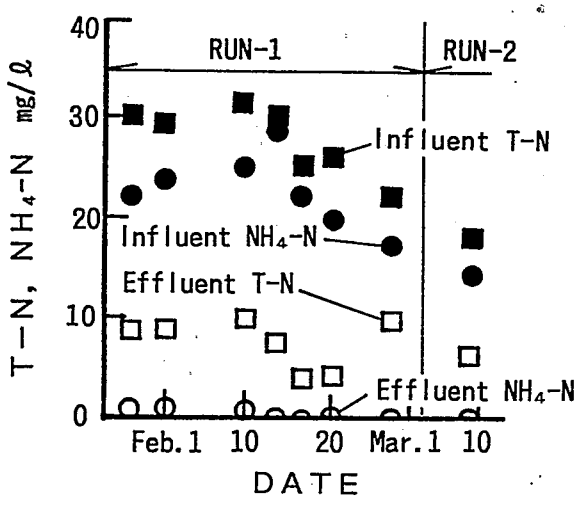


Figure 5.10 Performance of Nitrogen Removal

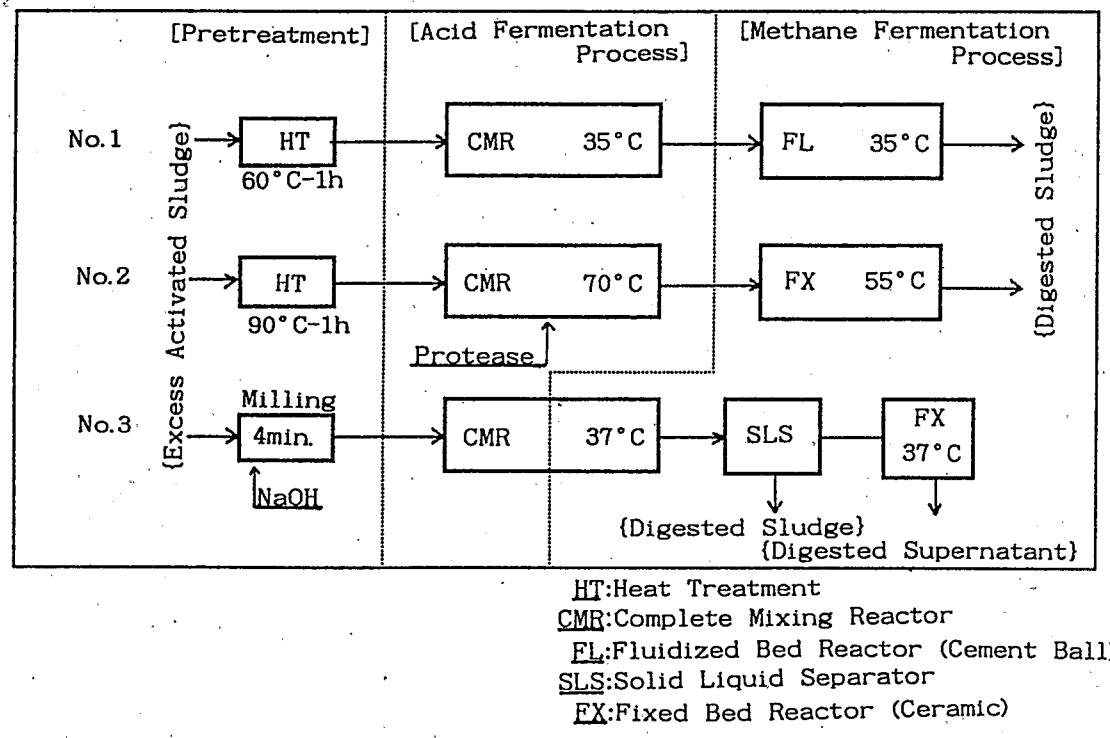


Figure 6.1 Flowsheets of Pilot Plants on Anaerobic Sludge Digestion

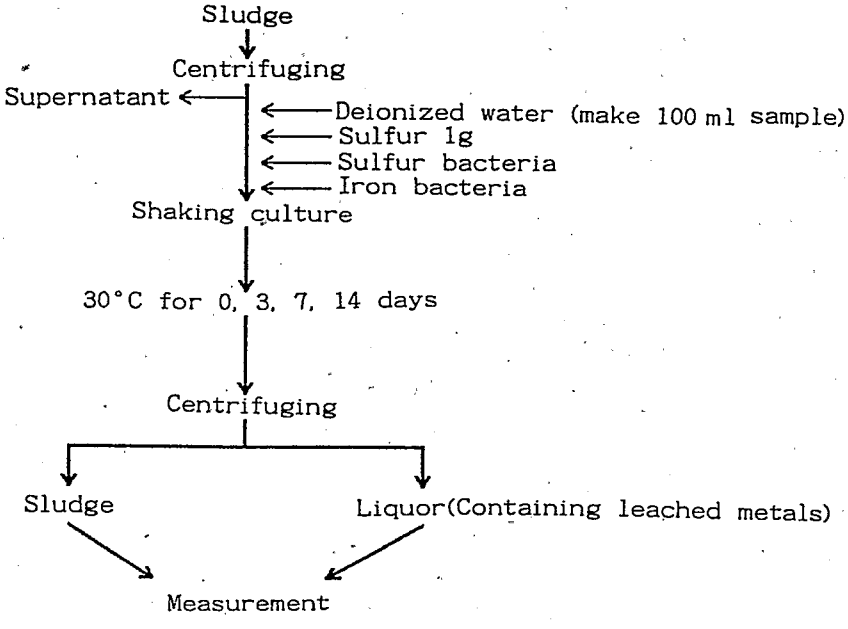


Figure 6.2 Procedure of Bacteria Leaching of Sludge

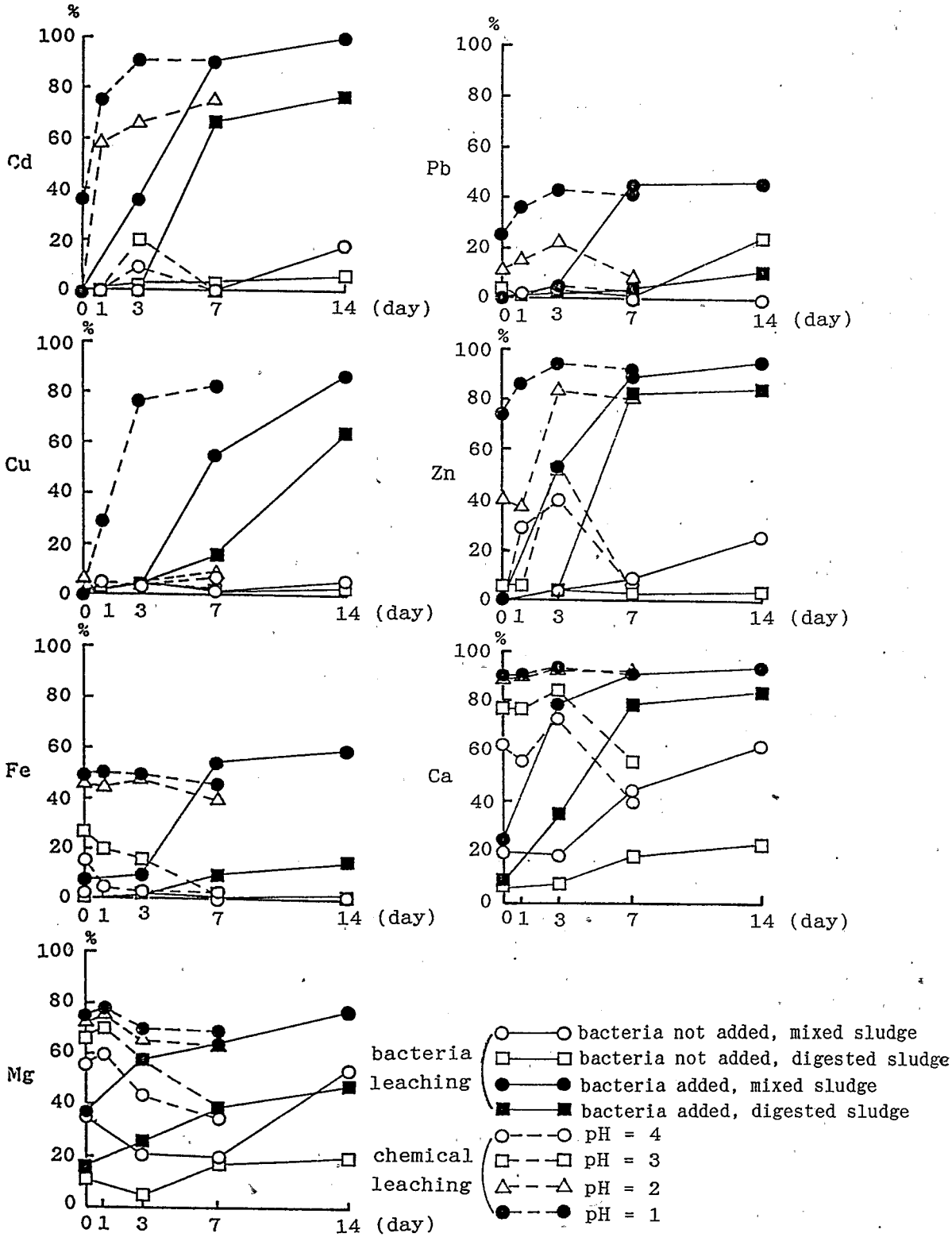


Figure 6.3 Changes of Leaching Ratios of Heavy Metals

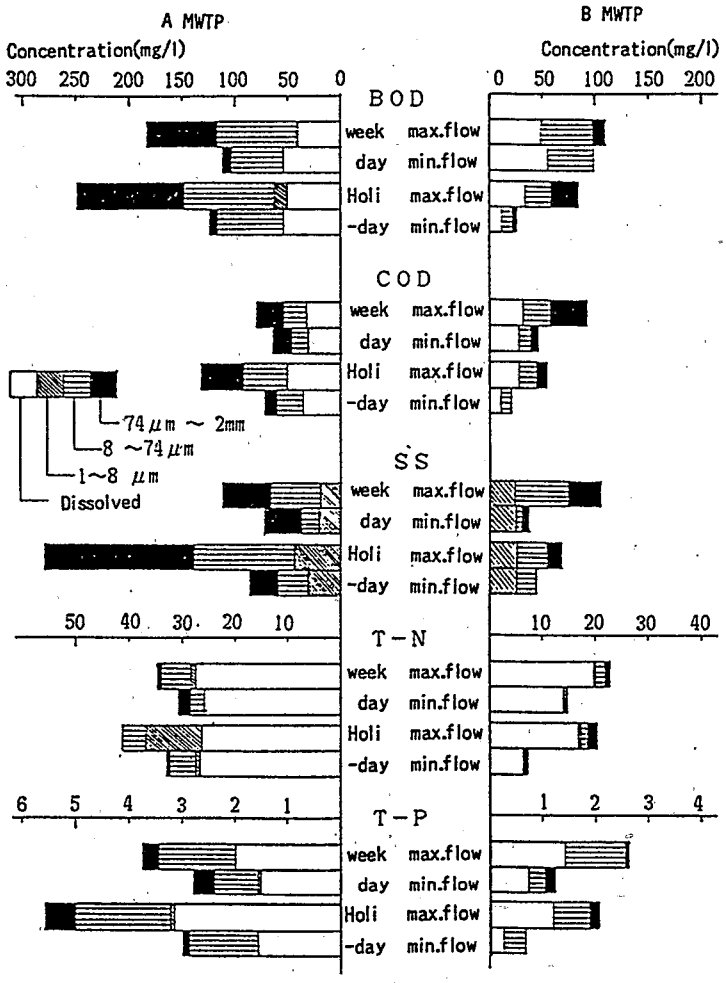


Figure 7.1 Distribution of Particulate Contaminants in Raw Wastewater

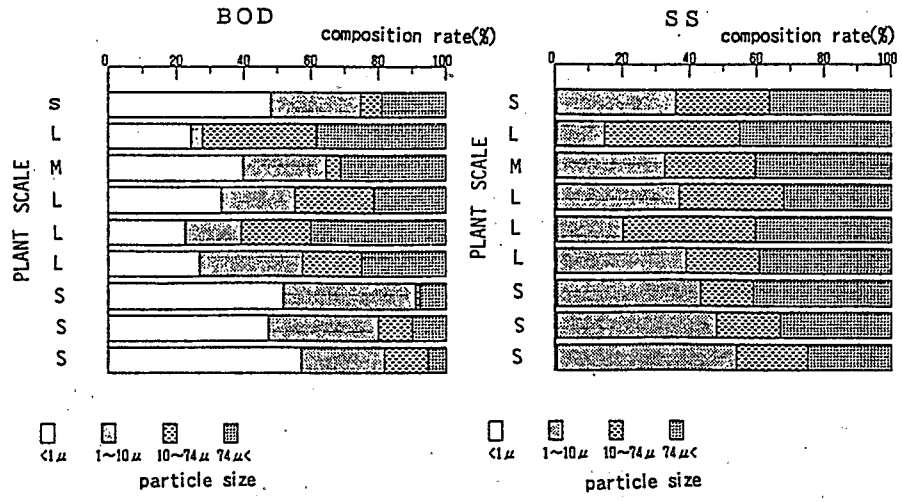


Figure 7.2 Comparison of Distribution of SS and Particulate BOD in Raw Wastewater by Plant Size

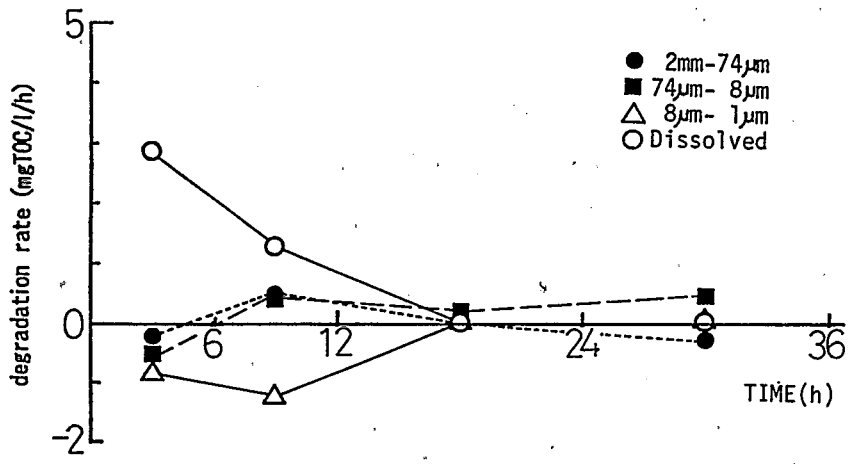


Figure 7.3 Difference of Biodegradation Rates of TOC due to Particle Size of Contaminant in Aerobic Condition

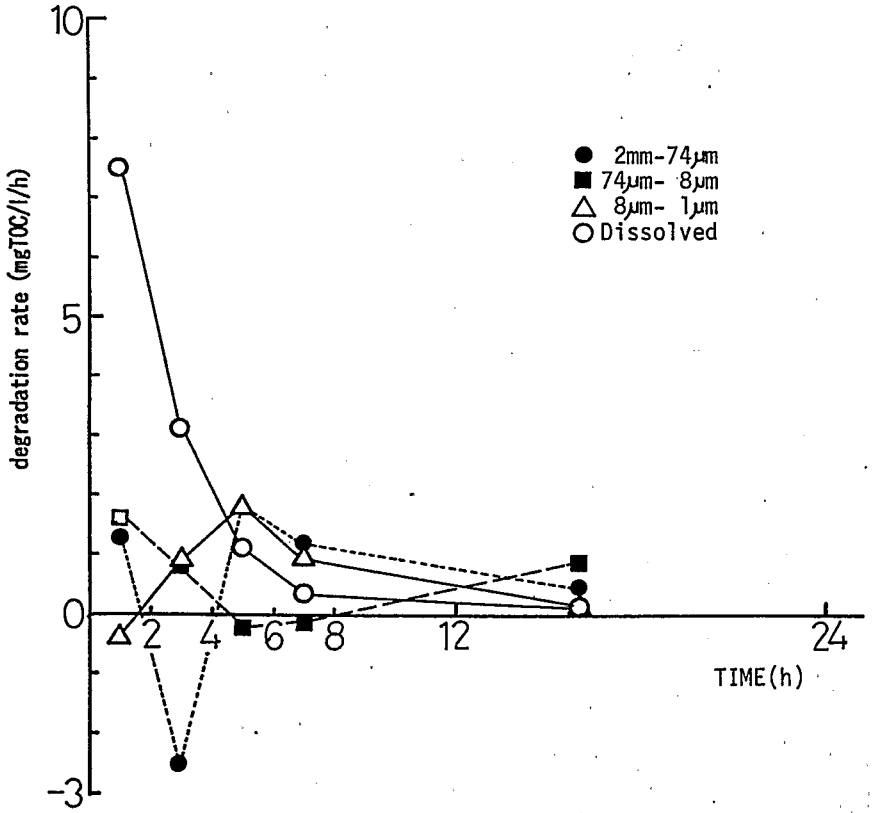
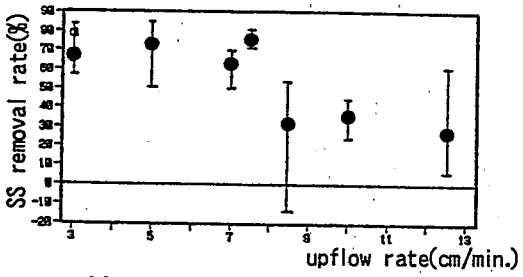
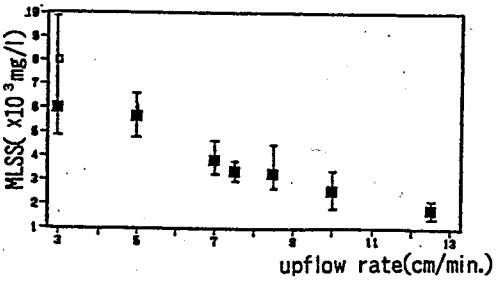


Figure 7.4 Difference of Biodegradation Rates of TOC due to Particle Size of Contaminant in Anaerobic Condition



(1) SS removal rate and upflow rate



(2) SS concentration in sludge blanket and upflow rate

Figure 7.5 Performance of Upflow Sedimentation Column

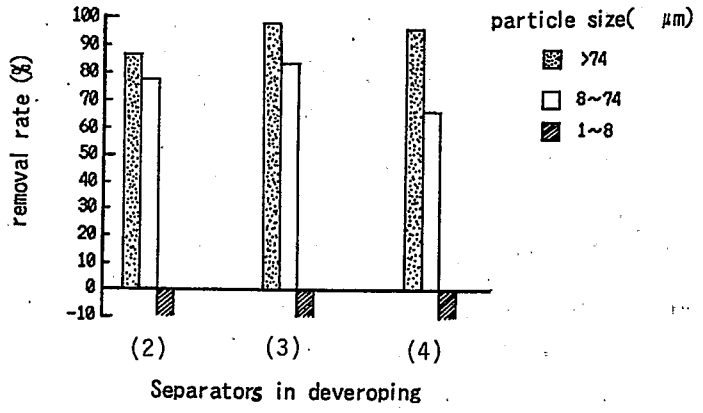


Figure 7.6 SS Removal by Three Solid-Liquid Separation Processes

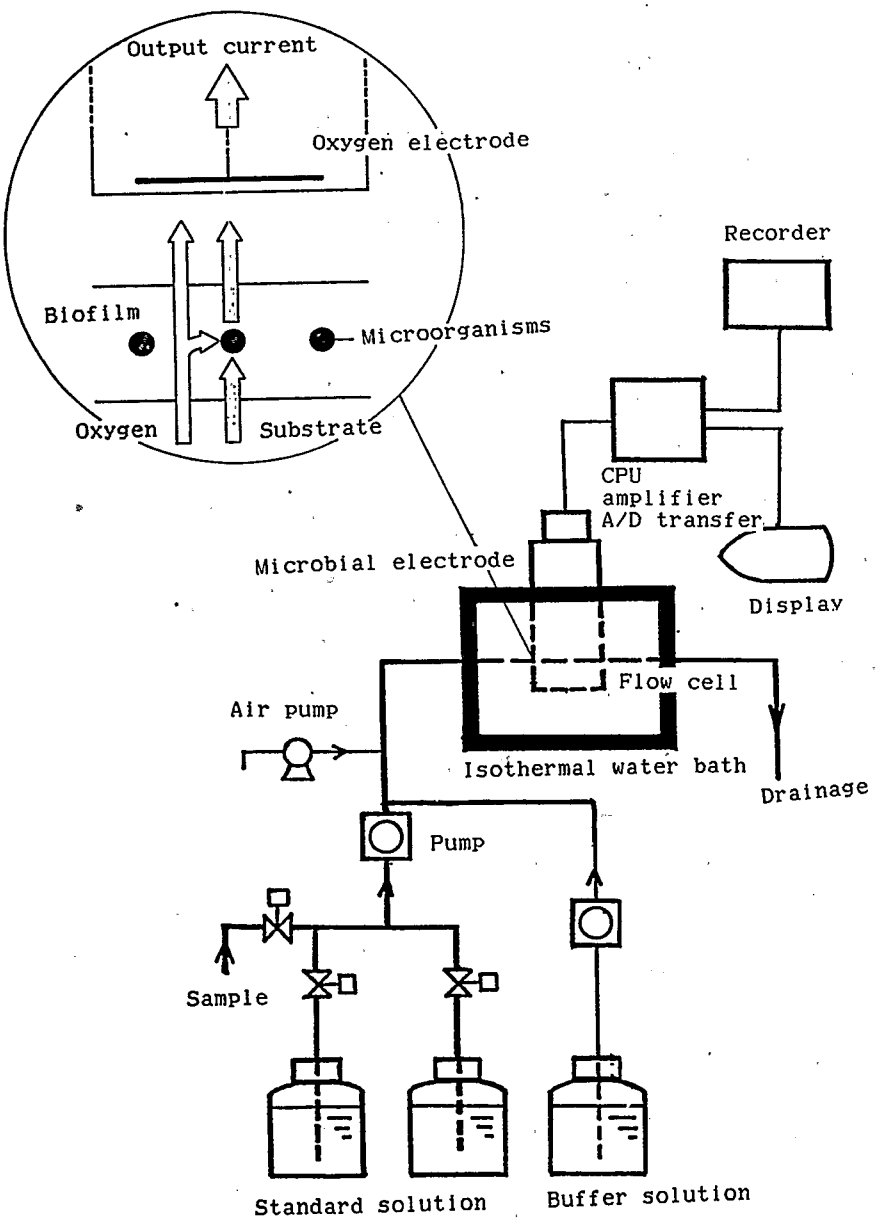


Figure 8.1 Flow Chart of Biosensoring System

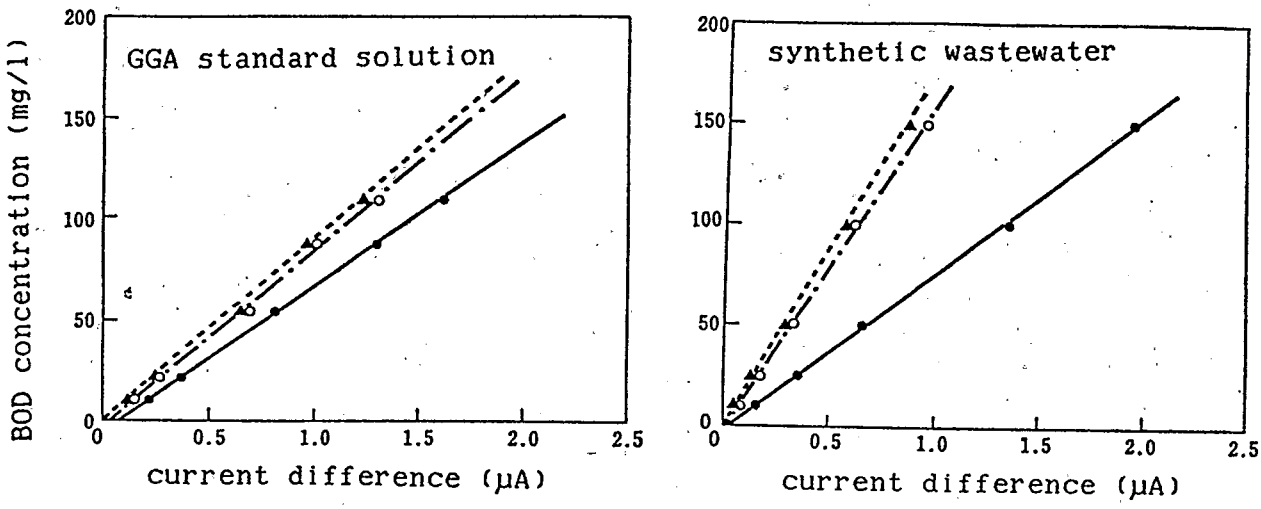


Figure 8.2 Performance of Three Selected Biofilms

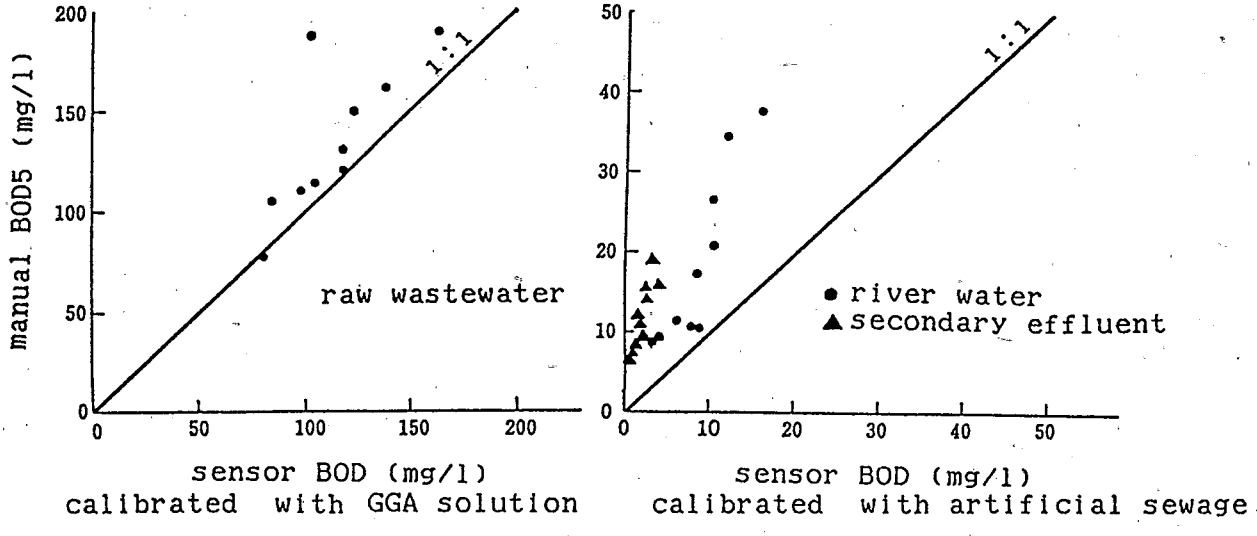


Figure 8.3 Relationship between Sensor BOD and Manual BOD

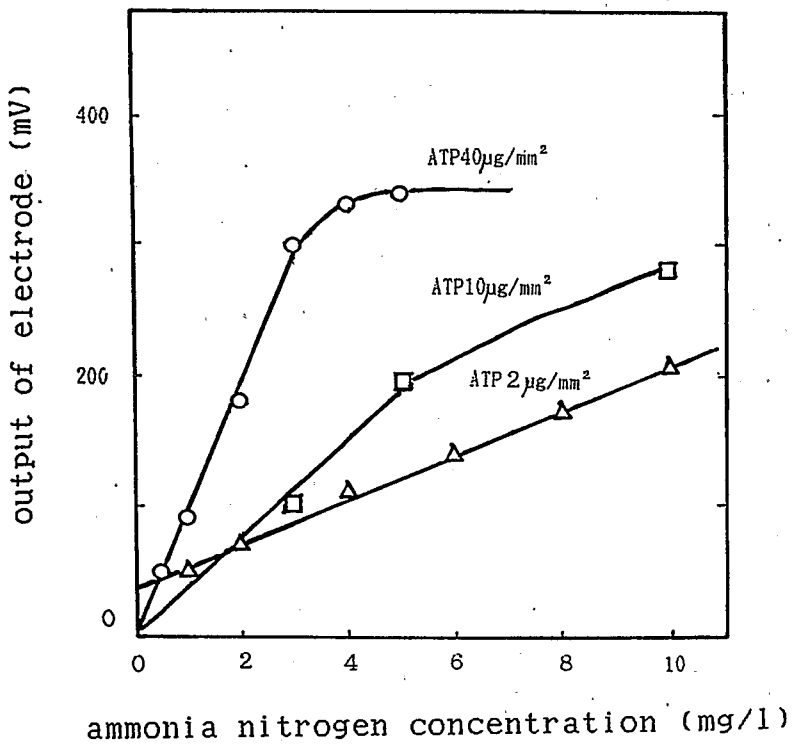


Figure 8.4 Relationship between Sensor Ammonia and Manual Ammonia

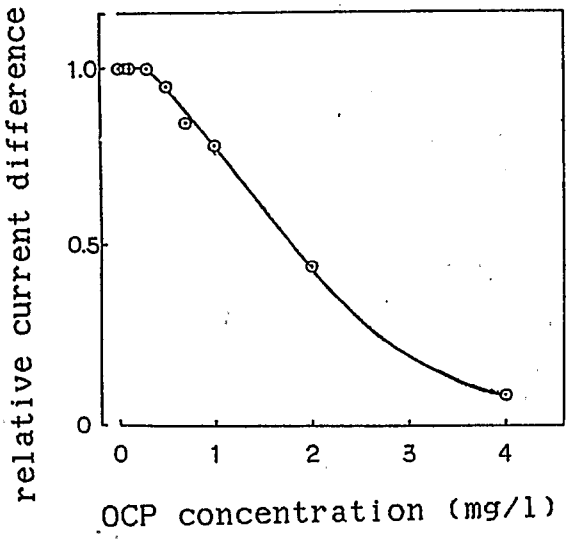


Figure 8.5 OCP concentration and Current Difference of Ammonia Sensor

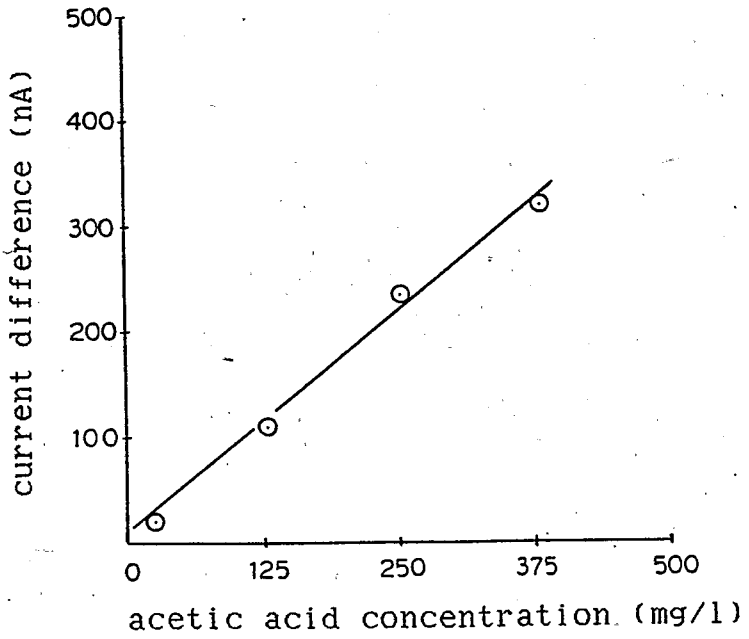


Figure 8.6 Calibration Curve of Organic Acid Sensor in Acetic Acid Standard Solution

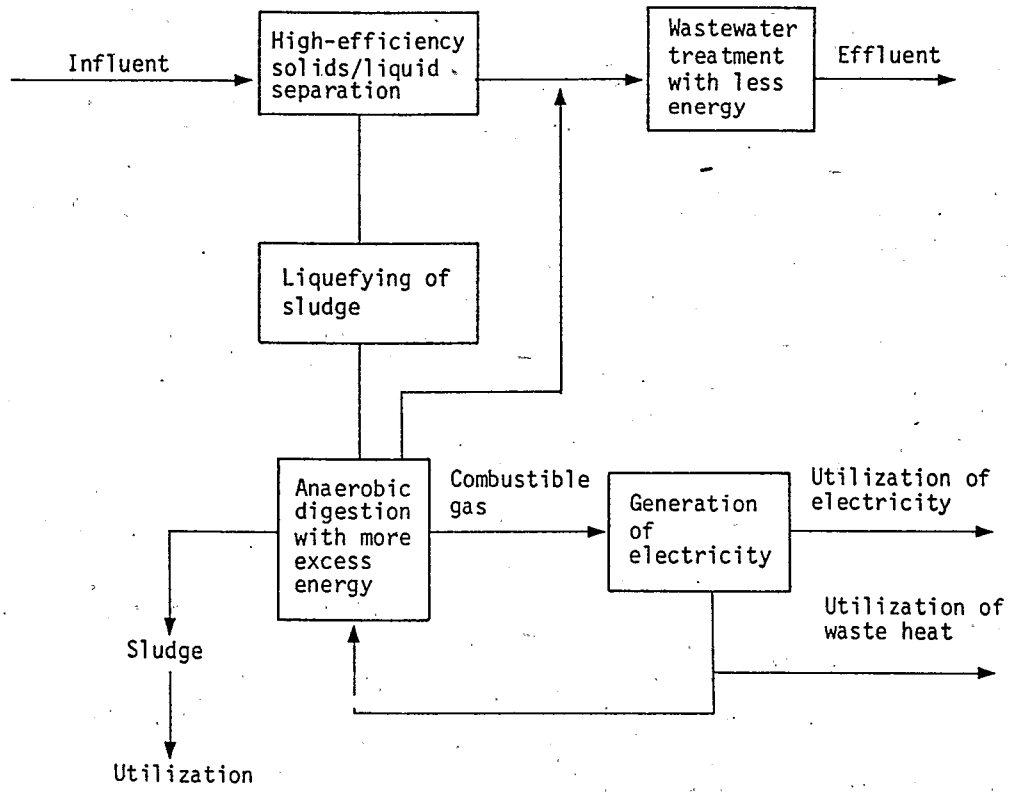


Figure 9.1 Self-energy Sustaining Wastewater Treatment System

FUNDAMENTAL STUDY ON APPLICATION OF HIGH GRADIENT MAGNETIC SEPARATION TO MUNICIPAL WASTEWATER TREATMENT SYSTEM

by

Kazuo HOYA, Director of Engineering Division, Regional Sewerage Center
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Sewerage Bureau
Tokyo Metropolitan Government

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Sewage Treatment Technology

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Abstract

A research was carried out on the development of small land required treatment process which is applicable for both combined sewer overflows treatment in wet weather and tertiary treatment of secondary effluent in dry weather using high gradient magnetic separator. Experiments with the bench scale and pilot scale apparatuses of high gradient magnetic separation process exhibited that the pollutants were removed sufficiently for both combined overflow treatment and tertiary treatment. The results of the experiments suggested that an optimum flow velocity of the magnetic separator was about 2000 to 3000 m/day and appropriate magnetite concentration was about 25 mg/l for combined sewer overflow treatment and about 5 mg/l for tertiary treatment. Preseparation of large flocs by high rate sedimentation prior to magnetic separation was considered to be effective measure to lengthen the filter run of the magnetic separator for combined sewer overflows treatment. Optimum overflow rate of preseparation was about 800 to 900 $m^3/m^2 \cdot day$. Sludge produced from the treatment process could be thickened and dewatered easily. Preliminary design was carried out for a demonstration size plant with a 10,000 m^3 /day treatment flow and full scale plant with a 100,000 m^3 /day treatment flow.

1. INTRODUCTION

In recent years, several kinds of technological improvements has been required for wastewater treatment plant in Tokyo Metropolis. One is the improvement of effluent water quality at storm weather. Collection system in the ward area of Tokyo Metropolis is combined type, hence large volume of wastewater is discharged from treatment plants at storm weather after only primary treatment. This causes the pollution of public water body at storm weather. The other is the introduction of advanced treatments for the prevention of eutrophication and for the reuse of effluent for aesthetic, recreastional and subpotable water supply purposes. Removal of phosphorus, BOD, COD, turbidity, chromaticity and coliform bacteria is especially important for these purposes. However, treatment plants in Tokyo do not have enough space to install many new facilities, hence developement of a treatment technology which is small land required and applicable to above all purposes is important nowadays.

High gradient magnetic separation using seeding material seems suitable technique for satisfying above technological requirements because of its high processing rate, high effluent water quality and adaptability for wide range of influent water quality. The theory of "High Gradient Magnetic Separation" was established by Oberteuffer (Oberteuffer, 1973). Currently, a high gradient magnetic separator (HGMS) based on this theory is applied for removing magnetic suspended solids in discharge from steel industry and for eliminating collosion products in the feed water of power plant. High gradient magnetic separation is also applicable to municipal and industrial wastewater treatments for the removal of non-magnetic contaminants using magnetic seeding method. In this method, non-magnetic pollutants in water are coagulated with magnetic seeding material by coagulant and subsequently removed by HGMS. Magnetite (Fe_3O_4) has been usually used as seeding material in the seeding method because of its large magnetic moment. Chemical coagulation of the magnetite and pollutants is made by alum and polyelectrolyte. Principal advantage of the seeding method is its high processing rate and high removal rate of suspended and colloidal matters.

Seeding method has been investigated by many researchers. Allen et al. (1977) and Allen (1978) investigated about application of the seeding method for combined sewer overflows treatment. According to them, the seeding method was effective for reducing most forms of pollutants present in a combined sewer overflows. They emphasized that the treatment system has several additional benefits, namely, high processing rate, small land requirements, and lower chlorine demand. Several studies have been also carried out on the removal of phytoplankton in eutrophic seawater or freshwater (Kurinobu & Uchiyama, 1982; Kaneko et al., 1981; Kaneko, 1982). These studies

confirmed the excellent efficiency of the seeding method for removing phytoplankton.

The seeding method, however, has not received practical application because of following two problems. First is high cost of chemicals, especially for magnetite. Reducing the concentration of magnetite or using low price seeding material instead of magnetite are important to reduce the chemical cost. In addition to this, selection of the most suitable polyelectrolyte may be important for reducing the required amount of seeding material because characteristics of polyelectrolyte strongly affect the mechanical strength of coagulated floc. Second is that filter run of the magnetic separator are impractically short when the feed water contains high concentration of aggregates. Preseparation of large aggregates before magnetic separation or increase in the matrix thickness should be required for lengthen the filter run.

Tokyo Sewerage Bureau has been carrying out the research about the seeding method since three years ago. The main objectives of the research are determination of optimum treatment conditions at low magnetite concentration and establishing the treatment flow applicable to both combined sewer overflows treatment and tertially treatment. The present paper outlines the results obtained by this research.

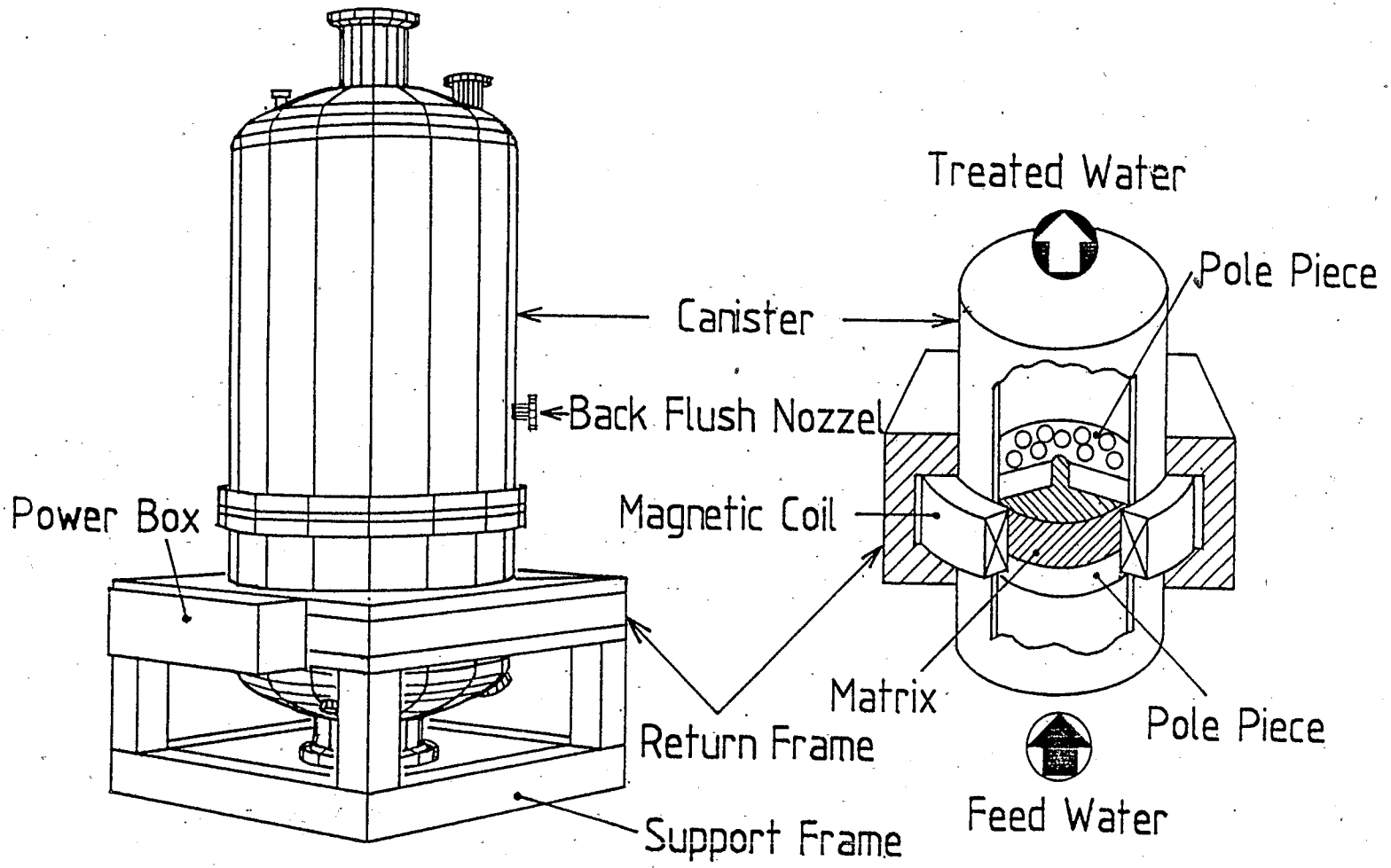

Fig.1. High Grad'ent Magnetic Separator

2. COMBINED SEWER OVERFLOWS TREATMENT

2-1. Bench Tests

Scope of testing

Bench tests were carried out to obtain basic data to about an optimum dosing rate and treatment conditions.

Bench test apparatus

Fig.2 shows the appearance of a batch type apparatus used for the bench test. The flowsheet of the apparatus is indicated by Fig.3. Flocculation tank was 5 litter in capacity. Size of mixing blade in the flocculation tank was 15 cm in length and 6 cm in width. The matrix for the high gradient magnetic separator consists of 100 pieces of SUS 430 expanded metal disc which were 16 mm in diameter (Fig.4). The matrix is 80 mm in thickness and 80% in void ratio. Electromagnets are on both sides of the matrix.

Test procedures are described as follows. Magnetite powder (1 to 4 μm in diameter) were added to 5 litter of raw water in the flocculation tank and mixed by the mixer at 150 rpm for one minute. Then, alum was added and flush mixed at 150 rpm fro 15 seconds followed by gentle mixing at 50 rpm for 1 minute. Finally, polyelectrolyte was added and stirred at 150 rpm for 15 seconds and 50 rpm for 2 minutes successively. The flocculated slurry was pumped to the magnetic separator by a filter pump which was located downstream from the magnetic separator to avoid the disruption of the flocs. Effluent from magnetic separator was sampled continuously for chemical analysis untill the breakthrough. After the breakthrough, matrix was backflushed with tap water.

Alum concentration

Ordinary jar test was performed prior to the experiments with the test apparatus to clarify the relationship between the alum requirement and SS value of raw sewage. Fig.5 shows one of the results of the jar test. In this case, optimum concentration was estimated for about 5mg/l because the transparency of supernatant reached to a plateau at this concentration. Relationship between SS and optimum alum concentration is shown by Fig.6. Optimum alum concentration fluctuated in narrow range (3 to 7 mg Al/l) while the SS values fluctuated considerably (100 to 600 mg/l).

Magnetite concentration

Fig.7 shows relationship between the magnetite concentration and the removal rate of total organic carbon (TOC). Magnetite concentration had almost no effect on the removal rate and also on the filter run in the range between 50 to 400 mg/l, hence magnetite requirements should be less than or equal to 50 mg/l. Further tests showed that magnetite concentration can be reduced to about 25 mg/l. This optimum value is considerably lower than that recomended by Allen (1978) (200mg/l). Lower flow velocity (2000m/day) comparing with that of Allen (5000 m/day) should be cause of low magnetite

demand in the present study. In Japan, price of magnetite is considerably higher than that in the United States, hence it may be unavoidable to reduce the flow velocity. Reduction of magnetite concentration is also important from the viewpoint of cutting down the amount of sludge generated from the process.

Polyelectrolyte concentration

To select suitable sort of polyelectrolyte, filter run and the transparency of effluent were compared for many kinds of polyelectrolyte. As shown in Table.1, anionic polyelectrolytes were generally superior to amphoteric and cationic ones. Among the anionic polyelectrolytes, P-720 had especially excellenct performance, hence it was used for all of remaining tests.

Fig.8 shows the effects of P-720 concentration on the filter run and effluent water quality. Polyelectrolyte concentration had almost no effect on the effluent water quality within the range of 0.5 to 3 mg/l. The filter run, however, decreased markedly under the concentration of 2 mg/l. It is concluded that the optimum concentration was 2 mg/l.

Flow velocity of high gradient magnetic separator

Fig.9 shows the effects of the flow velocity on the filter run and effluent water quality. There was no observable effect of flow velocity on the effluent water quality, except a slight increase of total phosphorus over 1500 m/day. On the other hand, the filter run markedly dropped at the velocity over 1500 m/day.

Magnetic field strength

As shown in Fig.10, magnetic field strength has little effect on the effluent water quality, while filter run increased markedly with the strength. Electrical power consumption increases with the magnetic field strength, hence optimum magnetic field strength must be determined by the consideration of total running cost.

2-2. Pilot Plant Tests

Scope of testing

To obtain the fundamental data for the design of full scale plant, a continuous flow pilot plant was operated on optimum conditions which had been determined by the bench test.

Pilot plant facility

Fig.11 and Fig.12 shows a flowsheet and side view of the pilot plant. Raw sewage taken by submerged pump was applied to automatic flushing strainer with 0.5 mm opening to remove rubbish, and then pumped to flush mixing unit. After addition of magnetite and alum, the mixture was stirred for 3 minutes. Coagulated slurry was fed to a floc preseparation tank (preseparator) after addition of the polyelectrolyte (P-720). Overflow rate of the preseparator was about

800 m3 /m2 • day. The preseparator consisted of three layers; settling layer in which large sized flocs settle down, helically flowing flocculation layer in which feed slurry flew into and flocculation proceed, and sludge blanket layer by which fine flocs was removed, in order of height. Remaining flocs in the effluent from the preseparator were removed completely by the magnetic separator. Matrix of the magnetic separator was 150 mm in diameter and 150 mm in thickness. SUS 430 expanded metal plates which mesh size was 13 mm in length and 5.1 mm in width were used for the matrix. Operating conditions of the magnetic separator were 2000 m/day for flow velocity, 3 KG for magnetic field strength and 1 hour for filter run. Backflush of the magnetic separator was made with mixture of pressurized water and air for 10 seconds with flow velocity of 2000 to 3000 m/hr.

Optimal parameter runs

Table 2 shows the results from an optimized pilot plant run of 6 hours continuous operation using raw sewage at dry weather as influent. The results show that high removal rates (more than 90%) of SS, turbidity, phosphorus, coliform bacteria can be achieved by the seeding method. Removal rates of BOD and COD were lower than those values. However, it was confirmed by bench test that higher removal rates of BOD and COD could be obtained in wet weather, probably because the ratio of soluble COD and BOD to suspended ones decreased markedly in wet weather. Fig.13 shows the counts of coliform bacteria in effluent for 19 runs operation. The counts were much lower than 3000 no./ml, which is the regulation number in Japan, hence addition of chlorine to the effluent should be avoidable.

The approximate amount of sludge generated from the process was estimated for about 140 to 310 g-dry weight per 1 m^3 of the influent. Seventy eight percent of the total sludge was generated from the preseparator and remainings from the magnetic separator. As shown in Table 3 and Fig.14, the generated sludge had excellent thickenability and dewaterability.

It was confirmed that about 99% of trapped flocs in the matrix were washed out by the backflushing which was made with the mixture of effluent water and two times volume of pressurized air. Little amount of cellulose fiber however remained in the matrix after the backflushing. It has not been clarified yet whether these fibrous materials causes the clogging of matrix after long time continuous operation. Ball shaped matrix seems to be more convenient than the expanded metal matrix to prevent clogging because tangle of cellulose fiber should not occur. Test of ball matrix is carring out now.

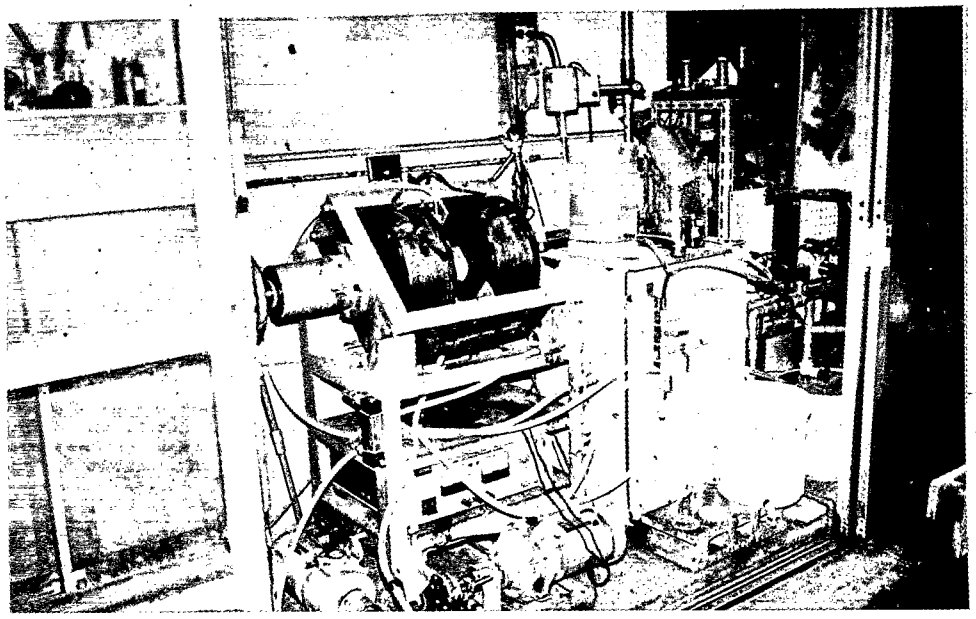

Fig.2. Bench test apparatus.

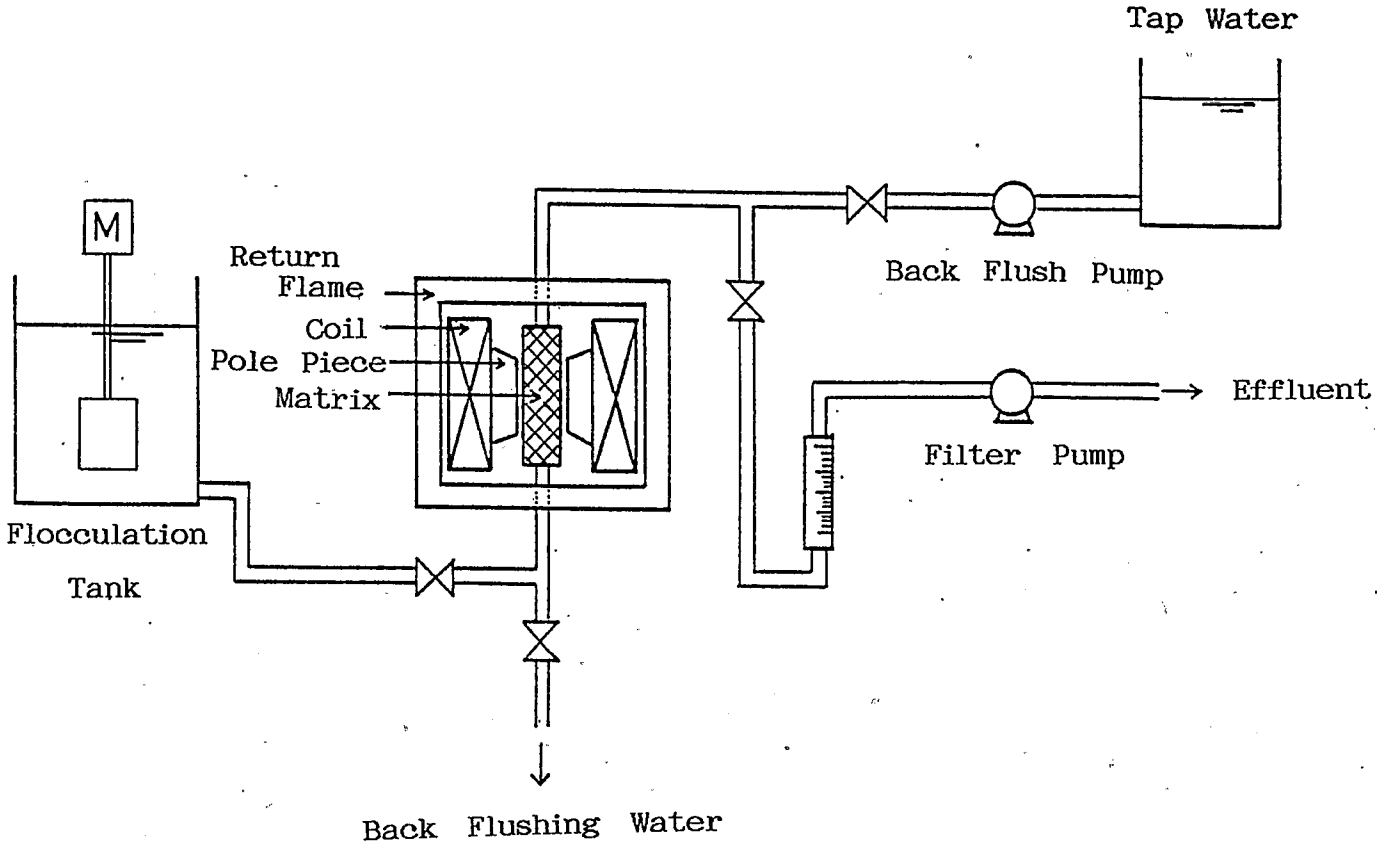

Fig.3. Flow sheet of bench test apparatus.

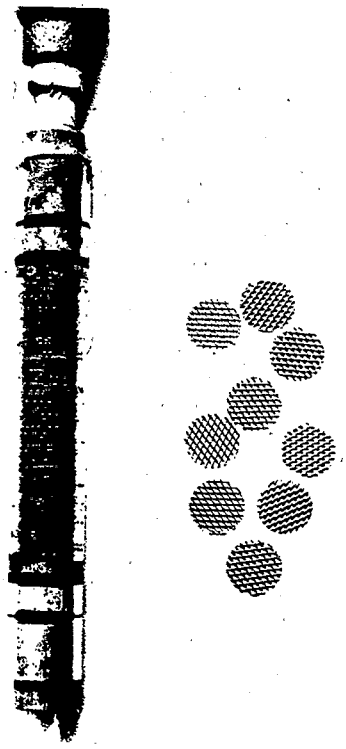

Fig.4. Matrix (16 mmϕ x 80 mm H) of high gradient magnetic separator used for bench test.

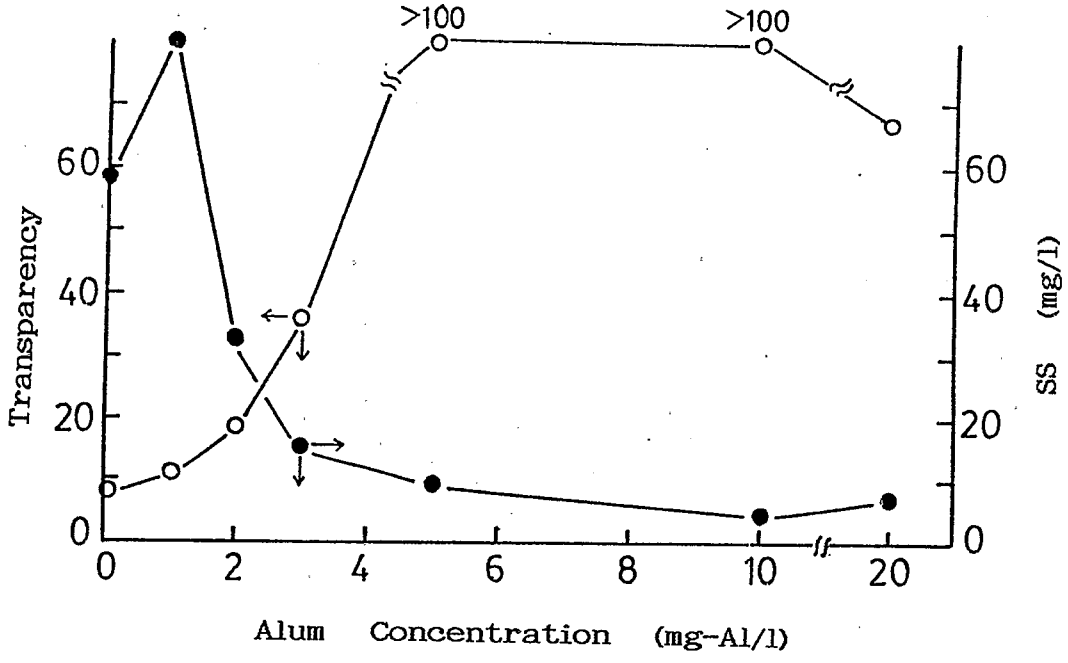

Fig.5. Effect of alum concentration on transparency and SS of supernatant in jar test of raw sewage. (magnetite: 25 mg/l, polyelectrolyte: 2 mg/l)

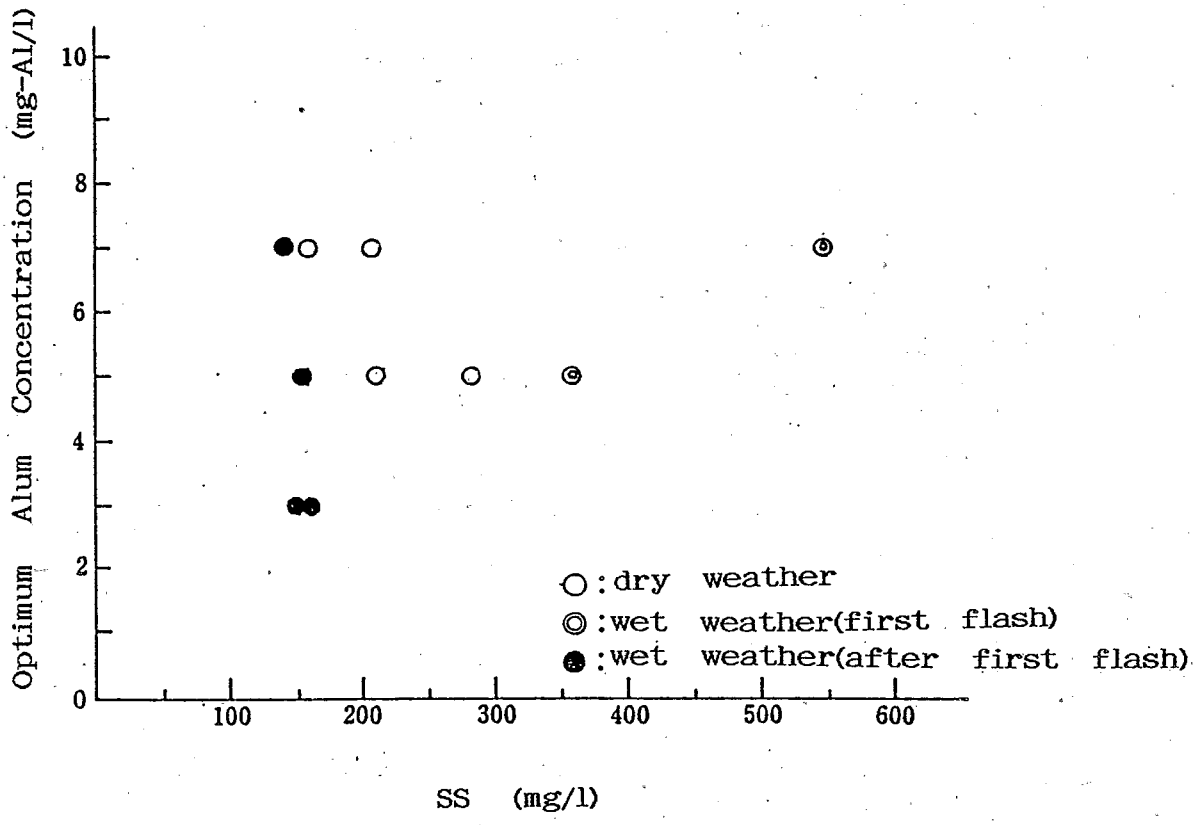

Fig.6. Relationship between SS concentration of raw sewage and optimum alum concentration.

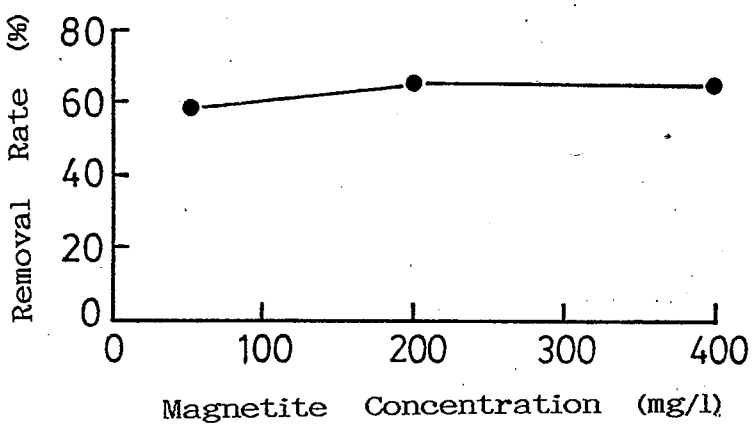

Fig.7. Effect of magnetite concentration on removal rate of total organic carbon (TOC).

Test conditions
Magnetic Field Strength: 5KG
Flow Velocity: 2000 m/day
Alum Conc.: 8.6 mg/l
Polyelectolyte Conc.(C-599-1P): 1 mg/l
Influent TOC: 73 mg/l

Table 1. Polyelectrolyte comparison.

Test conditions

Magnetic Field Strength : 5KG

Flow Velocity : 1400 m/day

Magnetite Conc.: 25 mg/l

Ferrous Alum Conc.: Al-8mg/l,Fe-3mg/l

Polyelectrolyte : 2 mg/l

Table 1-1. Anionic Polyelectrolyte

Name	Characteristics							Filter run (relative value)	Transparency (relative value)
	Chemical composition (mole%)						M.W. ($\times 10^4$)		
	DMC	DM-Ac	AAm	AAc	DAC	others			
P-700			100				1500	59	63
P-710			85	15			1500	88	72
P-713			80	20			1500	100	94
P-720			70	30			2000	100	100
P-730			60	40			2000	50	97
P-790				100			600	17	—
ⓐ			80	20			1300	45	82
ⓑ			70	30			2600	94	102
ⓒ			90			※1 10	500	90	48
ⓓ			90			※1 10	600	87	56
ⓔ			70			※1 30	500	70	57
ⓕ			80	20			1600	65	81
ⓖ						※2 2.2	10	74	49

Table 1-2. Amphoteric Polyelectrolyte

Name	Characteristics							Filter run (relative value)	Transparency (relative value)
	Chemical composition (mole%)						M.W. ($\times 10^4$)		
	DMC	DM-Ac	AAm	AAc	DAC	others			
ⓘ	80			20			300	34	—
ⓜ	80			20			500	34	—
ⓝ	40		55	5			300	34	—
ⓞ	40		55	5			500	29	—
ⓟ		80		20			300	47	85
ⓠ		80		20			500	28	—
ⓡ		40	55	5			300	44	—
ⓢ		40	55	5			400	55	83
ⓘ			60		40		200	46	75
ⓤ			60		40		400	66	89
ⓥ			80		20		500	49	75
ⓦ			90		10		300	84	85
ⓧ			90		10		600	136	60
ⓨ			80		20		700	29	—
ⓩ			95		5		600	19	—

Table 1-3. Cationic Polyelectrolyte

Name	Characteristics							Filter run (relative value)	Transparency (relative value)
	Chemical composition (mole%)						M.W. ($\times 10^4$)		
	DMC	DM-Ac	AAm	AAc	DAC	others			
C – 100	100						100	19	–
C – 200	100						200	21	–
C – 201		100					200	23	–
C – 202						※3 100	200	28	–
C – 252			50			※3 50	200	26	–
C – 270	70		30				200	32	–
C – 300	100						300	23	–
C – 301		100					300	33	–
C – 340	40		60				300	60	72
C – 341		40	60				300	48	–
C – 343	20	20	60				300	45	–
C – 360	60		40				300	50	–
C – 373	50	20	30				300	43	–
C – 380	80		20				300	36	–
C – 455			50		50		400	71	83
C – 500	100						500	28	–
CA						※4 100	1	19	–
DC-802P						※5 100	1	13	–
DC-753P						※5 100	8	14	–
DM-883P	100						7	17	–
DM-254P	100						25	25	–
ⓗ			40			※6 60	500	76	68
ⓘ						※7 100	20	18	–
ⓙ	95	5					200	12	–
ⓚ	95	5					400	17	–
C-599-1P								27	68
CP-658							1000	73	64

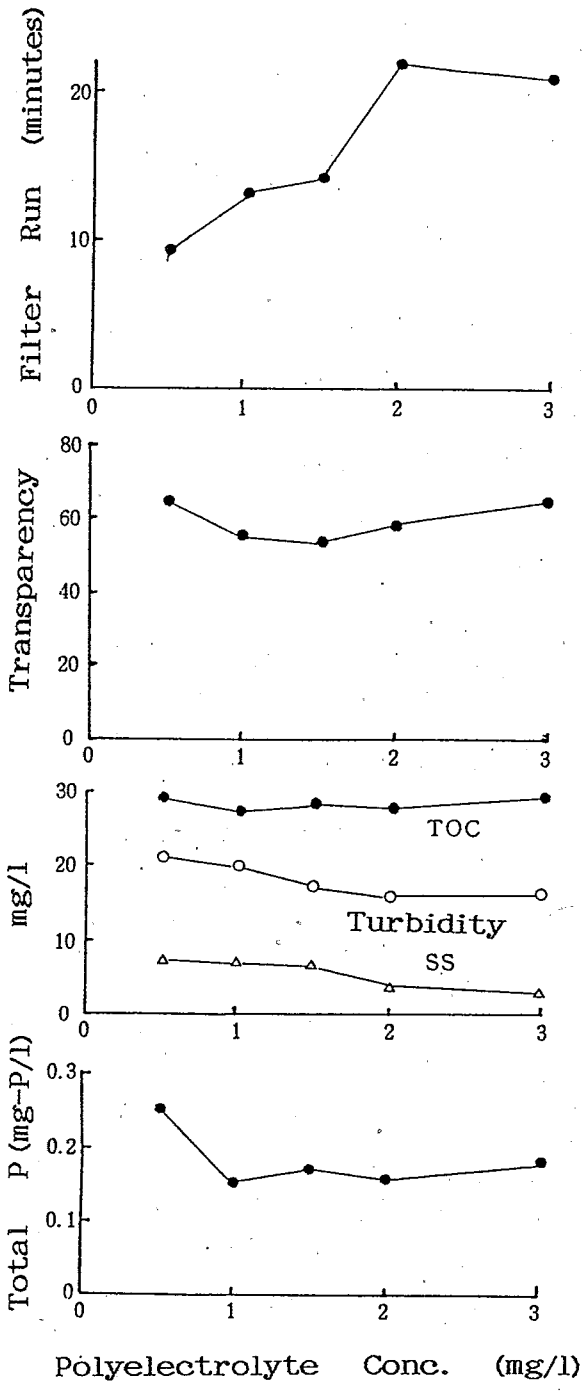

Fig.8. Effects of polyelectrolyte concentration on filter run and effluent water quality.

Test conditions
- Magnetic Field Strength:5KG
- Flow Velocity : 1400m/day
- Magnetite Conc. : 25mg/l
- Alum Conc. : 8mg/l
- Polyelectrolyte : P-720

Influent(Settled Sewage)
- Transparency : 5.5
- Turbidity : 95
- SS : 34 mg/l
- TOC : 77 mg/l
- Total P : 5.3 mg-P/l

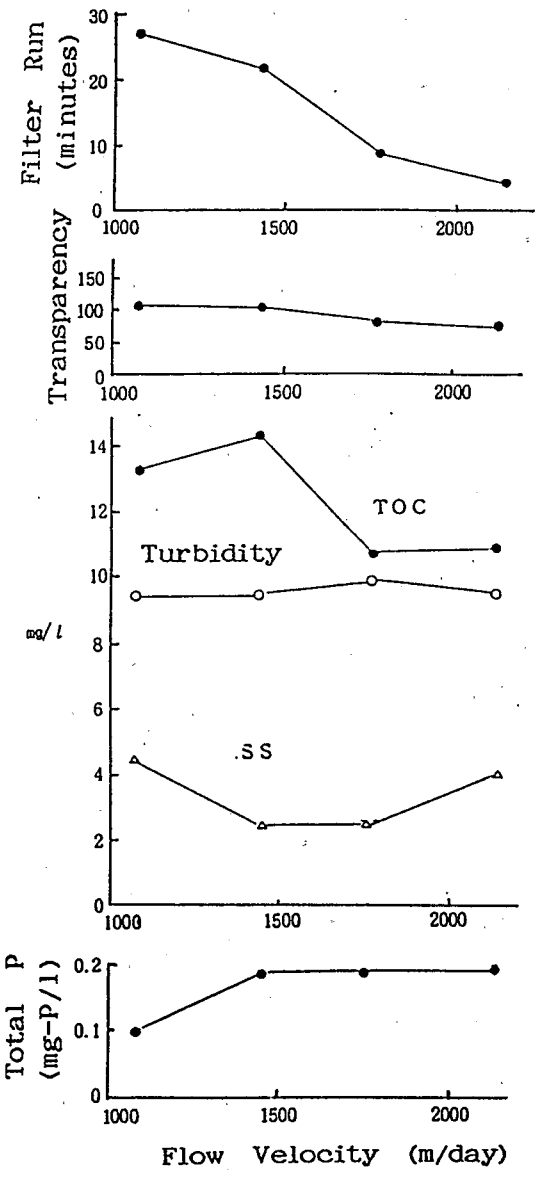

Fig.9. Effects of flow velocity on filter run of high gradient magnetic separator and effluent water quality.

Test Conditions
Magnetic .Field Strength : 5KG
Magnetite Conc. : 25mg/l
Alum Conc. : 8mg/l
Polyelectrolyte Conc. (P-720) : 2mg/l
Influent (Settled Sewage)
Transparency : 5.5, Turbidity : 93, SS : 62mg/l
TOC : 52mg/l, Total P : 3.8mg/l

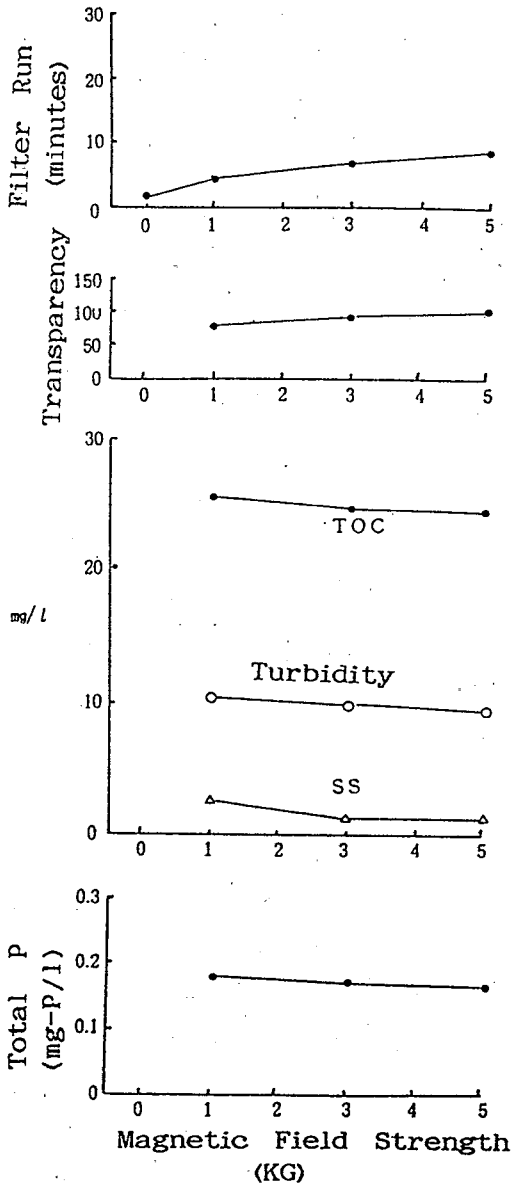

Fig.10. Effects of magnetic field strength on filter run of high gradient magnetic separator and effluent water quality.

Test Conditions
Magnetite Conc. : 25mg/l
Alum Conc. : 8mg/l
Polyelectrolyte Conc. (P-720) : 2mg/l
Influent (Settled Sewage)
Transparency : 6, Turbidity : 97, SS : 41mg/l
TOC : 68mg/l, Total P : 5.9mg/l

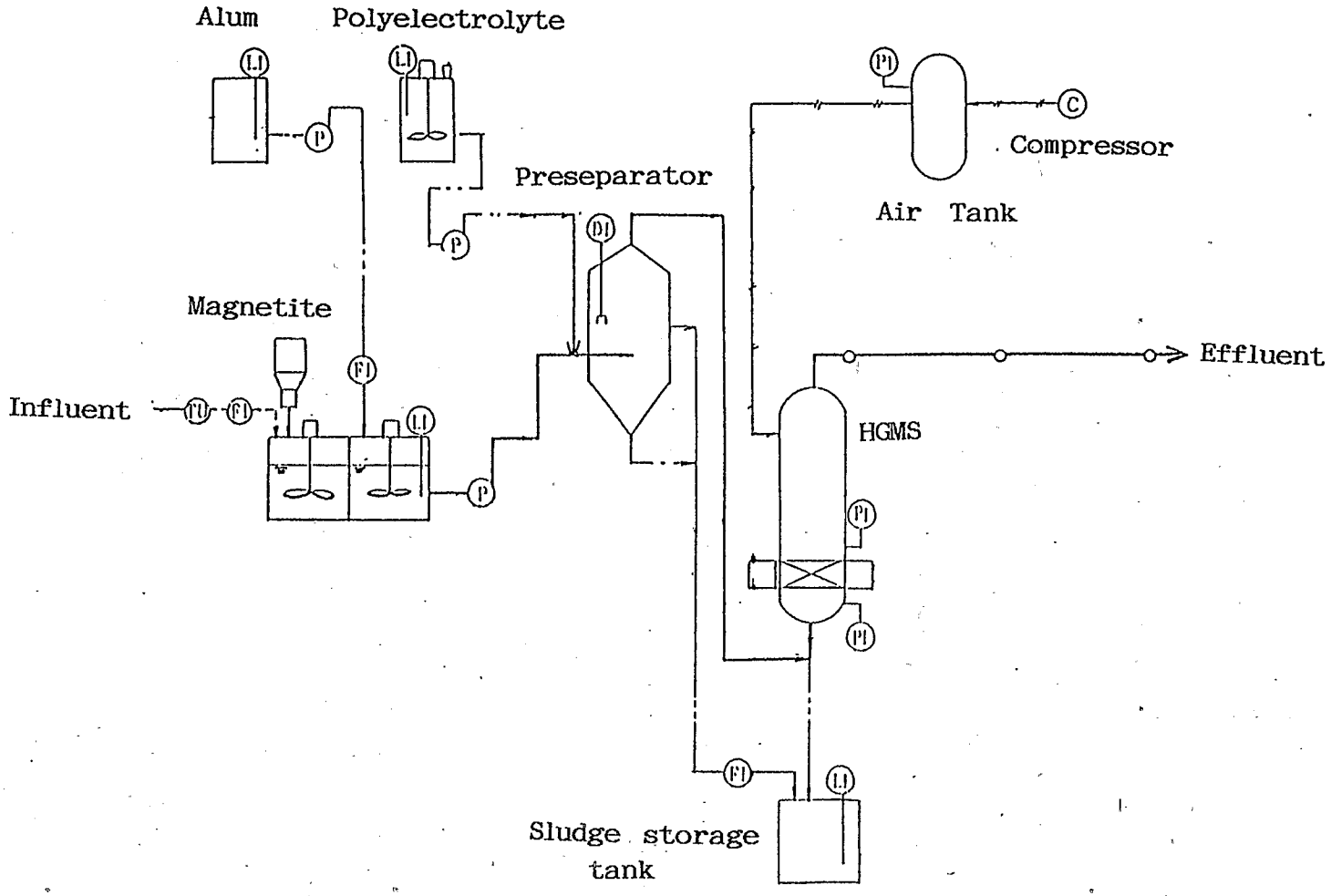

Fig.11. Flow sheet of pilot plant for combined sewer overflows treatment.
(HGMS : High Gradient Magnetic Separator)

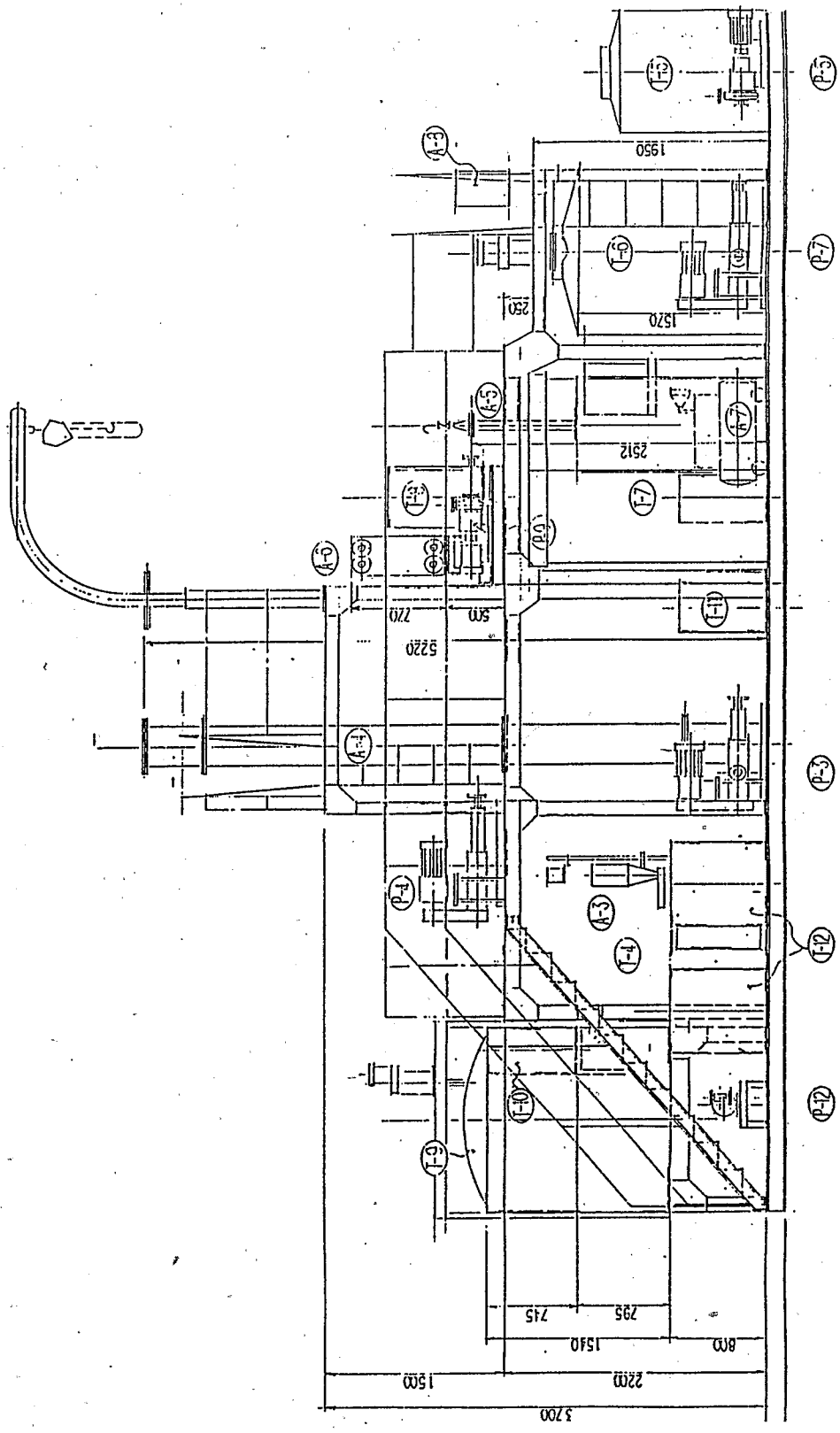

Fig.12. Side view of pilot plant for combined sewer overflows treatment.

Table 2. Results of continuous operation of pilot plant for combined sewer overflows treatment.

	Influent	Effluent
pH	7.3 - 7.8	6.7 - 7.0
Transparency	3 - 5	more than 150
Turbidity	81 - 202	2 - 7
Chromaticity	30 - 50	18 - 24
SS (mg/l)	108 - 190	less than 5
COD (mg/l)	74 - 109	21 - 50
BOD (mg/l)	106 - 177	21 - 51
Total P (mg-P/l)	25 - 4.2	0.1 - 0.2
Coliform bacteria (No./ml)	2×10^{5} - 3×10^{5}	13 - 16

(Operating conditions)
Influent : Raw sewage at dry weather
Magnetic Field Strength : 3KG
Flow Velocity : 2000 m/day
Alum Conc. : 7.6 mg-Al/l
Magnetite Conc. : 26 mg/l
Polyelectrolyte Conc. : 1.8 mg/l
Backflush cycle : 1 hour
Backflush Time : 10 seconds

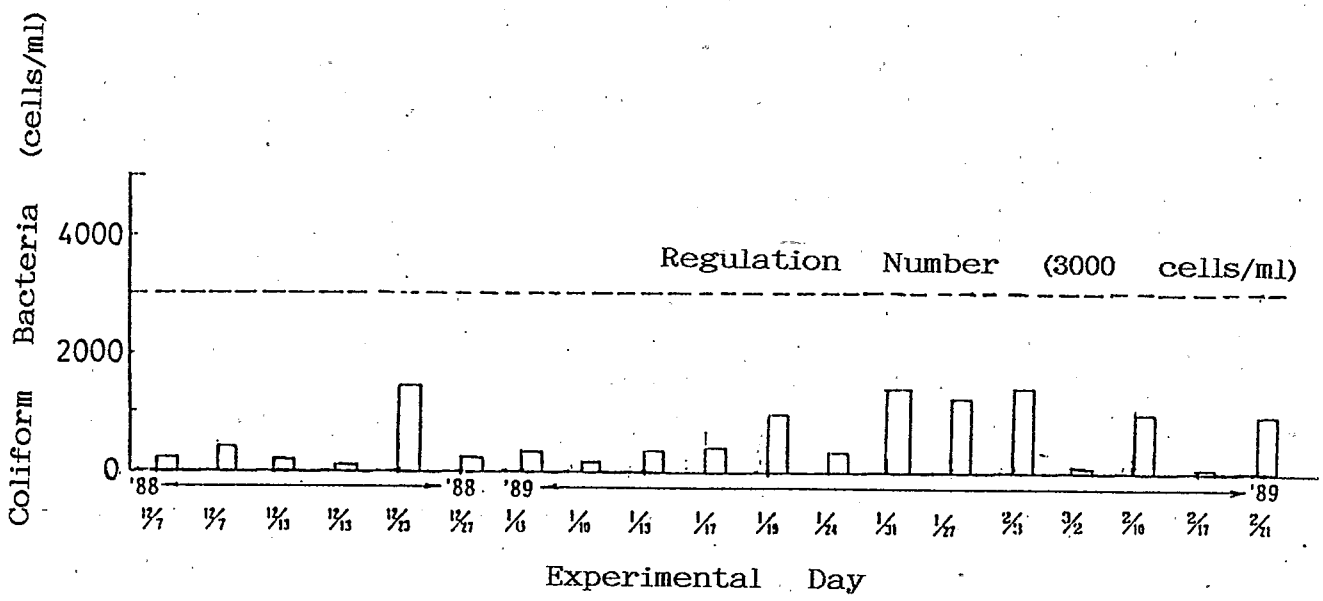

Fig.13. Number of coliform bacteria in effluent.
(Influent : raw sewage at dry and wet weather)

Table 3. Characteristics of sludge generated from pilot plant for combined sewer overflows treatment.

Raw sludge

SS	2000	mg/l

Thickened sludge

pH	6.5
TS	28600 mg/l
SS	26600 mg/l
VSS	17290 mg/l
Coarse suspended solids	24%
Calorific value	3690 Cal/g-dry weight

(gravity thickening for 12 hours)

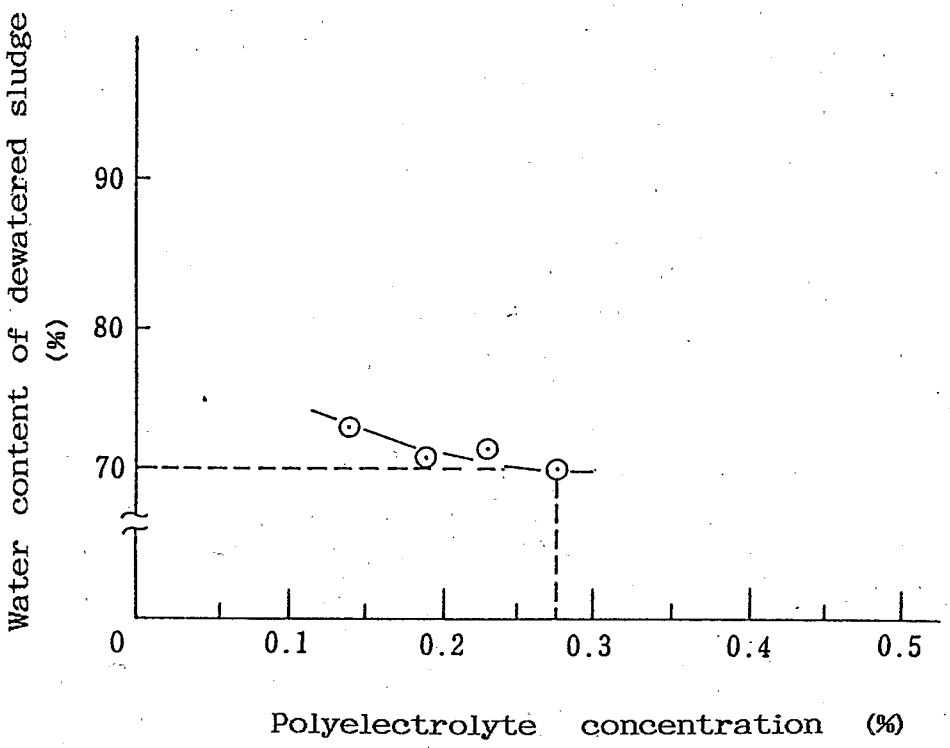

Fig.14. Effect of cationic polyelectrolyte concentration on dewatering of thickened sludge.

Test Conditions
Polyelectrolyte: Hitachi HK153B
Dewatering facility: Beltpress type
Dewatering pressure: 1 kg/cm^2

3. TERTIARY TREATMENT OF ACTIVATED SLUDGE PROCESS EFFLUENT

3-1. Bench Tests

Scope of testing

The bench test was carried out to obtain the basic data about the best dosing rate and treatment conditions.

Bench test apparatus

The bench test apparatus described above was used for also the experiment of tertiary treatment.

Alum concentration

Fig.15 shows the relationship between alum dosing rate and removal rate of total phosphorus. It is obvious that the total phosphorus concentration became lower than 0.2 mg/l when the mole-ratio of alum to total phosphorus is higher than 2.5.

Magnetite and polyelectrolyte concentrations

Fig.16 and 17 shows the effects of the magnetite and polyelectrolyte (C-599-1P) concentrations on the removal rate of TOC and total phosphorus. In the case of poor water quality influent (Fig.16), removal rate increased markedly with concentrations of polyelectrolyte and magnetite. On the other hand, in the case of improved water quality influent (Fig.17), high removal rate was obtained regardless of the concentrations of polyelectrolyte and magnetite. Optimum magnetite concentration should be ranges between 5 to 10 mg/l.

Fig.18 shows the effect of polyelectrolyte (P-720) concentration on the filter run and effluent water quality. Effluent water quality improved with polyelectrolyte concentration in the ranges between 0 and 0.5 mg/l, and became stable in the ranges from 0.5 to 1 mg/l. On the other hand, filter run increased steadily with polyelectrolyte concentration in the ranges between 0 and 1 mg/l. These results suggest that appropriate concentration is 0.5 to 1 mg/l.

Flow velocity of high gradient magnetic separator

As shown in Fig.19, flow velocity had almost no effects on the effluent water quality. The filter run however dropped markedly with increase of the flow velocity. It is estimated that filter run will be about 30 minutes at flow velocity of 2000 m/day when using the matrix of 15 cm thickness, which is standard value of full scale magnetic separator. This filter run should be practical for the operation of full scale magnetic separator, hence it is assumed that optimum flow velocity is 2000 m/day.

Magnetic field strength

Fig.20 shows the effects of the magnetic field strength on the

effluent water quality and filter run. Though the effluent water quality was not affected by magnetic field strength, filter run increased considerably with magnetic field strength. Filter run became rather stable over 3 KG, hence this value should be appropriate for practical operation.

3-2. Pilot Plant Tests

Scope of testing

A continuous flow pilot plant was operated with the optimum conditions determined by the bench tests to obtain a design parameters of full-scale plant.

Pilot plant facility

Fig.21 shows the flowsheet of the pilot plant. Magnetite and alum were dosed successively and mixed by a static mixer. The mixture was gently stirred by helical flow in flocculation tank for 4 minutes. The flocculated slurry was fed to the magnetic separator after addition of polyelectrolyte. Magnetic separator was operated with the flow velocity of 2000 m/day and the magnetic field strength of 3 KG. The matrix of magnetic separator was 10 cm in diameter and 15 cm in thickness. When the turbidity of the effluent increased markedly, backflushing was carried out with 40 litter tap water and 120 litter of pressurized air for ten seconds. Fig.22 shows the side view of the pilot plant.

Optimal parameter runs

Table 4 shows one of the results of the water quality analyses. High removal rate was obtained for SS, BOD, total phosphorus and coliform bacteria. Chromaticity and COD also removed for about 40%. These results prove that the seeding method has high performance of tertiary treatment.

The amount of sludge generated was determined about 30 to 50 g-dry weight per 1 m^3 of influent. This production rate is one-fifth of that for combined sewer overflow treatment. As shown in Table 5, the sludge was easy for thickening and dewatering. The caloric value of the dewatered sludge was almost half of that from the combined sewer overflows treatment. Incinerated ash contained ferromagnetic material for about 7%, hence it seems possible to reuse the ash as seeding material after ore dressing by magnetic separation.

The matrix was washed almost completely by the backflushing described above. In contrast to the combined sewer overflows treatment, fibrous material did not accumurate in the matrix.

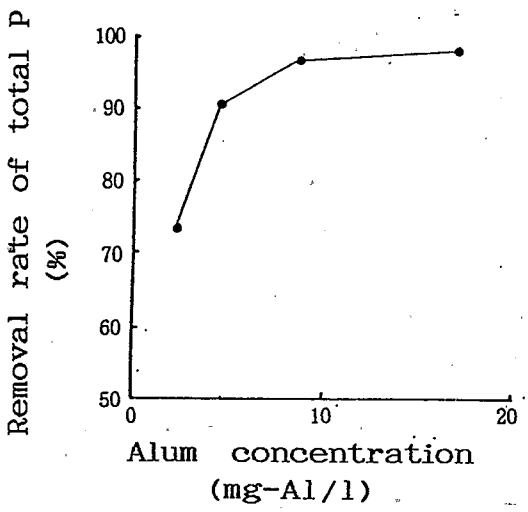

Fig.15. Effect of alum concentration on removal rate of total phosphorus.

Test Conditions
Magnetic Field Strength : 5KG
Flow Velocity : 1000 m/day
Magnetite Conc. : 200 mg/l
Polyelectrolyte(C-599-1P) : 1mg/l

Influent (activated sludge process effluent)
Total P : 4.6 mg/l

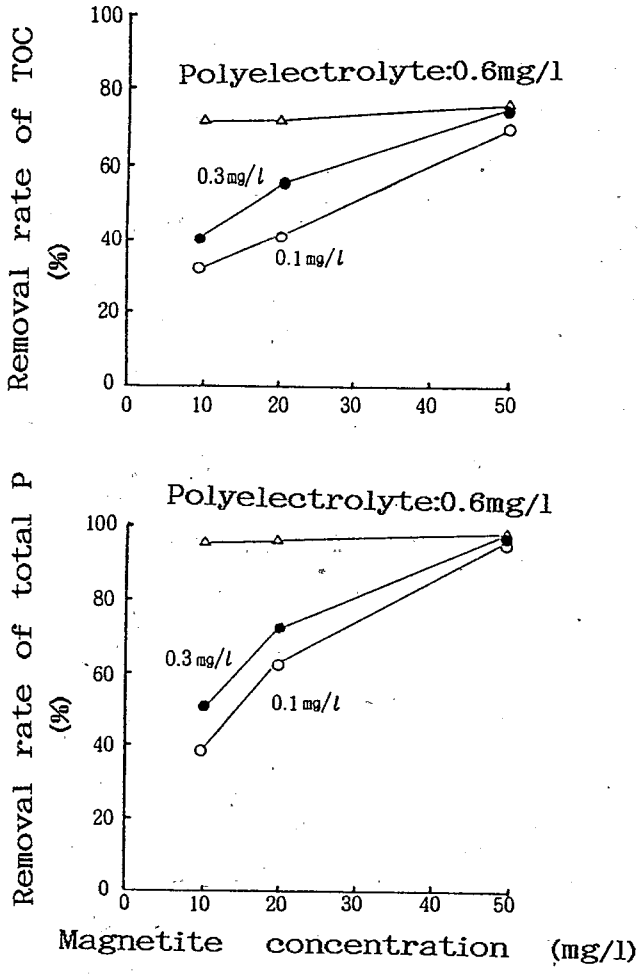

Fig.16. Effect of magnetite and polyelectrolyte concentrations on removal rates of TOC and total phosphorus.
(The case of poor water quality influent.)

Test Conditions
Magnetic Field Strength : 5KG
Flow Velocity : 1000 m/day
Alum Conc. : 8mg/l
Polyelectrolyte : C-599-1P

Influent (Activated sludge process effluent)
TOC : 28mg/l
Total P : 4.7 mg-P/l

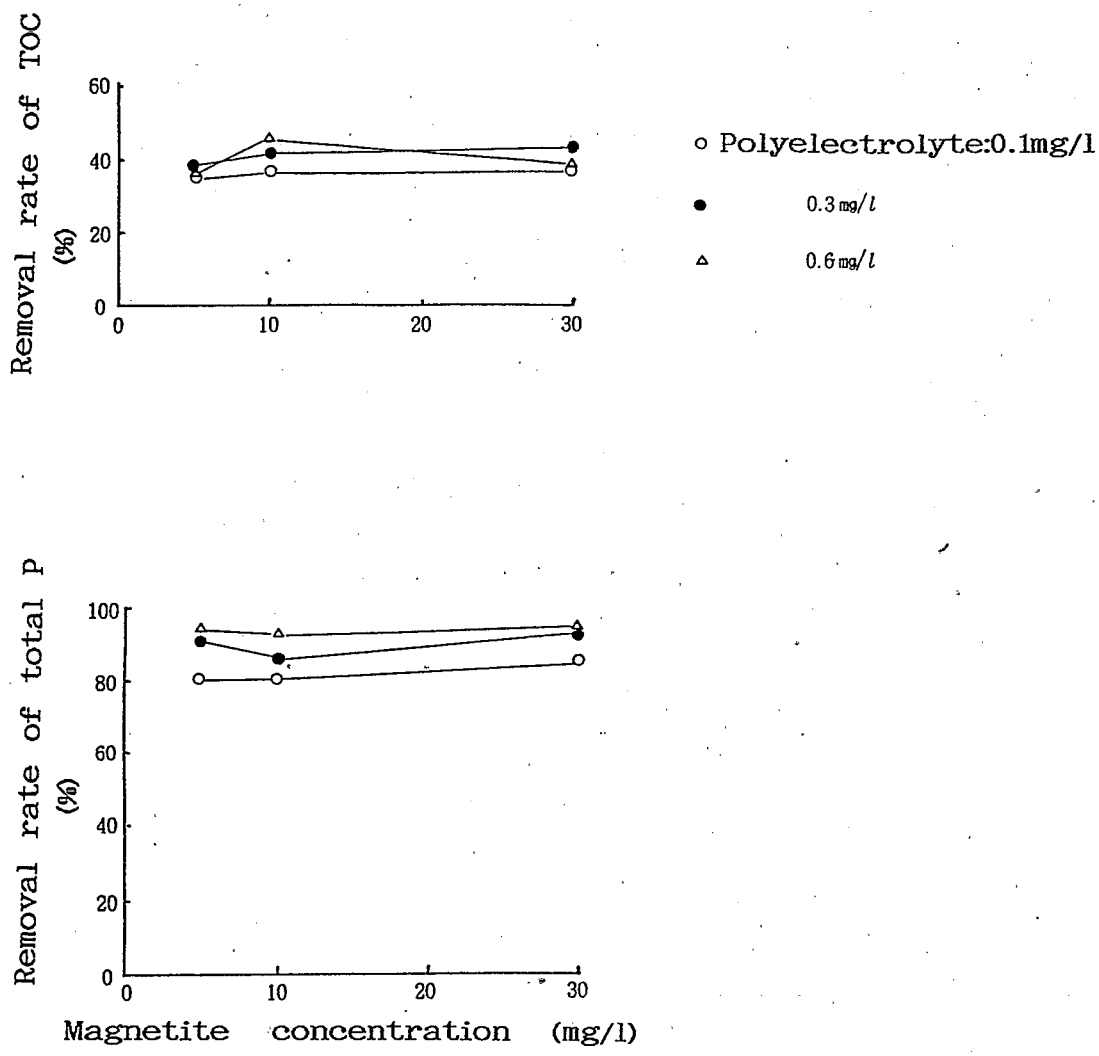

Fig.17. Effect of magnetite and polyelectrolyte concentrations on removal rates of TOC and total phosphorus.
(The case of the influent of high grade water quality)

Test Conditions
Magnetic Field Strength : 5KG
Flow Velocity : 1000 m/day
Alum Conc. : 8mg/l
Polyelectrolyte : C-599-1P

Influent (Activated sludge process effluent)
TOC : 13mg/l
Total P : 0.57 mg-P/l

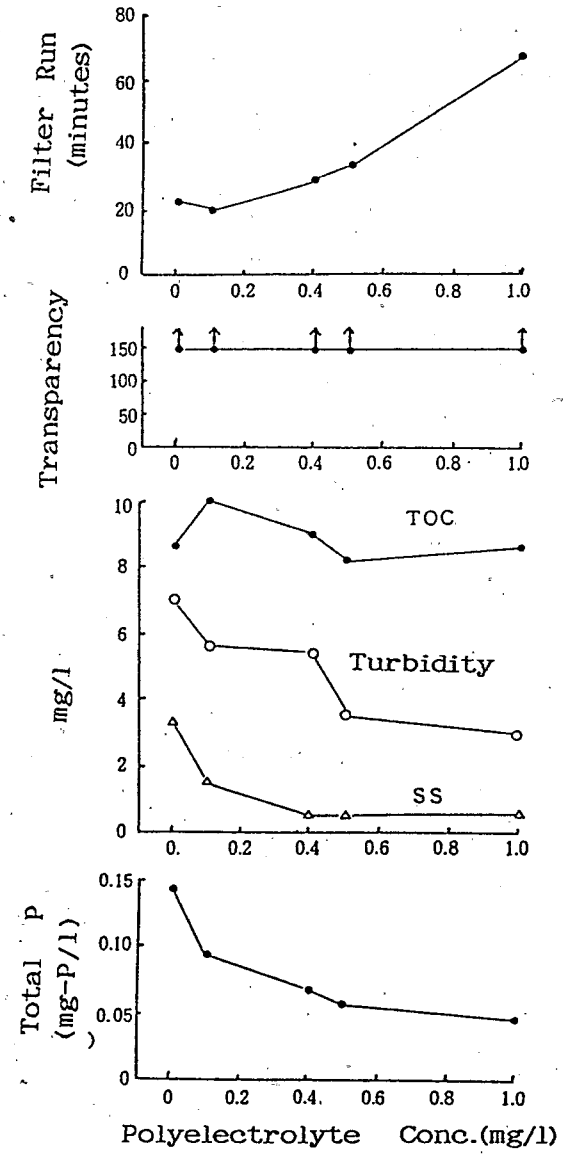

Fig.18. Effects of polyelectrolyte concentration on filter run of high gradient magnetic separator and effluent water quality.

Test Conditions
Magnetic Field Strength : 5KG
Flow Velocity : 1400m/day
Magnetite Conc. : 10mg/l
Alum Conc. : 4mg/l
Polyelectrolyte : P-720
Influent (Activated Sludge Process Effluent)
Transparency : 65, Turbidity : 8.7, SS : 4.3mg/l
TOC : 13mg/l, Total P : 0.54mg-P/l

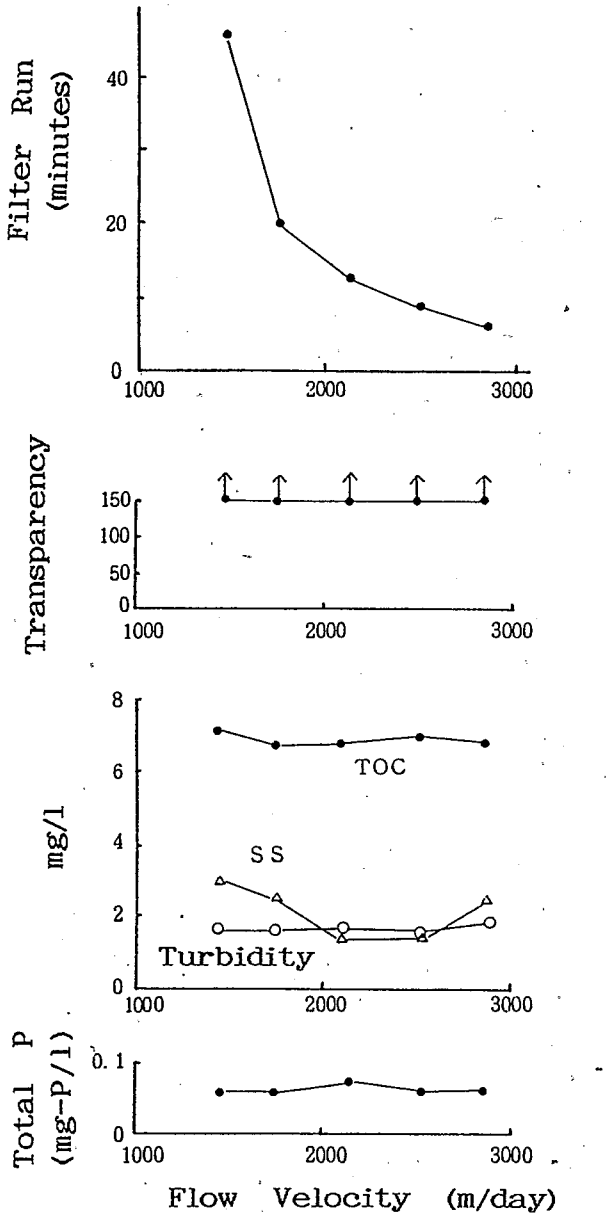

Fig.19. Effects of flow velocity on filter run of high gradient magnetic separator and effluent water quality.

Test Conditions
Magnetic Field Strength : 5KG
Magnetite Conc. : 10mg/l
Alum Conc. : 4mg/l
Polyelectrolyte Conc.(P-720) : 1mg/l
Influent (Activated Sludge Process Effluent)
Transparency : 36, Turbidity : 11.2, SS : 9.8mg/l
TOC : 18.4mg/l, Total P : 1.3mg-P/l

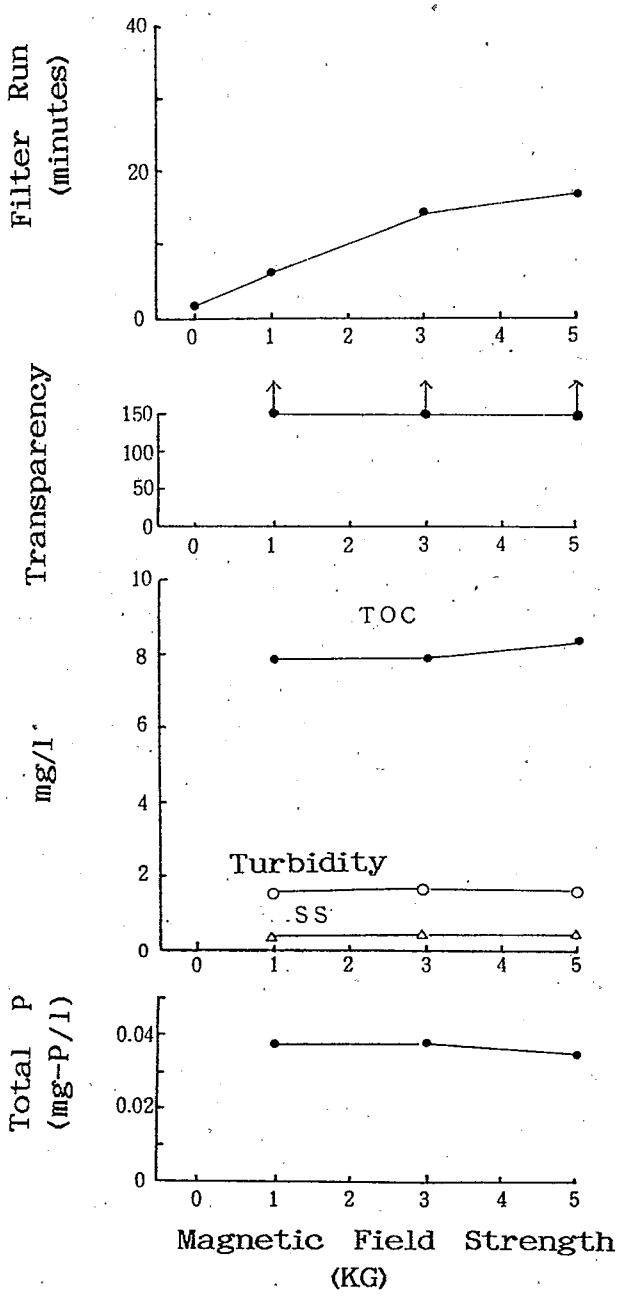

Fig.20. Effects of magnetic field strength on filter run of high gradient magnetic separator and effluent water quality.

Test Conditions
Flow Velocity : 2200m/day
Magnetite Conc. : 10mg/l
Alum Conc. : 4mg/l
Polyelectrolyte Conc.(P-720) : 1mg/l
Influent (Activated Sludge Process Effluent)
Transparency : 42, Turbidity : 9, SS : 9mg/l
TOC : 13.5mg/l, Total P : 0.67mg-P/l

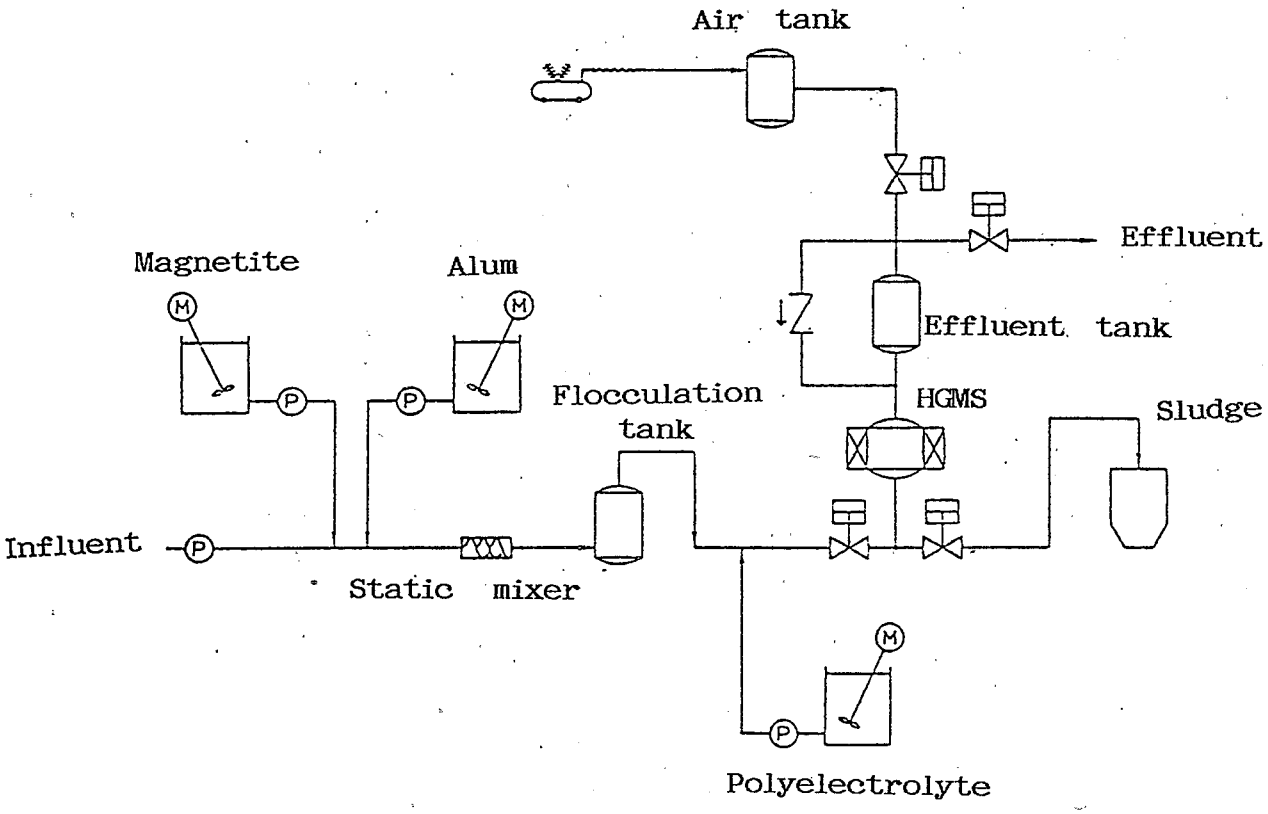

Fig.21. Flow sheet of the pilot plant for tertiary treatment.
(HGMS : High Gradient Magnetic Separator)

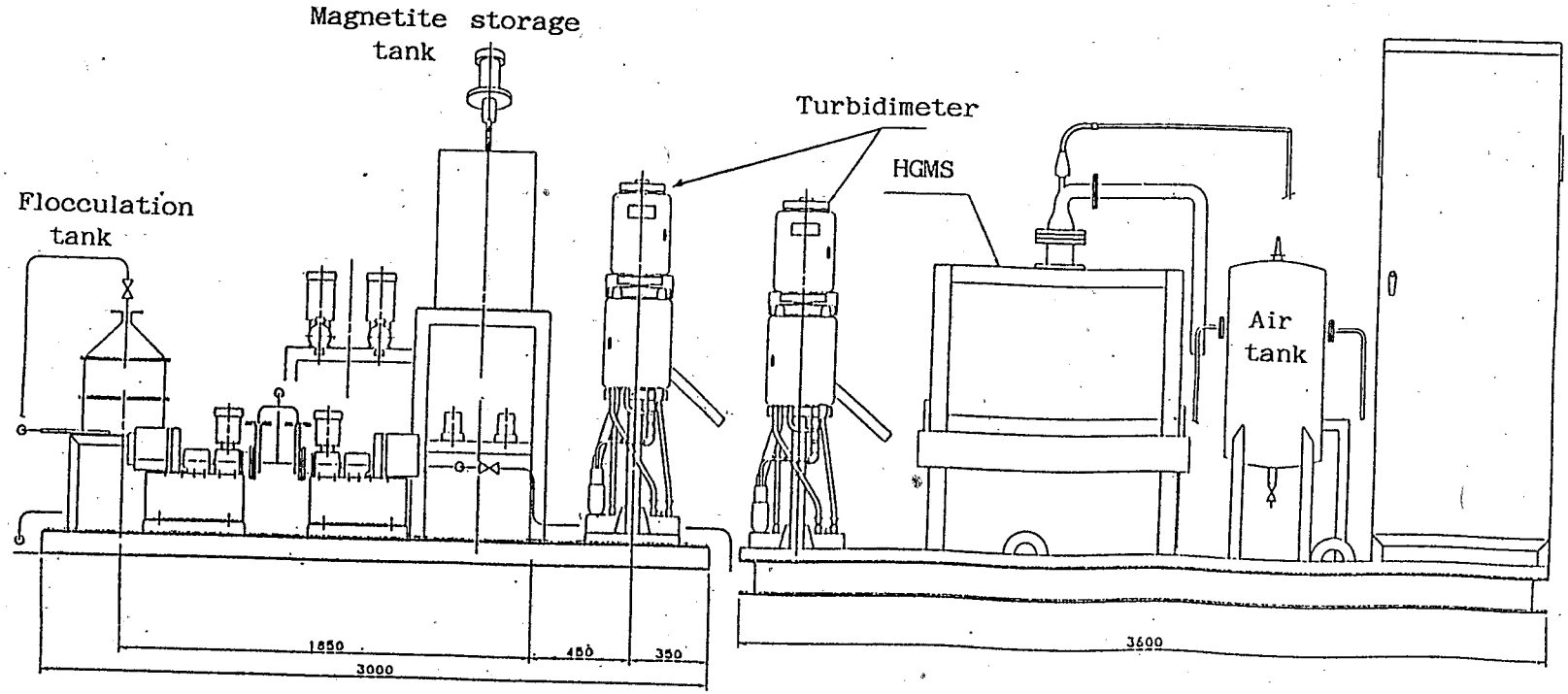

Fig.22. Side view of the pilot plant for tertiary treatment.
(HGMS : High Gradient Magnetic Separator)

Table 4. Results of continuous operation of pilot plant for tertiary treatment.

	Influent	Effluent	Removal rate
pH	7.8	7.4	-
Transparency	34	more than 150	-
Chromaticity	29	17	41%
SS (mg/l)	12	3.6	70%
COD^{Mn} (mg/I)	19	11	42%
BOD^{5} (mg/l)	7.5	2.3	69%
Coliform bacteria (No./ml)	340	14	96%
Total phosphorus (mg-P/ml)	1.7	0.33	81%

(Operating conditions)

Magnetic Field Strength : 3KG
Flow Velocity : 2000 m/day
Magnetite Conc. : 10 mg/l
Alum Conc. : 4 mg-Al/l
Polyelectrolyte Conc. : 1.5 mg/l

Table 5. Characteristics of sludge generated from tertiary treatment process.

Raw sludge

pH	6.9
SS	440 mg/l
VSS	200 mg/l
SVI	102

Thickened sludge

SS	9800 mg/l

(gravity thickening for 30 minutes)

Dewatered sludge

Water content	85%
Calorific value	2380 cal/g-dry weight

(Buchner funnel test with 760 mmHg for 15 minutes, no coagulant addition)

4. PRELIMINARY DESIGN OF DEMONSTRATION SIZE AND FULL SCALE PLANTS

Based on the results of the present study, preliminary design was carried out for a demonstration plant and a full scale plant which are used for combined sewer overflows treatment in wet weather and tertiary treatment of activated sludge process effluent in dry weather.

Fig.23 shows the flowsheet of the demonstration plant with capacity of 10000 m^3 /day. The plant consists the apparatus for dosing and flocculation, two preseparators of 2.75 m in diameter, and high gradient magnetic separator of 2.5 m in diameter. Preseparators are used only for treatment of wet weather flow. In dry weather, flocculated slurry is fed directly to the magnetic separator through by-path. Fig.24 shows a layout of the demonstration plant. Land required of this plant is 165 m^2. Fig.25 shows a layout plan of the full scale plant with treatment capacity of 100000 m^3 /day. The required area of this plant is estimated about 860 m^2. The preseparators occupied large area in the plants, hence omit of the preseparator is desirable from the view point for reduction of land required. Thickness of the matrix , however, must be increased considerably to omitt the preseparator, which increases electrical power consumption of the magnetic separator. Furhter careful consideration should be given to determine wheather adoption of the preseparator is profitable or not.

5. SEEKING FOR LOW PRICE SEEDING MATERIAL

Seeking for low price seeding material will be important from the view point of running cost reduction. In the present investigation, magnetite recovered from coal ash was tested as one of the promising low price seeding material. In power plants in Japan, coal was adopted as energy source in addition to oil and nucleous. Ash derived from some kinds of coal used in those power plants contains magnetite particles. Recovering technique of these magnetite particles by magnetic separator has been almost established. Preliminary pilot plant test showed that the coal ash magnetite was suitable for seeding material. In addition to the coal ash magnetite, industrial wastes from magnetite production plant and iron mill are seemed to be hopeful for utilizing as seeding material.

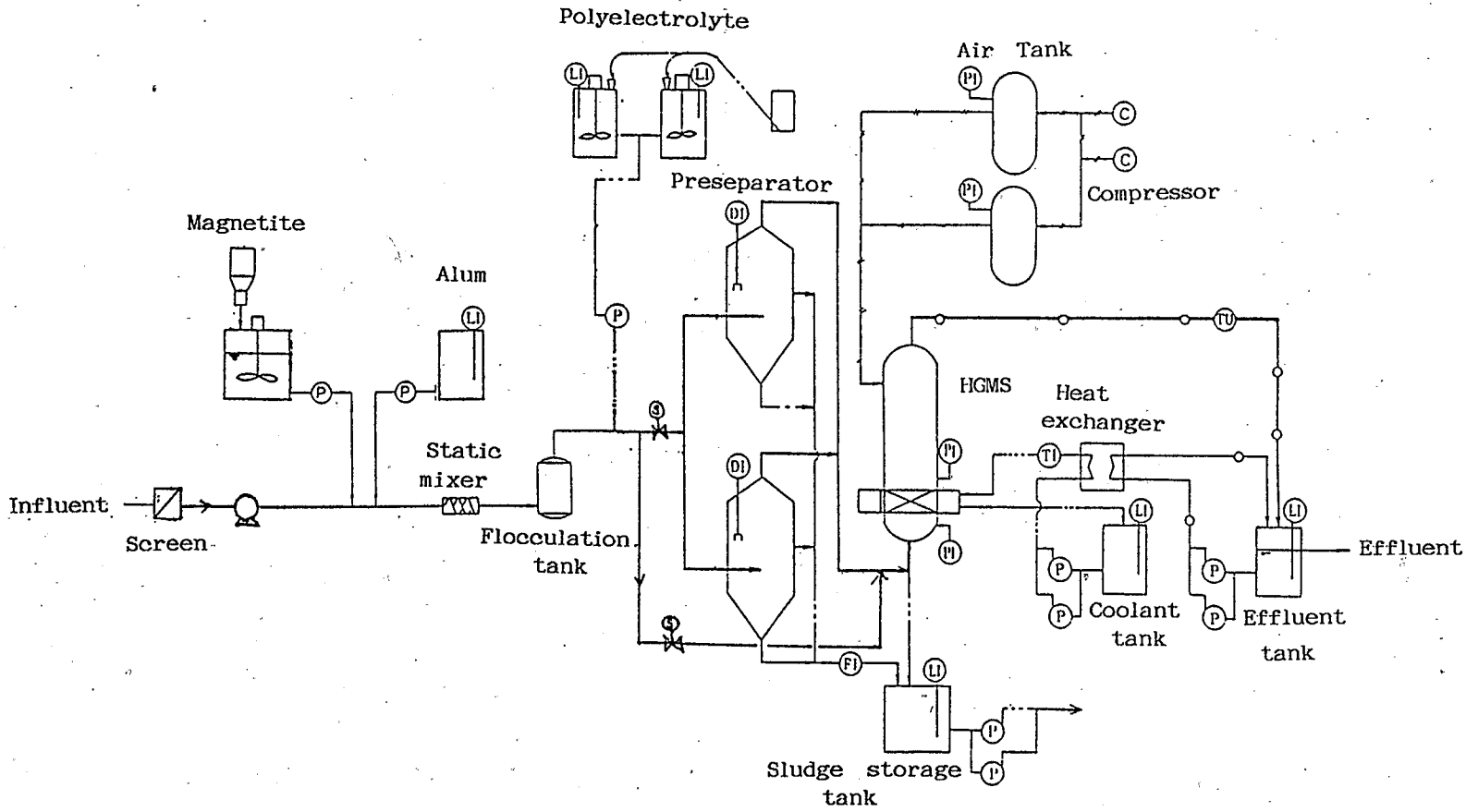

Fig.23. Flow sheet of demonstration scale plant for combined sewer overflows treatment and tertiary treatment.
(HGMS : High Gradient Magnetic Separator)

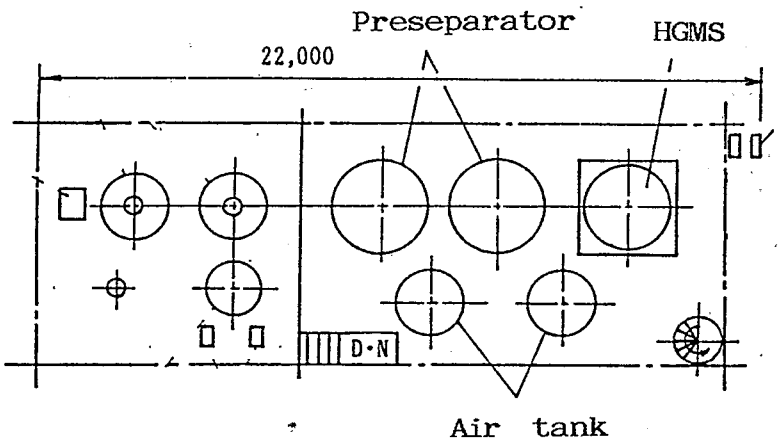

Fig.24. Arrangement plan of demonstration scale plant. (10000 m^3 /day)

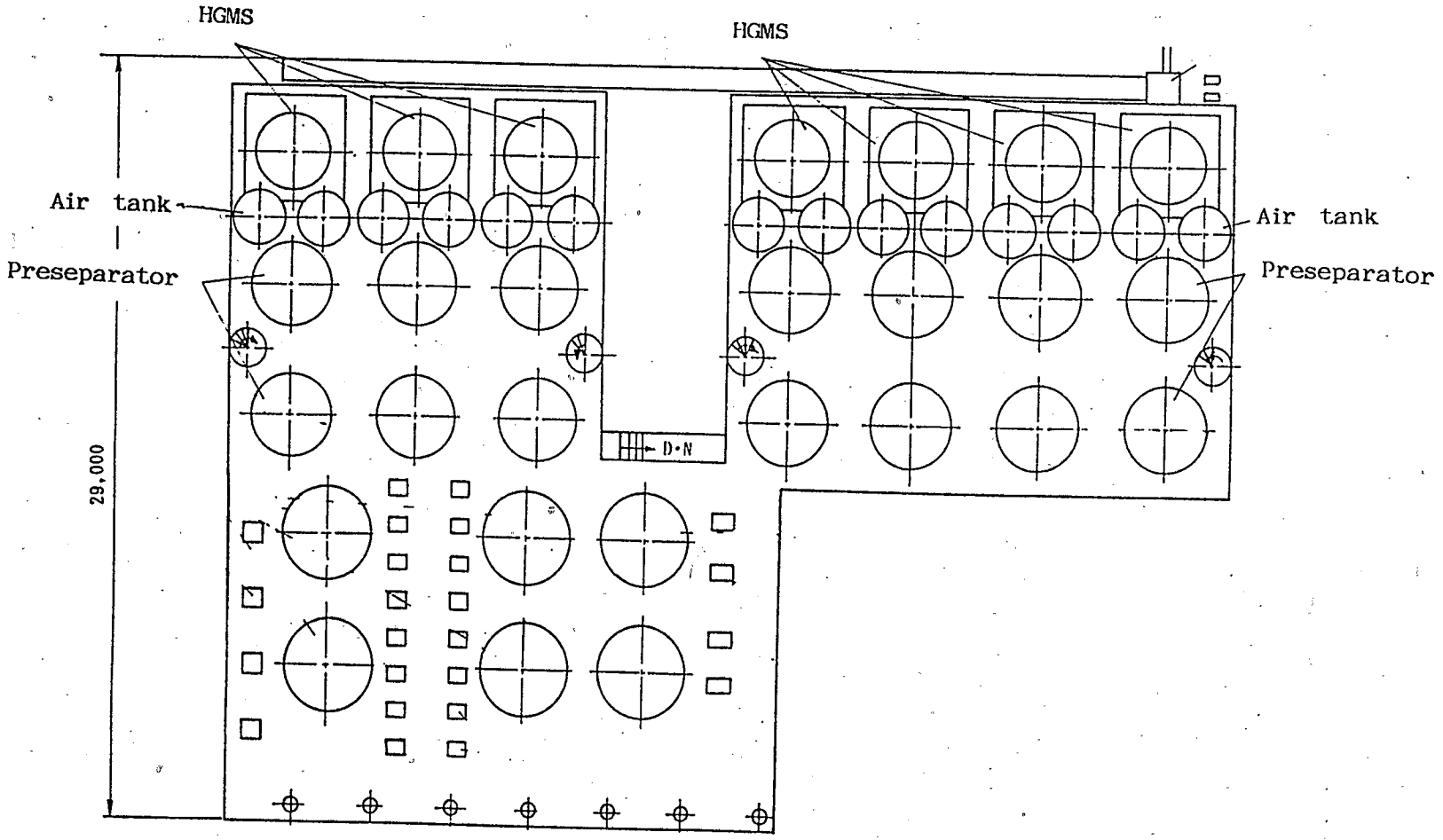

Fig.25. Arrangement plan of full scale plant.
(100000 m^3 /day)

6. CONCLUSIONS

The following conclusions were obtained from the bench test and pilot plant test of seeding method on the treatments of combined sewer overflows and secondaly effluent.

(1) Seeding method was effective for both combined sewer overflows treatment and tertially treatment.

(2) Optimum operating conditions of high gradient magnetic separator were 2000 m/day for flow velocity and 3 KG for magnetic field strength.

(3) Optimum concentrations of chemicals for combined sewer overflows treatment were 25 mg/l for magnetite, 3 to 7 mg-Al/l for alum and 1 to 2 mg/l for anionic polyelectrolyte.

(4) Optimum concentrations of chemicals for tertially treatment were 5 mg/l for magnetite, 2 to 4 mg-Al/l for alum and 0.5 to 1 mg/l for anionic polyelectrolyte.

(5) Sludge produced from the seeding process was easy for thickening and dewatering.

(6) Magnetite which was recovered from coal ash was possible to utilize as seeding material.

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NEW MOVEMENT IN SEWERAGE CONSTRUCTION PROJECTS IN JAPAN

by

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ABSTRACT

Huge investment in construction of public sewerage systems has been made in Japan based on the five year programs. Overall percentage of the sewered population still remains about 40%, but that in major cities has reached almost 90%. Under this situation, the scope of sewage works in Japan has been expanding to put more emphasis on such problems as more safety from inundation in urbanized aeras, combined sewer overflow problem, advanced wastewater treatment, wastewater reclamation and reuse, beneficial utilization of sewage sludge, multi-purpose utilization of sewers and treatment plant sites and so forth. This paper summarizes these movements in Japan.

INTRODUCTION

Human wastewater in Japan is basically treated by three different systems. These are public sewerage, septink tanks and collective human excreta treatment plants which treat human excreta collected by tank trucks. Public sewerage systems currently covers about 40% of the total population. Human wastewater from approximately 30% of the population is treated by septic tank systems. The remaining 30% is treated at collective excreta treatment plants. Most of the existing septic tank systems, as well as collective excreta treatment plants, treat only human excreta. Grey water from households served by these systems is discharged without any treatment.

Therefore, necessity of public sewerage systems is still very high. The sixth five year program for construction of sewerage systems plans to invest 12.2 trillion yen in construction of public sewerage systems during the period of five years from 1986 to 1990. In FY1989, the investment is to be approximately 2.3 trillion yen. About one third of the total investment, 800 billion yen is appropriated as a government subsidy.

In big ten cities in Japan, percentage of sewered population has reached almost 90% on the average, although there still remain problems to be solved. One of the problems is storm water discharge including the combined sewer overflow problem. In many places, the capacity for storm water discharge is not sufficient, and safety from inundation needs to be improved. Various measures to cope with this problem are beginning to be undertaken. In big cities where sewage works have rather long history, combined sewer systems are generally adopted. The combined sewer overflow is another problem which needs to be improved.

As sewerage systems with a secondary treatment level are installed, the next target of providing advanced wastewater treatment has become a practical one in many areas, particularly in the highly loaded areas or in the river basins where wastewater is finally discharged into lakes or other stagnant bodies of water. People's wishes to have better water environment and more water space amenity have been stronger than ever.

It has also been required to seek for additional functions in sewerage systems, such as wastewater reclamation and reuse, beneficial utilization of sewage sludge and multi-purpose uses of sewers and treatment plant sites. Interesting progresses have been made in these fields, and the movement toward this direction will be promoted further in the future.

Wastewater facilities in small communities are generally very poor. A number of small scale facilities will have to be constructed to improve the situation. Therefore, one of the current major concerns is to develop suitable technology for constructing economical and appropriate facilities in small communities.

1. STORMWATER DISCHARGE

1.1 More Safety from Inundation

Measures to prevent stormwater inundation in cities are generally based on improving sewer pipes and pumping stations to discharge stormwater to nearby rivers and the sea as quickly as possible. However, when such methods cannot be adopted because of the poor capacity of the river receiving the discharge and so forth, the following methods have to be applied.

(1) Bypath sewer

A bypath sewer line functions to divert stormwater flow to a river with a greater capacity, or directly to the sea, to cut the storm discharge in the upperstream region. This occurs when urbanization in the upperstream region has led to the occurence of inundation, but improvement in the sewer pipes and in the lower reaches of the river is difficult. One example is the Hirao-Takamiya region in Fukuoka City, where rapid urbanization caused frequent inundation in the lowlands, and it was difficult to improve the storm trunk sewers and rivers in the downstream region. However, a portion equivalent to two-thirds of the basin area, or 212 ha (34 m^3/sec), which equals the deficient capacity, will be cut. The stormwater is to be discharged directly to a river through the Hirao-Takamiya Trunk Sewer (3 to 4 m diameter, 2.5 km length) constructed along an abandoned railway line. The sewer is currently under construction.

(2) Stormwater retention tank

A stormwater retention tank functions to cut the peak discharge by holding the stormwater temporarily, and it may be constructed either on the ground or under the ground. Stormwater retention reservoirs have been constructed in many housing complexes in order to control the increase in storm runoff. Similar facilities have been built in sewer systems, too. The Jiyugaoka retention reservoir built in a residential area east of Nagoya City is a good example. The reservoir has a capacity of 10,000 m^3, and its top is used as a playing ground for a primary school. In urban areas where land use is very intensive, public spaces (such as roads and parks) are used to construct underground retention tanks. Good examples are the Imazu retention tank (26,000 m^3 capacity) in Osaka City, which was built under a baseball ground, and the Takatsuji retention tank (30,000 m^3 capacity) in Nagoya City, which was built under a tall residential building and a park. In addition to such large facilities, many small retention tanks have been constructed to cope with localized inundation disasters in urban areas.

In an area of Nagoya City, the citizens filed a suit against the city, claiming that the city did not respond properly. The court case lasted for 9 years. When a retention tank of 2,000 m^3 capacity was constructed under this street and inundation was successfully prevented, the citizens withdrew the suit.

A stormwater retention tank is an extremely effective means of preventing inundation. However, there remains the question of whether space for a stormwater retention tank can be ensured in congested urban areas. In the future, it will be necessary to examine the more effective utilization of public spaces, such as parks and schools.

(3) Large-scale storm trunk sewer

Where inundation is highly likely to occur in catchment basins, and the capacity of the receiving rivers is critically short, large-scale storm trunk sewers can be built to both bypass and hold stormwater. These may therefore be called "second rivers".

In Osaka City, such a facility, called the "Tennoji-Benten Trunk Sewer" (80,000 m^3 capacity), is already in service. In addition, the "Hirano-Suminoe Trunk Sewer" is under construction. Tokyo Prefecture is planning to construct a large-scale storm trunk sewer under the Loop 7 highway, and is expected to prevent inundation in the Kanda River and Shakujii River basins.

(4) Infiltration of stormwater

Along with the above measures to reserve and divert stormwater, various attempts to reduce stormwater discharge by the infiltration of stormwater have been made in many municipalities. This method maximizes infiltration into the ground by providing infiltration in yards, porous pavements, infiltration curbs and infiltration inlets. Furthermore, the runoff of stormwater flowing into pipes is delayed through pipe storage and pipe route bypassing.

1.2 Combined Sewer Overflow Problem

Combined sewer overflow should be basically controled by ensuring an appropriate intercept volume to a treatment plant. When this is difficult, measures based mainly on storage have been adopted. In addition, since storage and infiltration as inundation prevention measures are also effective in reducing the overflow volume, qualitative as well as quantitative measures are under examination.

(1) Sedimentation/storage of stormwater

The first facility for stormwater sedimentation/storage to reduce the overflow load was constructed at the Nakanoshima Pumping Station in Osaka City in 1974. Since then, many such facilities have been constructed. Most of them have the capacity to store a 5 mm rainfall, on average, within a drain basin.

(2) Treatment of wet weather sewage

The water volume and water quality of wet weather sewage fluctuates enormously and requires intermittent treatment. Consequently, the application of a biological treatment process is difficult. Thus, currently planned methods are based on physical-chemical treatment processes.

One of them is a swirl concentrator, which is characterized by having no moving parts and by ease of maintenance. Tokyo Prefecture and Matsuyama City are currently constructing swirl concentrators.

As for wet weather sewage treatment, plate settlement using chemical additives, and a super high-rate filtration, are under technological development.

2. ADVANCED WASTEWATER TREATMENT

2.1 Necessity of Advanced Wastewater Treatment

The sewage system plays an important role in preventing water pollution in public waters, and in preserving precious water resources. Depending on the conditions of the basin secondary treatment is not sufficient to protect water environment, making advanced treatment necessary. Currently, advance wastewater treatment is undertaken mainly in the basins of the rivers which are used as important sources of water supply, and lakes, in order to meet the water quality standards as quickly as possible. Typical lakes in Japan (including Lake Biwa and Lake Kasumigaura), have become considerably eutrophicated, so that nutrient removal is required.

2.2 Present state of Advanced Wastewater Treatment

The advanced treatment is under way at 22 sewage treatment plants throughout Japan. Each of them aims to meet or keep meeting the water quality standards. In some plants discharging their effluent into Lakes, biological denitrification and chemical addition into aeration tanks are practiced to remove nitrogen and phosphorus. As for advanced treatment processes in plants discharging their effluent into rivers, rapid filtration and chemical clarification are used to get higher removal of biochemical oxygen demand (BOD) and suspended solids (SS).

2.3 Future Direction

Advanced wastewater treatment has been practiced in areas where such treatment is in great demand. People's wishes, on the other hand, to have a better water environment and more water space amenity have become stronger and stronger. It is considered necessary, therefore, to take a policy to adopt advanced wastewater treatment more actively. There will be an increasing demand for advanced wastewater treatment, particularly at the following areas:

(1) Improvement of water environment in big cities and their surrounding areas

The sewage service has made great progress in big cities and their surrounding areas. However, due to too much concentration of pollutant loading, some rivers in the urbanized areas cannot be fully restored in water quality by secondary treatment. This may necessiate the employment of advanced treatment in the catchment areas of these rivers. In addition, in some big cities where water shortage is always a potential problem, the treated sewage has become precious water resources. The necessity of advanced treatment will further increase in relation to reuse of treated water. For example, many urban streams have experienced a reduced flow with the progress of urbanization. The restoration of stream flow to these rivers is greatly desired. For this purpose, the utilization of sewage water treated by the advanced treatment process is being planned in many municipalities.

(2) Protection of water resources for drinking water supply

In planning a sewerage project and examining the effect of effluent discharge on the water uses in the downstream reach, special attention has been paid to water intakes for drinking water supply. When the receiving river is used as a source for drinking water supply at the downstream, advanced treatment of sewage has been practised depending on the necessity.

Recently, people are demanding more safety in drinking water, and calling for better "taste" of drinking water. In this regard, necessity of advanced treatment is becoming higher than before.

(3) Prevention of eutrophication of lakes and reservoirs

Preventing eutrophication of lakes and reservoirs has become a significant problem these days. Effluent standards regarding phosphorus are established at 1066 lakes, and those regarding nitrogen at 78 lakes. Most of these standards do not necessarily require higher level of treatment than secondary treatment for domestic sewage. However, to restore a beautiful water environment and maintain good water quality for the lakes and reservoirs advanced treatment of sewage is the most effective measure to be taken, and it will be more significant in the future.

(4) Control of eutrophication of bays and the inland sea

Many big cities in Japan including Tokyo, Osaka and Nagoya are situated on bays and the Seto inland sea where exchange of water is little. Eutrophication of these seas has already intensified. To restore the water quality, nutrient loading should be minimized. At present, no quality standards concerning nutrients in the seas have yet been set forth. And, therefore, actually no advanced wastewater treatment is practised to remove nutrients.

However, more and more people are considering it quite necessary to restore water quality in the seas near big cities. In these areas, too, the necessity of advanced treatment is advocated.

3. MULTI-PURPOSE UTILIZATION

3.1 Reclamation and Reuse of Sewage

(1) Reclamed sewage as precious water resources

The annual demand for domestic water in Japan has been increasing by an average of 3%. In order to meet the increasing water demand, continual effort to develop new water resources should be made. At the same time, quality of water resources must be ensured.

The total volume of the effluent from about 700 sewage treatment plants in Japan reached approximately 8 billion cubic meters in FY1987. This volume corresponds to approximately one-fourth of the municipal water supply (the total of water for domestic and industrial water). Most of the effluent is discharged to the rivers and seas, although some is reused for various purposes.

With the expansion of industrial activities and the concentration of population into big citites, the municipal water demand is still increasing these days. Consequently, a system to reuse reclaimed sewage is being introduced as a means to cope with an intensive water demand in urban areas, and to achieve a stable water supply. Sewage can be a stable and abundant water resource obtained in a city, and its volume is expected to increase with the expansion of the sewerage systems.

(2) Model project for sewage reclamation and reuse

The Ministry of Construction has launched a model project for sewage reclamation and reuse since FY1979 in order to facilitate reuse of treated sewage. This project aims to supply reclaimed sewage to corporate buildings in a certain district as toilet flush water after implementing a necessary treatment of the reclaimed sewage. Currently, the projects are under way in eight districts throughout Japan. These include the Tenjin district of Fukuoka city, where the citizens suffered severe water shortage for over 287 days in 1978, and the subcenter of Shinjuku in the metropolis of Tokyo, where skyscrapers are abundant. Among them, five projects are already under operation. In FY1989, the Tokyo Metropolitan government is planning to start the project in a coastal subcenter.

(3) Flow augmentation, and ornamental streams and lakes

Many of the urban streams in large cites in Japan lose their flow in dry weather due to an increase in the impermeable area within their basins as the result of urbanization. On the other hand, people in cities have come to put more and more emphasis on a better environment, particularly an environment with clean water. Reclaimed water is a stable water resource within cities, and thus, there have been many projects to utilize reclaimed water for restoring live streams and creating clean water bodies within cities.

Restoring live streams utilizing reclamed water was practised at 6 locations in 1986. Among these 6 locations, reclaimed water was discharged year-round at 4 locations. The outline of these 4 projects is shown in Table 1.

Table 1. Typical examples of restration of live streams using reclaimed water

Name of city	Name of STP	Reclaimed Water flow (m^3/day)	Name of stream	Reclamation process
Tachikawa	Nishikimachi	212	Nekawa	Sand filtration
Metro. Tokyo	Tamagawa-Jouryu	20,720	Nobidome and Tamagawa	Sand filtration
Osaka	Nakahama	3,000	Outer Moat of Osaka Castle	Sand filtration
Osaka	Hirano	7,074	Imagawa and Komagawa	Sand filtration

In addition, a model project named "restoration of live streams in stormwater drains" was initiated in 1985. Most of the open storm sewers used to be natural storm drains. This project is to restore live streams in these storm sewers utilizing reclaimed water. Currently 8 projects of this type are being carried out as shown in Table 2.

Table 2. Effluent reuse model projects "Restoration of live streams in storm water drains"

Year of project start	Name of prefecture	Name of city	Name of storm drains	Design flow (m^3/S)
1985	Kyoto	Nagaokakyo	Shoryuji	0.06
1985	Oita	Oita	Nakajima No. 1	More than 0.1
1986	kanagawa	Yokohama	Saedo	0.12
1986	Osaka	Toyonaka	Houno-Nanbu	0.1
1987	Kanagawa	Yokahama	Terao	0.12
1987	Aichi	Nagoya	Arakogawa	0.23
1988	Aichi	Nagoya	Shirako-Sakae	0.002
1988	Hyogo	Nishinomiya	Edakawa No. 2	

Furthermore, reclaimed water from sewage treatment plants is used at many locations for ornamental lakes and streams within treatment plants and in parks nearby.

(4) Snow disposal

In order to remove and dispose of snow in urban areas including private lots, it is necessary to provide suitable disposal sites. Open channels, including open storm sewers, are often used for this purpose. A problem usually encountered is the lack of sufficient flow to flush away the disposed snow.

The effluent from sewage treatment plants has good characteristics for this use, such as being stable in flow, being available within a city, and having a relatively high temperature in winter.

A model project named "The rapid removal of snow in urban areas using reclaimed water" was initiated in 1985. Currently nine projects are being carried out in cities in heavy snow areas, as shown in Table 3.

Table 3. Model effluent reuse projects "Rapid removal of snow in urban areas using reclaimed water"

Year of project start	Name of prefecture	Name of city	Name of drains storm utilized	Design reclaimed water flow (m^3/S)
1985	Hokkaido	Ebetsu	Ebetsu No. 32	0.15
1985	Aomori	Aomori	Namiuchi	0.12
1986	Akita	Akita	Kusouzugawa-Sagan No. 1	0.12
1986	Niigata	Tokaichi	Inaricho and Honmachi	0.04
1987	Hokkaido	Sapporo	Yasuharu	0.5
1987	Niigata	Yuzawa	Nunoba No. 1	0.238
1987	Ishikawa	Kaga	Jouganji	0.03
1987	Fukui	Tsuruga	Maizaki-Shimizu	
1988	Ishikawa	kanazawa	Ekiura	

Another project is being planned in Sapporo City to utilize a stormwater reservoir in winter time as a disposal site for the removed snow. The disposed snow is melted away by introducing effluent from a sewage treatment plant into the reservoir.

Snow removal in snowy area is equivalent to stormwater drainage in other areas. More contributions to sewage works from this field are expected.

3.2 Beneficial Utilization of Sewage Sludge

(1) Necessity of effective utilization of sewage sludge

Wastewater treatment is accompanied by the generation of sewage sludge. As the sewered population increases, the volume of sewage sludge generated increases. Advanced treatment may increase sludge production in the future. In 1987, when the percentage of sewered population was 39%, the sludge production in dry solids (DS) was approximately 1.25 million tons, and the annual volume of sludge disposed of was 2.1 million m^3. When the percentage of sewered population reaches 50%, it is estimated, that the sludge production will be approximately 1.7 million tons in DS, and that the annual volume to be disposed of will be 3.7 million m^3.

On the other hand, the sites for landfill are limited. As urbanization progresses, the acquisition of landifill sites is getting more difficult. And the associated costs are increasing. Therefore, further efforts must be taken to reduce the volume of sludge, and to facilitate beneficial utilization of sludge as resources.

Table 4. Status of sewage sludge disposal. (1987)

(Units: 1000 m^3)

Disposal method / Treatment	Landfill	coastal reclamation	Beneficial utilization	Others	Total (%)
Dewatered cake	762	306	274	114	1,456 (68)
Incinerated ash	156	105	43	6	310 (15)
Dried sludge	17	4	126	19	166 (8)
Digested and thickened sludge	7	0	0	191	198 (9)
Total (%)	942 (44)	415 (19)	443 (21)	330 (16)	2,130 (9) (100)

Table 5. Status the beneficial utilization of sewage sludge (1987).

(Units: 1000 m^3)

Classification	Treatment	Dewatered cake	Incinerated ash	Dried sludge	Composted sludge	Digested sludge	Total
Green farmland	Committed by the municipality	117	0	7	80	0	204
	Assigned to a fertilizer company	157	7	15	24	0	203
	Subtotal	274	7	22	104	0	407
Use as a construction material		0	36	0	0	0	36
Total		274	43	22	104	0	443

(2) Current status of beneficial utilization

There is a wide range of applications, such as for farmlands and green spaces, and for construction materials. Recovery of energy is another example of beneficial utilization.

The amount of sludge utilized for farmlands and green spaces in the form of dewatered sludge cake, composted sludge, and dried sludge, comes to 11.5% of the total amount of sludge generated or 140,000 tons (DS) per year. The amount of sludge (mainly as incinerated ash) utilized as construction materials is estimated to be 6.5% or 85,000 tons.

(i) Utilization for farmlands and green spaces

The amount of sewage sludge utilized for farmlands and green spaces was 407,000 m^3/year in 1987. The amount has been fairly constant in the last few years. Most of the sludge utilized for this purpose is applied to farmlands. Only a small amount of sewage sludge is used for green spaces. Sewage sludge is mostly used in the state of dewatered sludge cake. However, more and more composted sludge is being utilized recently. Sewage sludge is more efficiently utilized for farmlands and green spaces in Europe than in Japan.

Currently, 21 municipalities are running sewage sludge composting plants. The product is sold in bulk, or 20 kg package. The retail price varies greatly, depending on the districts. The heavy metal standards for the sludge utilized for farmlands are that the maximum concentrations of mercury, arsenic, and cadmium shall be 2, 50, and 5 mg/kg, respectively. In addition, a tentative guideline has been worked out, specifying that sludge application is allowed in the areas where the zinc content of the soil is less than 120 mg per kg dried soil. These values are rather strict compared with those stipulated in European nations. In particular, the guideline for zinc in soil is very strict in terms of the zinc concentration in sludge or soil. This is an obstacle, in a sense, to distributing composted sludge in Japan.

(ii) Utilization for construction materials

There are two utilization methods: One is the direct utilization of incinerated ash or melted slag, and the other is to use incinerated ash or melted slag as a part of the raw materials to produce building materials. Examples of the former include roadbed material for pavement, back-filling material for retaining walls, and intermediate cover for sanitary landfills. Recently, incinerated ash has been used to improve soil characteristics of the excavated material from sewer construction sites. In order to use the excavated material for refilling, it is often required to increase the kinetic strength by mixing an additive. Incinerated ash has good characteristics as an additive for this purpose. This is a newly development in the utilization of incinerated ash.

In the latter utilization of incinerated ash or melted slag, these materials can be used as raw materials for producing such products as ceramic pipes, light-weight aggregates, bricks, tiles, and permeable blocks. Some of these raw materials are manufactured on a commercial basis. The amount of sludge used for construction materials is limited at present, but is expected to increase with the encouraged employment of the melting process.

(iii) Recovery of energy

Anaerobic digestion are carried out mainly to reduce and stabilize sewage sludge. However, energy recovery by digester gas power generation, recently has been another motivation in intalling sludge digesters. Approximately 30% of the methane gas energy is converted to electric energy by digester gas power generation. When the waste heat from the engine is recovered in the form of hot water, 70 to 80% of the methane gas energy in total can be restored. Currently, 14 sewage treatment plants in Japan are undertaking digester gas power generation. This contributes to reduce the total power consumption in each sewage treatment facility by 5 to 28%. In addition, a new project is being undertaken to make dewatered sludge solids through a special procedure, and to use the sludge as fuel. A project to produce oil from sludge is also under development.

3.3 Multi-purpose Utilization of Sewers and Treatment Plant Sites

(1) Utilization of treatment plant sites

Treatment plant sites and pumping stations are precious open spaces in an overcrowded city. Multiple utilization of the area permits effective use of the land and the realization of pleasant urban scenery, thus providing the neighborhood with a place of recreation and relaxation. This kind of planning is becoming popular these days.

To be more specific, sporting facilities including tennis courts and baseball grounds, as well as parks and green zones are being constructed in many sewage treatment plants. In some cases, sewage treatment plant sites including the covered top of the facilities are used as a refuge at the time of disaster. In addition, buildings such as music halls and martial art gymnasiums may also be constructed above plant sites. This new trend of utilization is becoming noticeable.

(2) Effective utilization of sewers

Open stormwater channels have been used as places to commune with water by restoring the clean shallows and constructing a promenade. This kind of model projects have been carried out at many locations, and are popular among local residents.

On the other hand, underground sewers have not been used efficiently due to their structural properties. However, optical-fiber cables have recently been laid along the sewers to fabricate an information network. Since a network of sewers is spread all over a city, this utilization of sewers will become more intensive in the future.

4. SMALL-SCALE SEWERAGE SYSTEMS

In Japan, there are 2817 communities with a population of less than 50,000, in which the percentage of sewered population is as little as 6% on the average. Approximately 2200 of these communities have not started constructing sewage systems yet. When planning sewerage systems in these regions, difficulties are sometimes encountered with selecting a plant site because the community is located in a mountainous area, or on the sea, or the local residents fail to reach a consensus of opinion. In such a case, a cost-effective, small-scale sewage treatment plant may be an amicable settlement. Thus, efforts have been made to develop and study the associated technologies.

The following sections will introduce prefabricated treatment plants, floating-type treatment plants, tunnel-type treatment plants, and alternative sewer systems as typical examples of technology development for small-scale sewage treatment plants.

4.1 Prefabricated Treatment Plants

The Japan Sewage Work Agency has studied cost-effective methods of constructing small-scale treatment plants since 1982. Based on the studies, a prefabricated treatment plant has been developed and standardized.

For the prefabricated plant, an oxidation ditch process is adopted. This is because it can withstand inflow load variations, is easy to maintain, and generates less sludge.

The prefabricated treatment plant is a circular structure that a conventional ditch is laid out as an outer circle, with a secondary sedimentation tank at its center. As for the structure, pre-stressed concrete panels manufactured at the factory are assembled at the site. For mechanical and electrical facilities, standardized equipments are used to reduce the costs. To rationalize the design, standardized designs are prepared for the plants with capacities from 300 to 1,200 m^3/day with an interval of 100 m^3/day. Thus, when a certain number of plants have been constructed, the costs of construction as well as design can be considerably reduced. Furthermore, the construction period can be shortened, and the construction work can be easily performed at the site. The first prefabricated treatment plant was completed in Nakanojo, Gunma Prefecture, in 1988. There are currently nine plants in Japan which are under planning or construction.

4.2 Floating-type Treatment Plants (Plants Floating on the Sea)

A number of small communities are located on the hilly sea shores, where it is sometimes quite difficult to acquire suitable sites for constructing sewage treatment plants. In these cases, it has been a common practise to create the land by coastal reclamation. This is sometimes very costly and requires the long period of construction.

The Japan Sewage Works Agency has developed a floating-type treatment plant as an alternative technology, and the first plant of this type was constructed in Atami City, Shizuoka Prefecture in 1985.

Steel made, ark-structure plant constructed at a dockyard is taken in tow to the sea site. The plant is installed at a specified position, connected with the sewer, and then put into operation. The floating plant can utilize either the conventional activated sludge process, oxidation ditch process, or sequencing batch reactor process.

There are two installation methods: Grounding at the bottom of the sea, and moorage at a pier with an anchor. Compared with the method of constructing a plant after reclaiming the sea, the advantages of the floating method are that the construction period can be shortened, and the construction costs could be reduced, particularly for smaller scale plants.

4.3 Tunnel-type Treatment Plants

In areas where it is difficult to find a suitable plant site, one alternative is a tunnel-type treatment plant. In Sweden and other Scandinavian nations, the rock bed is tunneled so that a so-called "tunnel treatment plant" can be constructed inside. In Japan, too, feasibility studies have been made on tunnel treatment plants. Eventually, the first plant will see construction.

Since a treatment plant is built in a tunnel, the plant has a long structure. This is the only difference compared to a plant built above the land. In selecting a treatment process, no significant restrictions are imposed on the tunnel plant.

When a treatment plant is operated in a tunnel, no considerations are required regarding the foundation and covers. In heavy snow areas, where coverage of the facilities is required for maintenance or to prevent the water temperature from dropping, the tunnel method will prove effective.

4.4 Alternative Collection Systems

The cost for sewers occupies 70% of the total sewage construction costs of a sewerage system. Therefore, to reduce the construction costs, sewer construction costs must be minimized.

Conventionally, most Japanese sewer systems are of the gravity type. Now that sewerage systems are constructed in small communities, the gravity sewers are not necessarily economical due to geographical conditions. Consequently, the employment of pressure or vacuum sewers has been actively investigated. A pilot project is underway to acquire data concerning the costs and design. Based on the results, the pressure or vacuum sewers may largely be substituted for gravity sewers in the future.

Innovative Sludge Handling System Using Pelletization for Simultaneous Thickening and Conditioning

by

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ABSTRACT

Many sewage treatment plants in Japan have suffered from difficulties in sludge handling processes, especially in sludge thickening/dewatering processes, because of the changes in sludge characteristics in the past one or two decades.

This paper introduces an innovative sludge handling system, developed by the Japan Sewage Works Agency, to solve those difficulties.

The newly developed sludge handling system employs a metal coagulant aid and an amphoteric polymer that has the characteristics of flocculants that are both cationic and anionic. The system consists of a sludge-charge neutralization reactor and a sludge pelletizing thickener. The sludge from wastewater treatment process is directly fed to the system. In the sludge-charge neutralization reactor, the fed sludge is dosed with the metal coagulant aid to neutralize anionic sludge charge. In the pelletizing thickener, the amphoteric polymer is fed into the charge-neutralized sludge to pelletize and thicken it. The small sludge particles form spherical sludge masses of 1 to 2 cm in diameter, which have excellent mechanical strength and low water content.

The function of this system as a sludge thickener has eliminated the necessity of the conventional sludge thickening process. An SS recovery efficiency of 90%, or higher, is provided by this system. Compared with conventional sludge handling systems, this system achieves a higher filtration rate of 200 kg-DS/m/h and lower dewatered sludge cake moisture content of 75%, by using a belt-press filter.

Phosphate is also physicochemically fixed to the sludge through use of a metal coagulant aid. Hence, this system can be also characterized as a sludge handling system that is suitable for biological phosphorus removal processes.

Innovative Sludge Handling System Using Pelletization for Simultaneous Thickening and Conditioning

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1. Introduction

Sludge handling systems have the following objectives:

1) to reduce the volume of sludge to economize land space for disposal and costs of sludge disposal,
2) to improve the ease of handling during disposal, and
3) to change the characteristics of the sludge to a stable form.

In genreal, the sludge handling systems consist of thickening, digesting, dewatering, and incinerating processes. In the past one or two decades, the characteristics of sewage sludges have changed in Japan. Especially, the VSS/SS ratio of sludge has increased significantly. The causes of these changes in the sludge characteristics are due to the transition from combined sewer collection systems to separate sewer collection systems, and the changes in our way of life.

As a result of the changes in sludge characteristics, many sewage treatment plants have suffered from difficulties in sludge handling, especially, in sludge thickening and dewatering processes.

This paper introduces an innovative sludge handling process, developed by the Japan Sewage Works Agency, to solve those problems in sludge handling systems.

2. Sludge Handling Difficulties in Japanese Sewage Treatment Plants

1) Problems Related to Gravitational Thickeners

The status quo of the sludge handling difficulties are well pointed out in a research report prepared by the Tokyo metropolitan government.[1)]

The chronological changes in the Volatile Suspended Solid to Suspended Solids (VSS/SS) ratio of excess sludges are shown in Figure 1. Figure 1 reveals that the VSS/SS ratio of excess sludges in 1965 was around 60%, and increased to around 80% in 1985. The chronological changes in the solid content of thickened sludges are shown in Figure 2, which reveals that the sludge was thickened to about 5% of the solid content in 1965, and decreased to 3%, or less, of the solid content in 1985. The gravitational thickenability of sewage sludge has decreased as the VSS/SS ratio increased.

The increase in the volume of thickened sludge caused by the decrease in solid content means an increase in the volume of the sludge to be supplied to the dewatering machines. This is one of the reasons for the recent deterioration of the dewatering efficiency.

The chronological changes in the dewatering rate and in the coagulants ($FeCl_3$ and $CaCO_3$) dosing ratio (% to DS), for a vacuum-type dewatering machine, which had been widely employed in Japan are shown in Figure 3. As shown in the figure, the dewatering rate has been decreasing, and the dosing ratio has been increasing.

The changes in the characteristics of sewage sludges such as those in Tokyo, have been observed in almost all Japanese sewage treatment plants. These changes include the gravitational thickenability of sewage sludge, the dosing ratio of chemicals, the dewatering capacity of the dewatering processes, and the moisture content of dewatered sludge cakes.

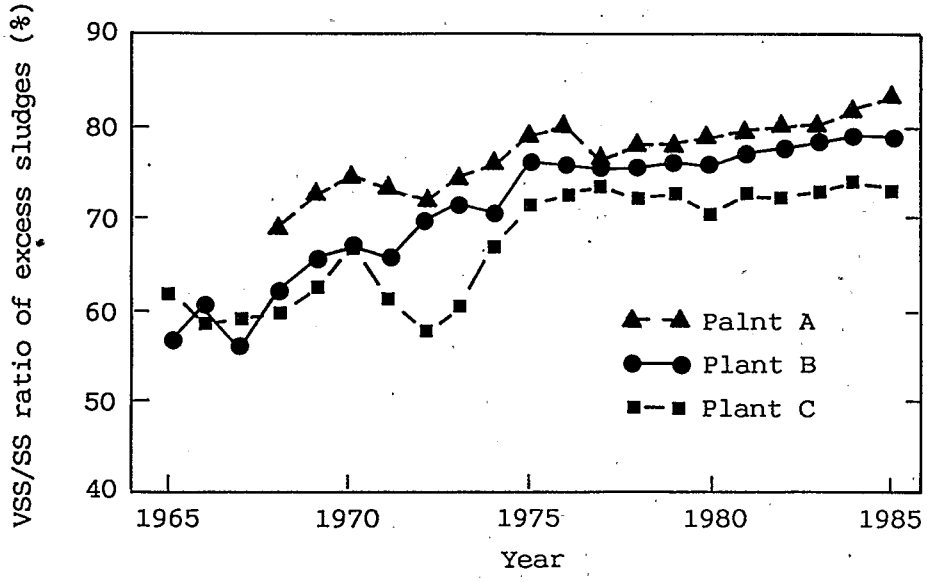

Figure. 1 Chronological change in the VSS/SS ratio of excess sludges. (Tokyo)

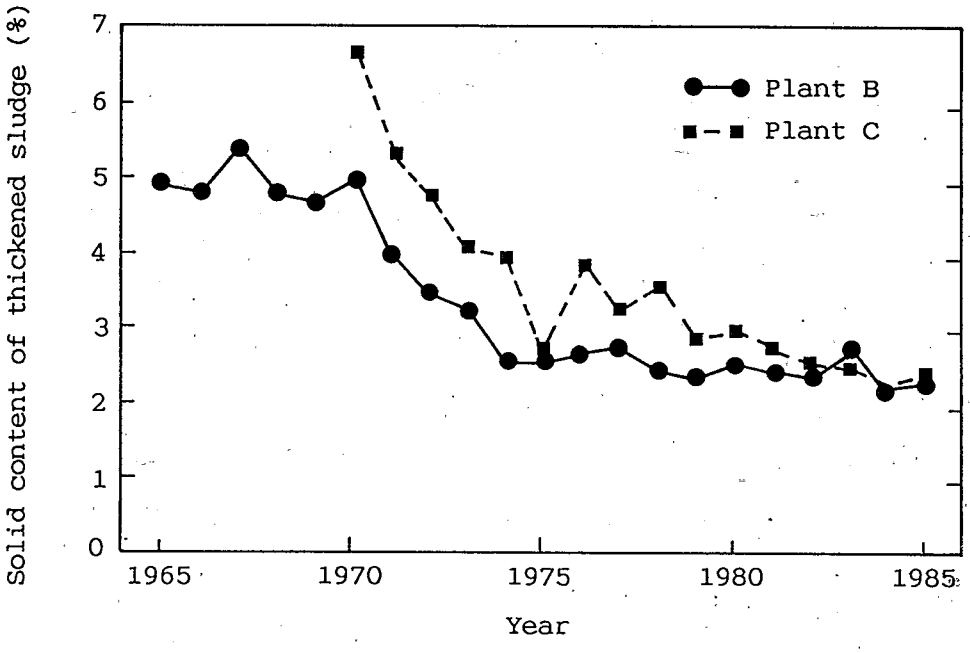

Figure 2. Chronological change in the solid content of graviationally thickened sludges. (Tokyo)

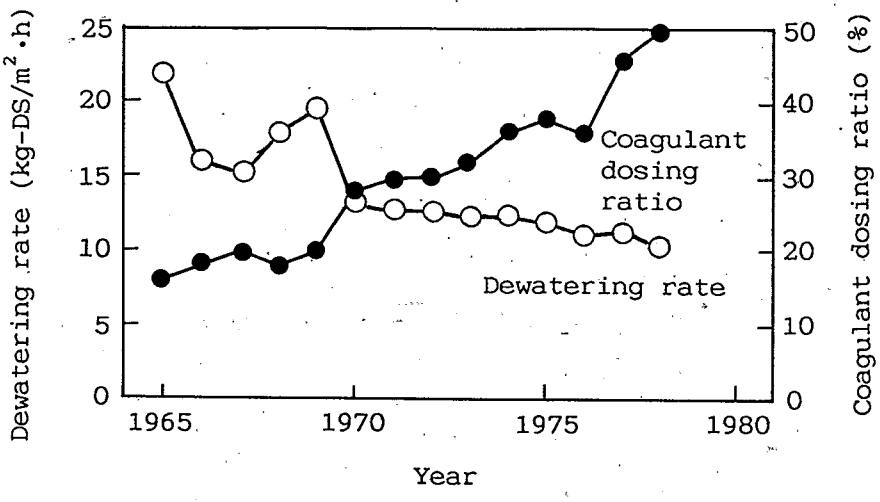

Figure 3. Chronological changes in the dewatering rate and coagulant dosing ratio (with vacuum-type dewatering machine). (Tokyo)

The increase of the moisture content of dewatered sludge cakes results in an increase in fuel consumption in incineration systems. Thus, to improve the performance of the total sludge handling system, it is very important to improve the performance of the sludge thickening and dewatering processes. Therefore, municipalities having difficulties with the sludge thickening processes have started to employ mechanical thickeners, such as centrifugal thickeners, instead of the gravitational thickeners.

2) Problems Related to Biological Phosphorus Removal

It is known that phosphate is released from sludges when a gravitational thickener is employed to handle the excess sludge in a biological phosphorus removal process. This results in a poor total phosphorus (T-P) removal efficiency.

An experimental study on the biological phosphorus removal efficiency was conducted in the S treatment plant. A schematic flow diagram of the plant is shown in Figure 4. The results obtained can well explain the relationship between the phosphate release in a sludge handling process and the effluent T-P. The operational conditions of the plant during the experimental study are as follows.

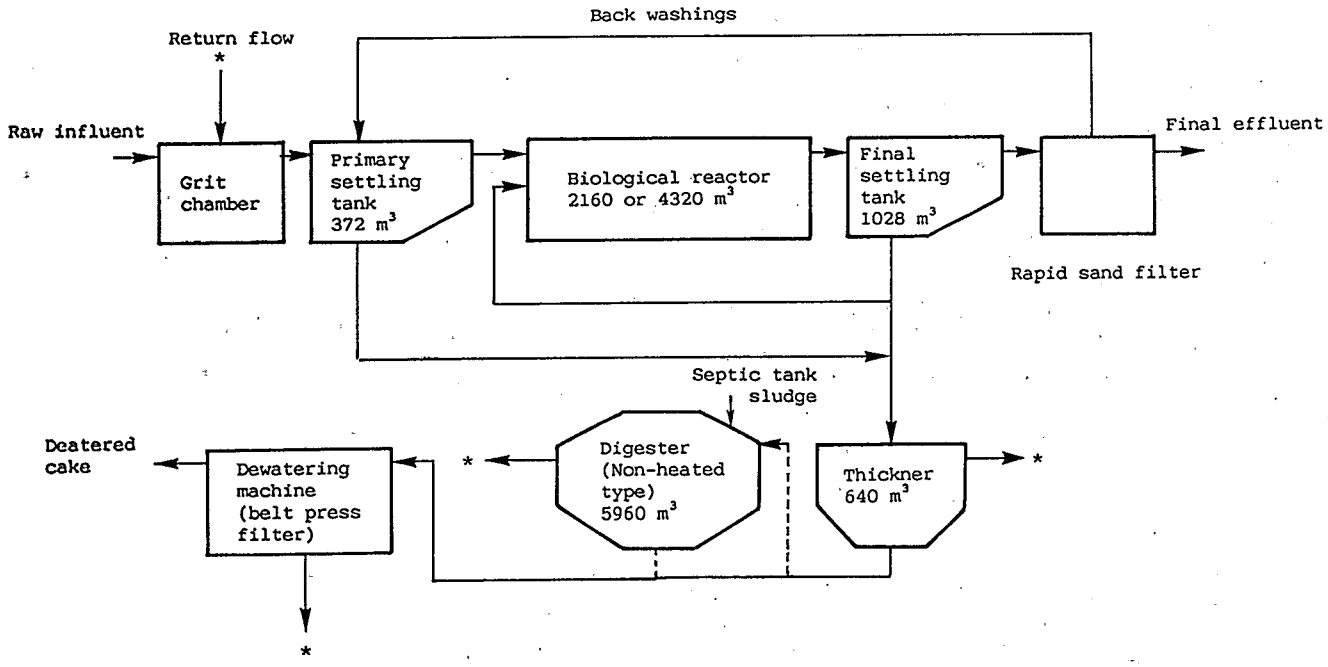

Figure 4. Schematic flow diagram of the S treatment plant

Influent Flow Rate	: 8,000 - 9,500 m^3/day
Hydraulic Retention Time of the Biological Reactor (1/4 was operated as anaerobic)	: 5.5 to 6.5 hours
MLSS	: 2,400 to 3,500 mg/L

The chronological changes in the sidestream phosphorus load ratio*) and the effluent T-P concentration are shown in Figure 5.

The sidestream phosphorus load ratio:

$$= \frac{\text{Sidestream phosphorus load}}{\text{Influent phosphorus load}} \times 100\% \quad (1)$$

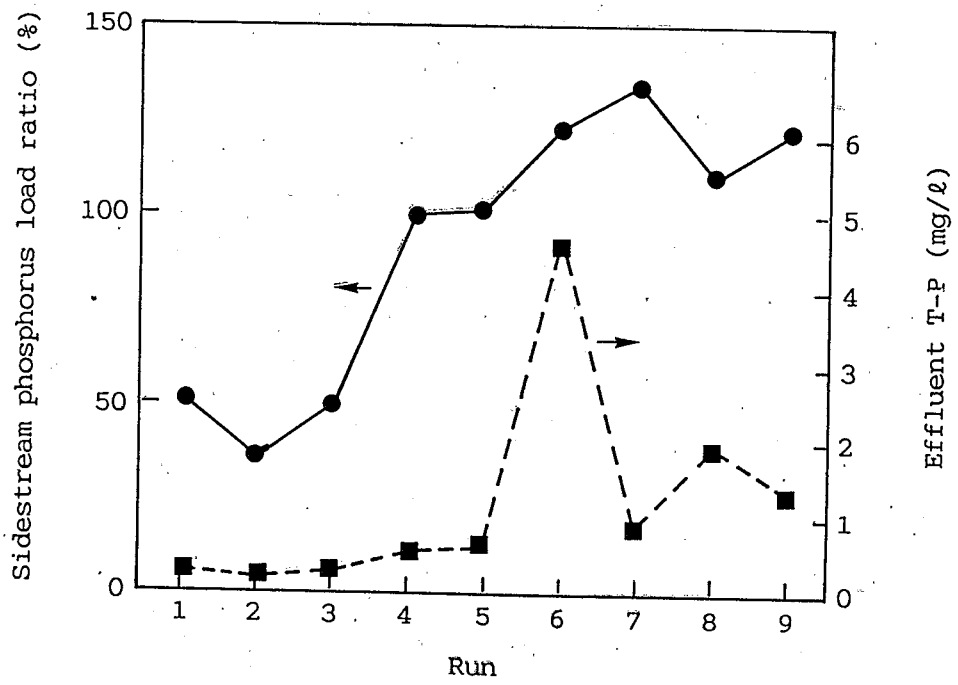

Figure 5. Chronological charges in sidestream phosphorus load ratio and effluent T-P.

As shown in Figure 5, the effluent T-P exceeded 1 mg-T-P/L when the sidestream phosphorus load ratio was more than 50%. The increase of the sidestream phosphorus load ratio resulted in an overloading of the T-P load in the biological reactor, which caused the increase in the effluent T-P.

In a gravitational thickening tank, the sludge can be easily placed in an anaerobic environment, where an active biomass easily releases phosphate. To reduce the sidestream phosphorus load and to achieve a high efficiency of biological phosphorus removal, it is necessary to develop a new sludge thickening process which handles sludge without releasing phosphate.

3. Targets in This Study

The Japan Sewage Works Agency decided to develop an innovative sludge handling system to solve the sludge handling difficulties. The following are its targets.

(1) Elimination of conventional sludge thickening systems

The difficulties in sludge handling systems come from conventional sludge thickening processes. One of the targets was the elimination of conventional sludge thickening processes to achieve a high efficiency of sludge handling.

(2) Reduction of the moisture content of the dewatered sludge cake with belt-press filter dewatering machines

An 80% moisture content of a dewatered sludge cake is common when the sewage sludge is dewatered with a belt-press filter. The target was to reduce the sludge cake moisture content of a dewatered sludge cake using a belt-press filter, from 80% to less than 75%.

(3) Reduction of phosphate release from excess sludges

The reduction of phosphate release from the excess sludge to attain an efficient biological phosphorus removal was another of the targets.

4. Theory of the Innovative Sludge Handling System

As sewage sludge particles are volatile-fraction rich, they form hydrophilic colloids because of the anionic charge. It is difficult to remove water from sludge particles under such conditions. In addition, as the particles are very small, there is high electric repulsion among the particles, and they are poorly agglutinative. Furthermore, small particles easily cause the filter cloth to be clogged.

To dewater sludges, usually, cationic polymer flocculants (cationic polymer/polymers) are dosed to the sludge to neutralize the anionic charge of the sludge, and thus the sludge particles are flocculated. However, as the flocculated sludge floc is still small and its mechanical strength is still weak, the moisture content of the dewatered sludge cake will be decreased up to 80% when dewatered with a belt-press filter. Also, poor separation of dewatered sludge cakes from the filter cloth is often observed.

The newly developed innovative sludge handling system employs a metal coagulant aid and a functional polymer, which has characteristics of flocculants that are both cationic and anionic (the amphoteric polymer). A schematic flow diagram of the system is shown in Figure 6.

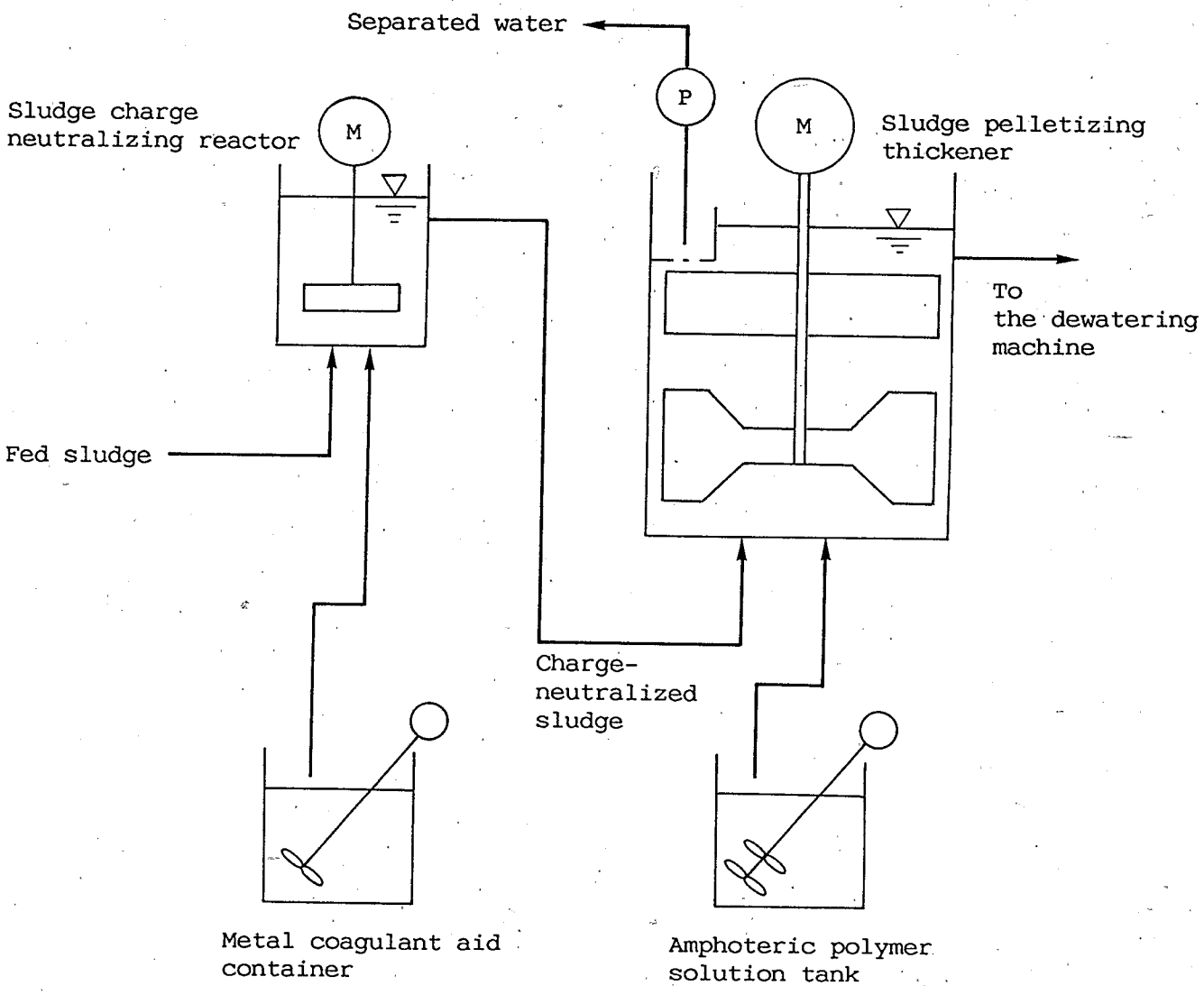

Figure 6. Schematic flow diagram of the system.

First, raw sludges withdrawn from sedimentation tanks are fed to the sludge-charge neutralizing reactor. Then, the metal coagulant aid is dosed to the sludge.

Next, the charge-neutralized sludge is fed to the pelletizing thickener with an amphoteric polymer. In the reactor, the sludge will be pelletized to spheres of 1 to 2 cm in diameter (the sludge pellets).

A conceptual model of the pelletization process is shown in Figure 7.

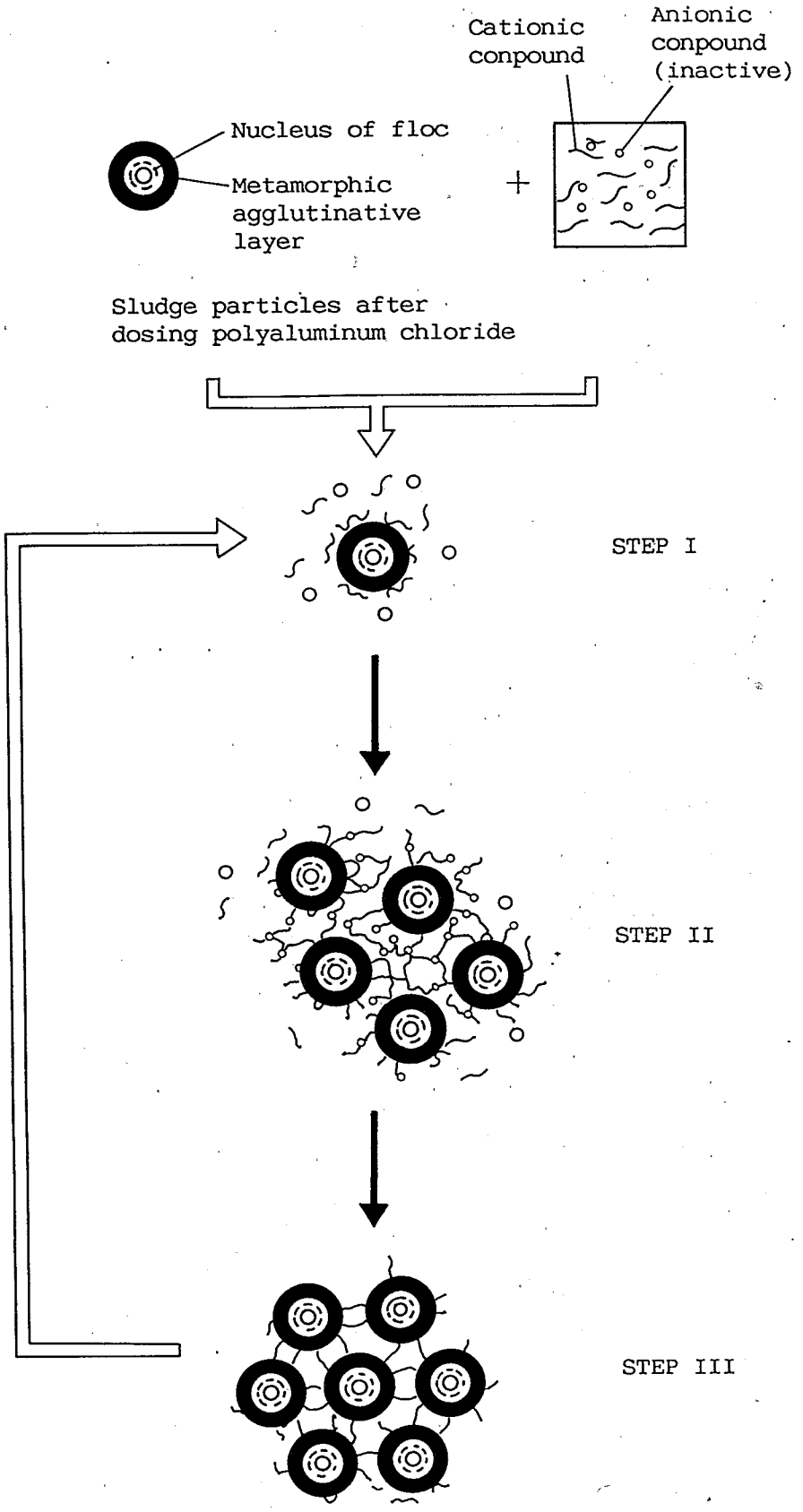

Figure 7. Conceptual model of the pelletization process.

The pelletization is carried out by the following 3 steps.

STEP 1 : The amphoteric polymer is fed to the charge-neutralized sludge. The cationic compounds of the amphoteric polymer adhere to the sludge particles. At the same time, inavtive anionic compounds of the amphoteric polymer will be charged due to charge in pH.

STEP 2 : The anionic compounds of the amphoteric polymer combine the cationic compounds which adhere to the sludge particles. The bridged sludge particles will start to form larger sludge flocs.

STEP 3 : The amphoteric polymers which bridge the sludge particles contract as the reaction continues, and the sludge pellets grow and become thicker and stronger. The growth of the sludge pellets will be accelerated by contact with growing sludge pellets and sludge particles.

An increase in the filtration rate, and a decrease in the moisture content of the dewatered sludge cake are expected, as a consequence of strong and thick sludge pellets being formed with this system.

From these system characteristics, the system can be explained as a sludge handling system which makes sludge particles into Ball-shaped sludge pellets and a very Effective Sludge Thickening system. Using these initial letters, the system has been named the BEST System.

5. Development of the Pelletizing Thickener

The development of the BEST system was originally begun to develop a sludge handling system suitable for biological phosphorus removal processes. At that time, a sludge handling system which combined metal coagulant aid dosing with polymer flocculant dosing was evaluated, for the physicochemical fixation of phosphorus. A schematic flow diagram of the prototype system is shown in Figure 8.

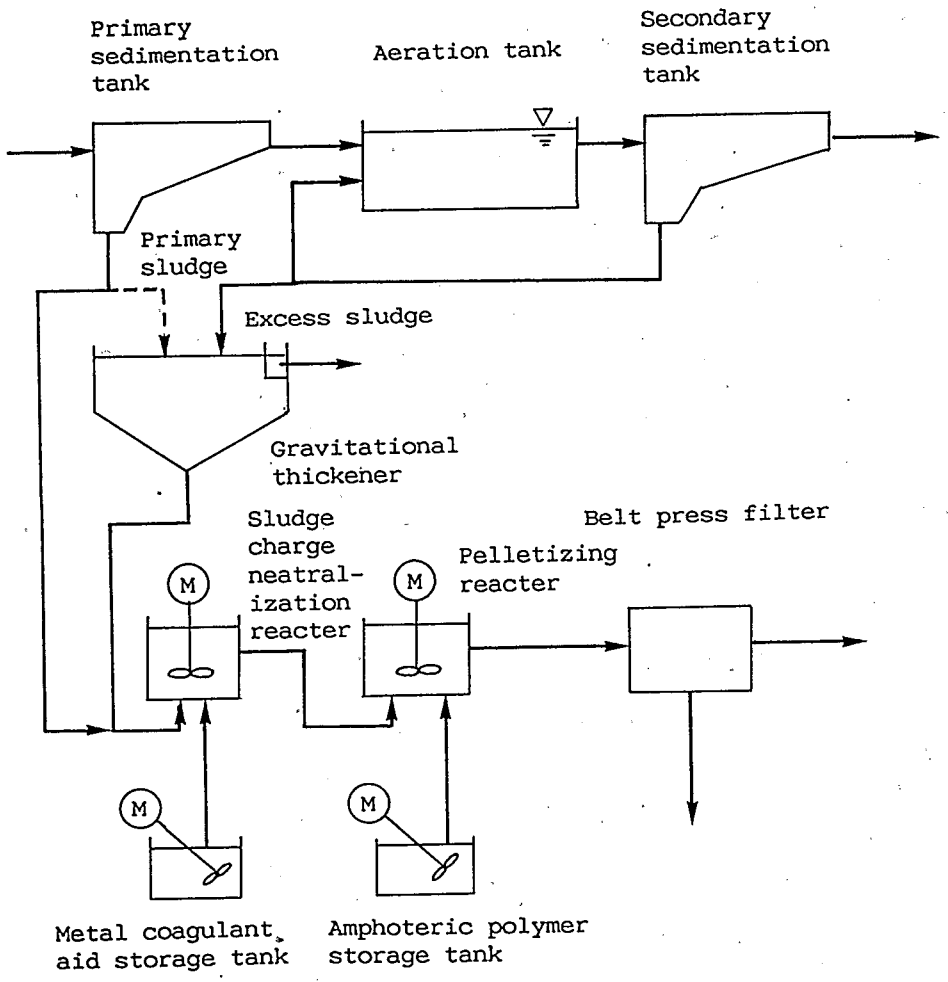

Figure 8. Schematic flow diagram of prototype system.

As can be seen in Figure 8, the prototype system consisted of a sludge-charge neutralization reactor and a pelletization reactor. The thickened/not thickened primary sludge and the gravitationally thickened excess sludge were fed to the system. The mixed sludge was charge-neutralized by the metal coagulant aid dosing, and pelletized by the amphoteric polymer dosing. The reason the amphoteric polymer was used is as follows. The sludge fed to the pelletization reactor was conditioned with a metal coagulant aid which is cationic, and the electrical charge condition changes to cationic or anionic, depending on the dosage of the coagulant aid. So, the amphoteric polymers were selected to be able to pelletize the conditioned sludge, regardless of whether it was charged as cationic or anionic. In the prototype system, the sludge thickening process was indispensable to making the following facilities as compact as possible.

Through the experiments, the development target to reduce phosphate release was achieved. In addition to this success, the observation of quite fine separation of liquid and solids in the pelletizing reactor suggested another possibility to develop an innovative sludge thickening process. This led the development of a sludge thickening/conditioning system, by adding a separated water withdrawer to the sludge pelletization reactor.

The structure of the separated water withdrawer was examined, using the following type of screens, which is installed at the bottom of the separated water withdrawer.

1) Metal nets,
2) Punching plates, and
3) Plates with concentric circular slits.

These screens are shown in Figure 9.

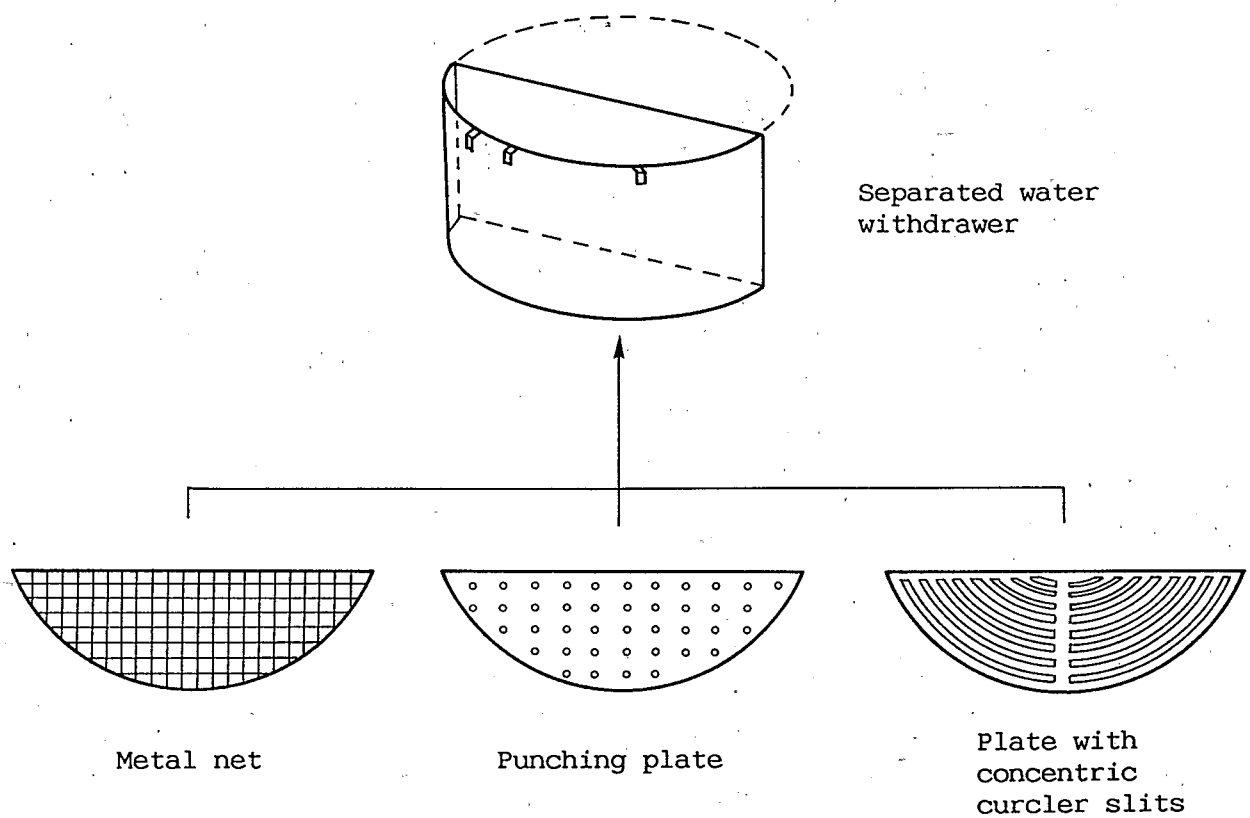

Figure 9. Examined screens.

The performance of each screen was examined in terms of the SS recovery efficiencies, clogging occurrences, and available flow rates through the screens. It was found that the plate with concentric circular slits was the most suitable for the separated water withdrawer, as it showed the best performance on the SS recovery efficiency, and no clogging troubles were observed. The clearance of the slit was 1.5 to 2 mm wide.

After gathering all the information about the favorable performances for the sludge pelletization thickener, the sludge pelletizing thickener shown in Figure 10 was developed.

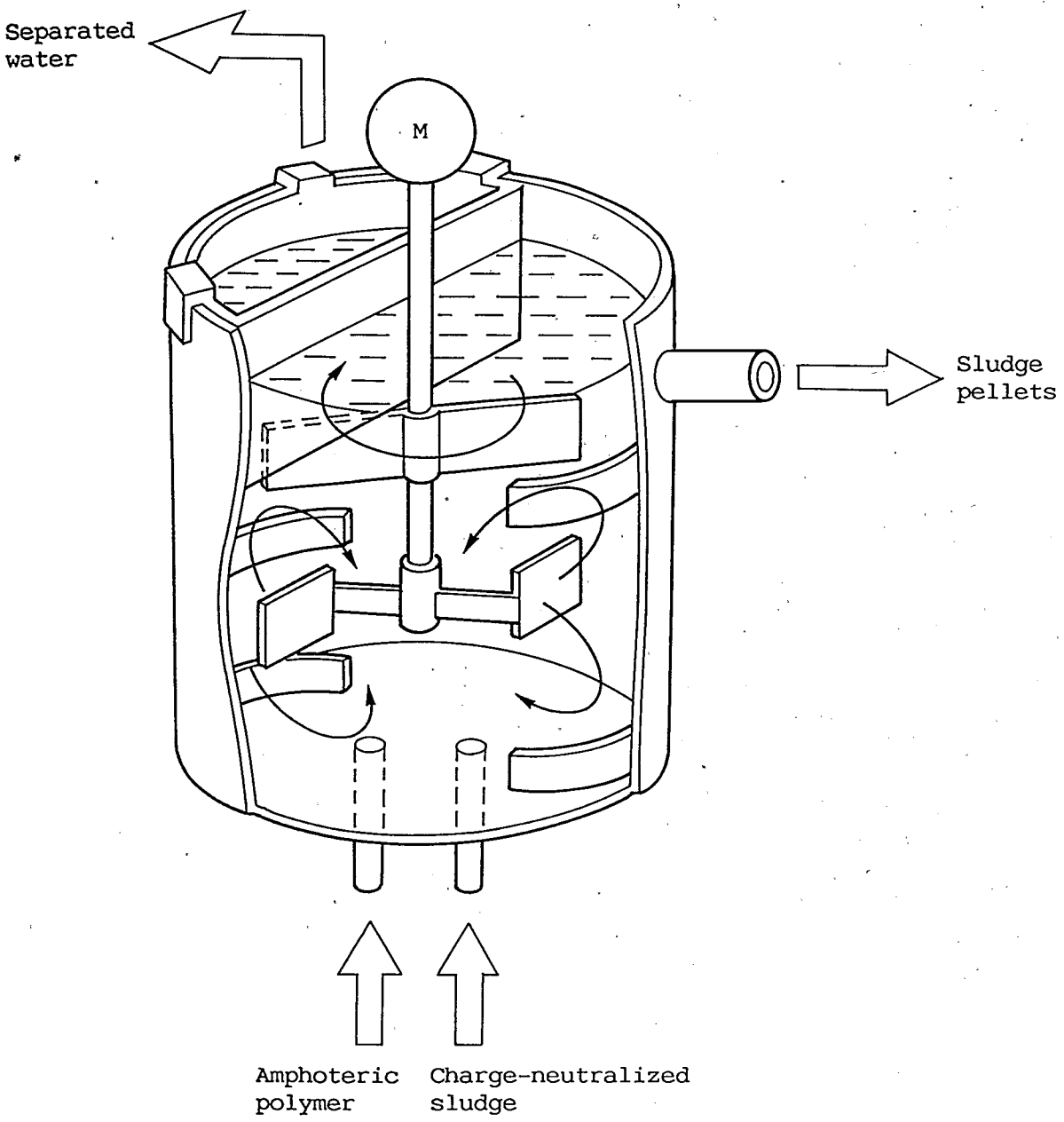

Figure 10. Structure of sludge pelletizing thickener.

The sludge from the sludge charge neutralization reactor is fed to the reactor at the bottom of the reactor, with an amphoteric polymer. There are two stirring paddles in the sludge pelletizing thickener. The fed sludge is mixed by the lower paddle with the amphoteric polymer. The lower paddle also establishes an internal recirculation of growing sludge pellets and fed sludge particles contact each other. The upper paddle propels the growing sludge pellets along the internal wall of the reactor. Then, the sludge pellets are consolidated and the water contained in the sludge pellets is removed from the pellets. Thus, the strength of the pellets increases. This is a benefit not only for improving dewatering performance, but also for improving the SS recovery efficiency of the system. As the sludge pellets contain a small amount of water, sludge pellets with a sufficiently high solid content will be supplied to dewatering machines.

The development of this sludge pelletizing thickener has made it unnecessary to employ conventional thickening processes. This elimination has made stable sludge dewatering possible, even when sludges were fed which showed poor thickenability or poor dewaterability to conventional sludge thickening, conditioning, and dewatering processes.

This system just requires tanks, stirring equipment, and pumps to feed sludges and coagulants. As all of them are very simple and compact, the construction cost of this system can be reduced compared with other conventional sludge thickening systems.

6. Experimental Procedures

The following four kinds of sludge were examined for the development of the BEST system.

(1) Mixed Raw Sludge of the Conventional Activated Sludge Process (the Mixed Raw Sludge),
(2) Excess Sludge of the Conventional Activated Sludge Process,
(3) Excess Sludge of the Anaerobic/Oxic Process, and
(4) Excess Sludge of the Oxidation Ditch Process
(Sludges described in (2) to (4) will be referred to as the Excess Sludges).

The characteristics of the above sludges are listed in Table 1. In general, the Mixed Raw Sludge is less difficult to thicken and to dewater, compared with other sludges. It is said that a moisture content of 80% would be the lowest, when the Mixed Raw Sludge is dewatered by a belt-press filter with the available maximum filtration rate of 130 kg-DS/m/h after single polymer conditioning. For the Excess Sludges, it is said to be very difficult to reduce the moisture content of dewatered sludge cake to less than 85% with an available maximum filtration rate of 50 kg-DS/m/h, when they are conditioned and dewatered in the same way as the Mixed Raw Sludge.

Table 1. Characteristics of four kinds of sludges

Kind of sludge	Characteristics
Mixed raw sludge of conventional activated sludge process	The fresher sludge has more fibers and shows better dewatering efficiency. However, this kind of sludge is generally corruptible. Once corruption progresses, the thickneability and dewatering efficiency deteriorate.
Excess sludge of anaerobic/aerobic process	Compared with the excess sludge of a conventional activated sludge process, this sludge shows a slightly better dewatering efficiency and higher phosphorus content. However, when this sludge is put into the gravitational thickener, or mixed with primarily sludge, phosphorus release occurs.
Excess sludge of conventional activated sludge process	In general, this kind of sludge shows poor dewatering efficiency. It is said to be very difficult to treat this sludge, even if the belt-press filtration is employed. The moisture content is as high as 85%.
Excess sludge of the oxidation ditch process	This sludge is considered to be poorly flocculated by the single polymer conditioning.

The characteristics of the sludges used in the experiments were measured on the items listed in Table 2.

Table 2. Measurement items and methods for the characteristics of sludges

Item	Unit	Notation	Method
pH	-	pH	JIS (Note 1)
Electrical conductivity	μs/cm	Cond	JIS
Methyl orange alkalinity	mg/ℓ	MA	JIS, Japanese standard method for the examination of wastewater quarity.
TS	%	TS	JIS, Japanese standard method for the examination of wastewater quarity.
VTS/TS	%	VTS/TS	Japanese standard method for the examination of wastewater quarity.
SS	%	SS	JIS, Japanese standard method for the examination of wastewater quarity.
VSS/SS	%	VSS/SS	Japanese standard method for the examination of wastewater quarity.
Fiber/SS	%	Fiber/SS	
CST	sec	CST	Toriton's CST measuring instrument and Wattman filter are employed.
Total phosphorus	mg-P/ℓ	T-P	(Mixed reagent method) (Note 2)
Orthophosphoric acid	mg-P/ℓ	O-P	(Mixed reagent method) (Note 2)

Note 1: Japanese Industrial Standards
Note 2: Methods for chemical analysis of water and and wastes (1971) of the Environmental Protection Agency.

Lab tests and pilot scale experiments were conducted to examine the basic performances of the BEST system with regard to pelletization and dewatering. The tests studied the relationship between the moisture contents of the sludge pellets after gravitational drainage, the achievable moisture contents of dewatered sludge cakes*) (AMC), and the operational conditions, such as chemical dosing ratios, etc.

*) The moisture content of a dewatered sludge cake decreases and the clogging level increases, as the dewatering pressure of the belt-press filter increases. Thus, the most preferable operation of a belt-press filter is done with the highest dewatering pressure, and the least filter cloth clogging. The achievable moisture content of the dewatered sludge cake is defined as the minimum moisture content of a dewatered sludge cake with less than 10% of the clogging area on the filter cloth in batch dewatering tests with sludge compression.

The tests were done to examine the dewatering efficiencies in the gravitational drainage zone, the primary pressure zone, the secondary pressure zone, and the final dewatering zone with the shearing force of a belt-press filter dewatering machine.

The dosage of the metal coagulant aid was examined by the capillary suction time technique, searching for the inflection point of the CST versus chemical dosage curve (the CST procedure).

The dosage of the amphoteric polymer was decided by examining the SS recovery efficiency for the sludge pelletization thickener and the mechanical strength of sludge pellets under various operational conditions.

7. Results and Discussions

7-1. Sludge Pelletization/Thickening Efficiency

The dosage of the metal coagulant aid, the dosage of the amphoteric polymer, the condensing ratio in the sludge pelletizing thickener, and the system's solid retention time in the pelletizing thickener are the operational conditions of the BEST system. The influence of these operational conditions on pelletizing and dewatering performances were examined.

1) Dosage of Metal Coagulant Aid (Polyaluminum chloride)

A metal coagulant aid is dosed to the fed sludge, with the objective of neutralizing the anionic sludge charge. The dosage of the metal coagulant aid is decided by the CST procedure.

The relationships between the CST value and the dosage of the metal coagulant aid for the mixed raw sludge and the excess sludges of a conventional activated sludge process are shown in Figure 11 and Figure 12, respectively.

Both Figures 11 and 12 show that the inflection points lay in almost the same dosage range. This fact suggests that proper dosage of the metal coagulant aid will range from 0.6 to 0.8% as Aℓ to SS.

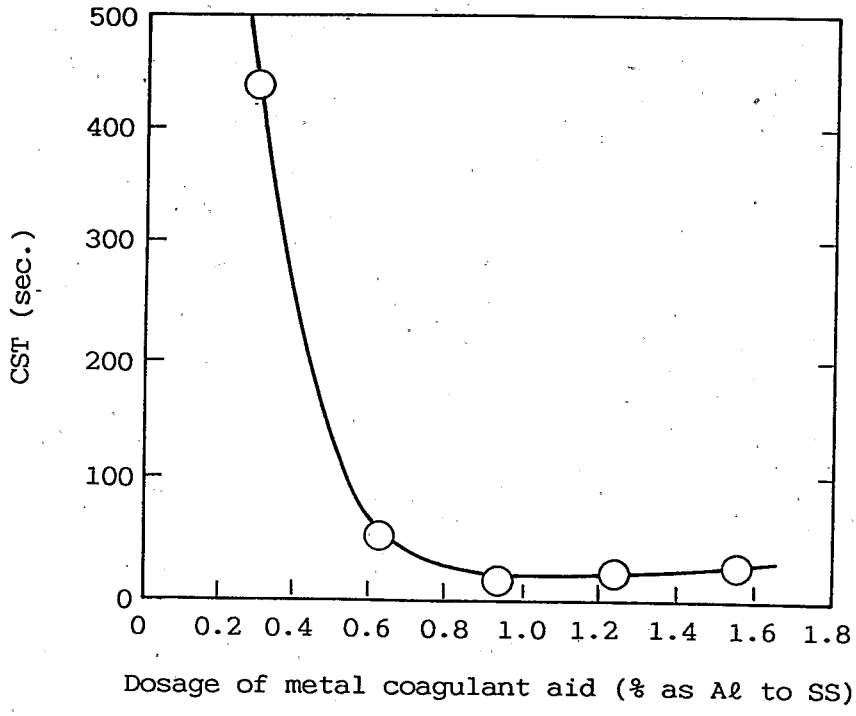

Figure 11. Relationship between the dosage of metal coagulant aid and the CST value. (Mixed raw sludge of a conventional activated sludge process.)

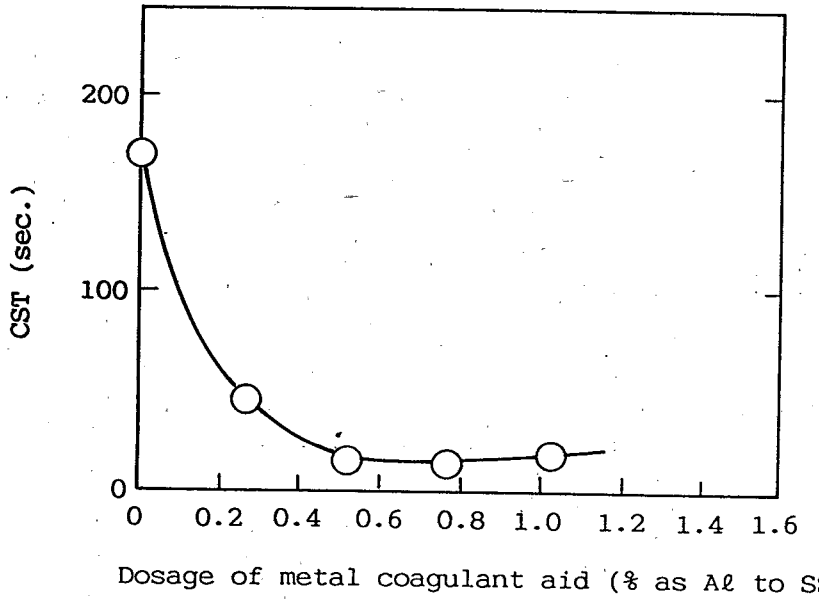

Figure 12. Relationship between the dosage of metal coagulant aid and the CST value. (Excess sludge of a conventional activated sludge process.)

The relationship between the AMC and the dosage of the metal coagulant aid is shown in Figure 13. As shown in the figure, the smallest AMC value is about 70%, with the metal coagulant aid dosage of 0.6 to 0.8% as Aℓ to SS. This means that the actual dewatered sludge cake separation will be satisfactory, when 75% of the moisture content of the cake is achieved, which was one of the targets.

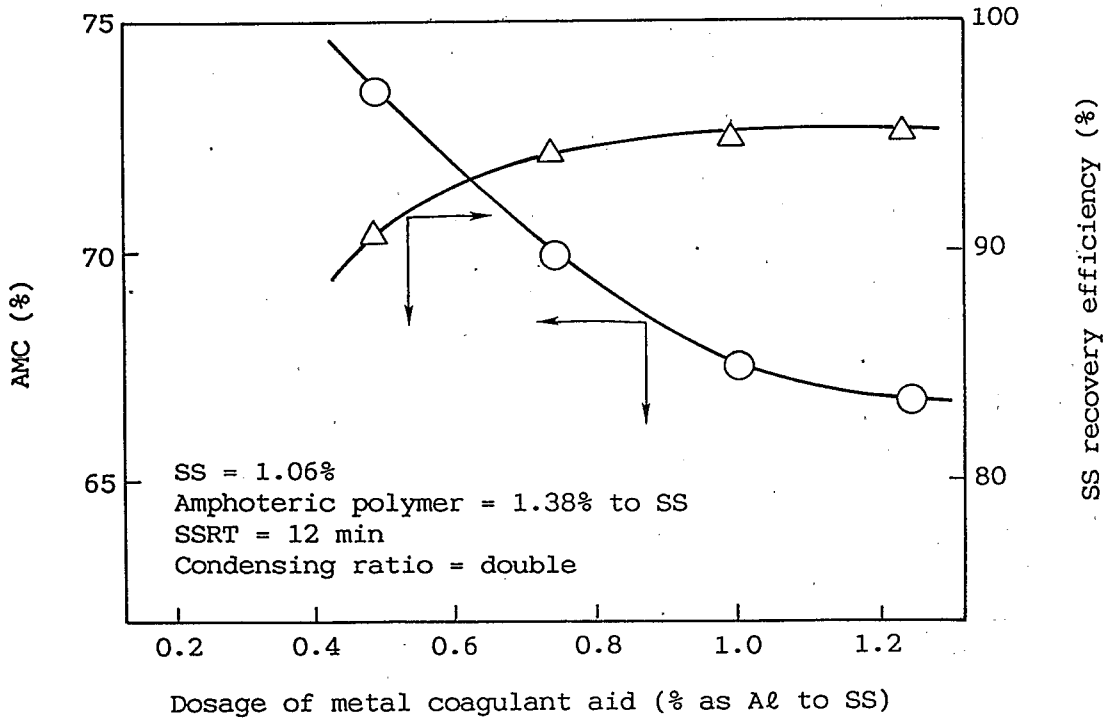

Figure 13. Relationship between the dosage of metal coagulant aid, and the AMC and SS recovery efficincy. (Mixed raw sludge of a conventional activated sludge process.)

From these results, the proper dosage of the metal coagulant aid is revealed to be in the range from 0.6 to 0.8% as Aℓ to SS.

2) Dosage of Amphoteric Polymer

As the dosage of the amphoteric polymer influences the mechanical strength of the sludge pellet, the relationship between the AMC, the SS recovery efficiency, and the dosage of the amphoteric polymer was examined. When the mechanical strength of the sludge pellet is strong enough, the AMC will decrease and the SS recovery efficiency will increase.

The relationship between the SS recovery efficiency and the dosage of the amphoteric polymer for the Mixed Raw Sludge is shown in Figure 14. The result indicate that the AMC showed far less than the target value of 75% when the dosage was 0.6% to SS, or more. And an SS recovery efficiency of 95% or more was achieved with the dosage of 0.7% to SS.

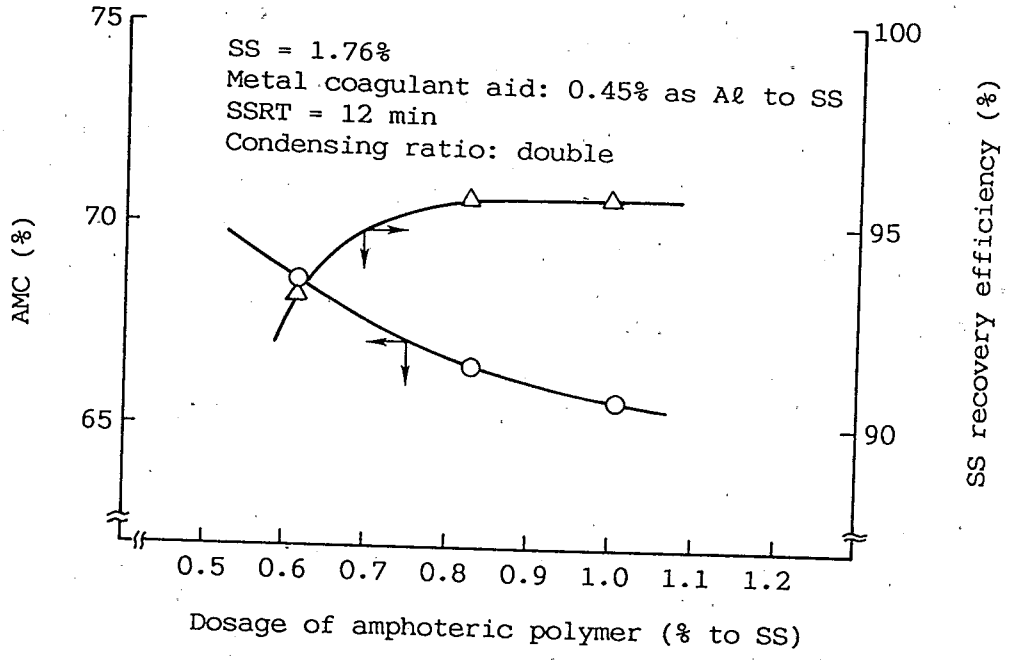

Figure 14. Relationship between the dosage of amphoteric polymer, and the mechanical strength of sludge pellet and the SS recovery efficiency.

The relationship between the same items as shown in Figure 14, for the excess sludge of a conventional activated sludge process, is shown in Figure 15. The result indicates that an amphoteric polymer dosage of 1% to SS provides 77% of the AMC and 95%, or more, of the SS recovery efficiency.

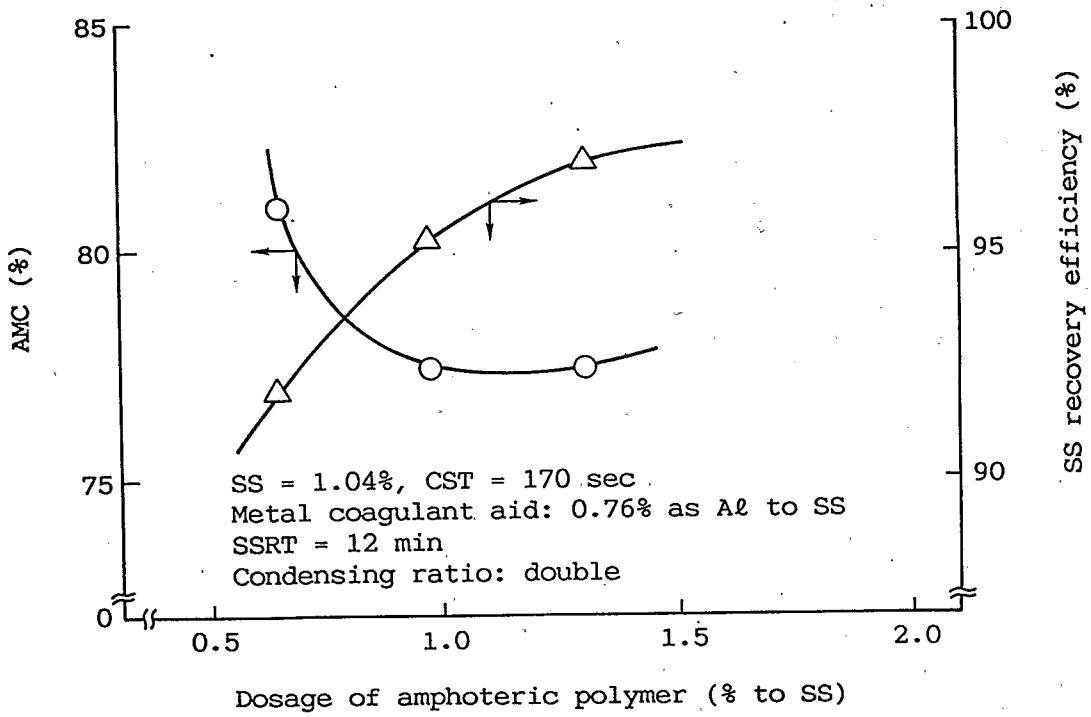

Figure 15. Relationship between the dosage of amphoteric polymer, and the mechanical strength of sludge pellet and the SS recovery efficiency. (Excess sludge of a conventional activated sludge process.)

From these results, the proper dosage of the amphoteric polymer is revealed to be 1% to SS, or less.

3) Condensing Ratio and SSRT of Sludge Pelletizing Thickener

As one of the functions of the BEST system is a sludge thickening process, the solid content of the sludge to be supplied to a dewatering machine from this system is very important. Therefore, to estimate the solid content of the sludge from this system, the condensing ratio and the SS recovery efficiency are defined as follows.

$$CR = \frac{Q_{in}}{Q_{in}-Q_t} \quad (2)$$

$$SSRE = \frac{C_A \times (Q_{in}-Q_t)}{C_{in} \times Q_{in}} \times 100\% \quad (3)$$

where CR : the Condensing Ratio (-)
SSRE: the SS Recovery Efficiency (%)
Q_{in} : Sludge feeding Rate (m^3/min)
Q_t : Separated Water Withdrawal Rate at the Sludge Pelletizing Thickener (m^3/min)

C_{in} : Solid Content of Supplied Sludge (mg/ℓ)
C_A : Solid Content in the Reactor (mg/ℓ)
C_t : Solid Content of the Separated Water (mg/ℓ)

By reforming Equation 3, the following equation is obtained.

$$CA = \frac{SSRE}{100\%} \times \frac{C_{in} \times Q_{in}}{Q_{in}-Q_t} = \frac{SSRE}{100\%} \times C_R \times C_{in} \qquad (4)$$

According to Equation 4, to increase the solid content of sludge fed to a dewatering machine, the condensing ratio should be kept high while keeping the SS recovery efficiency high. To examine the extent of the solid content of the sludge from this system, the relationship between the SS recovery efficiency and the solid content in the sludge pelletizing thickener when the Mixed Raw Sludge was fed to the system was examined, and is shown in Figure 16. It is suggested that the SS recovery efficiency tends to decrease when the solid content in the reactor exceeds 3 or 4%. This implies that the opportunity for the sludge pellets to clash increased (and then small sludge particles were produced) as the solid content in the reactor increased.

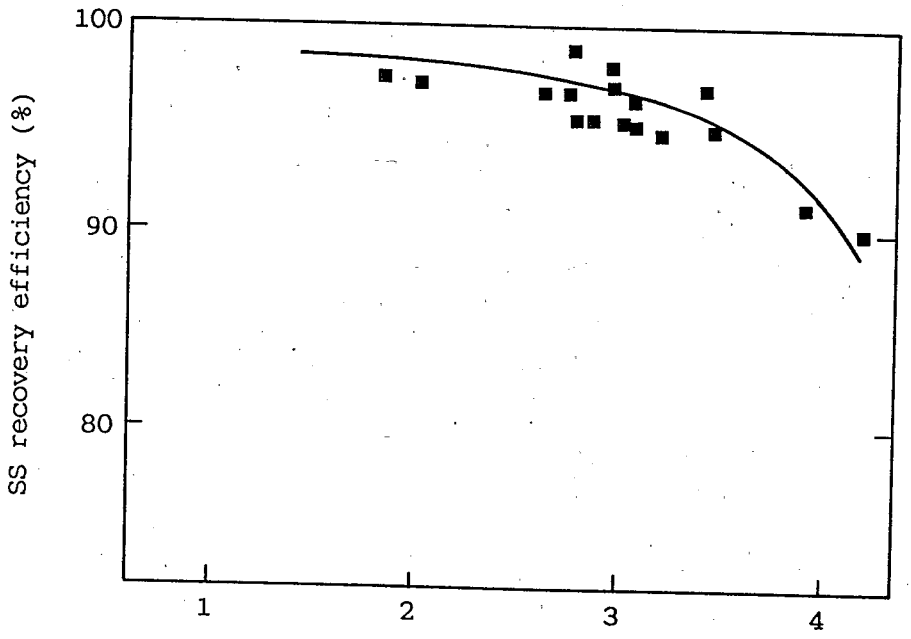

Figure 16. Relationship between the solid content in the sludge pelletizing thickener and the SS recovery efficiency. (Mixed raw sludge of a conventional activated sludge process.)

The same investigations were conducted for the Excess Sludges, and almost the same trend was observed. However, the critical solid content clashing was slightly smaller than that for the Mixed Raw Sludge. It was found to be about 3%, or less, of solid content.

From these results, it was found that the solid content in the sludge pelletizing thickener should be operated with 3 to 4% when the Mixed Raw Sludge is fed, and with 2 to 3% when the Excess sludges are fed.

Next, the relationship between SSRT*) and the moisture content of the dewatered sludge cake was examined. The relationship for the Mixed Raw Sludge is shown in Figure 17.

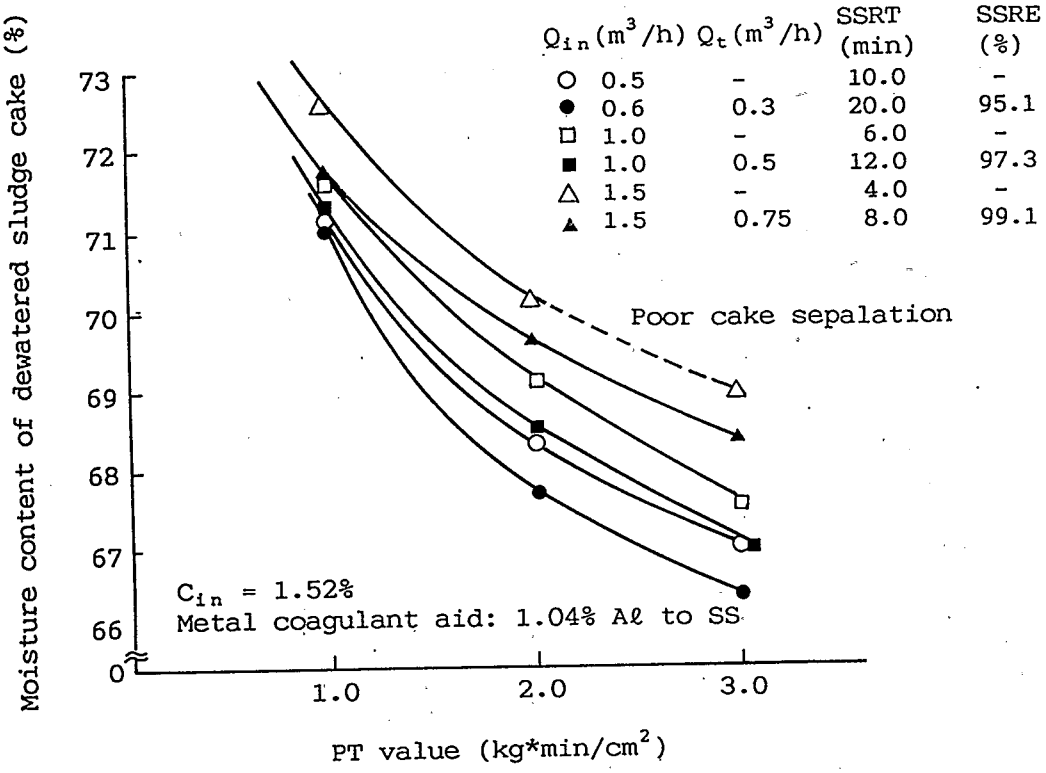

Figure 17. SSRT and pelletizing thickening performances.
(Mixed raw sludge of a conventional activated sludge process.)

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*) The system's solid retention time is described as the SSRT. The values of SSRT were calculated using the equation below.

$$\text{SSRT (min)} = \frac{\text{Total amount of Solid in the Reactor (kg-SS)}}{\text{Sludge feeding rate (kg-SS/min.)}} \quad (5)$$

*)PT = PxT
where PT : PT value (kg*min/cm²)
P : The dewatering pressure (kg/cm²)
T : Compression time (min.)

Figure 17 implies that the mechanical strength of the pellets will be higher, as the moisture content of a dewatered sludge cake is lowered, when the PT value*) is increased, and that a fine separation will be expected under these conditions.

Figure 17 also implies that the mechanical strength of the sludge pellets increased as the SSRT increased. However, the sludge cake separation was sufficient when the SSRT was 6 minutes, or longer. Thus, the SSRT in the sludge pelletizing thickener should be kept for a minimum of 6 minutes when the Mixed Raw Sludge is fed.

Also, it was revealed that the SSRT in the sludge pelletizing thickener should be kept for about 12 minutes for the Excess Sludges.

7-2. Dewatering Performance with a Batch Tester

Dewatering examinations were made employing a belt-press filter batch tester. Figure 18 describes the tester. The dewatering performances were evaluated with the procedure shown in Figure 19.

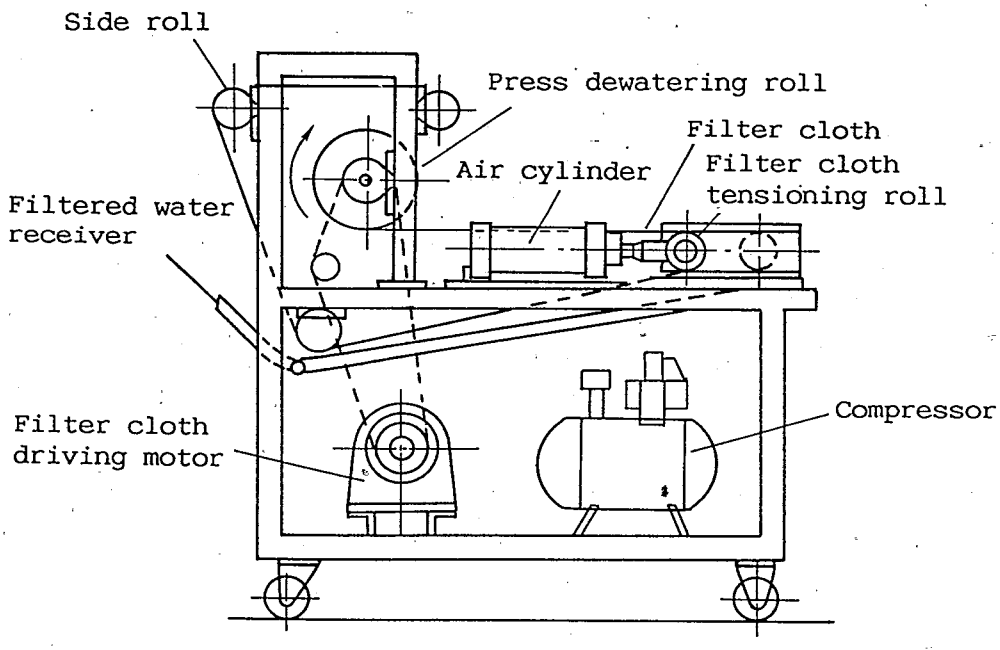


Figure 18. Belt press filter batch tester.

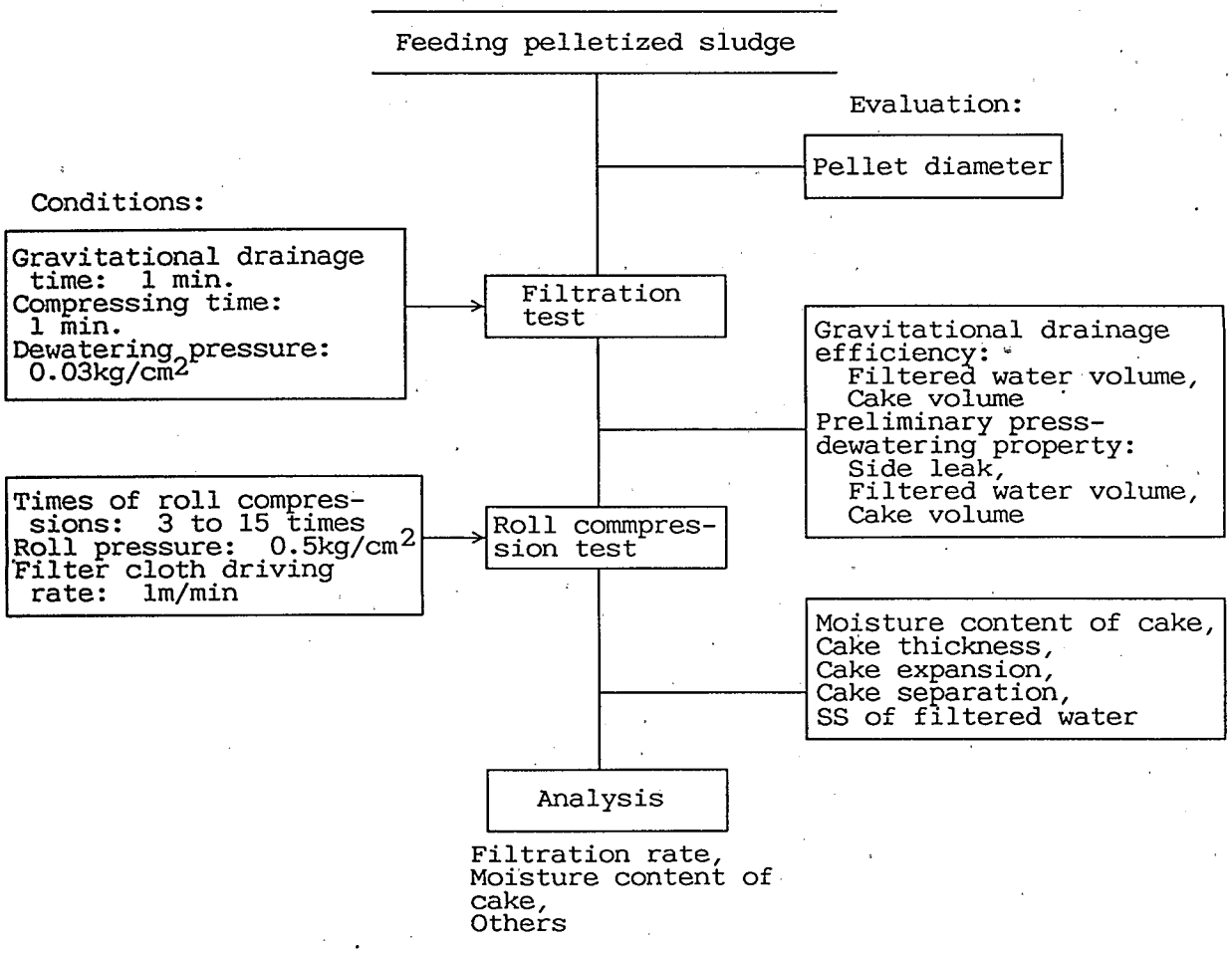


Figure 19. Procedure of batch dewatering tests.

The dewatering performances as a control were also examined with single polymer conditioning after centrifugal sludge thickening, to evaluate the dewatering performance by the BEST system.

The characteristics of the applied sludges are listed in Table 3.

Table 3. Characteristics of raw sludge, and centrifugally thickened sludges.

Sludge type	pH (-)	Cond (μS/cm)	TS (%)	VTS/TS (%)	SS (%)	VSS/SS (%)	Fiber/SS (%)	(VSS-Fiber)/SS (%)	T-P (mg/ℓ)	O-P (mg/ℓ)	M-AL (mg/ℓ)	SSCE (meq/g-SS)	Wo (%)	CST (sec)
Mixed raw sludge	5.6 (6.0)	3210 (2400)	3.67 (1.97)	62.7 (62.8)	3.52 (1.76)	62.5 (63.2)	18.2 (17.1)	44.3 (46.1)	683 (345)	18.36 (23.3)	796 (768)	0.209 (0.243)	87.6 (85.3)	1403 (875)
Excess sludge of anaerobic/aerobic process	6.4 (6.6)	1700 (1020)	3.83 (0.98)	76.8 (77.5)	3.15 (0.97)	75.1 (81.4)	16.1 (21.6)	59.0 (59.8)	690 (318)	105.42 (39.99)	952 (512)	0.237 (0.366)	90.6 (91.0)	147 (15)
Excess sludge of conventional sludge process	6.5 (6.6)	1710 (1380)	3.20 (0.52)	70.2 (65.9)	3.06 (0.47)	69.8 (70.2)	11.9 (8.4)	57.9 (61.8)	904 (158)	39.80 (21.90)	876 (704)	0.249 (0.337)	93.1 (92.6)	173 (24)
Excess sludge of oxidation ditch process	5.9 (6.5)	1600 (1440)	4.10 (2.36)	77.4 (79.1)	3.96 (2.20)	77.4 (80.4)	13.5 (12.7)	63.9 (67.7)	770 (448)	198.10 (91.95)	731 (518)	0.229 (0.282)	91.3 (92.0)	143 (44)

Note: The value in parentheses indicates the property of the sludge before centrifugal thickening.

The relationship between the filtration rates and the moisture contents of dewatered sludge cakes are shown for both the Mixed Raw Sludge and the Excess Sludges are shown in Figure 20 and Figure 21, respectively.

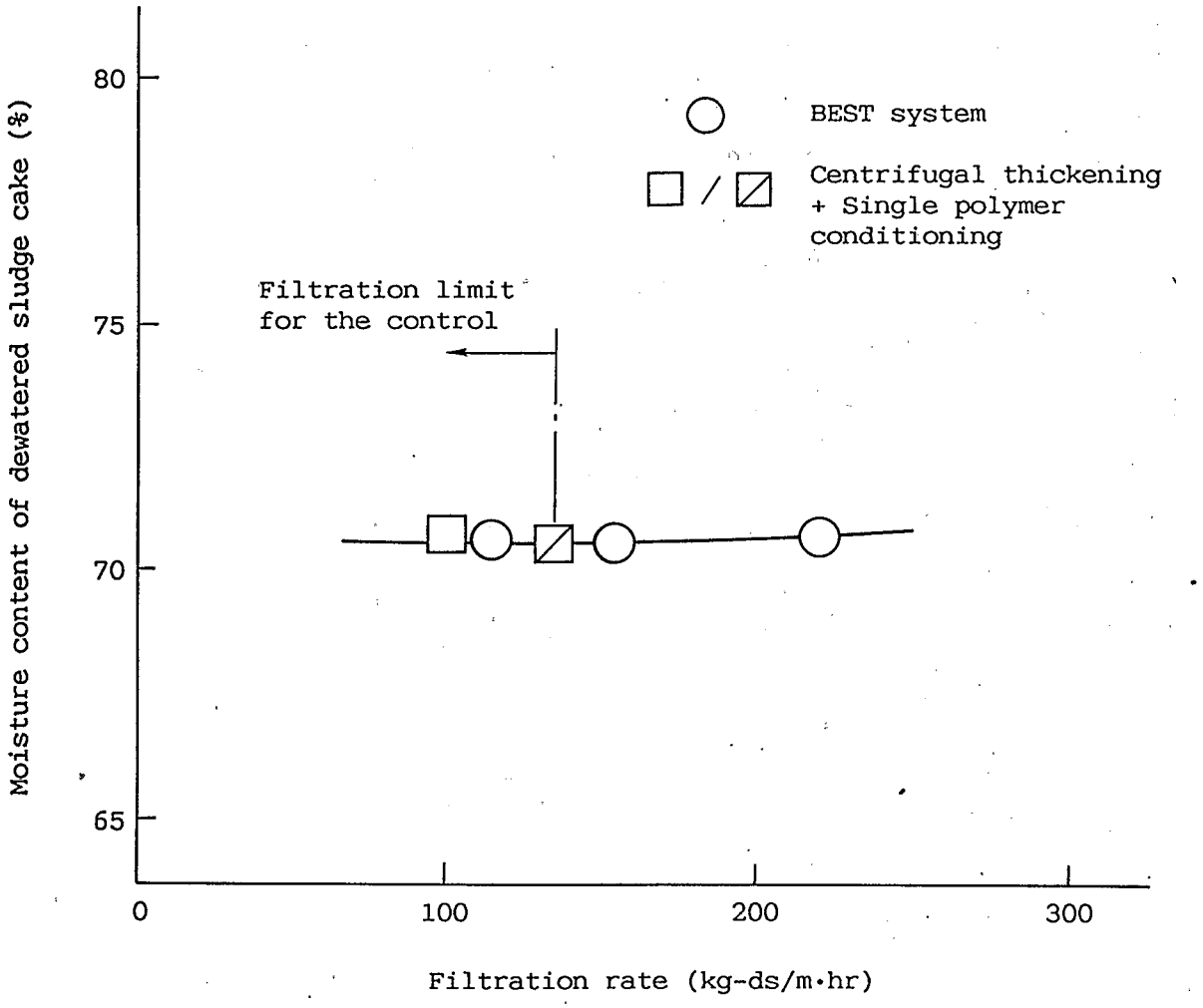


Figure 20. Relationship between the filtration rates and the moisture content of dewatered sludge cakes. (Mixed raw sludge of a conventional activated sludge process.)

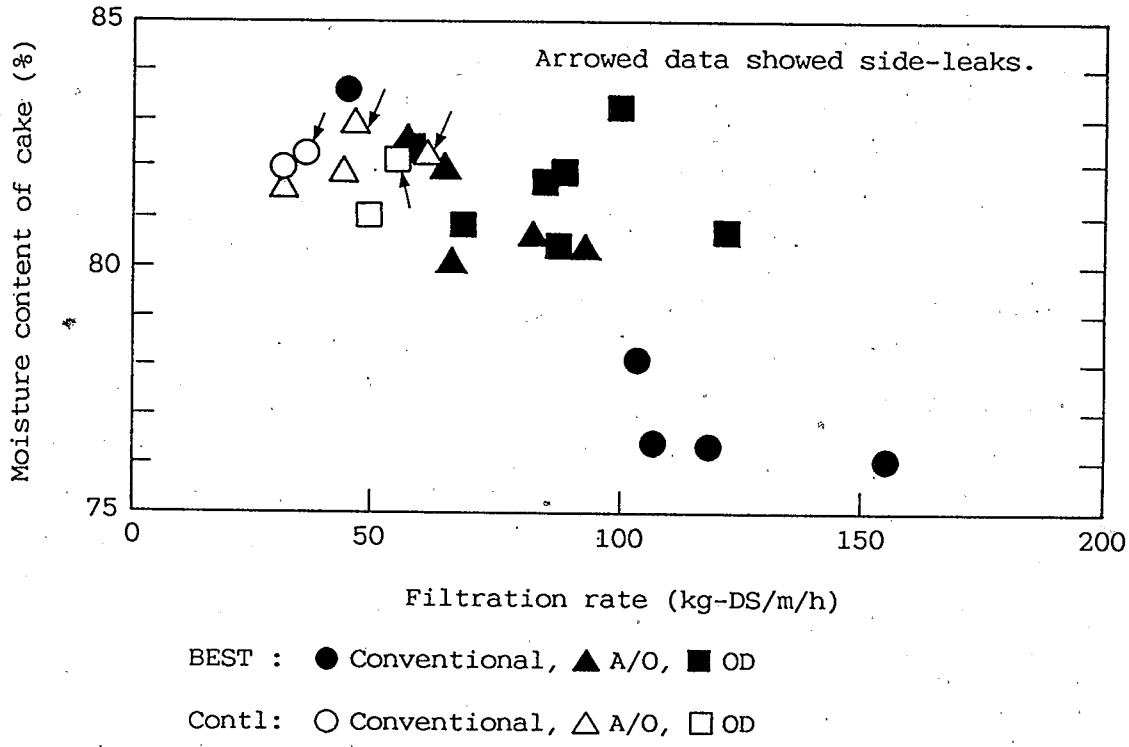


Figure 21. Relationship between the filtration rates and the moisture content of dewatered sludge cakes. (Excess sludges)

According to the results in Figure 20 for the Mixed Raw Sludge, the achievable maximum filtration rate for the control was found to be less than 130 kg-DS/m/h, as a side leak of sludge was observed when dewatered with this rate. Then, the moisture content of the dewatered sludge cake was 71%.

On the other hand, when the BEST system was employed, it was possible to dewater the Mixed Raw Sludge with a filtration rate of 220 kg-DS/m/h without any failure in dewatering. Then, the moisture content of the dewatered sludge cake was 71% in the examination. For the dewatering performance of a belt-press filter in Japan, it is generally said that 78% of the moisture content will be obtained with a filtration rate of 130 kg-DS/m/h when a gravitationally thickened mixed raw sludge is fed. The moisture content attained with the BEST system was lower than the conventional value, and the achieved filtration rate was almost twice that of the conventional performance.

The results for the Excess Sludges in Figure 21, also show that the BEST system could attain lower moisture content of the cake and larger filtration rate compared with the control system. For example, a sludge side leak was observed when sludges were dewatered with a filtration rate of about 50 kg-DS/m/h with the control method, and more than twice the filtration rate was achieved when the BEST system was used to dewater the excess sludge of a conventional activated sludge process. In addition, the moisture content for the BEST system was 2 to 5% lower than that for the control method.

7-3. Full Scale Evaluation of Dewatering Performance

Full scale run of the BEST system was tested in the T treatment plant with the operational conditions listed in Table 4. The BEST system was incorporated between the sludge storage tank and the belt-press filter as shown in the schematic flow diagram of Figure 22.

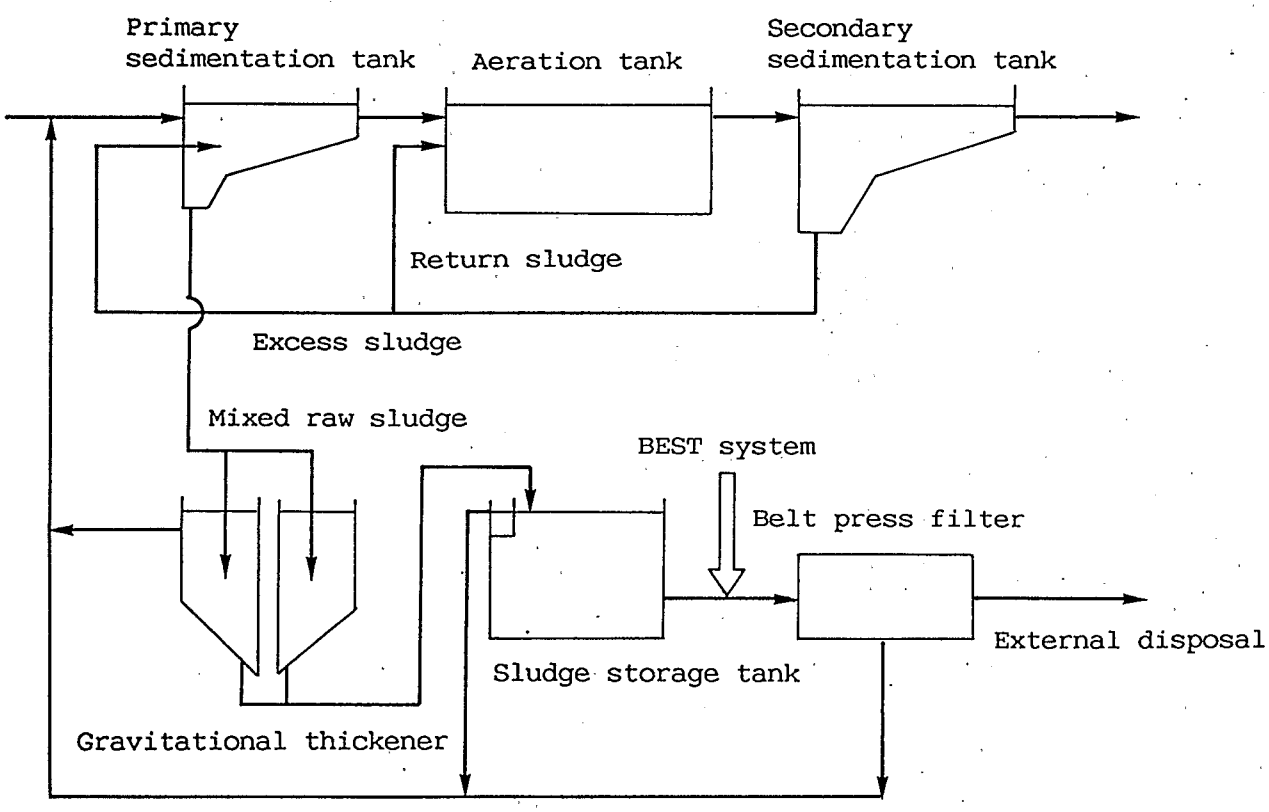


Figure 22. Schematic flow diagram of the T plant.

Table 4. Operational Conditions of the T Plant.

Wastewater Collection System	Separate Collection System
Wastewater Treatment Process	Conventional Activated Sludge Process
Influent Rate	3,800 to 6,400 m^3/day
BOD-SS Loading	0.64 kg-BOD/kg-SS/day
Sludge Handling System	Gravitational Thickening ==> Dewatering (Belt-Press Filter) ==> Disposal
Solid Content of Sludge	Mixed Raw Sludge = 0.8 to 1.0% of SS Thickened Sludge = 0.98 to 1.63% of SS

The test run was performed for one month, and the characteristics of the fed sludges during this period are shown in Table 5. The relationship between the filtration rates and the moisture content of the dewatered sludge cakes are shown in Figure 23. In the figure, the annual average values of the moisture content of the dewatered cake and the filtration rate are also shown.

Table 5. Characteristics of sludges during the experiment.

pH	Cond (μS/cm)	SS (%)	VSS/SS (%)	TS (%)	VTS/TS (%)	Fiber/TS (%)
5.6 to 6.3	1,560 to 1,900	0.98 to 1.63	86.2 to 88.3	1.11 to 2.30	82.3 to 86.4	15.5 to 29.3

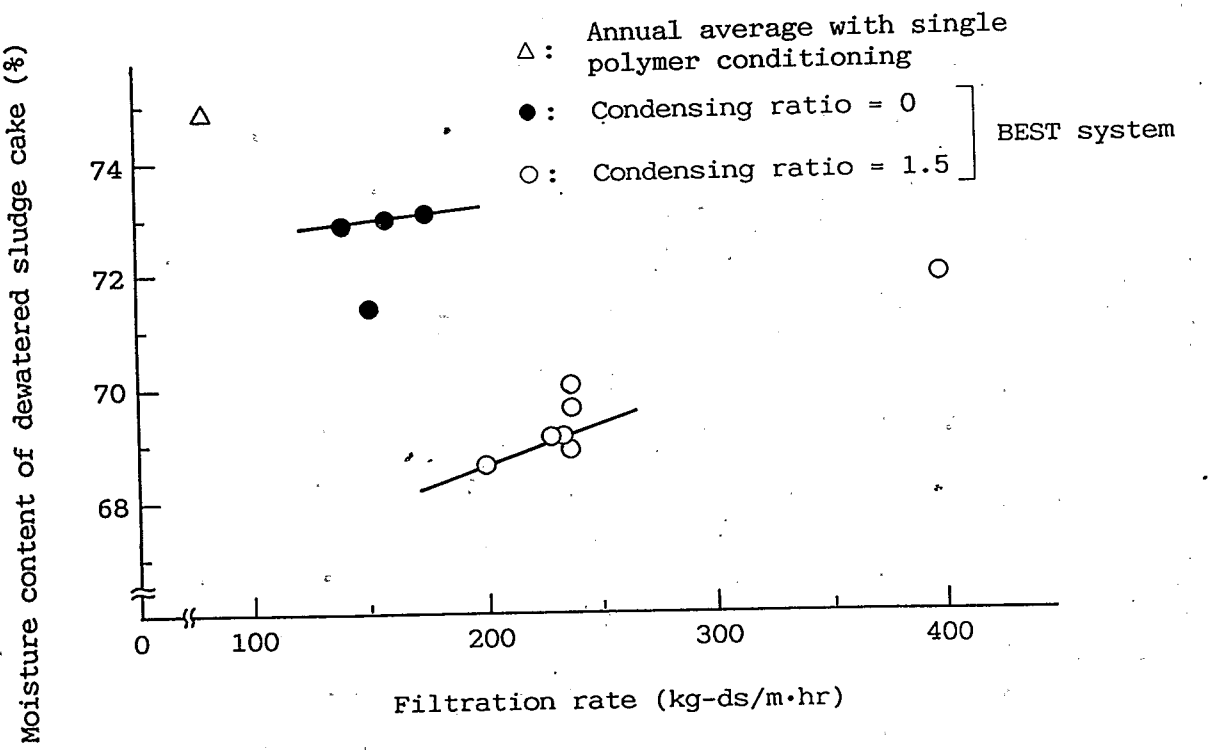


Figure 23. Relationship between the filtration rate and the moisture content of dewatered sludge cakes. (At the D plant; Mixed raw sludge of a conventional activated sludge process.)

As shown in Figure 23, the average filtration rate and the average moisture content of the cake were 90 kg-DS/m/h and 75%, respectively, when the single polymer conditioning had been employed. Improvements in the moisture content decreaseding to 72%, and in the filtration rate increasing to 150 to 160 kg-DS/m/h are shown in the figure, even when the BEST system was employed with a condensing ratio of zero. Also, the further improvement in the moisture content to 70% decreasing with an increased filtration of 220 kg-DS/m/h are shown when the BEST system was employed with a condensing ratio of 1.5.

The further improvements in the moisture content and the filtration rate are considered to be due to the following reason. The mechanical strength of the sludge pellets increased because of the elongated SSRT in the sludge pelletizing thickener brought about by the increase of solid content in the reactor as the result of withdrawing separated water. Thus, withdrawal of separated water from the sludge pelletizing thickener is found to be very effective in improving the dewatering efficiency of the system.

The conventional value of dewatering performance for a belt-press filter is described (in the specification code of the Japan Sewage Works Agency) as being able to make the moisture content of the dewatered sludge cake 75%, or less, of the solid content, with a filtration rate of 130kg-DS/m/h, when a mixed raw sludge is fed with 3% solid content after gravitational thickening. However, it is said that to attain these values have become extremely difficult in many municipalities in recent years. From these facts, the dewatering performance of the BEST system can be described as being very good.

7-4. Phosphate Fixation

The relationship between the ortho-phosphate concentration (O-P) and the dosage of metal coagulant aid, when a fresh biological phosphorus removal excess sludge was fed to the system is shown in Figure 24.

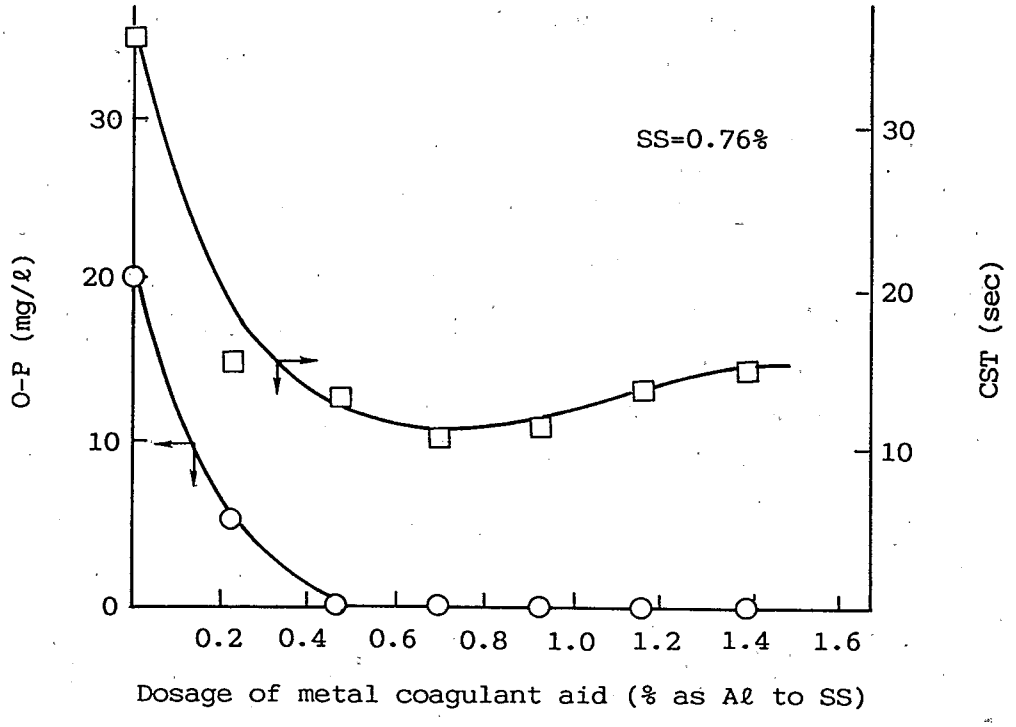


Figure 24. Relationship between the orthophosphate concentration and the dosage of metal coagulant aid. (Fresh excess sludge of an anaerobic/oxic process.)

The results in the figure indicate that O-P was almost completely fixed within the sludge with a metal coagulant aid dosage of 0.5% as Aℓ to SS. As discussed in Section 7-1, a metal coagulant aid dosage of about 1% as Aℓ to SS is required for charge neutralization, so the O-P in the sludge liquid will be almost completely fixed within the solids with this dosage.

Figure 25 shows the relationship between the same items shown in Figure 24, using excess sludge from a biological phosphorus removal process. The sludge was stored 5 days under anaerobic condition prior to the test. The results shown in the figure indicate that a dosage of 1.5% as Aℓ to SS was required to reduce O-P to the level attained with 0.5% dosage when the fed sludge was fresh. From the economic point of view, the sludge should be processed while it is fresh to fix O-P within the sludge.

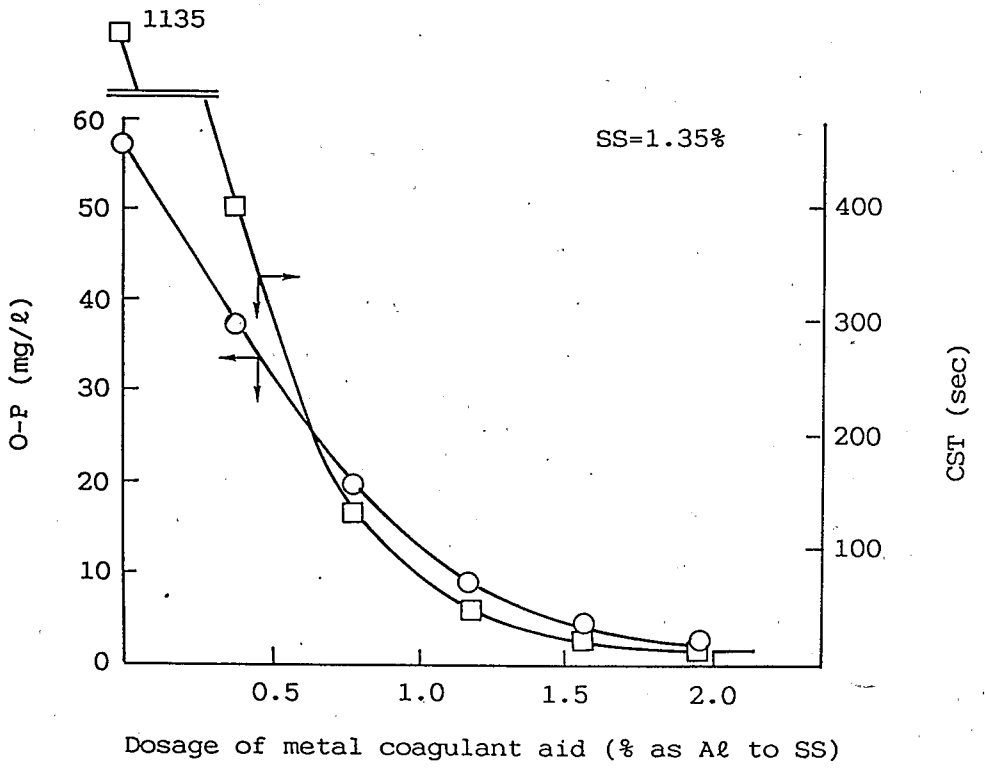


Figure 25. Relationship between the orthophosphate concentration and the dosage of metal coagulant aid. (The sludge was fed after 5 days storage)

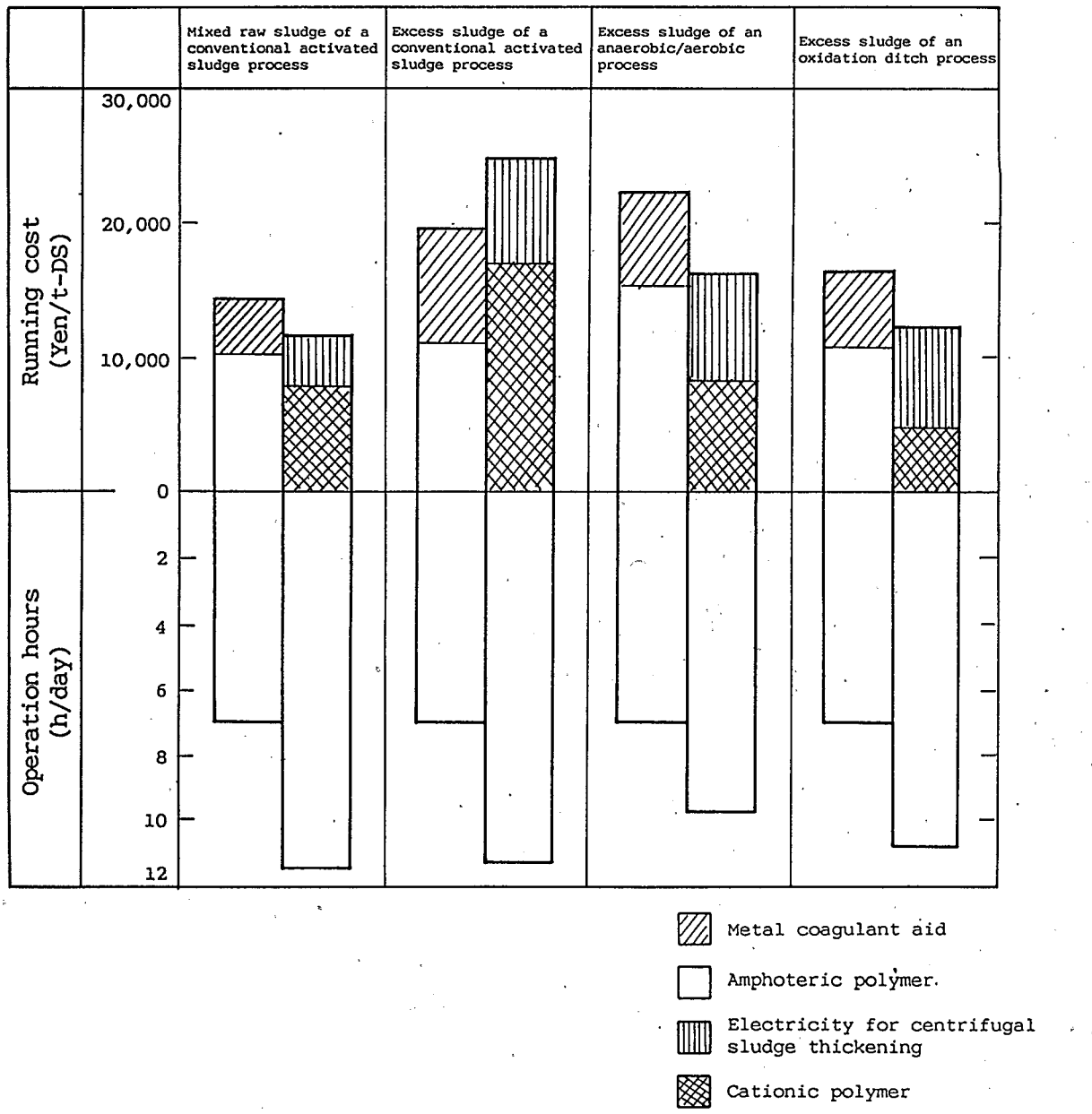
8. Operational Cost Estimation

The running costs were estimated for sludge conditioning and thickening when the BEST system or the conventional system is employed to dewater the Mixed Raw Sludge and the Excess Sludge. The estimation was made based on the conditions in Table 6.

Table 6. Conditions of cost estimation

Kind of Sludge	Moisture content of dewatered cake (%)	BEST system		Conventional system
		Dosage of metal coagulant aid (% as Aℓ to SS)	Dosage of amphoteric polymer (% to SS)	Dosage of cationic polymer (% to SS)
The Mixed Raw Sludge	70	0.46	0.78	0.60
Excess Sludge of convèntional activated sludge process	82	0.89	0.86	1.31
Excess Sludge of anaerobic/oxic process	82	0.74	1.18	0.63
Excess Sludge of oxidation ditch process	82	0.60	0.83	0.33

The estimates for the BEST system include costs for the metal coagulant aid and the amphoteric polymer. The estimates for the conventional system include costs for the cationic polymer and electricity for the centrifuge. Costs for electricity required to operate pumps and stirring equipment were not included in the estimates, as they were thought to be in significant in relation to the total costs. The results are shown in Figure 26. In the figure, operation hours are shown as well.



Note 1: The left bar represents the BEST system, and the right bar, the single polymer conditioning after centrifugal thickening.

Note 2: The operation hours for the conventional system indicate the hours required to dewater the same amount of sludge that the BEST system dewater within 7 hours.

Figure 26. Comparison of operational cost.

According to the results, the estimates for the BEST system are a little bit high compared with those for the conventional system when handling the Mixed Raw Sludge. The moisture contents of the dewatered sludge cake dewatered with both systems are the same, however, though the achievable filtration rates differ significantly between the two. When 7-hour operation of a belt-press filter is required to dewater the Mixed Raw Sludge with the BEST system, 11.4-hour operation of the same belt-press filter will be necessary to dewater the same amount of sludge processed with the control system.

9. Conclusion

The following is obtained in this study of the BEST system.

1) The BEST system makes it possible to dewater any of the mixed raw sludge of the conventional activated sludge process, the excess sludge of the conventional activated sludge process, the excess sludge of an anaerobic/aerobic process, and the excess sludge of an oxidation ditch process sludges, without any of conventional sludge thickening processes.

2) The required dosage of the metal coagulant aid for the BEST system ranges between 0.6 and 0.8% as Aℓ to SS of any sludges examined in this study.

3) The dosage of the amphoteric polymer for the BEST system is about 1% to SS for any of the fed sludges. The required dosage of the amphoteric polymer is almost the same, or slightly larger, compared with that of the cationic polymer dosage in a single polymer conditioning system.

4) The sludge pelletizing thickener should be operated with 3 to 4% of the solid content, and with 6 minutes, or more, of SSRT when the mixed raw sludge is fed into the system. When excess sludges are fed into the system, the sludge pelletizing thickener should be operated with 2 to 3% of the solid content, and for 12 minutes, or more, of SSRT. The system is very compact compared with a gravitational thickener.

5) The dewatering performance of the BEST system revealed that the moisture content of dewatered sludge cake can attain the target volume of 75%, or less, with a filtration rate of 200 kg-DS/m/h, when a mixed raw sludge is fed to the system and dewatered with a belt-press filter. When excess sludges are fed to the BEST system, the dewatering performance was improved compared with conventional system.

6) The running costs for this system are estimated to be almost the same, or slightly more compared with that for the mechanical thickening dewatering system, which employs a centrifuge for thickening, and the single polymer dosing to condition the sludge.

7) The release of phosphate in the system is almost completely cut off by dosing the metal coagulant aid.

For these characteristics of the BEST system, the following applications are now under consideration.

(1) Application to sewage treatment plants, where the gravitational sludge thickeners do not work efficiently.

(2) Application to small size sewage treatment plants, as an itinerant sludge handling system by trucking a belt-press filter combined with this system.

(3) Application to the sewage treatment plants where biological phosphorus removal processes are employed.

Reference

1. Annual Technical Report, Sewage Works Bureau of Tokyo Metropolitan Government, March, 1989.

Operation of Centralized Sludge Treatment Plant

by

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City of Yokohama

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on

Sewage Treatment Technology

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Abstract

The construction of sewage works in the city of Yokohama has progressed at a rapid pace ever since the first wastewater treatment plant went into operation in 1962. As of 1984, all eleven of the planned wastewater treatment plants were in operation. In March of 1989, the percent of sewered population reached 80 percent, or 2.54 million of the city's total population of 3.15 million.

The quantity of sludge that has to be treated and disposed of has, in the meantime, increased along with this construction. Recognizing the crucial importance of constantly keeping abreast of the ever-increasing quantity of sludge, the city has been pursuing technological development aimed at establishing more effective and efficient methods of sludge treatment and disposal.

In view of this city's rapid urbanization and the rising citizen concern for improving and preserving the environment, the city decided to modify the sewerage plan, shifting from the former system of decentralized sludge treatment at each of the eleven wastewater treatment plants to a centralized treatment system at two sludge treatment centers constructed within two waterfront area wastewater treatment plants.

One of these sludge treatment centers (Hokubu) began operating in September 1987. The center contains egg-shaped digesters, the first of their kind in Japan. To insure the efficient operation of these digesters, centrifugal thickeners are employed during pretreatment, enabling high-concentration digestion. In addition, the digestion gas energy is recovered as electrical power owing to a gas power generation system that transforms the energy. Furthermore, before incineration, the sludge cake is dried using the heat released from the incineration system, thus enhancing the efficiency of incinerator operations. These and similar measures demonstrate the city's efforts to incorporate technological development into treatment facilities in order to optimize the total system.

Full-scale operation records for approximately one year indicated that power generated from digestion gas supplied 64 percent of the center's total power needs. This figure amounts to an annual savings of ¥ 200 million for the present volume of sludge treatment, which is one-fifth of the planned volume. The record also confirms the advantages of centralized treatment; it was found that sludge treatment cost at Hokubu was less than half the sludge treatment cost for the same quantity of treatment processed at the individual wastewater treatment plants.

This report presents the center's operational record for the year in question, the actual status of energy recovery and recycling at the center, and current modes for effective utilization of incinerated sludge ash.

1. Foreword

The quantity of sewage sludge generated in cities has been steadily increasing with the expansion of sewerage systems. The increase is particularly eminent in major cities, which not only shoulder the enormous burden of costs required for treatment and disposal, but also are finding it increasingly difficult to acquire the needed disposal sites.

In the city of Yokohama, the construction of the sewerage system has progressed at a rapid pace ever since the first wastewater treatment plant went into operation in 1962. As of 1984, all eleven of the planned wastewater treatment plants were in operation. As of March 1989, the sewerage service rate reached 80 percent, or 2.54 million residents of the entire population 3.15 million.

The quantity of sludge generated in the city has been expanding each year with this progress. Efficient and effective treatment and disposal of sludge are becoming increasingly important issues. Moreover, the city experienced rapid urbanization during Japan's recent period of high-level economic growth. Subsequently, the improvement and preservation of the environment have become major concerns of city residents. In view of these factors, the city decided to shift from its former system of decentralized sludge treatment at each wastewater plant to a new system of centralized treatment at two sludge treatment centers constructed at two wastewater treatment plants located in the waterfront area.

One of these sludge treatment centers (Hokubu) went into operation in September 1987. This report presents the operational record of this facility and the effects of centralized sludge treatment.

2. Current status of sewerage system construction in Yokohama

Full-scale construction of sewerage system including wastewater treatment plants in Yokohama dates from 1957. This construction is being promoted in a series of five-year plans, the sixth of which is now underway, in a program called the Five-year Program for Sewerage Construction. In the process, the city's sewerage system has greatly expanded. Initially, the construction was confined to the oldest parts of the city in the waterfront area; it now covers all areas of the city. At present, the main priority is an increase in the sewered population in the city as a whole, and particularly in suburban areas.

In the sewerage system plan, the city is divided into nine treatment districts and includes 11 wastewater treatment plants, as shown in Figure 3-1 in the following section. With the initiation of the newest plant in 1984, all eleven of the planned wastewater treatment plants were in operation 22 years after the start-up of the first plant in 1962.

The original plan called for each wastewater treatment plant to carry out its own sludge treatment. Subsequently, however, it was decided to construct sludge treatment centers at two specific plants in the water front zone, where centralized treatment of the sludge generated at each of these plants could be done. This report concerns the operation of the Hokubu Sludge Treatment Center, which began operation in September 1987. The other center--the Nambu Sludge Treatment Center--is scheduled to go into operation during fiscal year 1989. Table 2-1 shows the current status of the city's sewerage system construction.

Table 2-1. Status of Sewerage System Construction (as of March 1989)

City area (ha)	43,082
Urbanization promotion area (ha)	32,473
Treatment district area (ha)	22,136
Total city population (thousands of persons)	3,158
Sewered population (thousands of persons)	2,542
Percent of sewered population (%)	80
Extended sewer length (km)	7,906
Number of WTPs (in operation)	11
Treatment capacity (10^3 m^3)	1,844
Number of pumping stations (in operation)	22
Number of stormwater retention reservoirs (in operation)	3

3. Centralized sludge treatment plans and an outline of the facilities

3-1. Plans and outline

3-1-1. Centralized sludge treatment plans

The city has formulated plans to construct two centralized sludge treatment centers in the waterfront area--the Hokubu and Nambu sludge treatment centers. As shown in Figure 3-1, the plans call for the treatment of sludge generated from the wastewater treatment plants in the northern section of the city, namely, the Midori, Kokubu, Hokubu No. 1, Hokubu No. 2 and Kanagawa plants, at the Hokubu Sludge Treatment Center, and for the treatment of sludge generated from those plants in the southern part, namely, the Chubu, Nambu, Kanazawa, Sakae No. 1, Sakae No. 2, and Seibu plants, at the Nambu Sludge Treatment Center. Table 3-1 shows the outline of the city's wastewater treatment plants.

As part of these plans, sludge is transported to the centers by means of pressurized pipelines. The sludge is not routed directly from the wastewater treatment plants to the centers, however. Wherever possible, it is routed through one or more wastewater treatment plants served by the same center.

Table 3-2 shows an outline of the sludge treatment center plans. Several treatment processes are planned for both centers in order to provide flexibility in sludge treatment and accommodate future technological innovation.

3-1-2. Current construction status of the centralized sludge treatment plant

As of March 1989, virtually all of the sludge pipelines had been completed. Routes 11 and 12 are still under construction, but routes 1 through 6 are already in operation, and routes 7 through 10 are scheduled for operation during fiscal year 1989.

Table 3-3 shows the construction status of the sludge treatment facilities at the Hokubu Sludge Treatment Center. In September 1987, the Center began full-scale operation based on a sequence process of thickening, digestion, dewatering, and incineration. The Nambu Sludge Treatment Center is scheduled for operation during fiscal year 1989.

Fig. 3-1. Map of Sludge Pipelines

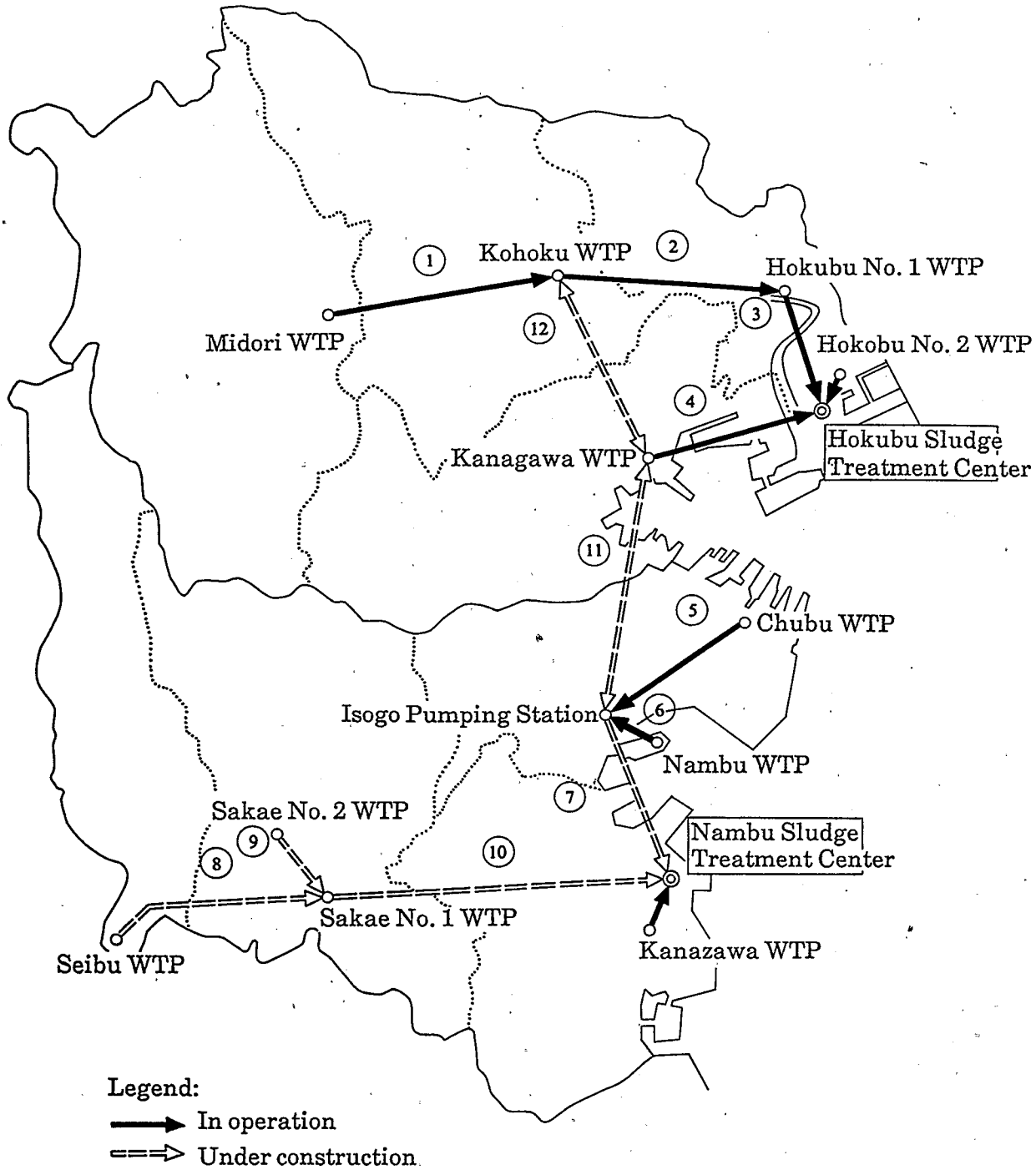


Table 3-1. Sewage Treatment Plants

(as of March 1986)

Treatment plant	Design treatment capacity (m^3/day) (population)	Current treatment capacity (m^3/day)	Operation started
1. Hokubu I	196,000 (377,000)	196,000	July '68
2. Hokubu II	86,400 (166,000)	86,400	Aug. '84
3. Kanagawa	543,200 (1,074,000)	380,200	Mar. '78
4. Chubu	96,300 (185,000)	96,300	July '62
5. Nambu	255,000 (433,000)	225,000	July '65
6. Kanazawa	345,000 (663,000)	230,000	Oct. '79
7. Kohoku	439,000 (844,000)	172,500	Dec. '72
8. Midori	433,000 (833,000)	126,000	May '77
9. Sakae I	124,000 (238,000)	62,000	Dec. '84
10. Sakae II	206,000 (396,000)	206,000	Oct. '72
11. Seibu	191,000 (367,000)	63,600	Mar. '83

Table 3-2. Outline of Centralized Sludge Treatment Plans

Center	Wastewater treatment plants covered	Capacity (TS: 1%)	Sludge treatment processes
Hokubu	Midori Kohoku Hokubu No.1 Hokubu No.2 Kanagawa	44,100 m^3/day	Thickening → anaerobic digestion → dewatering → incineration (gravity-type/centrifugal) (centrifugal) ↘ drying Thickening → wet oxidation → dewatering (centrifugal) (filter press)
Nambu	Chubu Nambu Kanazawa Sakae No.1 Sakae No.2 Seibu	30,000 m^3/day	Thickening → anaerobic digestion → dewatering → incineration (centrifugal) (belt press) Thickening → wet oxidation → dewatering (centrifugal) (filter press)

Table 3-3. Construction Status of Hokubu Sludge Treatment Center Facilities

(as of March 1989)

Facility	In operation	Under construction
Sludge intake tanks	1,500 $m^3 \times 2$	—
Thickening tanks	1,260 $m^3 \times 8$	—
Centrifugal thickeners	100 $m^3/hr \times 3$	—
Digesters	6,800 $m^3 \times 6$	6,800 $m^3 \times 6$
Centrifugal dehydrators	50 $m^3/hr \times 3$	—
Incinerators	100 wt/day $\times 1$	—
	150 wt/day $\times 2$	150 wt/day $\times 1$
Gas engine power generators	920 kW $\times 3$	920 kW $\times 1$
Gas holders	8,000 $m^3 \times 1$	8,000 $m^3 \times 1$

Note: The abbreviation wt refers to wet ton.

3-2. Outline of centralized sludge treatment (Hokubu Area)

3-2-1. Sludge transportation system

The sludge generated from the primary and final sedimentation basins at wastewater treatment plants is transferred, in mixed form, to concentration conditioning tanks, where the TS concentration is adjusted to about 1 percent. The sludge is then stored in a holding tank and transported through pipelines pressurized by pumps to the Hokubu Sludge Treatment Center.

The sludge generated at the Midori plant is first transported via the Kohoku plant to the Hokubu No. 1 plant. At each plant, the sludge is mixed with that generated from the wastewater treatment plant, and then transported to the Center.

The sludge generated at the Kanagawa plant is transported directly to the Center after its TS concentration has been adjusted. The sludge generated at the Hokubu No. 2 plant is also transported directly to the Center, in the form of primary sludge, i.e., sludge drawn from the primary sedimentation tank.

Ideally, the operation and management of the sludge transportation and intake system should be integrated with that of sludge treatment, the two forming a single system. Taking this into account, the city has adopted a system of centralized supervision and decentralized control.

3-2-2. Hokubu Sludge Treatment Center

As shown in Table 3-2, the Center has a daily treatment capacity of 44,100 cubic meters of sludge (TS: 1%) through a combination of three processes: anaerobic digestion/dewatering/incineration, digestion/dewatering/drying, and wet air oxidation /dewatering. At present, only the first process (digestion /dewatering/ incineration) is in operation. This section presents an account of the facilities involved in this process. Figures 3-2 and 3-3 present the layout of the Center and the flow of the digestion/dewatering/incineration process, respectively.

(a) Thickening process

The sludge transported from the wastewater treatment plants first enters the sludge intake tanks and is then transported to thickening tanks. In these tanks the sludge concentration is adjusted to about 2 percent and then increased to about 5 percent, by means of a centrifugal thickeners. In recent years, it is presumed that the thickening properties of sludge have deteriorated as a result of the rise of sludge VTS and decomposition during long-term transportation. Consequently, gravity thickening is supplemented with mechanical thickening to provide for the stability of, and alleviate the burden on, subsequent processes. The Center has adopted a mode of centrifugal thickening for this purpose, seeing that maintenance is relatively simple and the equipment's compact size facilitates installation.

(b) Digestion process

The Center has adopted an anaerobic digestion process in the interests of enhancing the overall flexibility of the facilities and making effective use of the gas generated in digestion. The egg-shaped tanks used for this digestion are the first of their kind in Japan and offer various advantages; they generate an extremely high agitation efficiency, ensure a minimal generation of scum, and facilitate the easy removal of sediment. In addition, the construction cost per unit of treated solids is virtually the same as for conventional tanks. There are six such tanks, each with a capacity of 6,800 m^3.

(c) Power generation process using digestion gas

The heat and power, derived as a result of the power generation using digestion gas, are effectively utilized for the purposes of heating the digesters and supplying power for various facilities within the Center. The system now consists of three generators in the 920-kilowatt class.

(d) Dewatering process

The Center uses centrifugal dehydrators to dewater the sludge after digestion. These centrifugal units are easy to maintain and offer a high solids capture rate and contribute to efficient operations. There are now three such units in operation, each with a capacity of 50 cubic meters per hour.

(e) Incineration process

To reduce the quantity of sludge for disposal, fluidized bed incinerators are used to incinerate the entire quantity of sludge cake. The temperature of the incinerator exhaust gas is about 800°C, and the exhaust gas is used for drying sludge cake before incineration; thus increasing the efficiency of the operation. There are now three such incinerators operating in the Center, one with a daily capacity of 100 wt and two with a capacity of 150 wt per day.

(f) Sidestreams treatment process

Various types of sidestreams are separated out in the sludge treatment processes (e.g.,thickener effluent, supernatant from digestion tanks, filtrate from dehydrators, and scrubber drain). There are plans to first pretreat these sidestreams at facilities constructed for this purpose within the Center, and then undergo secondary treatment process.

Fig. 3-2. Layout of the Hokubu Sludge Treatment Center

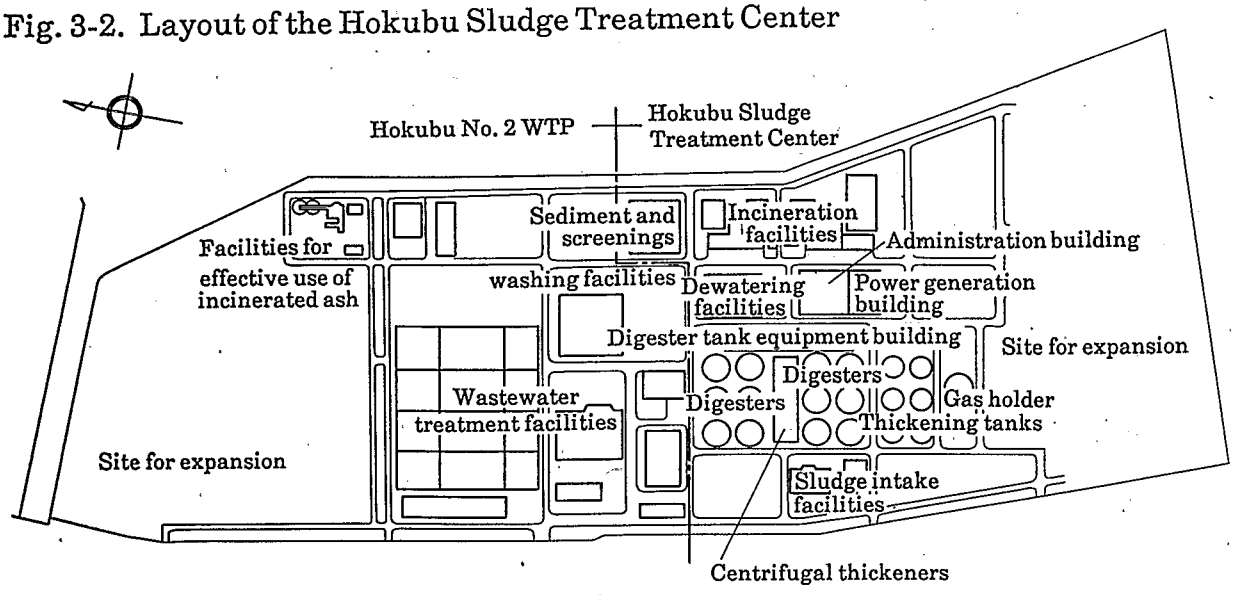
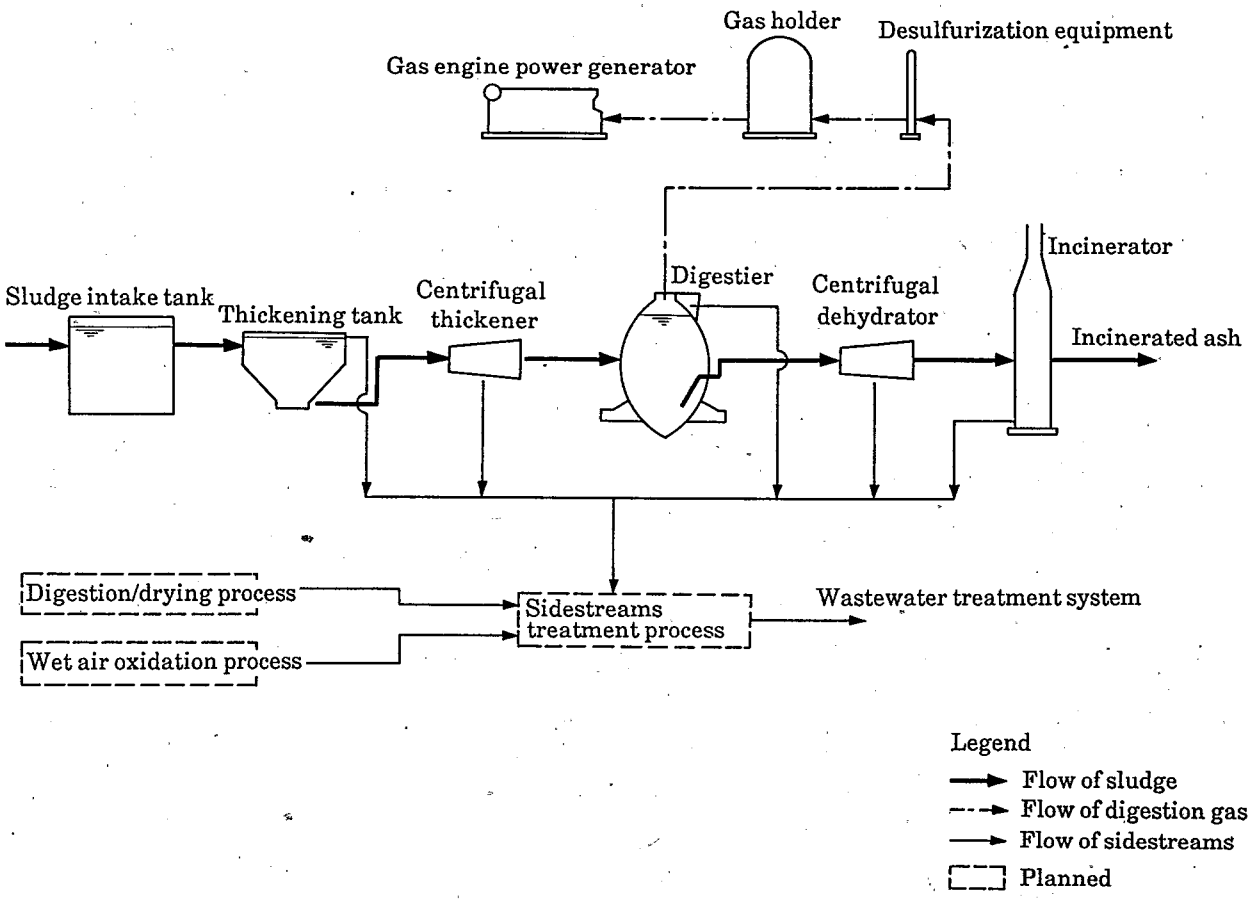


Fig. 3-3 Flow of Anaerobic Digestion/Dewatering/Incineration Process Flow Chart



At present, these plans are at the preliminary stages of process selection. In the meantime, the sidestreams generated at the Center are being sent back for treatment at three wastewater treatment plants (Hokubu No. 1, Hokubu No. 2, and Kanagawa), distributed according to the load produced by such treatment.

3-2-3. Instrumentation

(1) Outline

The instrumentation systems of the sludge transportation facilities at each wastewater treatment plant are linked to the centralized supervision and decentralized control system at the Center. This facilitates the integrated management of the sludge intake facilities, thickening and digestion facilities, and other sludge treatment facilities.

The main computer system, which is at the core of this instrumentation system, is equipped with a backup system for reliability and safety. The backup system assures operational continuity in the event of an unplanned shutdown or planned shutdown in accordance with the maintenance schedule of the other system. Even if both systems are shut down, the facilities can be monitored and operated through the monitor panel located within the central supervision and control room.

At the same time, each of the treatment facilities within the Center is equipped with a local station, microcontroller, and sequence controller. These devices can automatically control the facilities and exchange information with the main computer. Utilizing optical fiber cable, information is transmitted between the main computer and local stations through a high-speed dataway.

Figure 3-4 shows the Composition of Instrumentation at the Hokubu Sludge Treatment Center.

(2) Sludge transportation facility instrumentation

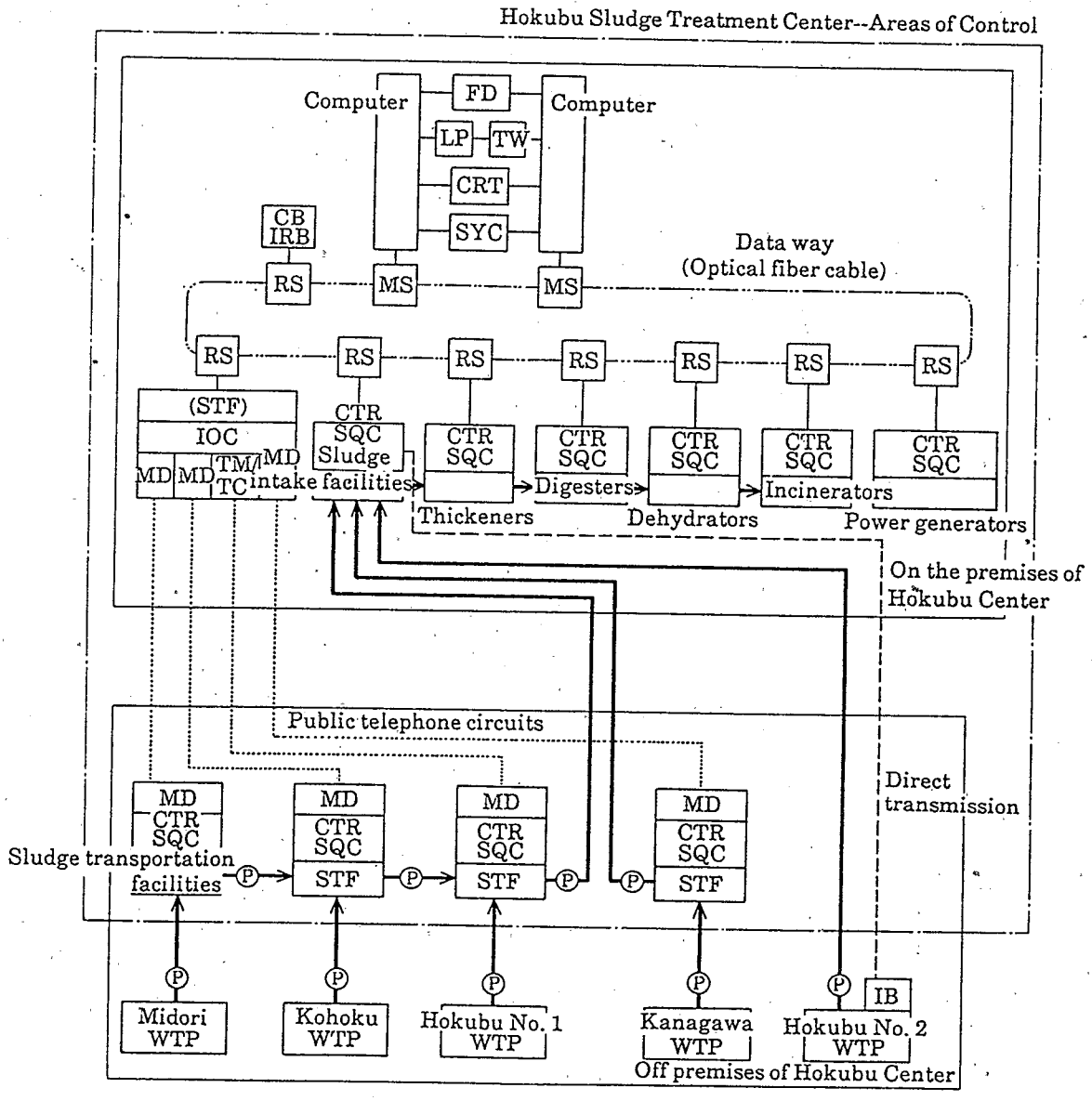
The function of controlling sludge transportation facilities is to assure the smooth transportation of sludge generated at each of the wastewater treatment plants to the Center. Sludge from numerous plants is transported through a single pipeline in a cascade. For this reason, coordinating the transportation time sequence at each transportation facility is a major part of the operation. To effect such coordination, the transportation facilities are controlled by a daily sludge transportation schedule.

The schedule is developed in accordance with the Center's general operational plans, which are based on estimates of the quantity of sludge generated at the wastewater treatment plants, and is formulated by main computer operators. The basic goals behind creating the plans are to maximize the total quantity of sludge transported per day, assure that the minimum required flow speed within the pipeline is maintained, and that a proper balance between the quantity of sludge entering the holding tanks and that of the transported sludge is upheld.

Sludge transportation is automatically initiated when the time specified in the transportation schedule is reached or when the level of sludge in the holding tank in the transportation facilities reaches its upper limit. Transportation is automatically terminated when the quantity specified in the schedule is reached (as calculated from the quantity and frequency of sludge transportation for the day in question) or when the level of sludge in the holding tank reaches its lower limit.

The sludge transportation facilities are also equipped with a timer control, which has a sequence controller as a backup system in the event of microcontroller failure. In this mode of control, the start-up and shutdown of the sludge transportation pump are based on the time of transportation initiation and the duration of pump operation in each case.

Fig. 3-4. Composition of Instrumentation at the Hokubu Sludge Treatment Center



Legend

FD:	Flexible disk drive		
CB:	Central monitor board		
LP:	Line printer	IOC:	Input/output controller
IRB:	Information retrieval board	CTR:	Microcontroller
TW:	Output typewriter	SQC:	Sequence controller
MS:	Master station	MD:	Modem
CRT:	CRT display	IB:	Instrumentation board
RS:	Remote station	TM/TC:	Telemeter/telecontrol
SYC:	System console	STF:	Sludge treatment facility

Flow of sludge
Direct transmission
Public telephone circuits
Data way

(3) Detection of pipeline abnormalities

Blockage: Sludge transportation pipelines are considered blocked when both the pump discharge pressure is higher than the set value and the quantity of sludge flow is lower than the set value or has rapidly decreased. When blockage is detected, the pump is immediately shut down and a blockage alarm is transmitted.

Leakage: Sludge transportation pipelines are deemed leaking when both the pump discharge pressure is lower than the set value and the quantity of sludge flow is higher than the set value or has rapidly increased. When leakage is detected, the pump is immediately shut down and a leakage alarm is transmitted.

(4) Instrumentation of sludge intake facilities

Because sludge residue within pipelines can cause blockage or generate gas, they must be cleaned after transportation. As a general rule, the pipelines are cleaned once every day. The cleaning is performed by flushing the pipelines with pressurized water, using the sludge transportation pump. After the sludge transportation for the day has been completed, the operator in the central supervision and control room in the Center orders the commencement of the cleaning.

On the intake side, separate tanks are used to receive the sludge and cleaning water. A near infrared-type sludge densitometer is attached to the pipeline on this side to constantly measure the concentration. When the concentration of fluids transported through the pipeline falls below the set level, the intake is automatically switched to the cleaning water tank. The operation makes use of the comparatively well-defined interface between sludge and cleaning water within the pipeline, and of the sharp drop in concentration when sludge is replaced by cleaning water.

4. Operation of Hokubu sludge treatment center facilities

4-1. Operation of sludge transportation system

The Hokubu Sludge Treatment Center treats sludge transported by pipeline from five wastewater treatment plants.

The sludge transportation system consist of transportation facilities (sludge holding tanks and sludge pumps), sludge pipelines, and sludge intake facilities (screenings removal units and sludge holding tanks).

Table 4-1 presents the quantities and the TS and VTS concentrations of sludge generated at each of the five wastewater treatment plants during fiscal year 1988.

Table 4-1. Sludge Quantities and TS and VTS Concentrations

STP / Item	Midori	Kohoku	Hokubu No. 1	Hokubu No. 2	Kanagawa
Quantity of sludge generated (m^3/day)	540-870 (680)	250-600 (360)	630-1,000 (760)	160-660 (480)	1,750-2,770 (2,120)
TS (%)	1.0-3.4 (2.4)	1.9-4.0 (3.4)	1.5-5.5 (3.3)	0.8-2.2 (1.4)	1.4-3.2 (2.0)
VTS (%)	49-88 (80)	51-79 (69)	52-84 (67)	66-83 (75)	62-85 (78)

Note: First figures indicate the minimum and second figures indicate maximum average weekly values over a one-year period. Figures in parentheses indicate total yearly averages.

The sludge is transported from the Midori plant to the Kohoku plant and further to the Hokubu No. 1 plant with treated wastewater through a pipeline. In order to separate the sludge from the water, a near infrared reflective-type sludge densitometer (TS meter) is attached to the pipeline in front of the sludge intake tank at the latter two plants. The line of separation is a TS concentration of 0.1 percent; the pipeline is emptied into the sludge intake tank if the concentration is over 0.1 percent and into the treatment plant grit chamber if it is under 0.1 percent. Change in TS concentration within the pipeline is well defined, and the separation of sludge and treated wastewater could be effected even at a higher TS concentration setting on the TS meter. However, a setting of 0.1 percent was decided upon to reduce the burden on the wastewater treatment plants.

Only sludge is transported in the pipelines leading to the Center from the Hokubu No. 1, Kanagawa, and Hokubu No. 2 plants. These pipelines are cleaned as required.

Sludge transportation management is based at the Center, and is a system of centralized supervision and decentralized control.

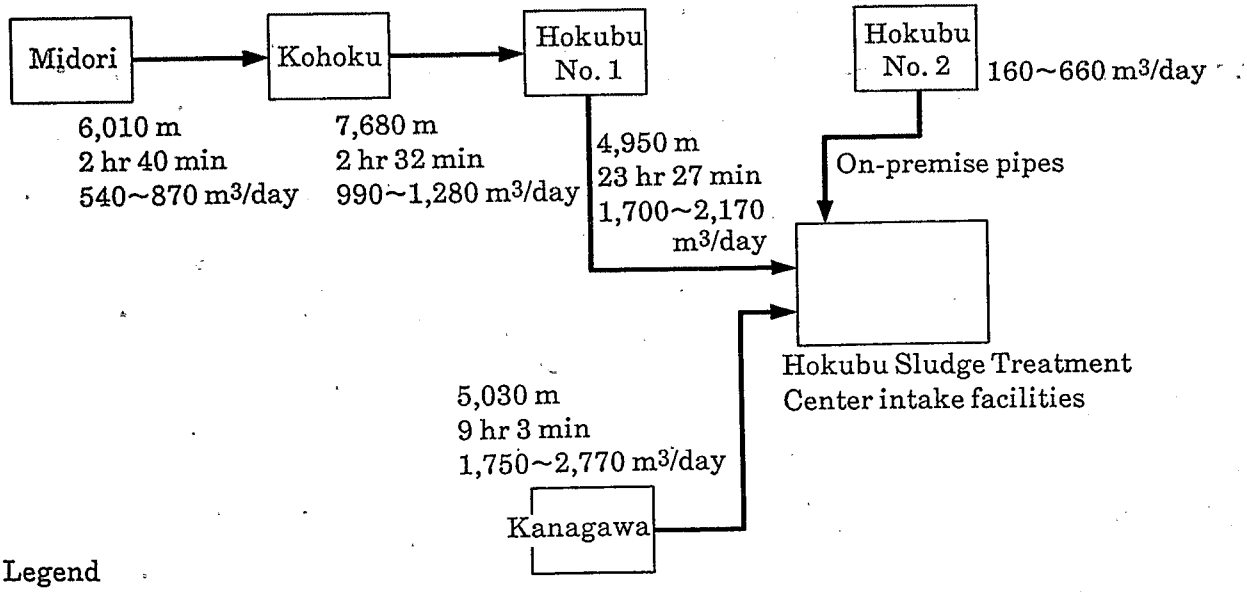
4-1-1. Operation and control of sludge transportation facilities

The time required for the movement of sludge and the sludge transportation time schedule are shown in Figures 4-1 and 4-2, respectively.

Maintenance and inspection of the various transportation facilities is the responsibility of the Center.

Sludge transportation is carried out three times per day. As shown in Figure 4-2, the transportation is initiated at prescribed times and terminated when the prescribed quantity has been transported or when the level of sludge in the holding tank reaches the prescribed quantity. In the event of an increase of sludge due to storms or other developments, agreements are reached with personnel at the various wastewater plants to adjust the aforementioned upper limits and increase the quantity of sludge transported.

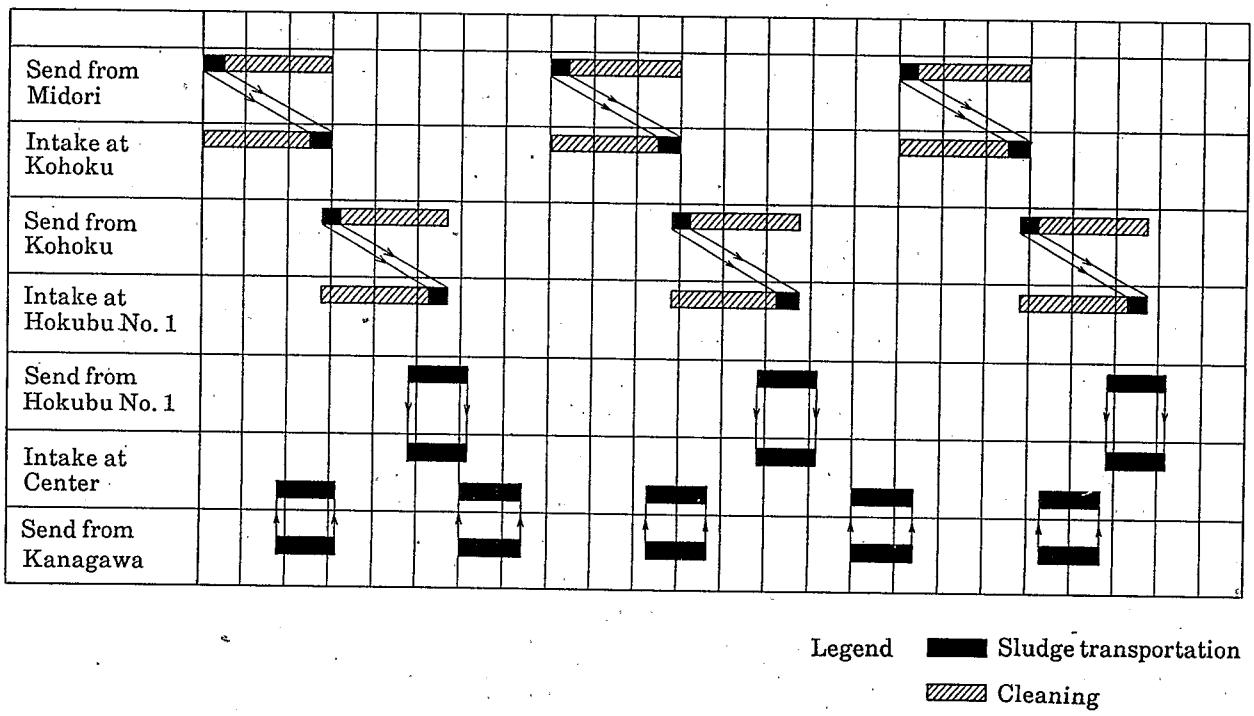
Fig. 4-1. Movement of Sludge from Transportation Facilities to the Center



Legend

First line: section distance
Second line: transportation time
Third line: transportation quantity

Fig. 4-2. Sludge Transportation Time Schedule



4-1-2. Management of sludge pipelines

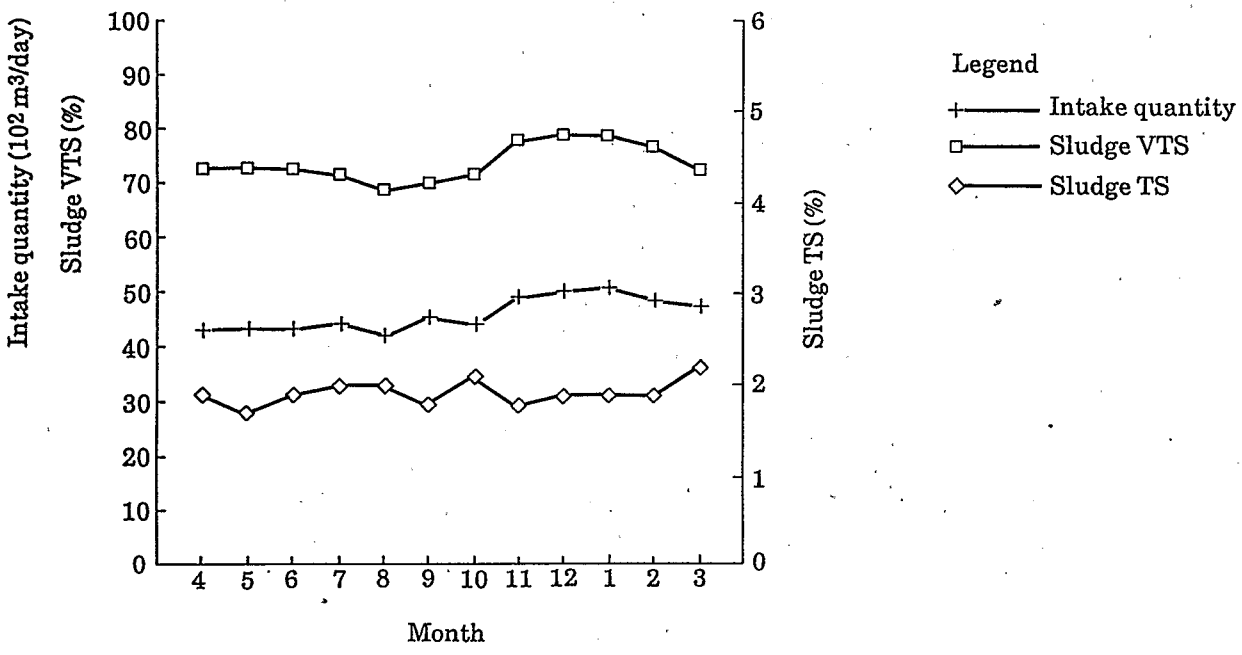
The sludge transportation pipeline is basically composed of ductile castiron pipe (DCIP) in sections that are buried and of fiber glass reinforced plastic mortar (FRPM) pipe in sections that are installed in public utility ducts. As part of the maintenance work, the operation of the pipeline's air valves, sludge discharge valves, and other equipment is checked periodically four times a year.

However slight, pipeline leaks must be immediately detected and repaired since the effluent is sludge. Consequently, all sections are equipped with flow meters on both the sending and receiving sides. Leakage is detected by monitoring and checking for variation between these meters. Similarly, the correlation between the quantity sent and the discharge pressure is used to check for leakage and blockage. At hourly intervals, the quantities sent and received are integrated to check for leaks and to check that operation is proper. The Center also maintains a store of materials for repairing leaks (e.g., various types of short pipe sections, connecting bolts, packing materials, and coupling rings).

4-1-3. Operation of sludge intake facilities

During fiscal year 1988, the quantity of sludge received at the Center ranged from 4,110 to 5,290 m^3 per day. The corresponding ranges for TS and VTS concentration were 1.4-3.0 percent and 62-84 percent, respectively. The trends of sludge intake quantity and TS and VTS concentrations are shown in Figure 4-3.

Fig. 4-3. Monthly Trend of Sludge Intake Quantity, and TS and VTS Concentrations



As shown in Figure 4-3, the intake quantity gradually increases, and is in inverse proportion to that of the TS concentration. This is presumably due to the fact that the quantity transported is increased, in order to maintain the quality of the treated wastewater at the plants when the sludge thickening properties deteriorate due to meteorological changes. It can also be seen that the VTS concentration is lowest during the summer (around August) and highest during the winter (around January). This trend is thought to stem from sludge decay that accelerates with the rise in water temperature.

4-1-4. Changes in sludge properties

As shown in Figure 4-1, the duration of time from the start of sludge transportation to the start of treatment at the Center is longer than one day for the Midori plant, which is the remotest plant. Table 4-2 indicates the change in sludge properties during this time.

Table 4-2. Changes in Sludge Properties during Transportation

	Total quantity at the five wastewater plants (t/day) ①	Intake quantity at Hokubu Sludge Treatment Center (t/day) ②	Rate of change (%) (②−①)/①×100
TS	92.2	92.0	△0.2
VTS	69.5	67.2	△3.3
SS	89.3	73.6	△27.6
DS	2.90	18.4	534
Total nitrogen	6.26	5.34	△14.7
Ammonia nitrogen	0.47	0.63	33.8
Total phosphorus	1.23	1.16	△5.7
Total soluble phosphorus	0.15	0.28	84.0
Volatile organic acid	3.20	5.34	66.9

Note: Figures in column 2 indicate the values for sludge supplied to the centrifugal thickeners at the Center. The triangle △ indicates the decrease in quantity.

In the course of transportation, the sludge decays and solids within it are increasingly dissolved and changed into solubles. As a result, the suspended solids concentration declines and the dissolved solids concentration rises. Naturally, there is also a rise in the amount of organic acid, an indicator of the degree of decay. Nitrogen and phosphorus also dissolve, leading to a rise in the concentration of ammonia nitrogen and total soluble phosphorus.

With this progressive decay of the sludge over time, the thickening property and characteristics of the sludge deteriorate. For this reason, great care must be taken in the operation and management of the centrifugal thickeners and other equipment used in centralized sludge treatment.

4-2. Operation record of the Hokubu Sludge Treatment Center

4-2-1. Outline

Table 4-3 presents the operation record of the Hokubu Sludge Treatment Center during fiscal year 1988. The entire quantity of sludge intake was subjected to the sequential process of centrifugal thickening, digestion, dewatering, and incineration. The uses of the gas generated during digestion greatly depend on the volume of gas generated and the operational status of the incineration facilities. During fiscal year 1988, about 30 percent of this gas was used as auxiliary fuel for the incinerator; the remainder was used as fuel for the gas engine of the electrical power generator.

Table 4-3. Hokubu Sludge Treatment Center Operation Record (FY1988)

	Quantity treated	TS (%)	VTS (%)
Sludge intake	1,606×10³ m³/year 4,110~5,290 m³/D (4,400)	1.4~3.0 (2.0)	62~84 (74)
Centrifugal thickeners	1,793×10³ m³/year 4,220~5,560 m³/D (4,900)	1.6~2.3 (1.9)	64~83 (73)
Digesters	560×10³ m³/year 1,100~5,650 m³/D (1,530)	4.2~6.2 (5.2)	63~84 (72)
Digested sludge	589×10³ m³/year 1,100~2,000 m³/D (1,560)	2.6~4.0 (3.2)	60~65 (56)
Sludge cake	76,826 tyear 130~270 t/D (210)	18~24 (22)	50~67 (57)
Incineration	83,280 t/year 170~360 t/D (230)	--	--
Incinerated ash	15,270 t/year 24~55 t/D (41.8)	--	--
Digestion gas generated	10,500×10³ Nm³/year 21400~37000 Nm³/D (28,800)	--	--
Electrical power generated	13,362 kWh/year 23000~51800 kWh/D (36,600)	--	--

Note: The difference between the quantity of sludge cake and the quantity incinerated reflects the transport of sludge cake from other wastewater treatment plants for incineration. Figures in parentheses indicate averages.

4-2-2. Operation of thickening facilities

As noted in section 4-1-4, the sludge received at the Center is in a state of advanced decay, which would seriously lower the effectiveness of gravity-type thickening. As a result, thickening is carried out employing a horizontal-type continuous centrifugal thickener, after the sludge has been passed through a drum-screen-type racks. The sludge TS concentration was in the range of 1.6-2.3 percent before the thickening process and 4.2-6.2 percent after it. The screenings separated out through the mechanically cleaned bar racks (amounting to 860 tons for the year) were first dewatered using a screw press and then transported by pneumatic conveyors to the incinerator for incineration.

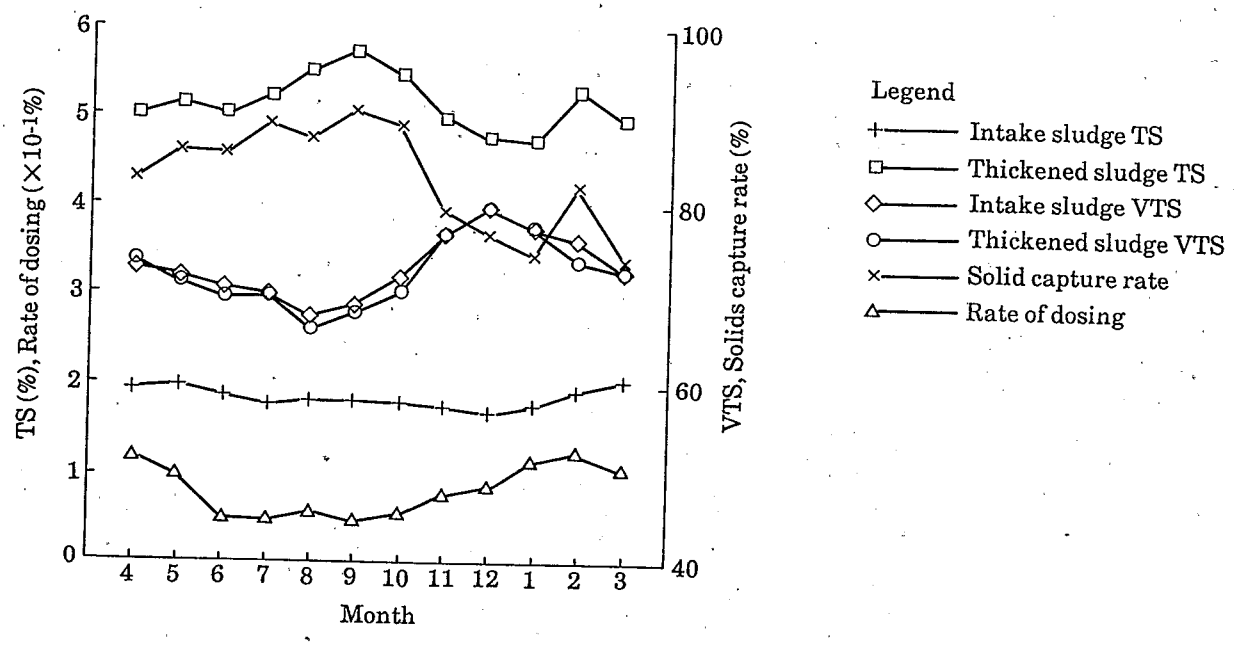
Table 4-4 shows the quantity and TS and VTS concentrations of sludge treated during fiscal year 1988. Figure 4-4 presents monthly changes in centrifugal thickenr operations.

Table 4-4. Quantity Treated and TS and VTS Concentrations (FY1988)

	Quantity treated per day (m^3)	TS (%)	VTS (%)
Intake sludge	4,220~5,560 (4,910)	1.6~2.3 (1.9)	64~83 (73)
Thickened sludge	1,100~1,900 (1,530)	4.2~6.2 (5.2)	63~84 (72)

Note: Figures in parentheses indicate yearly averages.

Fig. 4-4. Monthly Changes in Centrifugal Thickener Operations



The centrifugal thickneners recorded an operation rate in the range of 64-84 percent of full capacity. The centrifugal thickeners were operated mainly through dosing (chemical feeding), and the target capture rate for SS was 90 percent. In addition, since the thickener effluent was sent back to the wastewater treatment plants, the rate of dosing with polymer coagulants was controlled appropriately in order to improve the properties of the effluent. The centrifugal effect, weir radius, and differential speed were adjusted in accordance with the TS concentration of the intake sludge. The average values for centrifugal effect and differential speed were 1,016 G and 17.8 rpm, respectively.

There was a slight increase in the amount of intake sludge, but the quantity of thickened sludge remained fairly constant. As shown in Figure 4-4, there was little variation in the TS concentration of the intake sludge, but the VTS concentration rose during the winter. Even when the rate of dosing with polymer coagulants was increased, it was impossible to obtain a stable retrieval of solids during the winter. The average solids capture rate for the year was about 83.3 percent.

4-2-3. Operation of digestion facilities

The Center contains egg-shaped digestion tanks, each with a capacity of 6,800 cubic meters. The tanks are organized in series of three tanks each. During fiscal year 1988 two such series (six tanks) were used for anaerobic digestion. All tanks function as primary tanks. The operation specifications were as follows: a digestion temperature of 35 degrees centigrade, an average holding time of 27 days, and an average intake TS concentration of 5.1 percent. For the operation method employed there was high concentration and mesophilic digestion. The agitation was mechanically performed using a draft tube, and formed a downward flow within the tube. An outline of the digesters and their operational conditions are shown in Table 4-5.

Table 4-5. Outline of Digesters' Operation Condition

	Conditions	Comments
Quantity of sludge withdrawn	45 m^3	Automatic control within the scope of liquid level change
Quantity of sludge recycled	2 m^3/min	Continuous 24-hour operations
Heating	Recycled sludge heated by means of warm water; the warm water circulation pump is started at 34.5°C and shut down at 36°C.	External heating using a heat exchanger (heat source: waste heat from gas engine)
Agitator	400 rpm	Continuous agitation in form of a downward flow within the tube; agitating intensity is 3,000 m^3/hr (10 times tank volume per day)

All tanks were operated as primary tanks because of the poor solids-liquid separation in the secondary digesters. This is probably due to the fact that settlability of digested sludge was low and transferring the sludge from the primary to the secondary tanks, and with drawing the sludge out of the secondary tanks occurs so frequently that there is not enough time for the sludge to remain still. And also the difference in temperatures of the sludge in the primary tanks and that of the secondary tanks is such that heat convection occurs.

The intake pump was used to inject the thickened sludge into the sludge circulation pipe, where it was mixed with the recycled sludge. The average daily

intake quantity per tank was about 256 m^3, and the average quantity of sludge withdrawn per tank was about 260 m^3 per day. Scum generation has not been detected since the start of operation in September 1987. The average pH values and alkalinity within the digestion tanks were about 7.3 and 4,100 mg/l, respectively, with little variation.

Figure 4-5, 4-6 show the monthly trend of digester operation indicators.

Fig. 4-5. Monthly Trend of Digester Operation Indicators (I)

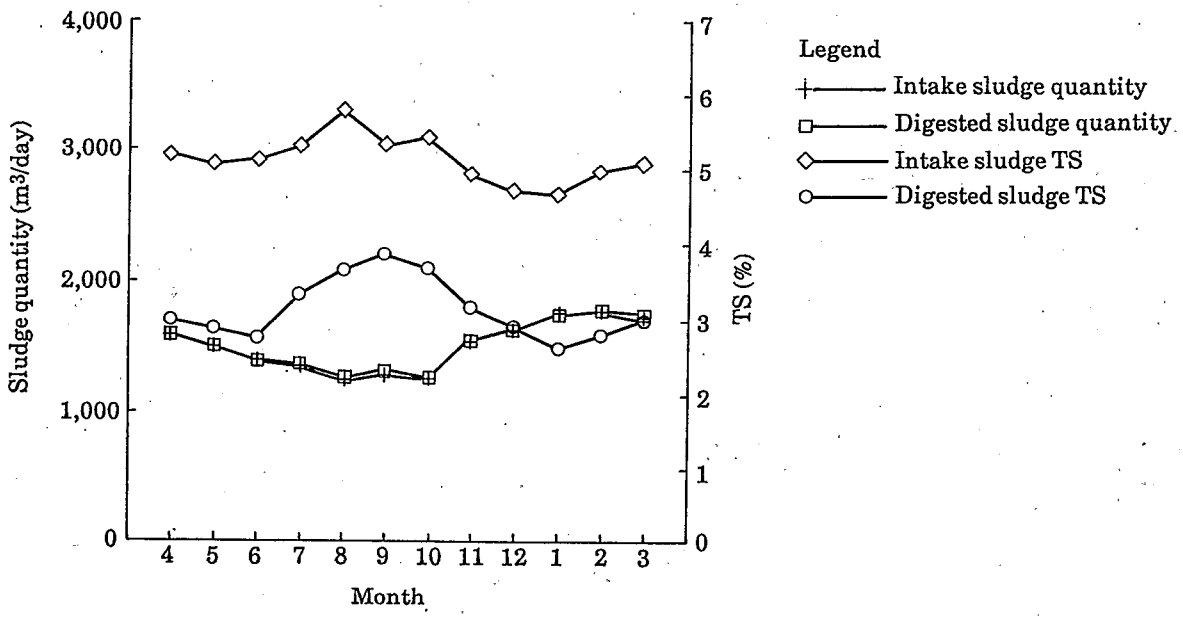
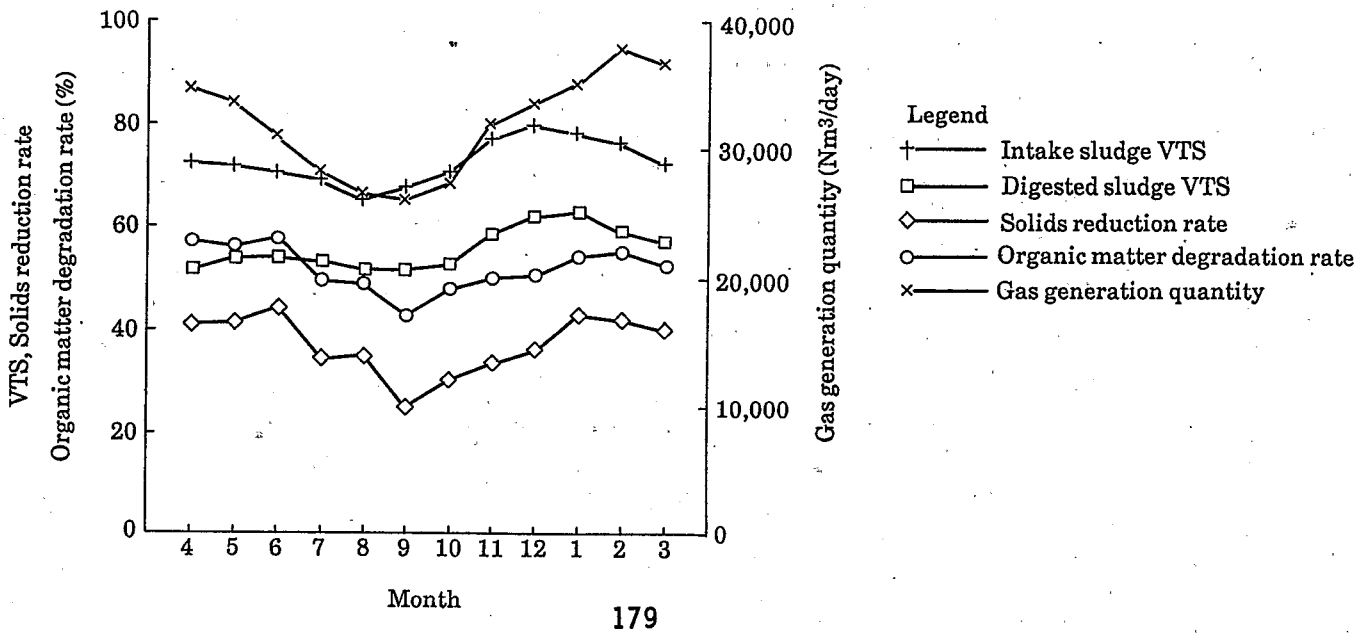


Fig. 4-6. Monthly Trend of Digester Operation Indicators (II)



As shown in Figure 4-5, 4-6, the trends of both the organic loading and the intake sludge quantity exhibit basically the same patterns of decrease in the summer and increase in the winter. Virtually the same patterns characterizes the trends of the VTS concentration of intake and withdrawn sludge. By contrast, the pattern for the TS concentration of intake and withdrawn sludge is one of increase in the summer and decrease in the winter--the inverse of the VTS concentration pattern. The trends of the respective rates of solid reduction and organic matter degradation resemble those of VTS concentration. The trend of the quantity of digestion gas generated resembles that of organic loading, and indicates that a sufficiently stable digestion was carried out at that load level. Since all of the tanks were operated as primary tanks without supernatant runoff, the quantity of intake thickened sludge was almost the same as that of withdrawn sludge.

Table 4-6 presents additional data for the operational record during fiscal year 1988.

Table 4-6. Digesters' Operation Record for FY1988

Item	Operation record
Quantity of thickened sludge intake (m^3/day)	1,100~1,900 (1,530)
TS (%)	4.2~6.2 (5.2)
VTS (%)	63~84 (72)
Quantity of digested sludge (m^3/day)	1,100~2,000 (1,560)
TS (%)	2.6~4.0 (3.2)
VTS (%)	50~65 (56)
Digestion period (number of days)	21~36 (27)
Digestion temperature (°C)	35.1~35.3 (35.1)
Amount of heat used for heating (1,000 kcal/m^3)	9.3~28.7 (18.2)
Volumetric loading (kgVTS/m^3/day)	0.96~1.9 (1.4)
Solids reduction rate (%)	28~45 (37)
Digestion ratio (%)	39~61 (51)
Digestion gas generation quantity (Nm^3/day)	21,400~37,000 (28,800)
Digestion gas generation rate (Nm^3/kg-VTS)	0.3~0.6 (0.5)
Multiplier for digestion gas generation (Nm^3/m^3·day)	16.7~21.0 (18.7)

Note: Figures in parentheses indicate yearly averages.

Table 4-7 shows the composition of the digestion gas. Hydrogen sulfide contained in the digestion gas was removed using a wet-type desulfurization unit, in order to enable using it as a source of heat for the incinerator and a source of power for the gas engine-driven power generator. The hydrogen sulfide concentration averaged 148 (30 - 600) ppm at the unit inlet and 0.9 (0 - 12.5) ppm at the unit outlet.

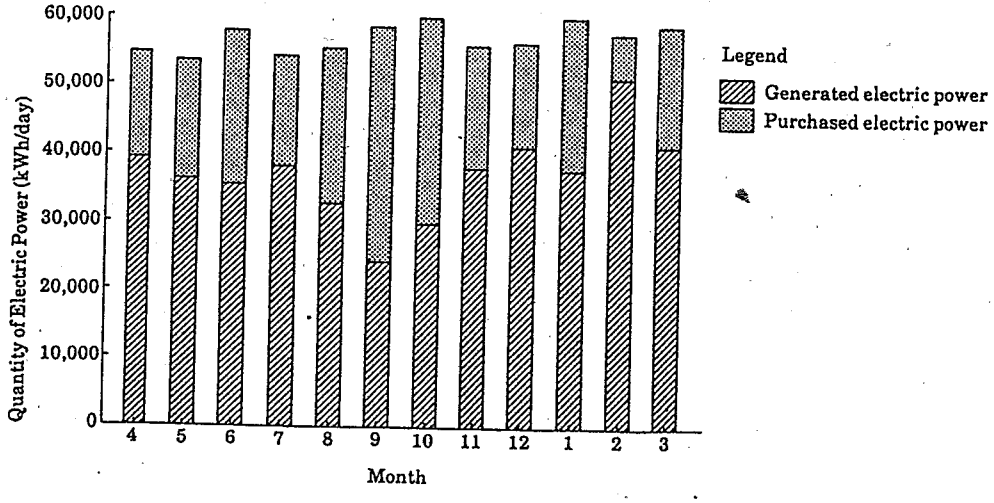
Table 4-7. Concentration of Digestion Gas Components

Constituent components	Concentration (v/v%)
Methane	59~62
Carbon dioxide	34~37
Hydrogen sulfide	30~600 ppm
Moisture	2.3~3.2

4-2-4. Operation of gas engine power generation facilities

After the digestion gas was removed from the gas holder, a gas booster was used to raise its pressure to about 1 kilogram-force per square centimeter. In this form, the gas was used as fuel for the gas engines driving power generators. The engines are water-cooled, 12-cylinder, boiling coolant gas models with a horsepower of 1,350 ps. There are directly connected to the generators, which are triphase current synchronized models (920 kW, 50 Hz, 6.3 kV). The generators constantly operated parallel with commercial power sources, and supplied a portion of the Center's power needs. The number of generators in operation were regulated in accordance with the quantity of digestion gas in the holding tank. The quantity of digestion gas used for power generation and of the generated power are shown in Figure 4-7.

Fig. 4-7. Monthly Trend of the Quantity of Digestion Gas Used for Power Generation and of the Generated Power



The daily average for the quantity of digestion gas used for power generation was 21,300 Nm^3, of the power so generated, the daily average was 43,000 kWh, and of the power used for the auxiliary generator unit, this was 6,400 kWh. The effective quantity of power generated (excluding the power required for operation of the auxiliary unit) averaged 36,600 kWh per day. This figure represents about 64 percent of the power used within the Center during the year in question.

The lower calorific value of the digestion gas was 5,500 kcal/Nm^3. The rate of power generated was 31.5 percent of the amount of heat supplied to the gas engine (26.8 percent when the power consumed by the auxiliary unit is deducted). The generated power per digestion gas unit was 2.02 kWh/Nm^3.

The waste heat from the generator (contained in the cooling water for the engine and in exhaust gas) was retrieved in the form of pressurized steam using a heat exchanger (the steam temperature averaged 121 degrees centigrade). The quantity of heat retrieved was equivalent to about 35 percent of that supplied to the gas engine. The pressurized steam was used as a source of heat for warming the digesters and for air conditioning. The quantity of heat used for heating the digesters ranged from 12,000 to 51,000 Mcal (averaged 28,000 Mcal) per day; the average utility rate was equivalent to 23.9 percent of the amount of calories supplied to the gas engine. The calorific value for heating the digesters per unit volume of intake sludge, varied greatly depending on the season. The respective maximum and minimum values per cubic meter were 28.7 Mcal (during March) and 9.3 Mcal (during August); the average value was 17.7 Mcal.

4-2-5. Operation of dewatering facilities

After digestion, the sludge was dewatered using a horizontal-type continuous centrifugal dehydrator unit. The TS concentration of the digested sludge before dewatering ranged from 2.5 to 4.2 percent. The moisture content of the sludge cake ranged from 76 to 82 percent. Table 4-8 shows the operational record of the dewatering process and Figure 4-8, 4-9 shows the monthly trend of dewatering operation indicators.

Table 4-8. Operational Record of Dewatering Processes

	Daily handling quantity	TS/ moisture content* (%)	VTS (%)
Digested sludge	1,100~2,000 m^3/day (1,600)	2.6-4.0 (3.2)	50~65 (56)
Sludge cake	130~270 t/day (210)	76~82* (78)	50~67 (57)

Note: Figures in parentheses indicate yearly averages.

Fig. 4-8. Monthly Changes in Dewatering Facilities Operations (I)

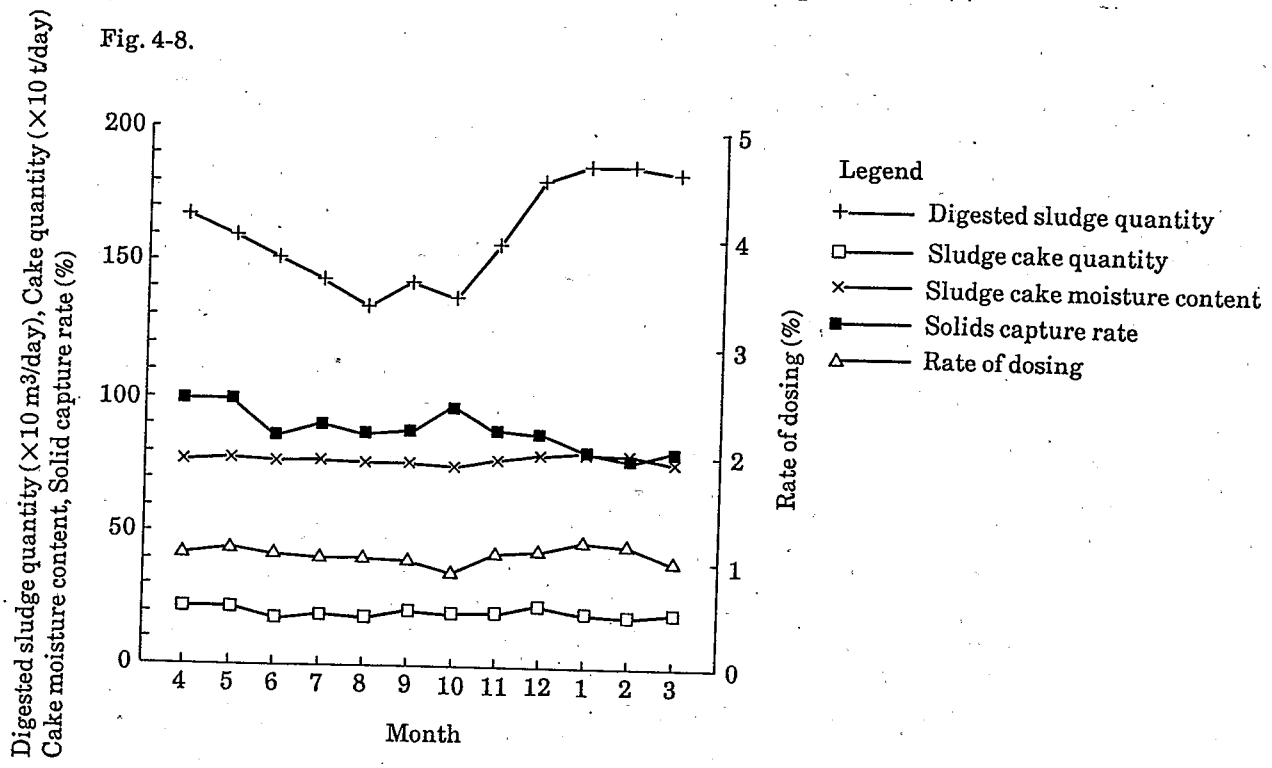
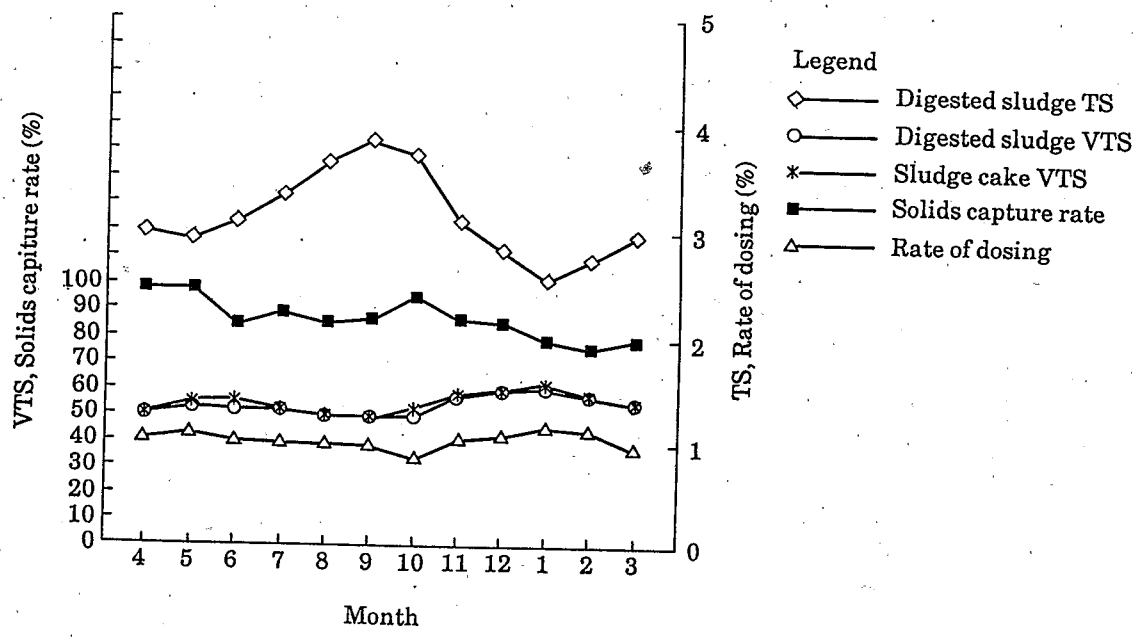


Fig. 4-9. Monthly Changes in Dewatering Facilities Operations (II)



As shown in Figure 4-8, 4-9, there is a basic correlation between the trend of the intake quantity of digested sludge and that of the sludge cake quantity; both exhibit a decline in summer. By contrast, the trend of TS concentration of digested sludge declines in winter. However, the trends of the VTS concentration of digested sludge and sludge cake both show an increase in winter.

This decline in TS concentration and rise in VTS concentration during winter is presumed to be the cause of the rise in the sludge cake moisture content and decline in the solids retrieval rate during the same season. There is a close correlation between the trend of the sludge cake moisture content and that of the digested sludge VTS concentration. When the VTS concentration was high, it proved difficult to maintain the moisture contents on a constant level even when the dosage was increased.

The annual average values for the solids capture rate and of dosage were about 90 percent and 1.06 percent per solid, respectively.

4-2-6. Operation of incineration facilities

The sludge cake generated at the Center was incinerated along with the sludge cake transported from the city's wastewater treatment plants. Incineration facilities consist of the incinerator proper (an erect hollow cylindrical structure), a sludge cake feeder, cake drier, heat recovery unit, and an exhaust gas treatment unit. The incinerators proper--fluidized bed furnaces--are made of fire-resistant brick and castable materials, which are lined on the outside with a steel sheet shell.

The ash resulting from incineration is composed of small particles with a diameter ranging from about 50 to 300 microns. The ash was removed from the incinerator along with the combustion gas and collected by cyclone separators and electrostatic precipitators. The collected ash was moisturized and then disposed of in industrial waste disposal sites within the city.

The quantity of sludge cake fed into the incinerator was maintained on a constant level. The incinerator was supplied with additional heat using auxiliary fuels in order to assure a temperature of about 800°C on top of the fluidized bed. The incinerator construction proceeded in a sequence (Nos. 1 through 3), and the thermal efficiency rises in correspondance with the newness of the incinerator in question, due to the advanced heat recovery. Table 4-9 shows the respective modes of incineration, heat recovery, and exhaust gas treatment adopted for each.

Table 4-9. Modes of Incineration, Heat Recovery, and Exhaust Gas Treatment for Each Incinerator

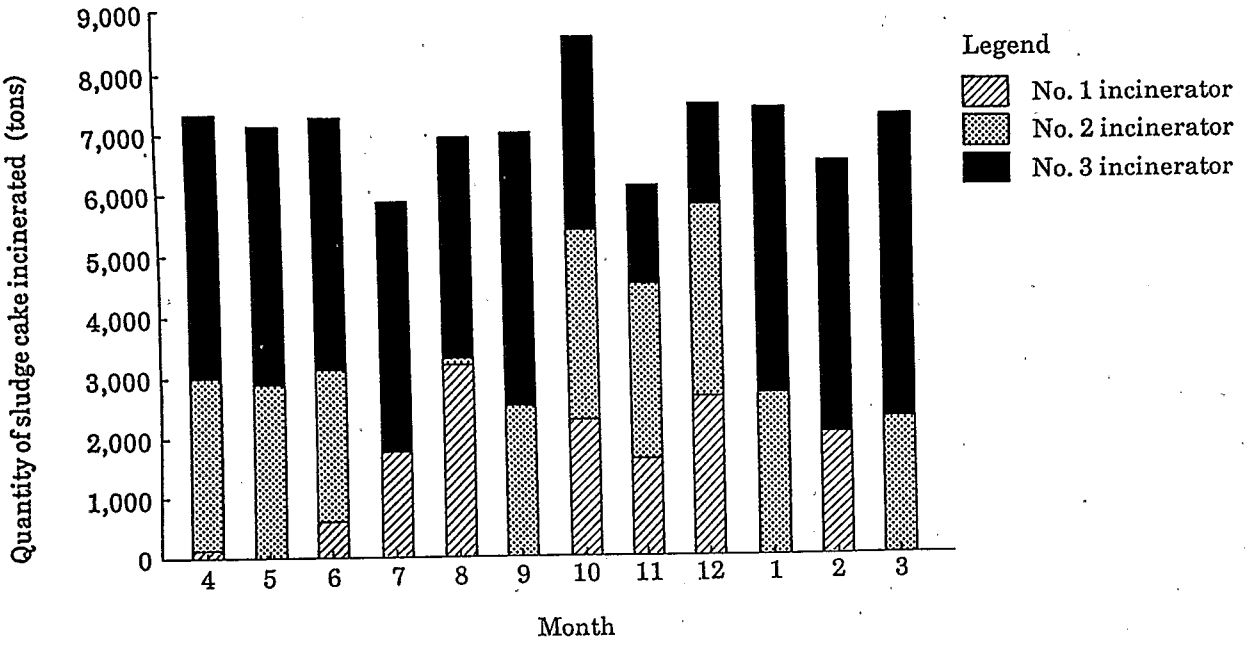
Item	No. 1 incinerator	No. 2 incinerator	No. 3 incinerator
Mode of incineration	Fluidized bed incinerator	Fluidized bed incinerator	Fluidized bed incinerator (with drier)
Auxiliary fuel	Pulverized coal, kerosene	Pulverized coal, kerosene, digestion gas	Digestion gas, city gas
Heat recovery	Heating of air for fluidization, waste heat boiler	Heating of air for fluidization, waste heat boiler	Heating of air for fluidization, heating of heat medium oil used for dryer
Mode of exhaust gas treatment			
SO_x treatment	$CaCO_3$-based dry desulfurization	$CaCO_3$-based dry desulfurization	NaOH-based wet desulfurization
NO_x treatment	Control of internal temperature and combustion air ratio	Control of internal temperature and combustion air ratio	Control of internal temperature and combustion air ratio
Smoke and dust treatment	Cyclone separator, electrostatic precipitator	Cyclone separator, electrostatic precipitator	Cyclone separator, electrostatic precipitator

The incineration operation centered around the No. 3 incinerator (capacity of 150 tons per day) offers the best thermal efficiency. Table 4-10 and Figure 4-10 show the record of operation and monthly trend of the quantity of sludge cake incinerated, respectively.

Table 4-10. Incinerator Operation Record (FY1988)

—	Number of days in operation	Annual quantity of sludge cake incinerated	Annual quantity of incinerated ash (with 30% moisturization)
No. 1 incinerator	151	14,510 t/year	3,430 t/year
No. 2 incinerator	259	24,850 t/year	5,110 t/year
No. 3 incinerator	297	43,820 t/year	6,730 t/year
Total		83,180 t/year	15,270 t/year

Fig. 4-10. Monthly Trend of Quantity of Sludge Cake Incinerated



The incinerators were operated in accordance with quarterly plans formulated on the basis of estimates of the sludge cake quantity to be generated during the quarter in question. This enabled each incinerator to operate continuously for at least one month. Repair work was scheduled to coincide with the legally stipulated boiler inspection (once annually) and conducted intensively along with preventive maintenance work. The repair periods, each lasting about one month, were May-June for the No. 1 incinerator, February for the No. 2 incinerator, and November-December for the No. 3 incinerator.

4-3. Treatment of sidestreams from the sludge treatment facilities

4-3-1. Properties of sidestreams

Figure 4-11 shows the sources of sidestreams out in the sludge treatment process , and Table 4-11, the related quantities and quality.

Fig. 4-11 Sources of Sidestreams

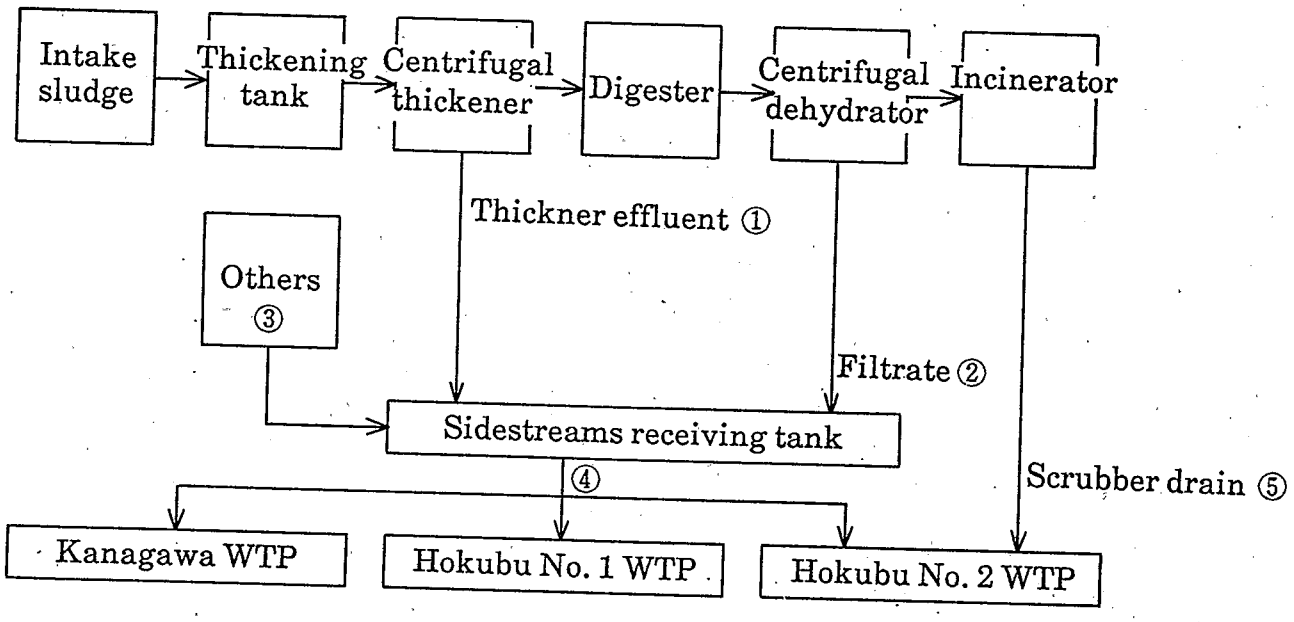


Table 4-11. Quantities and Quality of Sidestreams by Source (actual, FY1988)

	Thickener effluent ①	Centrate ②	Others ③	Total ④	Scrubber drain ⑤
Water quantity (m^3/day)	3449	1607	676	5730	1934
pH	6.0	8.0	--	7.3	8.7
TS (mg/l)	3900	1100	1500	2700	1200
VTS (%/TS)	71	36	4.9	66	9.6
SS (mg/l)	2000	72	1300	1400	4.1
COD-Mn (mg/l)	1200	150	590	820	29
BOD (mg/l)	2600	68	2400	1700	13
Nitrogen (mg/l)	360	980	150	500	68
Phosphorus (mg/l)	76	45	15	75	15

Note: The quantity for number three was calculated as ④ - (① + ②).

The sidestream sources are centrifugal thickeners, centrifugal dehydrators, scrubber drains and others. During fiscal year 1988, the first three sources accounted for 91 percent of the entire quantity of sidestreams. The single greatest source was thickener effluent, which accounted for about 45 percent of the total. It can also be noted that the quantity of the centrate is relatively high compared to the conventional-type sludge treatment system. This is due to the fact that the Center does not remove supernatant during digestion, meaning that the entire quantity undergoes the dewatering process.

As shown in Table 4-11, the concentrations of the thickener effluent are extremely high in virtually every category of water quality. Regarding nitrogen, however, the centrate concentration (980 mg/l) is higher than that of the thickener effluent (360 mg/l). This is because the nitrogen compound is decomposed during digestion and ammonia is present in the centrate.

Table 4-12 presents the calculation results (based on the above data) of the sidestreams loading for each category. In virtually all categories, the percentage of the total load occupied by thickener effluent is extremely high, from 75 to 90 percent.

As these percentages suggest, this is the source of the bulk of the sidestreams load. An exception is in relation to nitrogen, where the centrate accounts for just over 50 percent of the entire load. Since the thickener effluent accounts for another 40 percent, these two taken together account for over 90 percent of the total load in this category. By contrast, scrubber drains account for only a few percentage points of the entire load in every category, with the exception of TS.

Table 4-12. Composition of Sidestreams' Loads

Classification / Sidestreams	Thickened effluent		Centrate		Other processes		Scrubber drains		Total	
	(t/day)	(%)	(t/day)	(%)	(t/day)	(%)	(t/day)	(%)	(t/day)	(%)
TS	13.5	75	1.8	10	0.2	2	2.4	13	17.9	100
VTS	9.6	91	0.63	6	0.01	1	0.23	2	10.5	100
SS	6.9	87	0.12	2	0.9	11	0.01	0	7.9	100
COD-Mn	4.1	85	0.24	5	0.4	9	0.06	1	4.8	100
BOD	9.0	84	0.11	1	1.6	15	0.03	0	10.7	100
Nitrogen	1.2	40	1.7	53	0.1	3	0.13	4	3.0	100
Phosphorus	0.26	76	0.07	21	0.01	3	0	0	0.34	100

For the time being, the Center has been sending the sidestreams (thickener effluent, filtrate, and miscellaneous drainage) to three wastewater plants (Hokubu No. 1, Hokubu No. 2, and Kanagawa), where they are mixed with wastewater

influent and treated. Since scrubber drains account for only a small load (as noted above) they are not distributed among these three plants. Instead, it is sent separately from the other fluids to the Hokubu No. 2 plant (see Figure 4-11). The distribution ratio of the sidestreams among the three plants in the aforementioned order was 3:1:6.

Although the system is only temporary, the Center is currently sending sidestreams out in the course of sludge treatment to the wastewater treatment plants. It strictly controls the quality of the separated fluid, analyzing it at least three times per week. A major focus in water quality management is proper operation of the centrifugal thickener, since thickener effluent accounts for the bulk of the sidestream load.

4-3-2. Impact of sidestreams on wastewater treatment

(1) Treatment properties of sidestreams

In order to ascertain the treatment properties of sidestreams, the Center conducted gravitational sedimentation tests and biodegradability tests on it. Concerning SS, COD, and BOD, the removal rate by gravitational sedimentation was low (below 35 percent). Since the removal rate of SS and BOD in primary treatment of ordinary influent is generally in excess of 50 percent, this suggests that the sidestreams contained a large quantity of substances (e.g., coloids) that have low settleability.

Similarly, the biodegradability of sidestreams was also found to be low. The degree of biodegradation over a 24-hour period in the case of ordinary municipal wastewater was in excess of 90 percent for BOD and 70 percent for COD. In the case of sidestreams, however, the respective figures were 61 percent for BOD and 43 percent for COD after four hours, and 81 and 43 percent after twenty-four hours. These figures also indicate the difficulty of treating sidestreams.

(2) Impact of sidestreams on wastewater treatment

Table 4-13 shows the share of the total wastewater inflow for the three wastewater treatment plants receiving sidestreams from the Center (actual shares for fiscal year 1988).

Although sidestreams account for only 1.3 percent of the total quantity of water treated at the plants, they are the source of an extremely high share of the SS loads (13 percent), BOD (18 percent), and nitrogen (24 percent). Regarding quality, sidestreams contain many ingredients that cannot easily be removed through sedimentation or activated sludge treatment.

Table 4-13. Sidestreams' Share of Total Loads

	Sidestreams ①	Total influent at the 3 WTPs ②	Share (①÷②)×100
Water quantity (m^3/day)	5,730	442,000	1.3
SS (t/day)	7.5	57	13
BOD (t/day)	11	61	18
COD (t/day)	4.3	34	13
Nitrogen (t/day)	2.9	12	24
Phosphorus (t/day)	0.43	1.6	27

Table 4-14 supplies estimates of the extent to which water quality improvement would be made at the three plants now receiving the sidestreams if they were no longer receiving this.

In terms of the average value for the three plants, sidestreams account for an estimated 42 percent of the BOD, 42 percent of the COD, and 34 percent of the nitrogen of the respective secondary effluent load. When this amount is added, the effect on BOD is even greater. Furthermore, the figures are estimates based on annual averages; the sidestreams' impact would presumably increase if the treatability declines due to the decreased water temperature during winter or the deterioration state of activated sludge.

At present, no particular problems are experienced at the plants, in part, because the pumping time and distribution ratio among the three plants are adjusted in accordance with the treatment situation at each plant.

Table 4-14. Impact of Sidestreams on Wastewater Treatment Plants

Item	WTP	Treated water quality ①	Sidestreams' shares of treated water ②	Separated fluids impact (%) ②/①×100
BOD	Hokubu No. 1	12	7.4	62
	Hokubu No. 2	12	5.0	42
	Kanagawa	19	6.7	35
	Average	16	6.7	42
COD (Mn)	Hokubu No. 1	11	5.6	51
	Hokubu No. 2	11	3.8	34
	Kanagawa	12	5.1	42
	Average	12	5.1	42
Nitrogen	Hokubu No. 1	15	6.1	41
	Hokubu No. 2	16	4.1	26
	Kanagawa	17	5.5	32
	Average	16	5.5	34

Note: (1) = actual FY1988 figures: average figures are weighted averages.

4-4. Environmental pollution control

4-4-1. Background and regulations pertaining to pollution control

Japan's fast-paced economic development during the postwar period was accompanied by an ever worsening enviromental pollution resulting from the smoke and wastewater discharged by factories and other sources. The pollution eventually began to affect local communities, stimulating campaigns aimed at winning compensation through litigation for those whose health had been affected by it.

Structuring a regulatory framework to prevent such health problems began with the enactment of the Basdic Law for Environmental Pollution Control in 1967. This law was followed in 1968 by the Air Pollution Control Law (amended most recently in 1974) and by regulations pertaining to the enforcement of the same law, and then in 1978 by pollution control bylaws enacted by Kanagawa Prefecture. For its part, in 1975, the City of Yokohama devised general plans for administrative guideance to curtail the release of SO_x, soot and dust and in 1977 of NO_x.

Several laws related to aid for pollution victims were also enacted. The Pollution-related Health Problem Compensation Law, which was enacted in October 1973, has played a major role in providing aid for those suffering with health problems stemming from pollution. Revised in September 1987, when it was retitled: Law Concerning Compensation and Other Matters Pertaining to Pollution-related Health Problems. This law provides for the cancellation of so-called first-class regions in the event of violation and allows for projects to be carried out pertaining to the prevention of health hazards, funded by endowments created for that purpose. In accordance with this law, the Center pays pollution load charges as well, commensurate with its total quantity of SO_x discharged.

In all, discharge of soot and dust, SO_x , and NO_x are regulated by the Air Pollution Control Law and pollution control bylaws. The Air Pollution Control Law contains specific regulations concerning the permissible concentrations of SO_x and NO_x. The total quantity of such pollutants is regulated by this law and by pollution prevention bylaws and regulations.

Industrial waste is regulated by the Waste Disposal and Public Cleansing Law. The main sources of industrial waste generated by the Center are incinerated ash, ash removed from the cyclone separators and electrostatic precipitators.

The Center's modes of treatment for the substances covered by antipollution regulations are as follows:

Soot and dust:	Removal by cyclone separators and electrostatic precipitators

SO_x: Desulfurization in exhaust gas treatment units (removal of SO_x by showering with a caustic soda solvent)

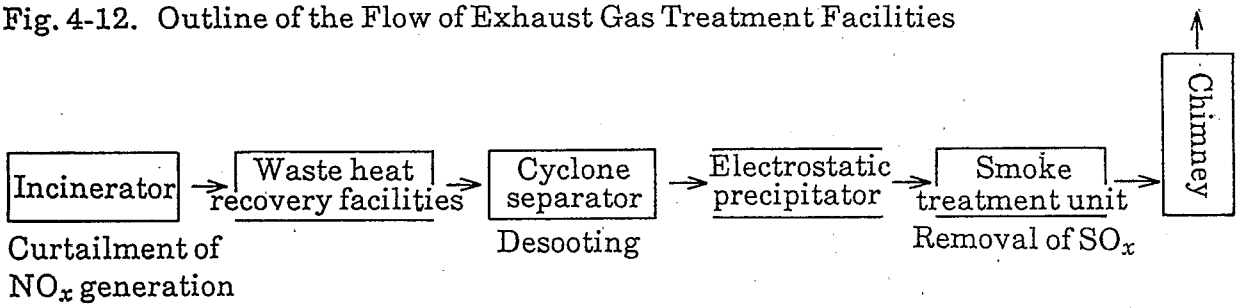
NO_x: Curtailment of generation through low-oxygen combustion

Incinerated ash: Disposal in inland landfill sites after boosting moisture content to 30 percent, to prevent scattering

Subrubber drain: Treatment with influent wastewater at wastewater treatment plants

Figure 4-12 shows the flow sheet of Exhaust Gas Treatment Facilities.

Fig. 4-12. Outline of the Flow of Exhaust Gas Treatment Facilities



4-4-2. Discharge of SO_x, NO_x, and soot and dust

Table 4-15 shows the pollutant concentrations and exhaust gas quantities generated by each incinerator during fiscal year 1988.

Table 4-15. Concentration of Pollutants Within Incinerator Exhaust Gas

Place of measurement	SO_x concentration (ppm)	NO_x concentration (ppm)	Soot concentration (g/Nm^3)	Exhaust gas quantity (1000 Nm^3/hr)
EP outlet of No. 1 incinerator	19~39	16~35	0.018~0.039	15~17.7
EP outlet No.2 incinerator	<5~41	34~56	0.018~0.039	9.2~15.7
Treatment unit outlet of No. 3 incinerator	<15~19	<10~15	0.001~0.002	10.5~12.5

Note: The O_2 conversion factor for the SO_x and NO_x concentration was 12 percent.

As noted in the previous section, SO_x and NO_x are regulated in terms of the entire quantity, and soot and dust are regulated in terms of the discharge concentration. The regulation standards and the maximum discharge levels recorded during the year in question are shown in Table 4-16.

Table 4-16. Pollutant Regulation Standards and Center Discharge Levels

	Maximum quantity of SO_x discharge (Nm^3/hr)	Maximum quantity of NO_x discharge (Nm^3/hr)	Soot and dust concentration (g/Nm^3)
No. 1 incinerator	0.69	0.62	0.039
No. 2 incinerator	0.64	0.88	0.039
No. 3 incinerator	0.24	0.19	0.002
Total	1.57	1.69	—
Regulation standard	7.44	8.33	*0.15

Note: Soot and dust discharge is regulated with reference to concentration.

As shown in Table 4-16, the Center's discharge of air pollutants fell far below the standard values (for both permissible discharge volume and concentration).

4-4-3. Odor countermeasures

Regulatory standards concerning offensive odors as contained in the national Offensive Odor Control Law and in the City of Yokohama's regulations stipulate permissible concentrations (ppm) for eight types of malodorous substances. There are also odor concentration standards contained in the city's provisional administrative guidelines concerning malodor countermeasures. The "odor concentration" is based on the odor perceptible by the human olfacotry sense. It is expressed as a multiplication rate relative to dilution, meaning the dilution multiple needed to reach threashold value.

Table 4-17 shows the regulatory standards for the eight types of malodorous substances on the boundary line of the Center, which is located in an exclusive industrial zone. Since the prospects for lowering the odor concentration through dilution and diffusion are uncertain, the actual deodorization unit design aims for a reading at the odor discharge outlet that is lower than the standards at the premise boundary, for both material and odor concentration.

The main malodorous substances of concern at the Center are four sludge related-substances: hydrogen sulfide, methyl mercaptan, methyl sulfide, and methyl disulfide. Their respective concentrations vary depending on the generation site. This is shown by the readings for one of the most typical malodorous substances--hydrogen sulfide. The readings (material and odor concentration, respectively) were 3 ppm and 10,000 - 40,000 at the sludge intake and sludge thickening tank, and 40 ppm and about 100,000 at the centrifugal thickener.

Table 4-17. Malodorous Substance Standards and Actual Data

Malodorous substances	Regulatory standards (at premise boundary)	Base odor level*	Odor level at outlet of deodorization facilities*
Hydrogen sulfide (ppm)	0.02	2.06	0.0003
Methyl mercaptan (ppm)	0.002	2.554	0.0049
Methyl sulfide (ppm)	0.01	0.145	0.0341
Methyl disulfide (ppm)	0.009	0.068	0.025
Ammonia (ppm)	1	0.4	0
Trimethylamine (ppm)	0.005	—	—
Acetaldehyde (ppm)	0.05	—	—
Styrene (ppm)	0.4	—	—
Odor concentration	32	23700	208

Note: Figures with an asterisk indicate an average of four readings taken at sludge thickening tanks and other deodorization facilities during fiscal years 1986 and 1987.

Efforts are made to completely seal the sources generating offensive odors, and steps are taken to reduce the deodorization airflow. In addition, leakage of odors from such facilities is prevented by absorbtion of odors and by maintaining a negative pressure within them. The deodorization airflow varies depending on the location; facilities are ventilated anywhere from once to ten times per hour.

The centralized treatment of sludge has enabled resolution of the problem of odor transmission to communities in the vicinity of wastewater treatment plants. It has also enabled centralized treatment of the generated odors. The Center's main mode of deodorization is soil deodorization, which also provides an effetive use of natural tracts. It is incorporating countermeasures for odors with a high concentration (50,000 or more) in its plans for pretreatment facilities.

5. Effects of centralized treatment

5-1. Effective use of energy

There are various methods of sludge treatment, each with their own advantages and disadvantages. The type of process adopted will depend on the disposal method and the purpose. This section presents an account of the effects of the sludge treatment process adopted at the Center, based on actual operations for fiscal year 1988, focusing on the use of energy, the energy balance, and on comparisons with other plants.

5-1-1. Energy use

One of the major objectives of the process adopted by the Center involved both the recovery and effective use of the latent energy harbored by sludge. The unit processes which make up centralized treatment are organically integrated during implementation. The overall flow of energy use and the proportions occupied by each process are shown in Figure 5-1 and Table 5-1, respectively. The processes in question are all those intermediate to and included in the sludge intake and incineration processes.

As shown in Table 5-1, the use of energy by process varies considerably depending on whether the focus is electrical power or fuel. As such, figures for electrical power were converted into calories to indicate the total quantity of energy use. It can be seen that sludge transportation, centrifugal thickening, and dewatering each account for less than 2 percent of this total. By contrast, the respective percentages occupied by digestion and incineration are roughly 5- and 30-fold that level. Incineration fuel accounts for about 50 percent of the total quantity of energy use.

Table 5-1 also takes into account the energy recovered as digestion gas in the centrifugal thickening, digestion, and gas engine generation processes, which form the nucleus of the entire treatment process. These three processes can basically be regarded as yielding a positive net energy. Collectively, these three processes generated about 59,000 Mcal per day of surplus energy, which can be effectively utilized for other uses.

Fig. 5-1. Overall Flow of Energy Use

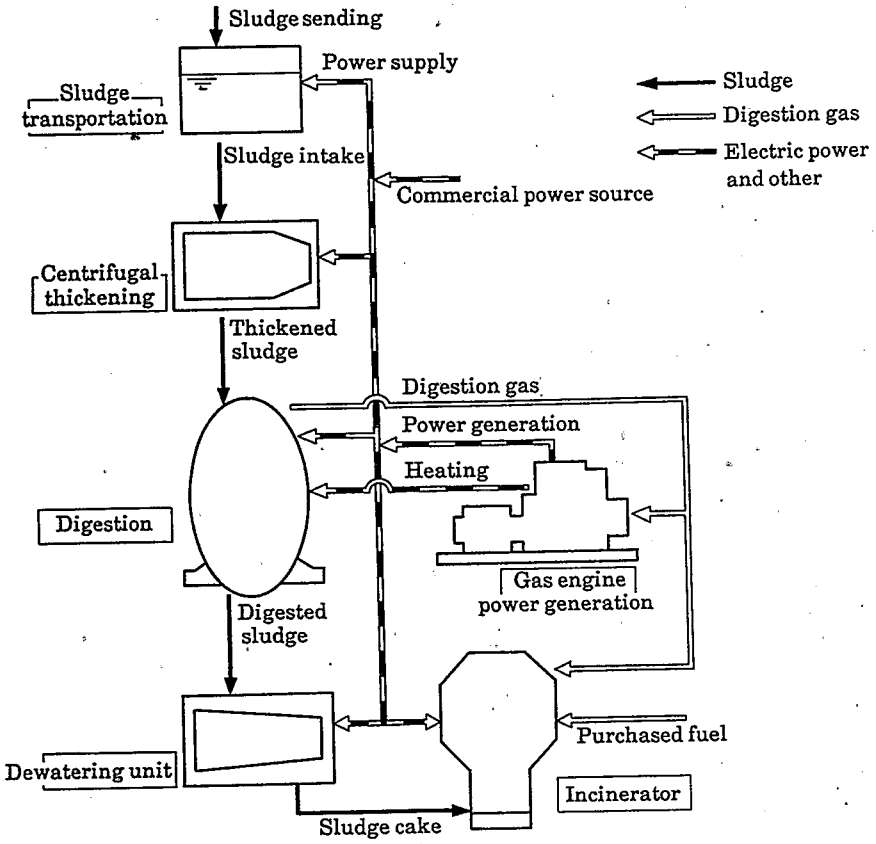


Table 5-1. Allotment of Energy Use by Process

	Electrical power (kWh/d)		Fuel (Mcal/d)		Quantity of energy use (Mcal/d)	
	①Quantity used	②Quantity used −quantity generated[1)]	③Quantity used	④Quantity used −quantity generated[1)]	①+③	②+④
Sludge transportation	6,503 (9.8%)	6,503	0 (0%)	0	5,593 (1.4%)	5,593
Centrifugal thickening	8,432 (12.7%)	8,432	0 (0%)	0	7,252 (1.8%)	7,252
Digestion	7,063 (10.6%)	7,063	28,079 (7.9%)	−130,123	34,153 (8.3%)	−124,049
Gas engine power generation	6,352 (9.6%)	−36,632	117,282 (33.2%)	89,203	122,745 (29.9%)	57,699
Dewatering	7,855 (11.8%)	7,855	0 (0%)	0	6,755 (1.6%)	6,755
Incineration	30,207 (45.5%)	30,207	207,992 (58.9%)	207,992	233,970 (57.0%)	233,970
Total	66,412 (100%)	23,428	353,353 (100%)	167,072	410,468 (100%)	187,220

Notes: 1. Figures for quantity used minus the quantity generated indicate the difference after the deduction of electrical and thermal (fuel) energy quantities generated by each process from the amount of electrical and thermal (fuel) energy (or equivalent) which each used.
2. Figures in parentheses indicate percentages of total.

5-1-2. Unit process energy balance and unit energy consumption

In order to ascertain the energy use in each process, the unit energy consumption was derived from annaul average figures as shown in Table 5-1.

By raising the concentration of the thickened sludge, it is possible to reduce the input quantity without reducing the load. This enables saving of energy used to heat the digesters, and the amount conserved can be used for other purposes. These considerations motivated the installation of centrifugal thickeners that would both boost concentration and could be artificially controlled. In addition, the combination of the egg-shaped digestors and mechanical agitators serves to assure complete mixing within all parts of the tank as well as to reduce both the required surface area and quantity of heat radiation.

Table 5-2. Units of Treatment Energy Use in Each Process

	Treated sludge quantity DSt/d	Quantity of electrical power use kWh/d	Quantity of fuel use Mcal/d	Consumption units of electrical power kWh/DSt	Consumption units of thermal power Mcal/DSt
Sludge transportation	92.0	6,503	—	70.7	—
Centrifugal thickening	92.0	8,432	—	91.7	—
Digestion	78.0	7,063	28,079 1)	90.6	360
Gas engine power generation 2)	(21,324)	(36,632)	—	(1.72)	—
Dewatering	50.0	7,855	—	157	—
Incineration (No. 3) 3)	33.3	15,967	Total: 24,591 Gas: 21,283	479	Total: 738 Gas: 639

Notes: 1. Figures for fuel in the digestion row indicate the quantity of heat used in warming.
2. Figures in the gas engine power generation row indicate the quantity of digestion gas supplied (21,324 Nm^3/d) and the effective generated power (36,632 kWh/d).
3. Figures for total and gas in the incineration (No. 3) row indicate the total quantity of heat supplied by incineration fuels and the subtotal of that supplied by digestion gas, respectively.

In using the surplus energy digestion gas obtained in this fashion, priority is accorded to its use as an auxiliary heat source for the incinerators; the remainder is used in the gas engine power generators and recovered as electrical power. Although it may appear complicated when compared to the standard procedure, this process of first extracting energy in useful form and then returning only the amount required for combustion to the incineration process, enables a more effective use of energy than autogeneous combustion incineration of the entire quantity of thickened sludge without digestion.

From the perspective of the digesters, the gas engine generation facilities are a source of heat for warming; from the standpoint of the gas engine generation facilities, the digesters are cooling appratus. At the very least, for the system to be

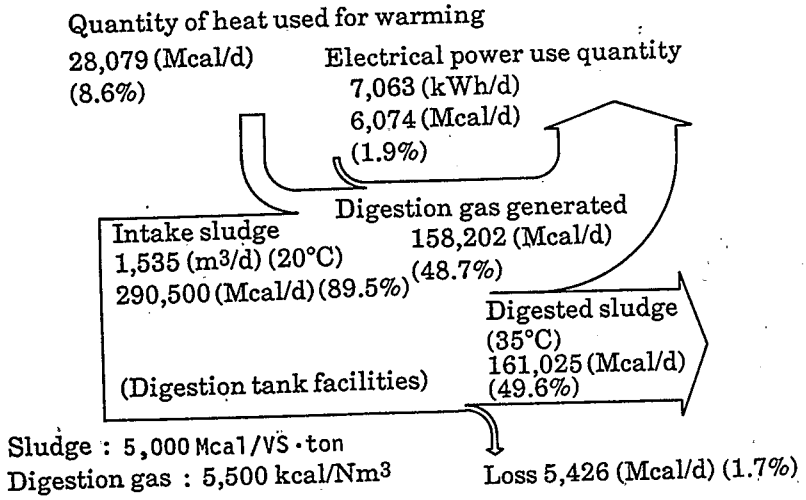
regarded as effective, the quantity of power derived from these generation facilities must exceed the power requirements of the centrifugal thickening, digestion, and power generation facilities. Therefore, these three processes must always function as a unit.

The energy balances of the digesters, gas engine generation facilities, and incineration facilities are described below.

(1) Energy balance of digesters

Figure 5-2 shows the energy balance of the digesters.

Fig. 5-2. Energy Balance of Digesters



As shown in Figure 5-2, approximately half of the intake sludge calories are retrieved as digestion gas. In the digestion process the energy retrieved as digestion gas is about 4.6 times as great as the input energy (electrical power use plus heat used for warming).

(2) Energy balance of gas engine power generation facilities

The energy balance of the gas engine power generation facilities is somewhat more complicated. The input energy (calculated in terms of calories) consists of digestion gas and electrical power used by the facilities; the output energy consists of the electrical power generated and exhaust heat. Since precedence is accorded to incineration utilizing digestion gas, the quantity of electrical power generated also varies depending on the incinerator operation. Table 5-3 shows the energy use in power generation facilities.

Table 5-3. Energy Use in Power Generation Facilities

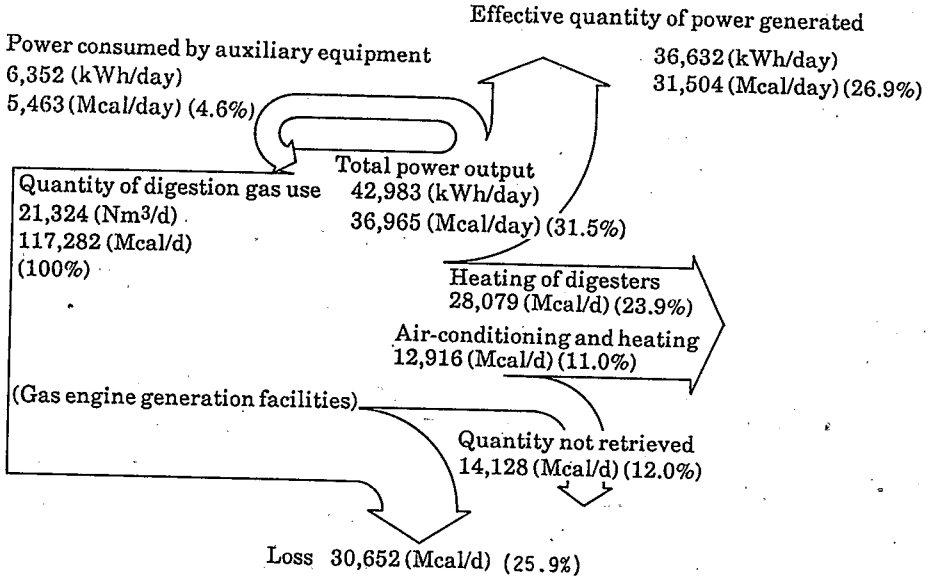
Year/ month	Quantity of digestion gas generated Nm³/d	Digestion gas use quantity			Gas engine generation facilities			Quantity of heat used for warming digestors Mcal/d
		Incinerator Nm³/d	Gas engine generation facilities		① Quantity of power generation kWh/d	② Quantity of power used kWh/d	①－② 1) kWh/d	
			Nm³/d	Mcal/d				
'88/4	32,565	9,919	22,647	124,557	45,694	6,343	39,351	31,187
'88/5	31,810	11,090	20,720	113,960	42,059	6,093	35,966	22,794
'88/6	28,013	7,675	20,338	111,861	41,124	6,287	34,837	17,765
'88/7	25,262	3,464	21,798	119,889	44,175	6,583	37,592	16,488
'88/8	23,352	4,021	19,331	106,320	37,809	6,871	30,938	12,035
'88/9	23,313	9,265	14,049	77,268	28,000	5,070	2,930	17,264
'88/10	24,923	7,426	17,497	96,234	34,619	5,835	28,784	19,591
'88/11	29,440	7,480	21,960	120,780	43,920	6,410	37,510	30,210
'88/12	31,140	7,720	23,420	128,810	47,630	6,580	41,050	36,450
'89/1	31,510	10,060	21,450	117,975	43,610	5,980	37,630	42,885
'89/2	32,210	3,180	29,030	159,665	59,370	7,560	51,810	39,696
'89/3	31,530	7,980	23,650	130,075	47,790	6,610	41,180	50,583
Average	28,764 (100%)	7,440 (24.9%)	21,324 (74.1%)	117,283	42,984	6,352	36,632	28,079

Note: Under gas engine generation facilities ① – ② indicates the balance after power consumed by the generation facilities is deducted from the power generated. This difference is the effective quantity of power generated. The unit of effective power generation amounts to 1.72 kWh/Nm³.

A major factor in the fluctuation of the digestion gas quantity generated reflects the change in the quantity of organic substances contained in the thickened sludge. For the year in question, this latter quantity ranged from a maximum of about 70 tons (in February) to a minimum of about 50 tons (in September), and averaged about 58 tons. There was a close correlation between this quantity and that of the digestion gas.

Figure 5-3 shows the balance of energy in the gas engine power generation facilities.

Fig. 5-3. Energy Balance of Gas Engine Power Generation Facilities



The generative efficiency obtained from the effective quantity of power generated amounted to 26.9 percent, but the efficiency of the generator itself was 31.5 percent. The digester's heating needs were met using about one-fourth of the digestion gas supplied to the generation facilities. A certain quantity of energy (expressed in terms of calories in the figure) was not retrieved, only because an effective use for this energy was lacking. However, this quantity can be effectively utilized.

(3) Energy balance of incineration facilities

As noted in previous sections, the Center operates three incinerators. The first two (No. 1 and No. 2) began operation before the digesters did. The No. 2 incinerator can use both digestion gas and other types of auxiliary fuel, but the No. 1 incinerator, which is the oldest, cannot use digestion gas. By contrast, the No. 3 incinerator, which is the newest, was basically designed to use digestion gas.

Table 5-4 shows the specific differences among the incinerators regarding energy use.

Table 5-4. Energy Balance of Incinerators

Incinerator (No.)	Sludge incinerated quantity ① (DSt/d)	Incinerated ash quantity (DSt/d)	Quantity of energy use					②/① (Mcal/DSt)
			Electrical power (kWh/d)	Auxiliary fuel (Mcal/d)			② Total (Mcal/d)	
				Digestion gas	Other auxiliary fuels	Subtotal		
No. 1	22.7	17.5	6,191	0	62,480	62,480	67,804	2,987
No. 2	21.3	15.2	8,048	28,986	49,496	78,482	85,403	4,010
No. 3	33.3	17.4	15,967	21,283	3,308	24,591	38,322	1,151

As shown in Table 5-4, the No. 3 incinerator has the lowest energy consumption units per DS tons, and digestion gas can be used to meet about 87 percent of its auxiliary fuel needs. In view of this advantage, which suggest that incinerators of the No. 3 type will constitute the main mode of incineration in future centralized sludge treatment, it was decided to base the consumption units for incineration on the values for the No. 3 incinerator for the following discussion.

5-1-3. Energy balance comparison with other processes

The aforementioned findings provide the basis for Figure 5-4, which shows the energy balance per unit of solids received for treatment at the Center. For the purposes of comparison, calculations were also made of the energy balance in the conventional digestion/incineration process in which a boiler is used to warm the tank and effect separation of supernatant and calculations were also made in raw sludge dewatering/incineration processes. In these calculation, the values applied for the various units are based on the actual figures obtained in the Center's operation, with the exception of the units for digesters in the conventional digestion/incineration process, which are nationwide averages for fiscal year 1986.

Table 5-5 shows the energy balance comparison between the three above-mentioned processes.

Fig. 5-4. Energy Balance per Unit of Treated Solids at the Center (No. 3 Incinerator)

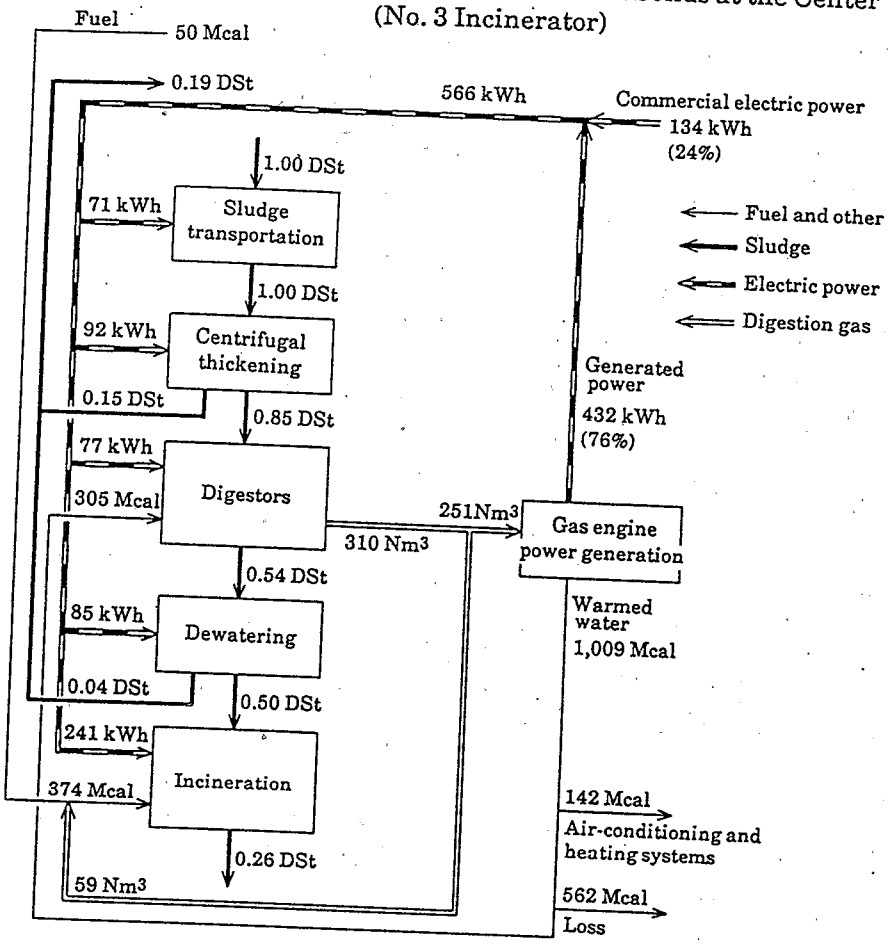


Table 5-5. Energy Balance Comparison

	Unit	① Digestion/incineration process in Yokohama (Centralized treatment)	② Conventional digestion/incineration process	③ Raw sludge dewatering/incineration process
Treated sludge quantity	DSton	1.00	1.00	1.00
Total energy use				
Electrical power	kWh	566	634	708
Fuel	Mcal	1,755	1,894	0
Total (in terms of calories)	Mcal	2,242	2,439	608
Outside energy introduced				
Electrical power	kWh	134	634	708
Fuel	Mcal	50	161	0
Total (in terms of calories)	Mcal	165	706	608

Notes: 1. Total energy use refers to the total quantity of energy used for treatment; outside energy introduced refers to that portion of the total energy use and is equivalent to the balance after the energy retrieved from the sludge in the course of treatment is deducted from the total.
2. Total (in terms of calories) refers to the sum of the electrical and fuel energy converted into calories.

The figures for total energy use reveal that processes (1) and (2) require about the same amount of energy for treatment (expressed in terms of calories), but that process (3) requires only about one-fourth the energy that processes (1) and (2) do. In this respect, process (3) is the most advantageous. (The calculation for process (3) assumed autogeneous combustion bacause it excluded the digestion process.)

However, process (1) is the most advantageous in respect to outside energy introduced, where the quantities for processes (2) and (3) are virtually the same. The determinant behind this difference is the generation of digestion gas, which reverses the advantages of process (1) and (3) and also almost equalizes the values for process (2) with those of process (3). In process (2), the generated digestion gas must be applied directly in order to heat the tanks, which rule out the installation of power generation facilities fueled by this gas.

Process (1) is clearly superior, in view of the fact that it offers energy retrieval in the form of both electrical power through power generation using the digestion gas and thermal power for heating the digesters using the generation unit exhaust heat.

5-2. Construction costs

5-2-1. Hokubu Sludge Treatment Center construction costs

Table 5-6 shows the composite construction costs for the Hokubu Sludge Treatment Center by unit process.

Table 5-6. Hokubu Sludge Treatment Center Construction Costs

Facility	Construction costs (millions of yen)
Sludge intake	1,330
Gravity thickening	1,770
Centrifugal thickening	2,590
Digestion	10,490
Dewatering	2,440
Digestion gas power generation	3,270
Incineration	12,990
Other facilities	7,340
Total	42,220

5-2-2. Comparison of the construction cost

It is not possible to make a fair comparison of the construction costs for centralized treatment and the conventional method because of the differences in the processes themselves and the level of technological development involved. Nonetheless, the following may be presented as relevant data calculated on the basis of the next three points:

1) The evaluation is based on a comparison with the Kohoku Wastewater Treatment Plant. This comparison was used because the sludge treatment process varies from plant to plant, and the main process of the Kohoku plant closely resembles the process at the Center.

2) The process sequence that is the subject of comparison is thickening, digestion, dewatering and gas power generation; incineration is not included.

3) The cost of land acquisition was not taken into consideration.

Following is the evaluation method that was employed. First, a calculation was made of the construction costs per unit of treatment capacity in each unit process, at the Kohoku WTP and the Center. These figures were then totaled. Next, the cost of the facilities at Kohoku WTP (conventional decentralized treatment) was adjusted to reflect the rise in prices. Also, the construction cost of facilities associated with sludge transportation were added into the total for the Center (centralized treatment facilities). The two figures were then compared. The procedures and results are shown in Table 5-7.

Table 5-7. Comparison of Construction Costs of Sludge Treatment Facilities

(Units: thousands of yen per DS tons)

	Conventional treatment system	Centralized treatment system
Unit construction costs		
Thickening	6,100	39,660
Digestion	51,810	78,280
Dewatering	91,430	34,130
Other processes	7,270	17,630
Gas power generation	—	42,280
Subtotal	156,610	212,520
Deflator	1.440	—
Sludge transportation facilities	—	26,530
Total construction costs	225,520	239,050

Note: The deflator is the price increase ratio between 1975-1985.

It can be seen that the unit construction costs for centralized treatment are about 6 percent higher than those of a conventional treatment system. This difference is due to the inclusion of gas power generation and environmental protection facilities in the centralized treatment facilities. If comapred after removing the gas power generation facilities' expenses to equalize conditions with decentralized treatment, the construction costs for centralized treatment are about 13 percent lower, clearly indicating the benefits of centralized treatment.

5-3. Sludge treatment costs

5-3-1. Treatment costs at the Center

Table 5-8 shows the sludge treatment costs for the Hokubu Sludge Treatment Center during fiscal year 1988.

Table 5-8. Treatment Costs at the Hokubu Sludge Treatment Center (FY1988)

(Unit: thousands of yen)

	Transportation and intake	Digestor dewatering	Incinera-tion	Total	Breakdown of treatment costs (thousands of yen/DSt)
Personnel	34,543	132,139	132,943	299,675	10 (6)
Power	24,377	57,830	46,626	128,833	5 (3)
Chemical agents	18	232,954	39,181	272,153	10 (8)
Fuel	0	0	118,808	118,808	4 (0)
Repair	9,996	108,372	77,085	195,453	7 (4)
Other costs	19,077	55,234	144,728	219,039	8 (3)
Total treatment costs	88,011	586,579	559,371	1,223,961	44 (24)
Quantity of treated solids (DSt)	28,580	28,580	28,580	(28,580)	
Treatment costs (thousands of yen/DSt)	3	21	20	44	

Note: Other costs include comissions and utilities (light, heat, water). Incineration costs do not include the incineration cost of cake transported from other treatment plants. Figures in parentheses indicate the costs which do not include incineration cost.

This table presents the breakdown of treatment cost per unit of solids in sludge transportation and intake, digestion/dewatering, and incineration, which are the processes involved in the comparison with the conventional treatment process sequence in the next section. These costs per DS·tons were 3,000 yen for sludge transportation and intake, 21,000 yen for digestion and dewatering, and 20,000 yen for incineration, totaling 44,000 yen.

5-3-2. Treatment costs in conventional processes

A calculation was made for the fiscal year 1988 treatment costs in the digestion/dewatering process sequence in the Chubu, Nambu, and Kohoku wastewater treatment plants, in order to estimate the treatment costs in the standard process prior to the introduction of centralized treatment. In the Kohoku plant, this cost was obtained by adjusting the treatment costs in fiscal year 1983 by the price increase ratio over the intervening period The findings are presented in Table 5-9.

Table 5-9. Treatment Costs in Conventional Processes (FY1988)

(Unit: thousands of yen)

	Chubu (WTP)	Nambu (WTP)	Kohoku (WTP)	Total	Breakdown of treatment costs (thousands of yen/DSt)
Personnel	109,621	123,664	54,991	288,276	25
Power	7,208	49,730	17,641	74,579	7
Chemical agents	11,954	85,511	14,988	112,071	10
Fuel	647	235	35	917	—
Repair	27,658	60,093	43,466	131,217	11
Other costs	17,129	79,398	74,829	171,356	15
Total treatment costs	173,857	398,631	205,928	778,416	68
Quantity of treated solids (DSt)	1,672	7,203	2,573	11,448	
Treatment costs (thousands of yen/DSt))	104	55	80	68	

Note: Other costs include commissions and utilities (light, heat, water). Incineration costs are excluded.

As shown in Table 5-9, the total quantity of solids treated at these three plants was 11, 448 tons. Treatment of this quantity cost ¥ 778 million, or ¥ 68,000 per ton of solids.

5-3-3. Comparison of sludge treatment cost

Table 5-10 compares the treatment costs of the Center with the estimated costs in the conventional sludge treatment system.

Table 5-10. Comparison of Sludge Treatment Costs

(Unit: thousands of yen, DS tons)

	Conventional system	Centralized system
Treatment costs per ton of treated solids	68	24
Treatment costs (up to and including dewatering)	778,416	674,590
Quantity of treated solids (DS tons)	11,448	28,580
Incineration costs per ton of treated solids	23	20
Total treatment costs per ton of treated solids	91	44

Note: Figures for incineration costs in the conventional process sequence are based on the assumption that a power generation process is not utilized.

As shown in Table 5-10, the costs of treatment, up to and including dewatering in the conventional system per ton of treated solids, are about three times greater than the corresponding costs of about 24,000 yen at the Center.

If the incineration process is also included, the costs per ton of treated solids in the conventional system would increase to an estimated 91,000 yen. This figure is almost double that of the corresponding center costs of about 44,000 yen. As such, the benefits of centralized treatment also appear in the comparison of total treatment costs, i.e., including the incineration costs.

These findings indicate that a rise in the expanse of facilities, while involving a rise in the total maintenance and operational costs, including power, fuel, and other expenses, also induces a greater economy of scale including considerable labor saving, meaning a decline in the treatment costs per ton of treated solids. There is also the benefit derived from a more effective use of energy due to the use of digestion gas for power generation.

The benefits of centralized treatment also appear in regards to transportation as well. The cartage costs of transporting sludge cake by truck are about ¥ 2,600 per ton. Transportation of the entire sludge cake quantity generated at the Center (about 77,000 tons) by truck would therefore cost about ¥ 200 million. Transportation of the sludge equivalent by pipeline costs only about ¥ 88 million.

5-4. Improvement of water quality in wastewater treatment plants

5-4-1. Improvement of water quality through sludge transportation

The shift from treating their own sludge to pumping it to the Center through pressurized pipelines has improved the quality of treated effluent in the wastewater treatment plants. This is due to the fact that the shutdown of the sludge treatment facilities has eliminated the load brought about in sidestreams. However, it is difficult to quantify this improvement, for the following reasons.

1) The wastewater treatment situation exhibits a broad fluctuation depending on such factors as climate, the amount of rainfall, and operating conditions during the year in question. Improved quantification would require comparing yearly data for at least several years before and after the shift to pipeline transportation.

2) The influent load has been fluctuating every year due to increases in the quantity of influent wastewater and the expansion of treatment facilities. This has made long-term comparisons difficult.

3) The Kanagawa and Hokubu No. 1 plants both transport sludge and receive sidestream water at the same time, making it difficult to isolate the influences of sludge transportation.

For these reasons, it was decided to compare the treated effluent quality for a short period of time before and after sludge transportation at the Midori and Kohoku plants, both of which are not subject to the influence of sidestreams loads. The period of time in question is about one year, during which the treatment facilities were not expanded. The results are shown in Table 5-11, just for reference.

Table 5-11. Treated Effluent Quality Before and After Sludge Transportation

WTP / Item	Midori		Kohoku		
	Before sludge transportation	After sludge transportation	Before sludge transportation	After sludge transportation	
Term of data collection	Oct. 1983 - Apr. 1984	Oct. 1984 - Apr. 1985	July 1984 - June 1985	July 1986 - June 1987	July 1987 - June 1988
Sludge transportation rate (%)	0	100	0	21	66
Number of facility trains	2	2	3	5	5
Influent wastewater quantity (m^3/d)	30000	41000	65000	62000	84000
BOD load (kg/m^3day)	0.33	0.32	0.42	0.21	0.24
Transparency (cm)	56	89	64	83	94
SS (mg/l)	6	3	6	5	4
COD-Mn (mg/l)	14	12	11	8.9	8.6

Notes: 1. The Midori plant initiated sludge transportation in October 1984 (treatment facilities were expanded in May 1985).
2. The Kohoku plant initiated sludge transportation in September 1985 (treatment facilities were expanded in May 1985).

The effect of transportation appears most clearly in the figures for the Midori plant, which show major improvement in transparency, SS, and COD after the shift to transportation, despite an increased influent. The effect cannot be clearly grasped in the case of the Kohoku plant, due to the fact that the transportation rate (the share of the total sludge generated occupied by that transported to the Center) was increased in increments from an initial 21 percent to 100 percent over a three-year period, during which the facilities were expanded. Nevertheless, a comparison of water quality at the transportation rates of 21 and 66 percent, when the BOD load was virtually the same, reveal a somewhat improved tendency in transparency and SS.

5-4-2. Other improvements

Various factors, such as an abnormally high influent of stormwater and industrial drainage as well as bulking, can impair treatment in wastewater treatment plants. Unlike the above, however, the deterioration of water quality

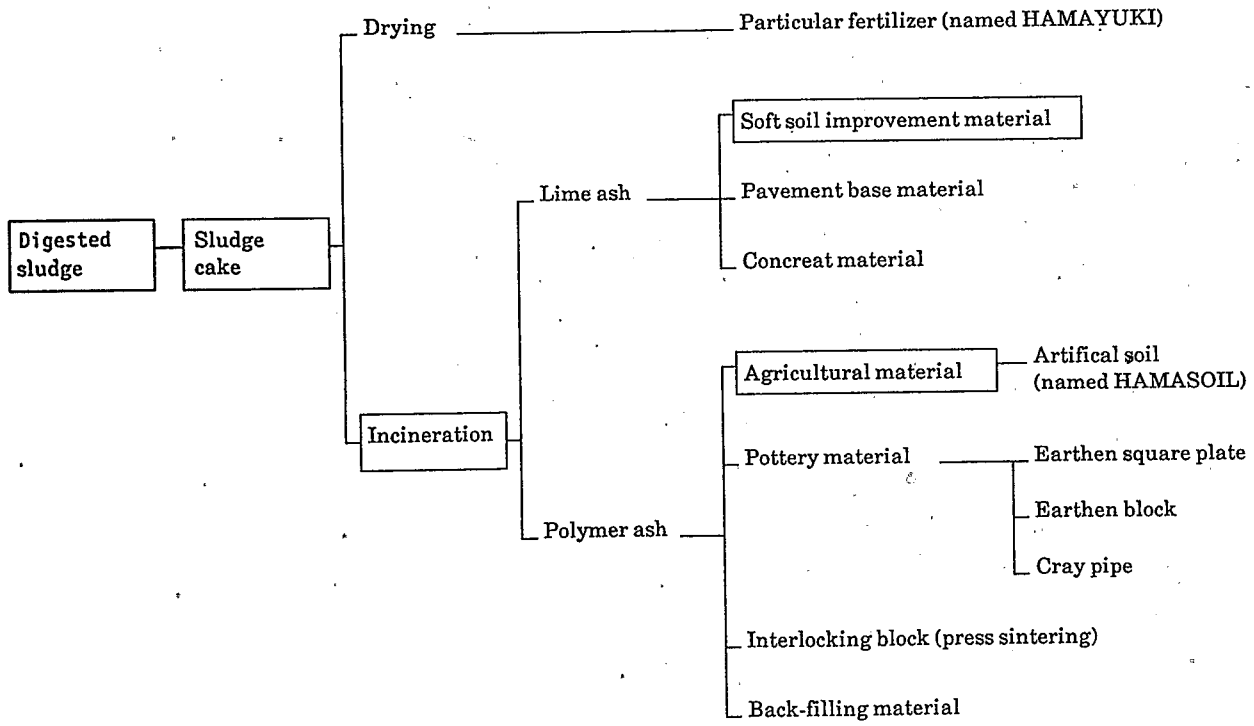
associated with sludge treatment trouble is generally not transitory, but exerts a direct impact on the activated sludge over a long period of time. For this reason, its impact on wastewater treatment is fairly serious.

The shift to centralized treatment has resolved many such problems, owing to the fact that the sludge is promptly sent off the plant premises for treatment. In this area as well, centralized sludge treatment is exerting a favorable influence on wastewater treatment.

5-5. Effective use of sludge.

In order to prolong the service life of Yokohama's existing disposal sites, virtually the entire quantity of sludge generated in the city is incinerated at the Hokubu and Nambu sludge treatment center. Nevertheless, the city has recognized sewage sludge as a useful resource, and is pursuing various applications of its effective use, as shown in Figure 5-5.

Fig. 5-5. Utilization of Sludge



The following is an outline of the two main types of usage that are the subject of the city's promotional efforts: soil improvement and nursery farming.

5-5-1. Soil improvement

Incinerated ash can be divided into two categories, lime ash and polymer ash. A polymer dewatering agent has been adopted for use in both of the city's centralized sludge treatment centers (Hokubu and Nambu). But the ash from the Hokubu center has a higher CaO content (approx. 25%). The reason for this is that the Hokubu center adopts a dry-type desulfurization method in which $CaCO_3$ is added to prevent SO_x generation during the incineration of the sludge cake.

Due to its high water absorption capacity and the chemical action of CaO, incinerated ash with a CaO content in excess of 20 percent can be mixed with soil to improve its suitability for uses relevant to civil engineering. It has been confirmed that lime-type incinerated ash has sufficient strength for use as material for backfill and roadbeds (especially the lower layer). In actual use, however, the powdered form of this ash causes handling problems.

The disposal of surplus soil resulting from public works projects is growing into a major problem in Japan as the number of disposal sites for it declines. Consequently, it was decided to promote the use of lime-type ash as soil improvement material in order to resolve the problem of the disposal of both this ash and surplus soil. Surplus soil generated during sewerage system construction is disposed of at considerable costs when it is of a poor quality. However, improving this soil, mixing it with incinerated ash for reuse as backfill material, would in effect resolve three problems: it would reduce the quantity of waste surplus soil and incinerated ash, and reduce the expenditure for their disposal and backfill use on sand.

Various tests confirmed the practicality of incinerated ash and raised hopes for the use of lime-type ash as soil improvement material. In view of these prospects, the city constructed a soil improvement plant at the Hokubu No. 2 Plant, adjacent to the Center, for effectively utilizing incinerated ash. This plant, the first of its kind in all of Japan, went into operation in April 1989.

Operation of this plant at a rated output would yield a savings of about 300 million yen annually in expenditures for disposal of incinerated ash and surplus soil and for the purchase of sand.

The plant may be outlined as follows:

1) Capacity: 30 m^3/hour (continuous operation)

2) Rate of incinerated ash admixture: no more than 30 percent relative to the dried weight of the base soil; quicklime and cement are used as additional agents if necessary

5) Quality standard for improved soil: an indoor CBR of at least 10 percent

6) Quantity of incinerated ash used: 22,000 m^3 per year

7) Quantity of improved soil produced: 49,000 m^3 per year

8) Construction cost: 1.08 billion yen

5-5-2. Use as nursery farming

Using polymer ash with a low lime content, the city developed the agricultural materials "hama soil." Production of hama soil begins with the addition and mixture of the incinerated ash and polyvinyl alcohol as a binder. After pelletization and chemical treatment with an ammonium sulfate solution, the mixture is then washed and dried. The following can be cited as the major characteristics of Hama Soil.

1) Excellent water retention, water absorption, and aeration qualities for plant cultivation.

2) A pH of 6-7, ideal for plant cultivation

3) Durability; retains structure

4) Change of color reflecting degree of dryness/wetness, which serves as a guideline for watering

5) Lack of odor; sterile

A pilot plant is now producing about 300 kilograms of this soil per day. Efforts are being made to reduce the production costs while promoting sample shipments to the market.

5-6. Evaluation

The following effects were anticipated in planning the centralized sludge treatment facilities.

(1) A more effective system of sludge treatment regarding facility construction, maintenance and operation.

(2) Promotion of retrieval and comprehensive use of the energy generated from sludge due to the greater quantities involved in centralized treatment.

(3) Improvement of the secondary effluent quality at the wastewater treatment plants due to the elimination of sidestream water from sludge treatment facilities.

(4) Improvement in the efficiency of sidestream treatment through centralized treatment of the same at the Center.

(5) Alleviation of the need for pollution control facilities at the individual wastewater treatment plants.

As has already been noted, to date, the operation of the centralized treatment facilities has achieved the anticipated degree of success. Effects (1) through (3) have clearly surfaced. The results regarding efficient retrieval and the comprehensive use of energy exceeded our expectations.

This performance rests on several major factors. First, the combination of the centrifugal thickeners and the egg-shaped digesters enable a high-concentration digestion that yields an increased quantity of digestion gas and enable the amount of heat for warming the tanks, per load, to be reduced.

Second, the expanded quantity of digestion gas generated is used on a priority basis as supplementary fuel for the incinerators, which need a considerable proportion of the energy consumption in the centralized treatment process. Third, any remaining digestion gas is used as a source of energy for power generation, which supplies 64% of the Center's own electrical power needs.

Finally, energy is conserved in the operation of incinerators by effectively using their exhaust heat to dry the sludge cake before incineration. In short, the successful operation of centralized treatment was made possible by the fruits of much research and extensive developments aimed at optimizing such treatment as a total system.

Presently, the Center sends back sidestream generated in the course of sludge treatment for treatment to certain wastewater treatment plants which send it sludge. While this system causes no particular problems, the Center is currently promoting detailed studies for the implementation of the original plans, which anticipate centralized treatment of sidestreams at the Center. This system will be implemented in the near future.

Nevertheless, research on the following subjects is needed to augment the effects of centralized treatment:

(1) Studies for operating applications of the centrifugal thickener during storms and in the winter; a concentration of about 5 percent was achieved by the thickener in mixed sludge.

(2) More efficient operation of the egg-shaped digesters (digestion periods, organic loading, agitation methods, gas generation quantities).

(3) Study of dewatering units and pretreatment to bring the moisture contents of digested sludge to a level permitting autogeneous combustion, and the development of a more efficient incinerator.

(4) Increase the Center's self-sufficiency rate in electrical power and fuel energy to 100 percent through promotion of number three above and the promotion of energy conservation in facilities.

(5) Utilization of low-temperature energy produced through cogeneration.

(6) Facilitation of sludge transportation, since the thickening characteristics deteriorate during transport.

6. Afterword

The record for the first year of full-scale operation of the centralized treatment facilities at the Hokubu Sludge Treatment Center reveals that electrical power generated using the digestion gas supplied 64 percent of the total electrical power consumed at the Center. This represents an annual savings of about 200 million yen in electrical power costs, despite the fact that the Center was still operating at only one-fifth of its planned capacity. It also reveals a decline in the amount of energy that had to be purchased from outside sources and labor saing stemming from the shift to centralized treatment at the Center. As a result, the Center's sludge treatment costs were less than half those of the same sludge quantity under the previous system of treatment at each wastewater treatment plant. These and other effects of centralized treatment have been described in the report.

At the same time, it must be acknowledged that it has only been a short time since the Center went into full-scale operation. For this reason, the current report must be termed an interim report as far as the operation and evaluation of centralized sludge treatment facilities are concerned. It is also clear that there is still room for improvement in several areas, including sludge transportation, effective incinerator operation, further energy conservation, and treatment of sidestream water. In addition, the Nambu Sludge Treatment Center will begin operation before the end of this year. As such, we are looking forward to preparing another report specifically on these developments and subsequent progress, at the next available opportunity.

In closing, the author would like to thank all those who assisted him in the preparation of this report, particularly Mr. Hachiro Shimomura and Mr. Tetsuro Ishii, whose help was instrumental in creating the project.

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DEVELOPMENT OF SLUDGE TREATMENT TECHNOLOGY IN BIOLOGICAL PHOSPHORUS REMOVAL PROCESS

by

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ABSTRACT

Fukuoka city has been developing an advanced wastewater treatment process, mainly based on the biological phosphorus removal, to remove large amounts of nutrients. Biological phosphorus removal process has many advantages as compared with the other phosphorus removal processes. However, it has one major disadvantage that phosphorus absorbed by the activated sludge may be released when the sludge is handled inadequately, and this makes it difficult to treat sewage efficiently. All the sewage treatment plants in Fukuoka city are provided with mixed sludge gravity thickening process and anaerobic digestion process. Phosphorus absorbed by the sludge may be released in these processes. In order for the biological phosphorus removal plants to run with some stability, it is necessary to decrease phosphorus concentration in the return water from sludge treatment process.

For this reason, Fukuoka city has been investigating the behavior of phosphorus in the sludge, and developing the sludge treatment technologies for prevention of phosphorus release.

This paper introduces several results of our investigation and development as follows.

1) Prevention of phosphorus release by adopting dissolved air floatation thickening.
2) Phosphorus fixation with lime and ferric chloride in digestion tank.
3) Phosphorus fixation with lime and ferric chloride for supernatants.
4) Phosphorus fixation by crystallization of Magnesium Ammonium Phosphate.

1. INTRODUCTION

Recently, eutrophication or aggravation of closed water with excessive quantities of in-flowing nutrients, is increasingly causing social problems. The sewage systems, which are largely dependent on the conventional secondary treatment, cannot cope adequately with eutrophication. This is because of the limited capacity of the conventional secondary treatment to remove nutrients, in particular nitrogen and phosphorus. However, the eutrophication is steadily progressing. And it is essential to abate nutrients flowing into closed water in order to preserve its quality.

Sewage, therefore, has been increasingly treated from the angle of solving eutrophication problems. Several sewage treatment plants around lakes are attempting to remove nutrients drastically, and more advanced wastewater treatment processes are being developed for cities facing closed sea water.

Fukuoka City has been recently developing an advanced wastewater treatment process, mainly based on biological phosphorus removal, to remove large amounts of nutrients from sewage. This paper describes some of important developments carried out so far for the sludge treatment process as the core of the biological phosphorus removal process.

PHOTO 1. FUKUOKA CITY AND HAKATA BAY

2. OUTLINE OF EXPERIMENTS OF ADVANCED WASTEWATER TREATMENT PROCESS

2-1. Background of Experiments

Fukuoka City is an economic and cultural center of the Kyushu Island located in the southwest part of Japan, and is the 8th largest city in Japan with a population of 1.2 million.

The city has been populated along the Hakata Bay with the city facing north. Most of the rivers running through the city drain into the bay, and treated water from the 6 sewage treatment systems operated by Fukuoka City flow into the bay either directly or via rivers. The bay is fairly closed with Umino-nakamichi peninsula and Shika Island in the east, and Itoshima peninsula in the west. Therefore, the sea water exchange with the open sea (Japan Sea) requires many days. As the population and production volume increase, the bay is increasingly becoming eutrophic, causing red tide to occur more frequently.The sea water quality of the bay is improving gradually according to the expanded sewage treatment facilities, but it is generally still behind the related environmental target levels of water quality preservation.

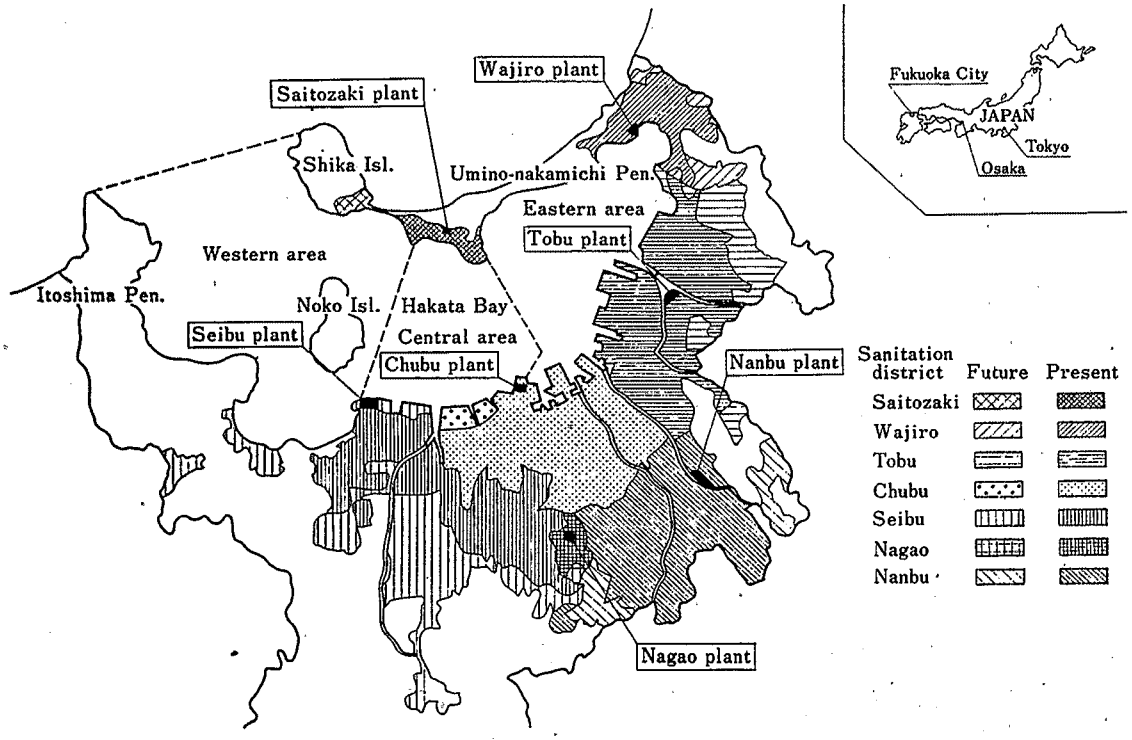


FIGURE 1. LOCATION MAP OF TREATMENT PLANTS AND SEWERAGE DISTRICTS

TABLE 1. SERVICE OF FUKUOKA CITY

(as of March 1989)

Total area of the city	33,818	ha
Total area of urbanizd districts	14,969	ha
Total population	1,200,204	
Total service area	11,346	ha
Service area in urbanized districts	75.8	%
Population in sewage service area	947,800	
Sewerage service ratio	79.0	%
Treatment capacities (total of 6 plants)	538,000	m^3/day

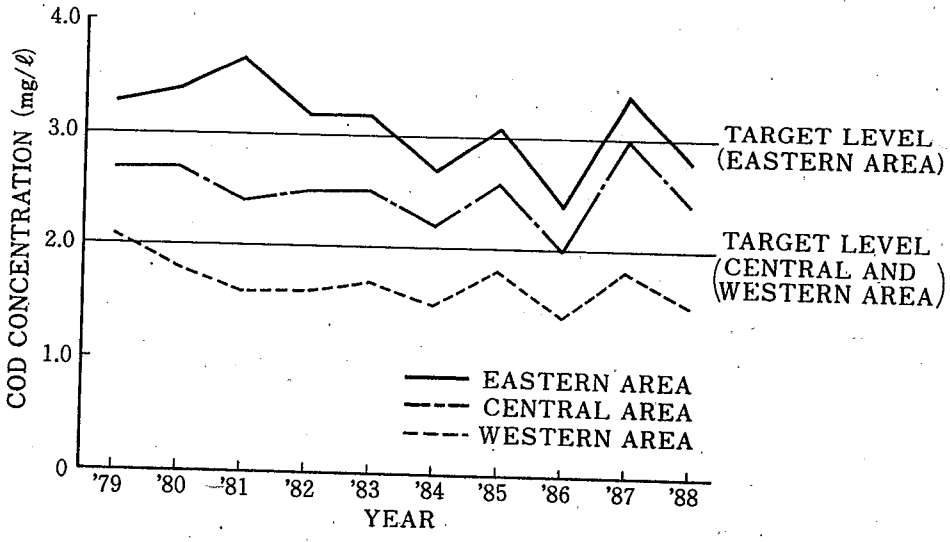


FIGURE 2. TREND OF WATER QUALITY IN HAKATA BAY (COD CONCENTRATION)

The city has decided to develop new advanced wastewater treatment techniques because the sewerage systems will expand to cover 79% of the city towards the end of 1989, and increasing quantities of the eutrophication causing materials cannot be treated efficiently using secondary treatment processes.

2-2. Object of Advanced Wastewater Treatment Experiments

The object of the advanced wastewater treatment experiments is to develop biological phosphorus removal techniques. The advanced processes for the prevention of eutrophication are generally designed to remove COD, phosphorus and nitrogen. The new treatment process of the city is aimed at removal of phosphorus. This is because the Environmental Bureau of the city has found through many surveys that phosphorus is the limiting material for eutrophication in Hakata Bay.

The phosphorus removal processes are three general and practical categories:

①The physico-chemical process represented by the coagulation -precipitation and crystallization processes;
②The combination process in which the physico-chemical and biological process are combined. represented by the activated sludge process combined with alum coagulation; and
③The biological process represented by biological phosphorus removal process.

Fukuoka City has selected the biological phosphorus removal approach for the following reasons:

①The existing activated sludge plants can be easily modified into the biological phosphorus removal plants.
②It is superior to the coagulation-precipitation process in running cost, and
③It produces less quantities of sludge than that of the coagulation -precipitation process.

One of the major disadvantages involved in the biological phosphorus removal process is that phosphorus absorbed by the activated sludge may be released when the sludge is handled inadequately, and this makes it difficult to treat sewage efficiently. The sludge treatment plants in all of the sewage treatment facilities are provided with anaerobic digestion tanks, because the sludge is finally treated into composts for farms. Therefore, phosphorus in the return water from the sludge treatment plants must be carefully controlled and reduced.

3. OUTLINE OF ADVANCED WASTEWATER TREATMENT EXPERIMENTS

3-1. Outline of Experimental Plant

One of the 6 activated sludge lines at the Tobu sewage treatment plant was modified into the experimental plant,which has capacity of 16,700 m^3/day. Tobu sewage treatment plant has total capacity of 100,000 m^3/day max with conventional activated sludge process. A schematic diagram of experimental phosphorus removal process is as shown in Figure 3. The sludge treatment plants consists of the mixed sludge gravity thickening, two-stage anaerobic digestion, elutriation, and filter press dewatering.

In the experiments, the aeration tank (herein refferd to as the biological reactor, when used for the biological phosphorus removal process) consisting of 7 compartments, devided by the partition walles, modified in such way that the first 2 compartments were made anaerobic stage by stopping aeration, the remainder left aerobic stage as before. The anaerobic stage was mixed by the agitators and the openings in the wall between the anaerobic and aerobic tanks were reduced, to limit backflow of water. The excess activated sludge generated at the experimental line was passed to the preaeration tank of another adjacent line, to prevent release of phosphorus in the primary sedimentation tank. The return sludge was passed back to the first compartment of the biological reactor as previously done.

The target level of the phosphorus removal was 0.5 mg/ℓ or less as the total phosphorus in the final sedimentation tank effluent. The target level both of BOD and SS were set equivalent to those for the standard activated sludge process.

The sludge treatment experiments were carried out by bench scale test plants because quantities of sludge generated from the experimental line were too small to be treated by the commercial line.

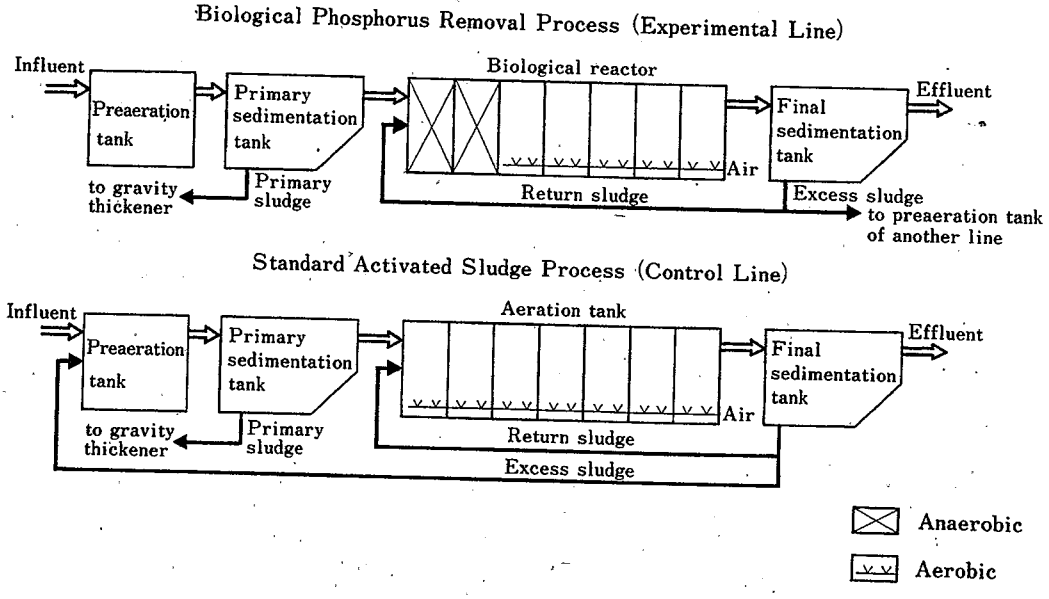


FIGURE 3. SCHEMATIC DIAGRAM OF EXPERIMENTAL PHOSPHORUS REMOVAL PROCESS

3-2. Results of Biological Phosphorus Removal Experiments

The experimental conditions and results are shown in Table 2 and Table 3, respectively. Total phosphorus in the treated water was in arrange of from 0.37 to 0.44 mg/ℓ, well below the target of 0.5 mg/ℓ, and also the BOD and SS levels were equivalent to those obtainable by the standard activated sludge process. Temporary increases in NOx and decreases in phosphorus removal efficiency, caused by decreased substrates concentrations, were observed during raining periods, but the plant conditions recovered well soon after rain stopped. As a result, the target level was reached for more than 90% probability through experimental period.

These favorable results, however, were with 5 to 6 mg/ℓ as total phosphorus in the influent wastewater. It was anticipated by the calculation by material balance, when all the lines were modified into the biological phosphorus removal process, that total phosphorus in the feed water would increase to 9 mg/ℓ, unless some countermeasures to prevent release of phosphorus were taken in the sludge treatment system. It was experimentaly proven that increasing total phosphorus in the feed wastewater to 8 mg/ℓ resulted total phosphorus in the effluent from final sedimentation tank to above 1 mg/ℓ.

In order for the biological phosphorus removal plants to run with some stability, it is, therefore, necessary to decrease phosphorus concentration in the return water from the sludge treatment process to a level comparable with that generally observed in the standard activated sludge process.

TABLE 2. EXPERIMENTAL CONDITIONS AT WASTEWATER TREATMENT PROCESS

			BPRP [1]		SASP [2]
			1986fy	1985fy	1986fy
Flow rate		(m^3/day)	12,105	12,139	12,143
Aeration/flow ratio			4.7	3.9	4.3
Return sludge ratio		(%)	86	82	77
Excess activated sludge		(m^3/day)	315	313	173
MLSS		(mg/ℓ)	2570	2380	2330
SVI			140	140	180
C-BOD·SS loading		(kg/kg-day)	0.16	0.15	0.11
SRT		(day)	4.9	6.1	6.8
MLDO [3]		(mg/ℓ)	3.6	4.0	2.8
Reactor water temp.		(℃)	21.5	21.0	21.0
HRT	Anaerobic	(hour)	1.3	1.3	
	Aerobic	(hour)	3.1	3.3	4.7

Note: 1 Biological Phosuphorus Removal Process
2 Standard Activated Sludge Process
3 at the end of reactor

TABLE 3. RESULTS OF TREATMENT (WASTEWATER TREATMENT PROCESS)

			BPRP 1986fy		BPRP 1985fy		SASP 1985fy	
			Influent[1]	Effluent[2]	Influent	Effluent	Influent	Effluent
T-P	Concentration	(mg/ℓ)	6.62	0.37	5.54	0.44	5.89	2.96
	Removal ratio	(%)	94		92		50	
C-BOD	Concentration	(mg/ℓ)	140	2.8	117	2.9	100	2.3
	Removal ratio	(%)	98		98		98	
SS	Concentration	(mg/ℓ)	135	4	111	4	97	5
	Removal ratio	(%)	96		96		95	
T-N	Concentration	(mg/ℓ)	39.8	17.3	38.2	19.8	36.5	17.5
	Removal ratio	(%)	57		48		52	
Nitrification ratio[3]		(%)	30		10		62	

Note: 1 Influent of Reactor

2 Effluent of secondary sedimentation tank

3 No_2-N and No_3-N/T-N

4. DEVELOPMENT OF SLUDGE TREATMENT TECHNIQUES

4-1. Sludge Treatment Process

All the sludge treatment processes at the city's sewage treatment plants consist basically of a mixed sludge gravity thickening, two stage anaerobic digestion, elutriation and filter press dewatering. The sludge from the biological phosphorus removal plant contains much more phosphorus than that of the standard activated sludge plant. The phosphorus may be released again into water when left under anaerobic conditions for a long time. Phosphorus release is an unavoidable problem in the sludge treatment process which operates mostly under anaerobic conditions. More phosphorus is released in the gravity thickning and digestion processes than in the others. Thus, prevention of phosphorus release and recovery of released phosphorus have been studied in the above two stages.

4-2. Composition of Excess Activated Sludges

The compositions of the excess activated sludge from the biological phosphorus removal process and the primary sludge is as shown Table 4. As shown, the excess sludge contains more organic matter than that of the primary sludge, and phosphorus content is 2.6 times of that of the primary sludge. Pottasium and Magnesium contents are also higher than those in the primary sludge, suggesting biological inclusion of phosphorus in the sludge.

TABLE 4. COMPOSITION OF SLUDGES

		Excess sludge(A)	Primary sludge(B)	A/B×100
TS	(%)	3.67	3.96	93
VS	(%)	70.8	63.3	112
pH		6.53	5.94	110
T-N	(mg/kg)	67,000	43,500	110
T-P	(mg/kg)	26,200	10,300	254
K	(mg/kg)	9,000	2,980	302
Ca	(mg/kg)	11,900	19,800	60
Mg	(mg/kg)	9,000	6,410	140
Cu	(mg/kg)	359	411	87
Zn	(mg/kg)	1,000	1,300	77
Fe	(mg/kg)	16,100	16,400	98
Cd	(mg/kg)	1.76	1.78	99
T-Hg	(mg/kg)	0.94	1.52	62

Note: Excess sludge is one thickend by dissolved air floatation
Primary sludge is one thickend by gravity thickener

4-3. Phosphorus Release in Gravity Thickening

Release of phosphorus from the excess sludge was investigated with storage tests. the results are shown in Figure 4. Release of phosphorus was completed in 24 hours, increasing ortho phosphorus concentration in liquid to 100 mg/ℓ, accompanied by the reduction in total phosphorus concentration in the sludge from 3.7% to 2.2 %. Amounts of phosphorus release in 12 hours was about 60% of the amounts released in 24 hours. From a standpoint of phosphorus release, the gravity thickening process generally has a detention time of about 12 hours, which indicates that gravity thickening is not suitable as a process for treating excess sludge in the biological phosphorus removal process.

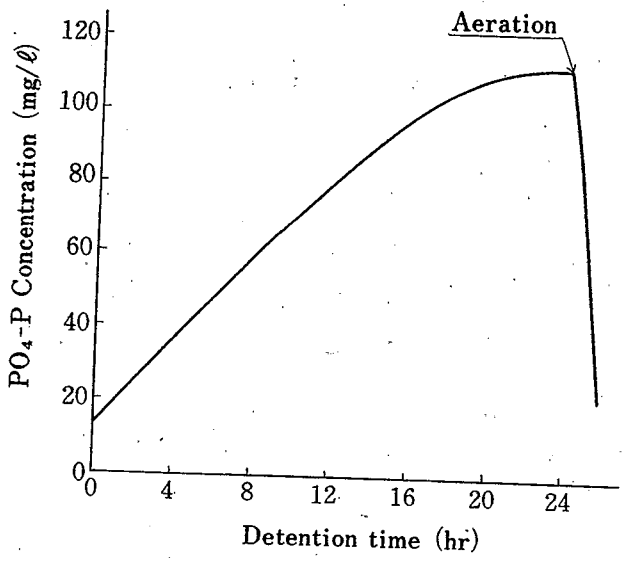


FIGURE 4. RELEASE OF PHOSPHORUS FROM EXCESS SLUDGE

4-4. Dissolved Air Floatation

The excess sludge which is high concentration of organic matter and difficult to thicken by gravity is generally supplied with mechanical concentration technique. The potential of dissolved air floatation was studied for the reason of its short thickning time and also aerobic conditions. The operational results are summarized in Table 5. The results were 3.3% as the thickened sludge concentration and 0.23 mg/ℓ as the phosphorus concentration in the separated water.

TABLE 5. RESULTS OF THICKENING TEST

		Excess sludge of BPRP	Floatation		Gravity	
			Thickened[1]	Effluent[2]	Thickened[1]	Effluent[2]
pH		6.5	6.3	6.5	6.2	6.5
SS	(mg/ℓ)	4,840	33,400	320	17,300	1,940
PO_4-P	(mg/ℓ)	1.91	5.46	0.23	9.80	5.64
T-P in sludge	(%)	3.1				

Note: 1 Thickened sludge
2 Thickener effluent

4-5. Phosphorus Release in Digestion Process

The digestion experiments were conducted, to investigate the behavior of phosphorus in the digestion process, where the excess sludge from phosphorus removal process, thickened by dissolved air floatation was digested, independently or mixed with the primary sludge (mixing ratio: 3 versus 2). The experimental condition and results are shown in Table 6 and 7, respectively.

In the case of excess sludge only, the sludge concentration and also organic matter concentration were too high to be efficiently treated within the 26 days detention time, and the solid-liquid separation was impossible.

In the case of the mixed sludge, on the other hand, solid-liquid separation was feasible, because their sludge and organic matter concentration were within adequate ranges.

Figure 5 shows changes in phosphorus concentration in supernatant. The phosphorus concentration in the case of excess sludge alone, increased gradually with time, up to 413 mg/ℓ after 2 months. The phosphorus concentration in the case of mixed sludge, however, was fairly constant at 200 to 250 mg/ℓ without a trend of the increasing for the experimental period. In any case, however, the supernatant contains very high concentrations of phosphorus and, when returned directly to the waste water treatment system, will invariably increase inflow phosphorus quantities drastically.

TABLE 6. EXPERIMENTAL CONDITION OF DIGESTION TEST

		Excess sludge	Mixed sludge[1]
Sludge volume before digestion	(ℓ/day)	5.0	5.0
after digestion	(ℓ/day)	3.03	2.97
Supernatant volume	(ℓ/day)	1.60	1.72
Digestion temperature	(℃)	37.0	36.1
Digestion period	(day)	26	26
Digestor gas generation ratio		7.46	7.94
Organic loading	($kg/m^3 \cdot day$)	1.05	0.91
Digestion rate	(%)	35	34

Note: 1 Excess sludge and Primary sludge (3:2)

TABLE 7. RESULTS OF DIGESTION TEST

		Excess sludge			Mixed sludge		
		Inf.[1]	Dig.[2]	Sup.[3]	Inf.	Dig.	Sup.
pH	(%)	6.5	7.4	7.4	6.3	7.2	7.4
TS	(%)	4.14	3.35	3.63	3.18	3.01	0.78
VS	(%)	66.6	56.3	56.7	73.9	65.2	63.0
T-P	(mg/kg)	1,130	960	1,040	791	1,042	417
PO_4-P	(mg/ℓ)	101	289	274	75.4	245	239
T-N	(mg/kg)	2,610	2,670	2,760	2,170	2,550	1,420
M-Alkalinity	(mg/ℓ)	282	3,520	3,520	386	2,950	2,930
VOA	(mg/ℓ)	12	112		135	137	

Note: 1 Influent
2 Digested sludge
3 Supernatant

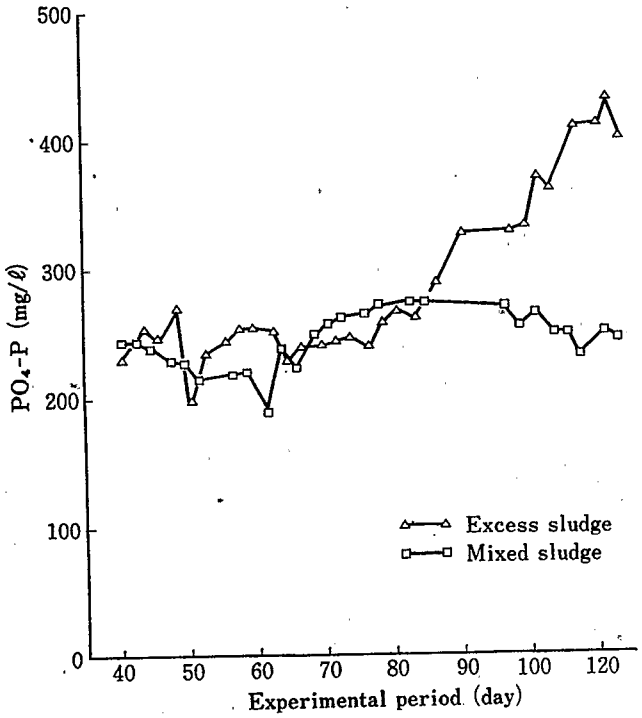


FIGURE 5. CHANGES OF PHOSPHORUS CONCENTRATION IN SUPERNATANT

Therefore, the phosphorus blockade is necessary. In an attempt to control phosphorus concentration in the return water from the digestion process, the city has studied three approaches as follows:

①Addition of lime (calusium oxide) and ferric chloride to the feed sludge for the digestion tank;

②Addition of lime and ferric chloride to the supernatant, to fix phosphorus present therein;

③Utilization of crystallization of Magnesium Ammonium Phosphate (MAP)

4-6. Phosphorus Fixation (Preventing Phosphorus Release) with Lime and Ferric Chloride in Digestion Tank

Figure 6 illustrates that the pilot plant with 2 tanks of 70 liter volume for the phosphorus fixation. The sludge used in the experiments was the excess sludge treated by the dissolved air floatation process. The fixing chemicals were lime and ferric chloride. Four cases are studied, where the lime addition ratio was set at 0, 4.8, 7.4 or 9.8% of the total solid weight, with neutralization-equivalent quantity of ferric chloride, to follow phosphorus fixation and digestion.

The results are shown in Figure 7 and Table 8. The extent of phosphorus fixation ascended in proportion to lime addition ratio, whereas that of digestion reached a maximum at about 0.7 mol/mol as the ratio of the fixing chemical to total phosphorus in the feed sludge, decreasing thereafter with lime addition ratio. This could be attributed to decreased activity of the digestive bacteria as a result of increased quantity of the inorganic coagulant. It is supposed that the optimum addition ratio of the fixing chemical will be around 1.0 mol/mol (about 5% as feed lime concentration) for the phosphorus fixation from consideration of digestion ratio, limiting the attainable fixing ratio to around 70% or phosphate concentration in the digested sludge to approximately 100 mg/ℓ.

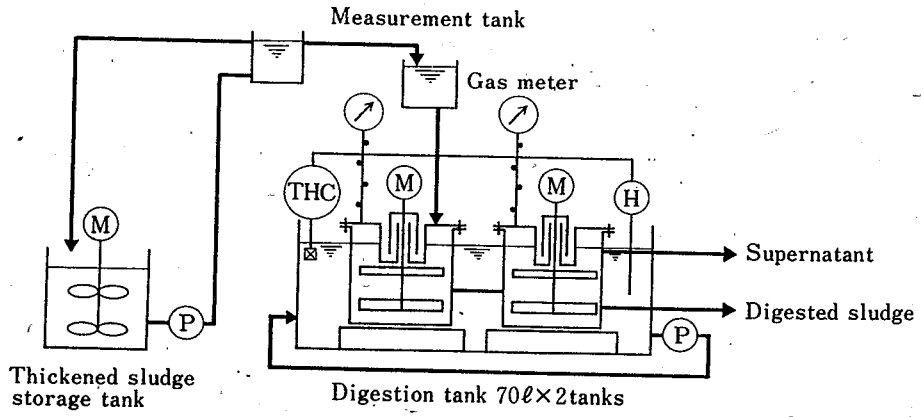


FIGURE 6. FLOW CHART OF DIGESTION EXPERIMENT PILOT PLANT

TABLE 8. RESULTS OF PHOSPHORUS FIXATION IN DIGESTION TANK

Dosage Molar ratio to T-P			TS	VS	T-P	PO_4-P Con.[1]	PO_4-P F.R[2]	Digestion ratio
Lime	$FeCl_3$		(%)	(%)	(mg/kg)	(mg/ℓ)	(%)	(%)
0	0	0	2.97	66.4	1,170	498	0	36
4.8	5.7	0.84	3.15	65.5	1,140	208	59	38
7.4	8.9	1.58	2.90	67.3	970	39.3	92	32
9.8	11.9	2.10	3.03	68.9	900	11.9	98	22

Note : 1 Concentration
2 Fix ing ratio

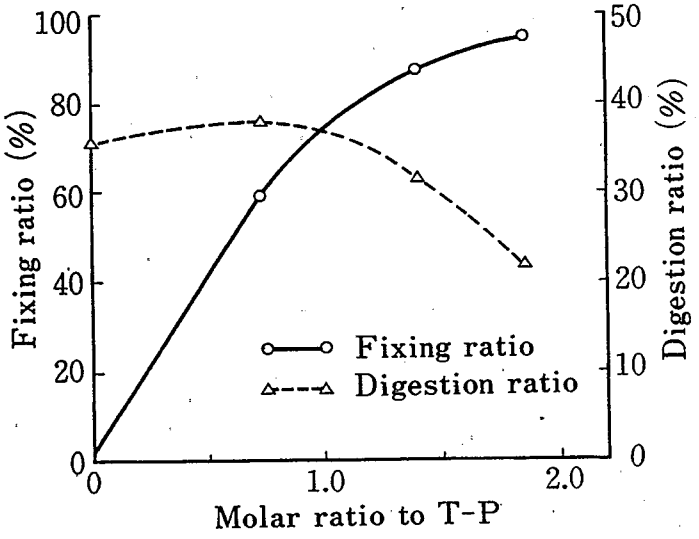


FIGURE 7. PHOSPHORUS FIXATION IN DIGESTION TANK

4-7. Phosphorus Fixation with Lime and Ferric chloride for Supernatant

Next, lime and ferric chloride were added to supernatant to prevent phosphorus release. Three cases were studied:

①lime alone
②ferric chloride alone
③a combination

The results are shown in Figure 8 ~ 10. Ferric chloride removed SS and phosphorus more efficiently than lime. The results indicated that molar addition ratio of the fixing chemical, to keep total phosphorus removal ratio at 90% or more, was about 1 mol/mol with ferric chloride and 4 mol/mol with lime. A combination of these chemical gave the most favorable results; phosphorus fixation ratio of 90% or more was attained at a dosage of about 1 mol/mol, accompanied with improved compositions of both treated water and also precipitated sludges.

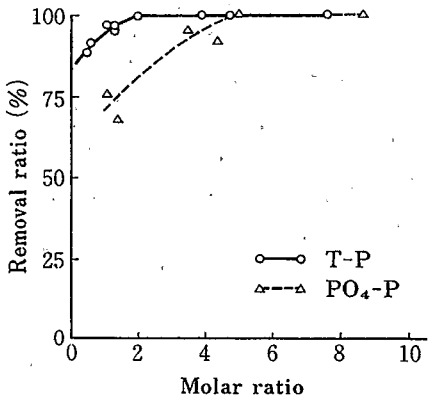


FIGURE 8. PHOSPHORUS FIXATION WITH FERRIC CHLORIDE

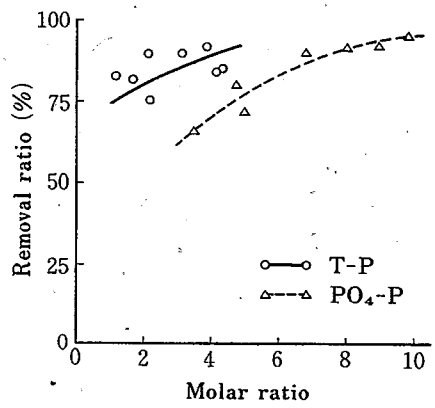


FIGURE 9. PHOSPHORUS FIXATION WITH LIME

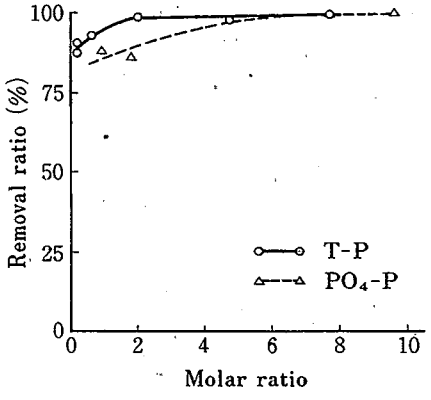


FIGURE 10. PHOSPHORUS FIXATION WITH FERRIC CHLORIDE AND LIME

4-8. Phosphorus Fixation by Crystallization of Magnesium Ammonium Phosphate (MAP) -Development of MAP Process

Many reports said that scales are frequently deposited on supernatant transfer pipes of digestion tank in a sewage treatment plants near sea, causing pipe plugging problems. J. Borgerding detailes that these scales consist mainly of magnesium ammonium phosphate (MAP). Namely, magnesium in sea water brought into a digestion tank reacts with phosphate and ammonium ions present in supernatant from a digestion tank to form insoluble $MgNH_4PO_4 \cdot 6H_2O$ crystals. Therefore, our attempts were made to fix phosphorus by artificially forming MAP.

4-8-1. MAP Forming Conditions

The nesessary conditions for the formation of MAP are:

① presence of magnesium
② slightly alkaline condition of the system .

Supernatant from the anaerobic digestion system contained 207 mg/ℓ of a phosphate, 7.0 mg/ℓ of soluble magnesium, and 756 mg/ℓ of ammonia nitrogen.

As the molar ratio of MAP is 1(P):1(N):1(Mg) from the moleculer formula of $MgNH_4PO_4 \cdot 6H_2O$, magnesium was deficient stoichiometerically to form MAP. Addition of magnesium to the supernatant to a Mg/P ratio of 1.05/1 caused formation of MAP crystals, increasing phosphorus removal ratio.

Solubility of MAP declines as pH ascent, as shown in Figure 11, and the system should be kept at pH 8 or higher, to reserve MAP efficiently. The supernatant alkalinity should be increased by some means, because its pH level was typically within a range from 7.3 to 7.5.

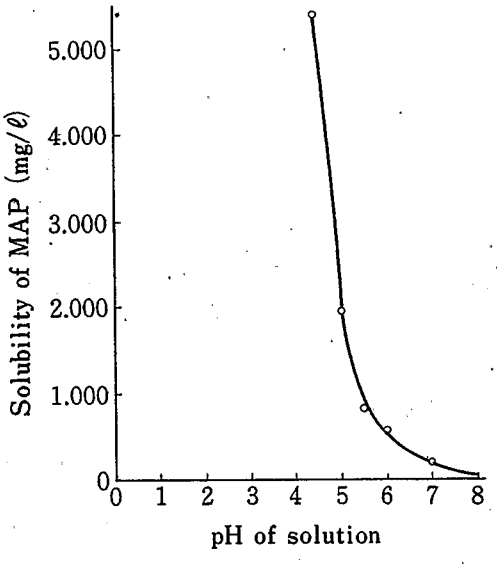


FIGURE 11. EFFECT OF pH ON SOLUBILITY OF MAP [1)]

The following four methods were studied to increase pH level of the supernatant:

① addition of sodium hydroxide
② addition of lime (calusium hydroxide)
③ addition of magnesium hydroxide
④ aeration to accelerate decarbonization

Result of lime was less effective than that of sodium hydroxide, requiring 3 times more quantity of feed chemical to keep the same pH level. Result of magnesium hydroxide failed to increase pH level efficiently. On the other hand, aeration was found to be an effective method to remove carbon dioxide dissolved in the supernatant at a high concentration, to go up pH level to 8.4 in about 2 hours. The pH level adjustment by aeration, however, was accopanied by the odor problems, and so the exhaust gases should be deodorized because of the presence of high concentrations of the materials causing offensive odor. The composition of exhaust gases are shown in Table 9.

TABLE 9. COMPOSITION OF EXHAUST GASES

		Concentration
Ammonia	(ppb)	1,900
Metyl mercaptan	(ppb)	150
Hydrogen sulfide	(ppb)	3.6
Methyl sulfide	(ppb)	37
Di-methyl di-sulfide	(ppb)	2.3

Based on the above results, it has been found that MAP can be produced efficiently by adding magnesium to the supernatant, and by increasing the system pH to at least 8 with addition of sodium hydroxide or aeration. The MAP process to continuously remove phosphorus from the supernatant has been then attempted with pilot plant experiments.

4-8-2. Pilot Plant Experiments for the MAP Process (Part 1)

Figure 12 and Table 10 show the flow diagram of the pilot plant and the experimental conditions respectively. The supernatant storage tank with 2 m³ volume was equipped with an agitator and also an aeration system capable of controlling air ratio arbitrarily with a Hertz converter to increase pH level. The feed water pumps was a variable type, with manual control covering a flow ratio from 4 to 0.5 m³/hour. A NaOH solution was added automatically to the feed water, to keep a pH level in a range from 8.3 to 9.0. $MgCl_2$ injection ratio was changed in accordance with the phosphorus concentration in the supernatant. The supernatant was passed from the storage tank to the rapid stirring tank, then transfered to the slow stirring tank for crystallization. And, finally, the crystalline precipitate were separated from the feed water in the settling tank, and some part of the precipitates were sent back as the seeds to the rapid stirring tank.

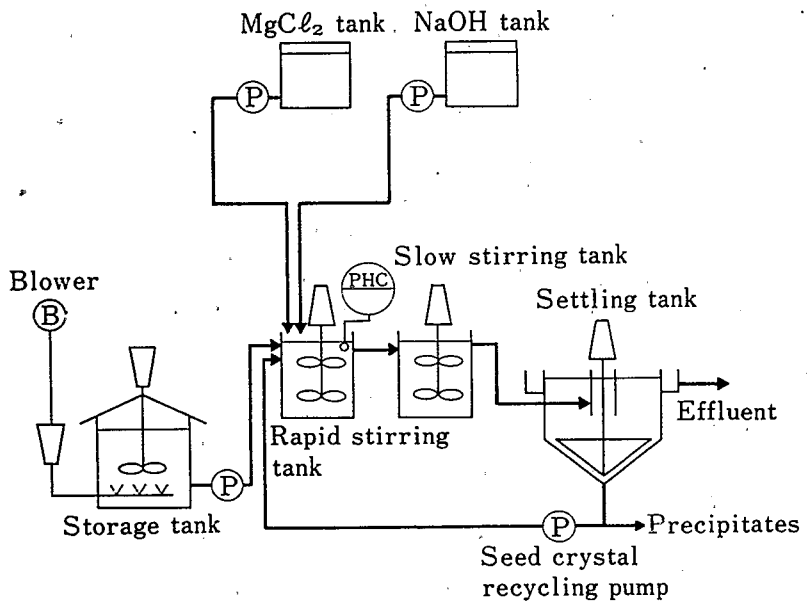


FIGURE 12. FLOW DIAGRAM OF MAP PROCESS PILOT PLANT (No.1)

TABLE 10. EXPERIMENTAL CONDITIONS OF PILOT PLANT (PART 1)

Flow rate	(m³/hour)	0.5
Dosage of $MgCl_2$ solution	(mℓ/min.)	6
Molar ratio to PO_4-P		1.05
Retention time in stirring tanks	(min.)	24
Retention time in settling tank	(min.)	144

Figure 13 shows the results of pilot plant study continuous treatment with capacity of 0.5 m³/hour of the supernatant from the digestion tank at the Tobu sewage treatment plant. Ortho-phosphate concentration of the influent fluctuated from 20 to 80 mg/ℓ, but the effluent Ortho-phosphate concentration ranged from 5 to 15 mg/ℓ (10 mg/ℓ on the average), that is, 60 to 70% as Ortho-Phosphate removal ratio. The crystals, thus, formed were very fine, mostly deposited on SS surfaces, and sometimes leaked into the effluent to cause high phosphorus concentration. The leaked crystalline precipitates may become soluble ortho-phosphate at a stage of reduced pH, and increase phosphorus load of the waste water treatment system. Therefore, it is very important to control SS in the settling tank. In the experiments, the effluent water ratio must have been controlled at 0.5 m³/hour (40m/day as LV) or less. Figure 13 includes test results with the seed crystal recycling. Recycling of the seed crystals, though sometimes increased SS in the effluent water, was an effective method of increasing size of magnesium ammonium phosphate crystals.

The problem observed in this pilot plant experiments were difficulty in aeration to adjust pH of the water in the storage tank, frequently causing precipitation of MAP in the storage tank, and/or decrease in the purity of MAP because the precipited MAP was withdrawn together with the sludge out of the settling tank. As a next step the new pilot plant was designed and constructed to solve the above problems.

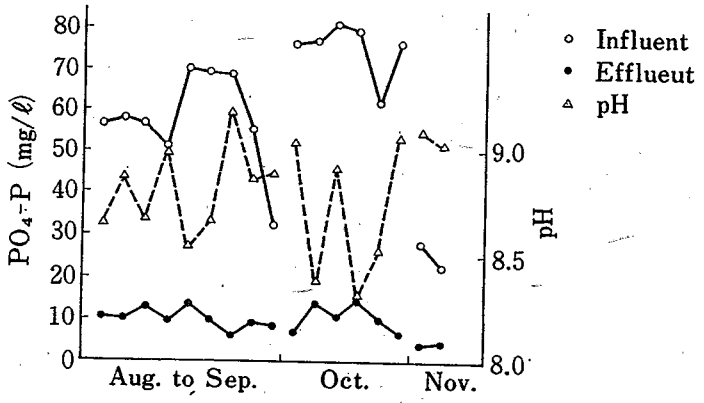


FIGURE 13. RESULT OF MAP PROCESS EXPERIMENT

4-8-3. Pilot Experiments for the MAP Process (Part 2)

Figure 14 illustrates the new pilot plant where aeration, stirring, and crystallization are carried out together in the same reactor tank. MAP precipitates formed in the reactor column, located in the middle of the reactor tank, are separated from SS and settled by gravity. The experiments are still ongoing, but the results obtained so far have indicated a prospect of producing high-purity MAP. However, phosphorus removal ratio is not high so far, because of low phosphorus concentration in the supernatant separated from the digestion tank of standard activated sludge process. As a following experiment, we will attempt the case that supernatant of higher phosphorus concentration is used.

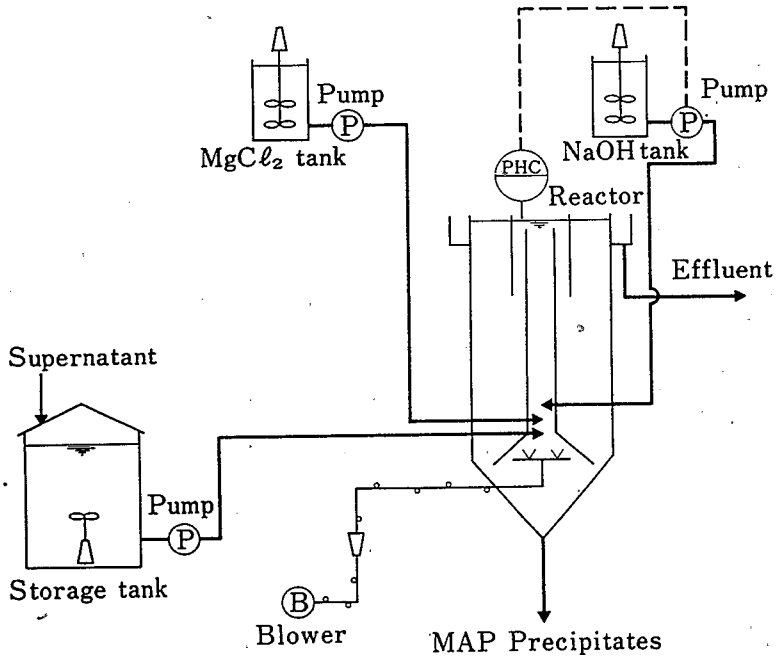


FIGURE 14. FLOW DIAGRAM OF MAP PROCESS PILOT PLANT (No.2)

PHOTO 2.
PILOT PLANT
(No.2)

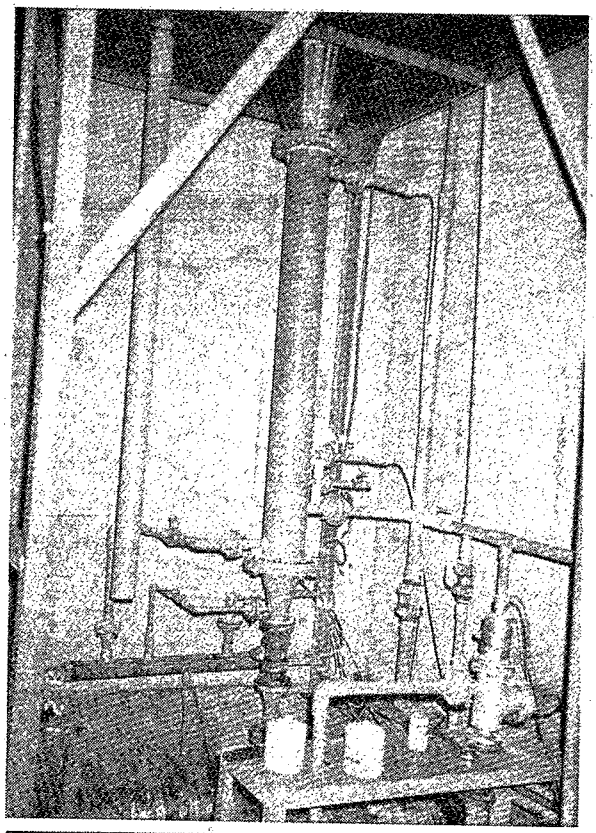
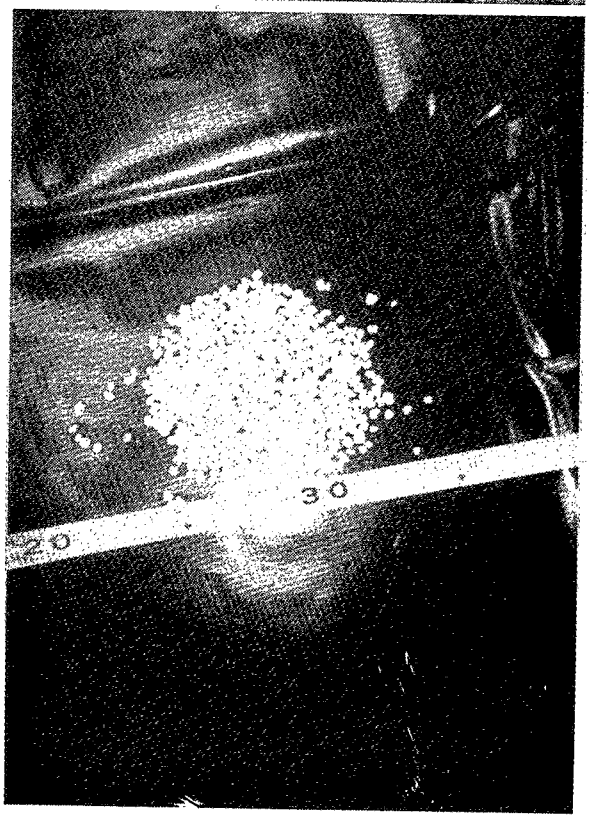


PHOTO 3.
MAP CRYSTALS
RECOVERED
IN THE PLANT



5. AFTERWORD

The pilot experiments are still under way, and there is room for further improvement especially in the case of MAP. Fukuoka City is planning to improve the sludge disposal process well-matched the biological phosphorus removal process. This sequence consists of mechanical thickening of excess sludge, anaerobic digestion , MAP phosphorus fixation, dewatering, and compost production processes.

The MAP process is desirable not only from the viewpoint of stabilization of the advanced wastewater treatment process, but also from the viewpoint of the effective utilization of resources. Magnesium ammonium phosphate, rich in phosphorus and nitrogen, is expected to find use as a product for slow-effect fertilizers. As a newer process, MAP process, therefore, will be developed further also as a phosphorus supplier to fully utilize the limited quantity of natural resources in Japan, because the phosphorus of necessary volume in Japan is covered with 100% by import from foreign countries.

REFERENCE: 1) J.Borgerding, Phosphate deposits in digestion systems, Journal WPCF, vol. 44, No.5, 1972

TOXICITY MONITORING METHOD BY BIOSENSOR

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ABSTRACT

Use of the ammonia biosensor as a tool to monitor toxicity of water is presented. Since Nitrosomonas are known one of the most sentive classes of microorganisms in the biological wastewater treatment processes, respirometry measurement (oxygen uptake) using Nitrosomonas gives more information on the toxic characteristics of water than using activated sludge.

Ammnoia biosensor was initially developed using Nitrosomonas europaea (ATCC 25978) to monitor the ammnoia concentration of water. The ammonia biosensor consists primarily of a receptor (immobilized nitrifier) and a probe of dissolved oxygen meter (DO probe). Depending on the concentration of ammonia in water, the rate of oxygen uptake in the receptor will vary. Comparison of the measurement of output current of a DO probe with the calibration curve which shows the relationship between the ammonia concentaration and the output current gives an estimation of the concentration of ammonia of a sample.

However, if a sample contains a toxic compound against Nitrosomonas europaea, the output current of a DO probe would be lowered depending on the level of toxicity of the compound. Therefore, the level of toxicity can be estimated by comparing the expected output current which is estimated by measuring the ammonia concentration with the actual output of a sample which contains a toxic compound.

Time required to measure the toxicity of a sample is found 10 to 15 minutes. According to the experimental results using ortho-chlorophenol, trichloroethylene and tetrachloroethylene, the ammonia biosensor detects their toxicity with the order of 0.01 mg/l. Although it was known that ammonia concentration would influence the relative magnitude of output, the level of toxicity could be expressed reasonablly by using the kinetic parameters. Required time for the sensor to recover from the damage caused by a toxic compound is found to depend on the magnitude of toxicity previously applied.

INTRODUCTION

A chemical specific approach to regulate the discharge of toxic compounds can only be possible if the levels of toxicity of all chemical compounds against selected indicator organisms are understood and all chemical compounds can be measured by available detection methods. However, it is practically impossible to understand the toxicity levels of all chemical compounds. In addition, available detection methods have not yet developed for thousands of toxic compounds (1). Therefore, use of the chemical specific approach to control the discharge of toxic compounds is limited to specific group of compounds such as heavy metals.

Compared to the chemical specific approach, a biological approach has several advantages to control the discharge of toxic compounds to the public water bodies. In the biological approach, living organisms are directly used to measure the level of toxicity of a compound, or of water even when the amount or name of compounds contained in the water are completely unknown. A variety of biological testing methods is already developed from a simple respirometry test to a complicated bioassay test (2). Although toxicity of a compound against specific living organism does not necessarily mean that the compound is toxic against all living organisms including men, use of a specific living organism as a tool to measure the level of toxicity is practically meaningful.

USEPA developed three types of toxicity test, namely microtox toxicity test, respirometry (oxygen uptake) test, and adenosine triphosphate (ATP) test, to measure the toxicity of wastewater as near real-time and inexpensive monitoring tools (3). Since the measurement of rate of oxygen uptake is a routine job for a plant operation, the respirometry test is the easiest among these methods. However, activated sludge is not necessarily the most sensitive group of microorganisms to toxic compounds. It is known, on the other hand, that nitrification process becomes upset when the influent contains excessive heavy metals or hazardous chemicals (4). Therefore, use of Nitrosomonas as a tool for a respirometry test may give more sensitive result of toxicity test than using activated sludge.

Ammonia biosensor is found capable to detect the level of toxicity of some toxic compounds, though the ammonia biosensor was initially developed to monitor ammonia. And also the ammonia biosensor is found to be potentially used as a continuous monitoring tool of toxicity. This paper describes the principal mechanisms of the ammonia biosensor and introduces some of the experimental results how the biosensor responds to a toxic compound.

DEVELOPMENT OF AMMONIA BIOSENSOR

Structure. Fundamental structure of the ammonia biosensor is shown in Figure-1. At the tip of a probe of dissolved oxygen meter (DO probe), a receptor is mounted where nitrifying bacteria (nitrifier) are immobilized by acetyl cellulose membrane filters. This sandwich structure of receptor is found easier to mount than other immobilizing methods such as gel immobilization by polyacrylamide. And also gel immobilizations are reported to bring a negative effect on the activities of some microorganisms (5). When a buffer solution which contains no ammonia is fed to the sensor, there is a little consumption of oxygen by endogenous respiration of nitrifier

resulting in rather high and steady output current of a DO probe. This steady output is considered as the base output. In recording the output, the current is converted to voltage. When a sample contains ammonia, oxygen is consumed in the receptor resulting in rather lower output. Since the difference of this output from the base output depends on the concentration of ammonia in the sample, ammonia concentration can be estimated. The difference between the sample output and the base output is simply expressed as "the output" unless otherwise specified.

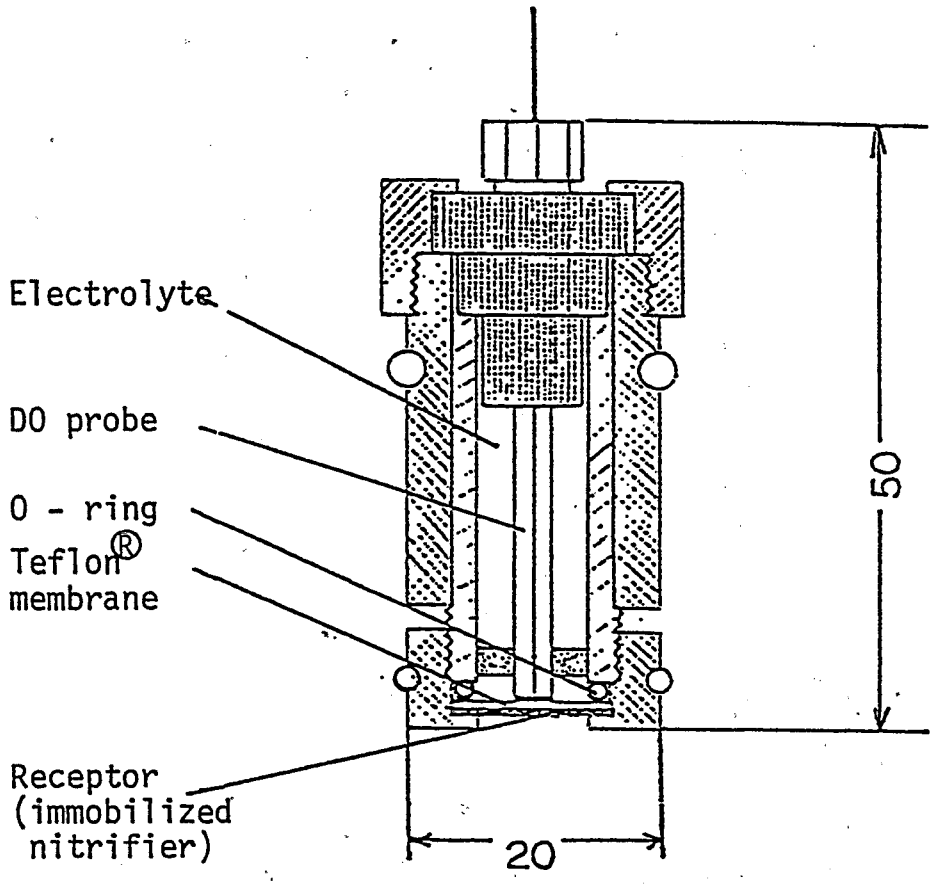

Figure-1 Fundamental Structure of the Ammonia Biosensor

Receptor(mixed culture). In the initial stage of development of the ammonia biosensor, various efforts were made to obtain nitrifier from activated sludge. Mixed liquors were sampled from a wastewater treatment plant where nitrification was well proceeded. Activated sludge was cultured by feeding substrate containing ammonium chloride and no organic compounds. After three month cultivation, nitrifier seemed well grown in the sludge. The sludge was immobilized by membrane filters and mounted to a DO probe. The results of this prototype sensor are shown in Figure-2. It seemes that the sensor responds quite accurately to the change of ammonia concentration. The relationship between the output and the ammonia concentration is shown in Figure-3. Although linearity of the relationship can not be maintained in the region of high ammonia concentration because of oxygen limitation, use of this prototype sensor seemed possible. However, the prototype sensor was

found to respond to organic compounds contained in a sample, because mixed culture of microorganisms were applied as a receptor. Figure-4 shows an example where the sensor responded to organic compounds. To overcome the effect of organic compounds, efforts were made using antibiotics and other chemicals in vain. The initial development became a complete failure.

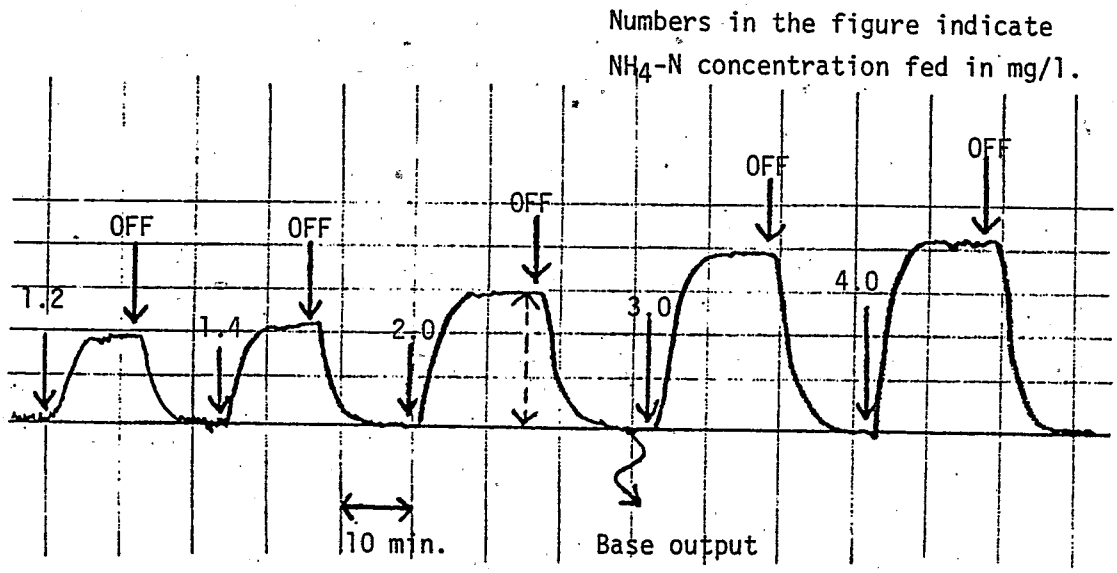

Figure-2 Recording Results of the Prototype Sensor

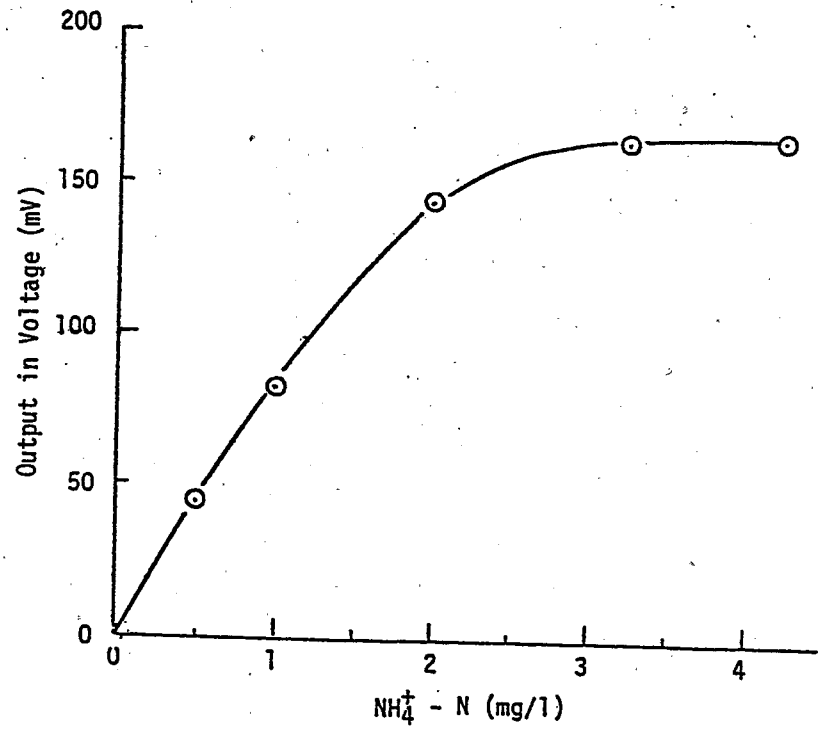

Figure-3 Relationship between the Output and Ammonia Concentration

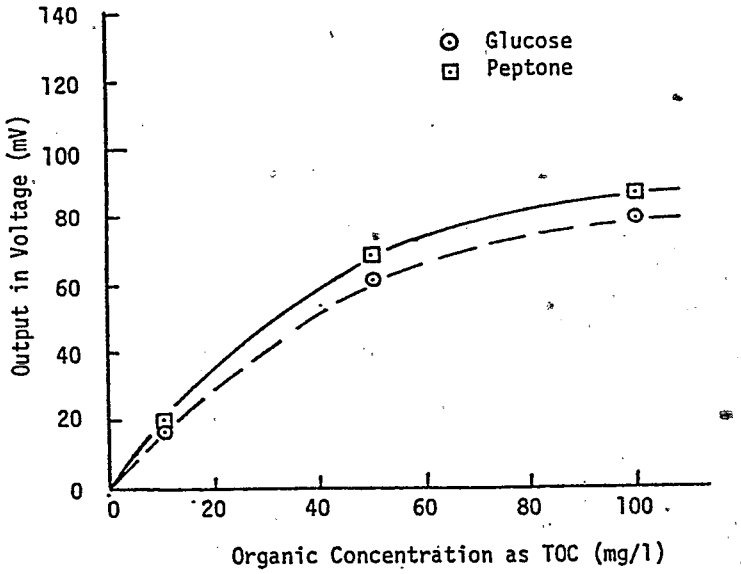

Figure-4 Output Response to Organics Using the Prototype Sensor

Receptor(pure culture): Use of pure cultured Nitrosomonas was considered from the experience of initial development. Nitrosomonas europaea (ATCC 25978) was cultured by using Pramer's culture medium (6) as shown in Table-1. Receptor structure is shown in detail in Figure-5. Although optimum pH and temperature where the maximum output was gained were found 9.2 and 35°C respectively, the sensor was tested under pH of 8.0 and temperature of 30°C to save required amount of a buffer solution as well as energy for heating. No response to organic compounds was identified by using glucose and peptone as organics. The amount of nitrifier was found to affect the relationship between the output and the ammonia concentration. Figure-6 shows how the relationship differs by the amount of immobilized nitrifier expressed in adenosine triphosphate (ATP) density. It can be said from Figure-6 that higher concentration of nitrifier is preferable to measure lower concentration of ammonia, and lower concentration of nitrifier is preferable to measure wider range of ammonia concentartion. Although durability test is still going on, no deterioration effect is identified for 50 days of use.

Table-1 Pramer's Culture Medium for Nitrosomonas

$(NH_4)_2\ SO_4$	2.5 g
$Na_2\ H\ PO_4$	13.5 g
$KH_2\ PO_4$	0.7 g
$Na\ H\ CO_3$	0.5 g
$Mg\ SO_4 \cdot 7H_2O$	0.1 g
$Ca\ Cl_2 \cdot 2H_2O$	5 mg
$Fe\ SO_4 \cdot 7H_2O$	30 mg
Distilled water	1,000 ml

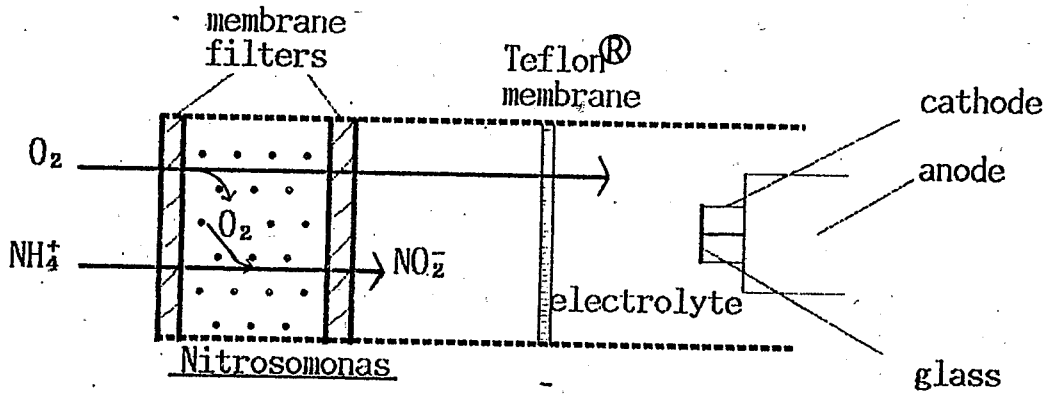

Figure-5 Detailed Structure of the Receptor of Ammonia Biosensor

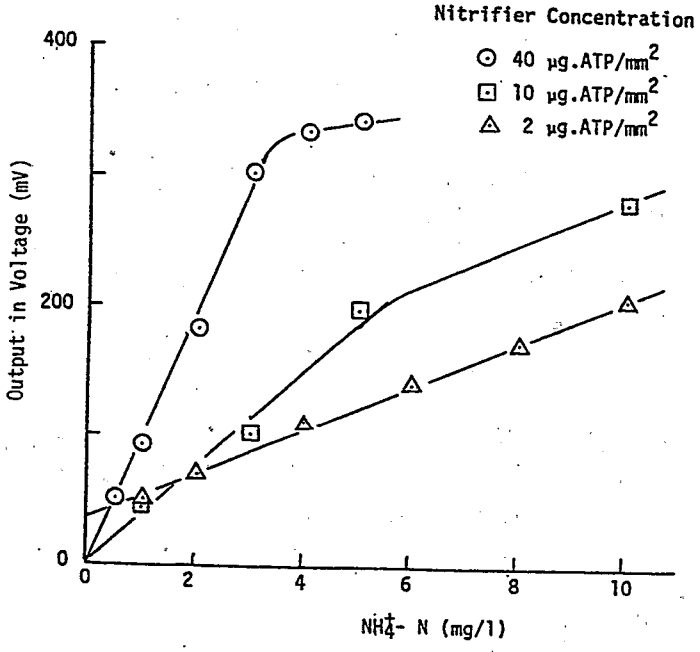

Figure-6 Effect of Nitrifier Concentration on the Output-Ammonia Concentration Relationship

Ammonia Biosensor System. To apply the ammonia biosensor as a monitoring tool of ammonia, supporting devices should be combined. Although sampling is an important part of whole monitoring processes, no effort has been made since an existing sampling procedure may be applied. During the analysis of ammonia, pH and temperature of the sample should be maintained constant. If a sample is lack of oxygen, oxygen should be supplied. Standard solutions of ammonia should be prepared for automatic calibration. The output should be expressed in concentartion, which requires the mount of a calculating system as well as a recording system. Figure-7 shows a schematic system of the monitor using the ammonia biosensor. A buffer solution is supplied to adjust the pH of a sample automatically. Air is supplied to a sample regardless to the dissolved oxygen concentration of a sample. Figure-8 shows an operational example of the monitor.

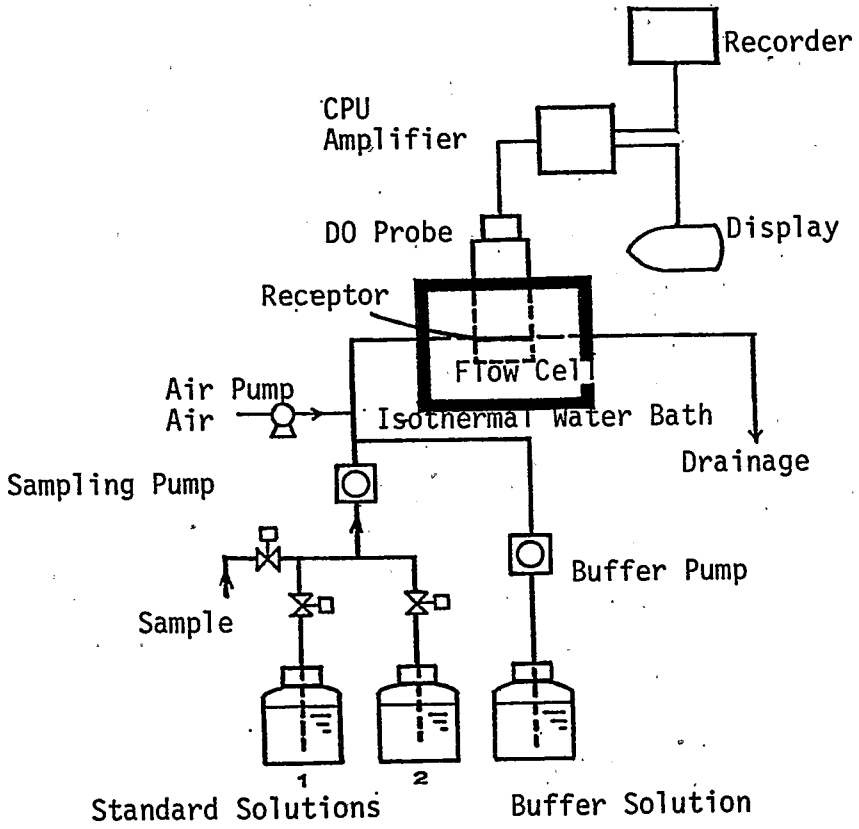

Figure-7 Schematic View of the Monitor Using the Ammonia Biosensor

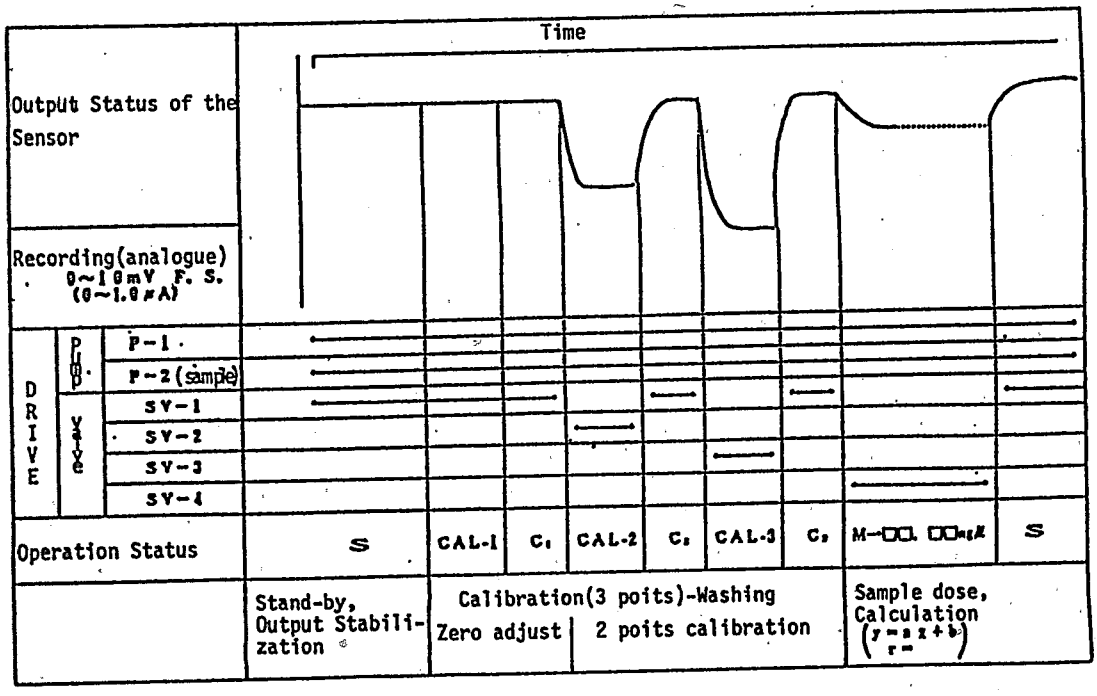

Figure-8 Operational Example of the Ammonia Monitor

MONITORING OF TOXICITY

Effects of toxic compounds. Since the ammonia biosensor uses Nitrosomonas europaea as a receptor of the sensor, it is expected to have some effects in the output when supplied with a toxic containing sample. Three chlorinated organic compounds, ortho-chlorophenol, trichloroethylene and tetrachloroethylene, were tested to measure their toxicity against Nitrosomonas europaea under the ammonia concentration of 1.0 mg/l. Since aeration of a sample is considered to have an effect to change the concentration of a test compound, a sample solution was previously aerated before dosing a toxic compound to a sample solution. Necessary time of recovery for the ammonia biosensor was examined for ortho-chlorophenol as a toxic compound.

Since maximum allowable concentaration (MAC) of each compound below which there was no effect on the rate of respiration was not precisely detected, the maximum concentration where the output was the same with the control was assumed as MAC. Figures 9, 10 and 11 show the results of experiment for ortho-chlorophenol, trichloroethylene and tetrachloroethylene, respectively. Relative magnitude of output was calculated by comparing sample output with control output. For ortho-chlophenol, MAC was considered as 0.034 mg/l. It can be recognized that there is a substantial decrease in the relative magnitude of output as the concentration of ortho-chlorophenol increases. Similar tendency is also recognized for other two compounds.

When the ammonia biosensor is fed with a sample containing toxic compounds, time is required for the sensor to recover from the effects of toxics. One of the experimental results using ortho-chlorophenol as a toxic compound is shown in Figure-12. Required time of recovery is increased proportionally to the increase in toxic concentration.

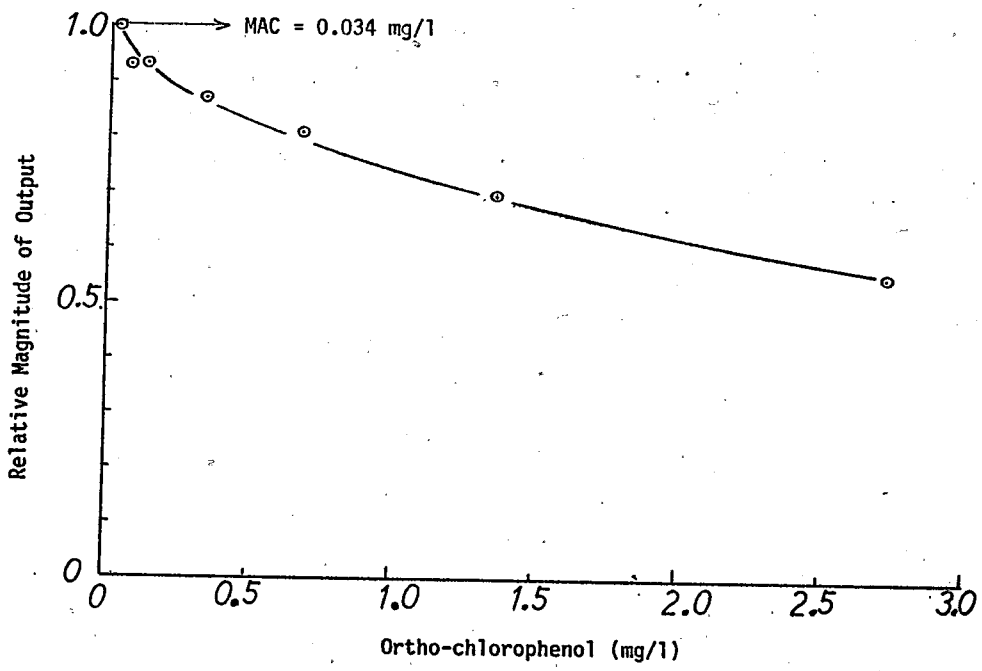

Figure-9 Effect of Ortho-Chlorophenol on Relative Magnitude of Output

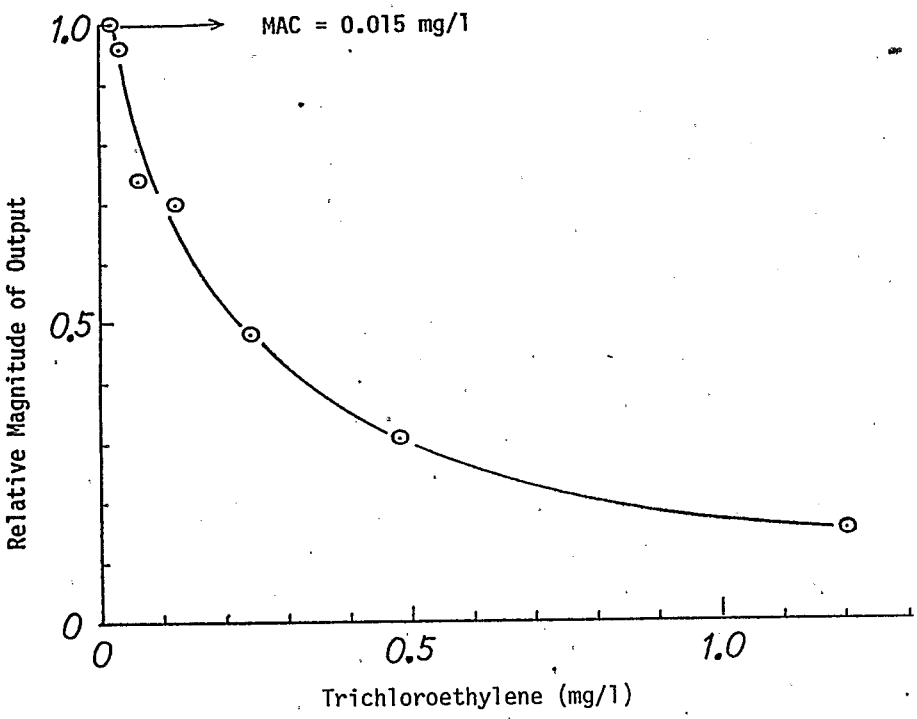

Figure-10 Effect of Trichloroethylene on Relative Magnitude of Output

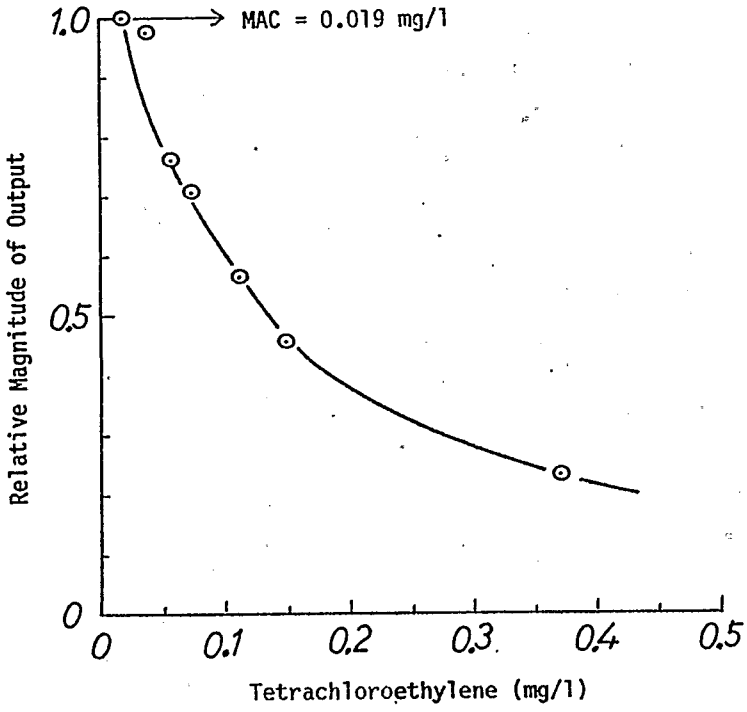

Figure-11 Effect of Tetrachloroethylene on Relative Magnitude of Output

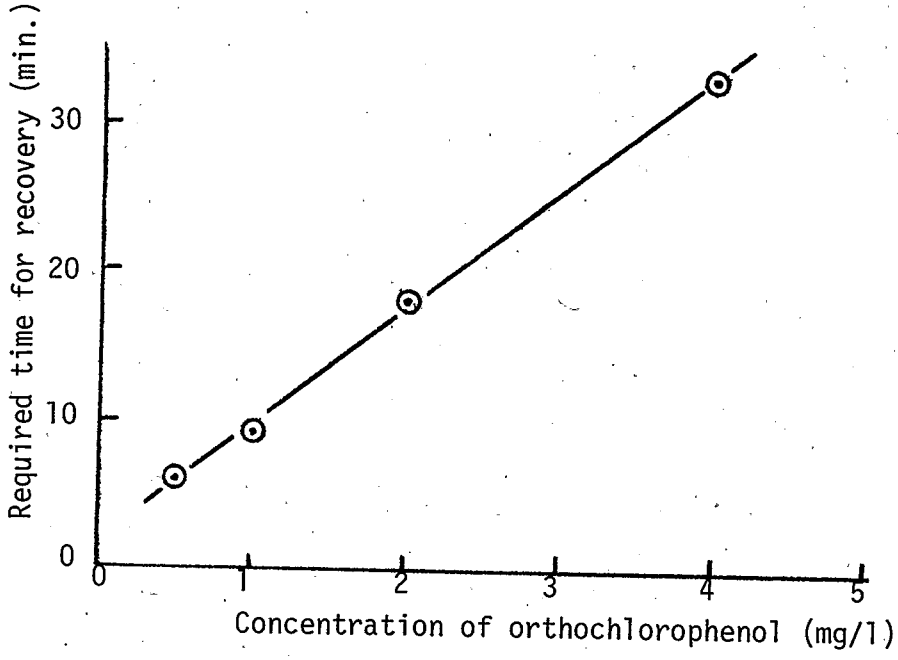

Figure-12 Required Time of Recovery for Ortho-Chlorophenol

Inhibition analysis. Assuming the inhibitory effect of tested organic compounds as fully competitive reversible inhibition, the rate of oxygen consumption in the receptor can be expressed as follows (7):

$$v = Vmax\ S\ /\ (\ S + Ks\ (1 + I\ /\ Ki)) \qquad (1)$$

where, Vmax is the maximum rate of reaction, S is substrate concentration, Ks is half velocity constant, I is concentration of an inhibitor, and Ki is an equilibrium constant for enzyme and inhibitor.

Since there is no effect detected if the concentration of an inhibitor is below MAC, the equation (1) may be rewritten as follows:

$$v = Vmax\ S\ /\ (S + Ks\ (1 + (I - MAC)\ /\ Ki)) \qquad (2)$$

where, (I - MAC) = 0 if I is lower than MAC.

According to the quation (2), toxicity can be expressed by measuring (I-MAC)/Ki. Assuming the relative magnitude of output of the ammonia biosensor is the same as the relative rate of reaction, toxicity can be expressed as follows:

$$(I - MAC)\ /\ Ki = ((1 - p)/p)\ ((Ks + S)\ /\ Ks) \qquad (3)$$

where, p is the relative magnitude of output as shown in Figures 9 to 11.

It is known from the equation (3) that the relative magnitude of output of the ammonia biosensor, p, is not an absolute expression of toxicity, since p will vary with substrate concentration, S, even when the toxicity expressed by (I-MAC)/Ki remains the same. Since the ammonia concentration used in the

toxicity experiments was 1.0 mg/l as nitrogen and Ks for Nitrosomonas can be assumed 2.35 mg/l as nitrogen for 30 C (8), toxicity of three organic compounds can be summarized as Tables 2, 3 and 4. The equilibrium constant for enzyme and inhibitor, Ki, for each compound is also shown in the tables. Although the coefficient of variation (CV) for ortho-chlorophenol is rather high, lower CVs are obtained for trichloroethylene and tetrachloroethylene.

Table-2 Toxicity of Ortho-chlorophenol

Concentration (mg/l)	Relative magnitude of output, p	Toxicity (I-MAC)/Ki
0.034	1.0	0
0.068	0.932	0.104
0.136	0.941	0.089
0.340	0.867	0.219
0.680	0.810	0.334
1.360	0.694	0.629
2.720	0.552	1.157
6.800	0.185	6.280

Ki = 2.184 mg/l, CV = 0.272

Table-3 Toxicity of Trichloroethylene

Concentration (mg/l)	Relative magnitude of output, p	Toxicity (I-MAC)/Ki
0.015	1.0	0
0.030	0.961	0.058
0.060	0.741	0.498
0.120	0.699	0.614
0.240	0.480	1.544
0.480	0.305	3.248
1.20	0.151	8.015

Ki = 0.217 mg/l, CV = 0.073

Table-4 Toxicity of Tetrachloroethylene

Concentration (mg/l)	Relative magnitude of output, p	Toxicity (I-MAC)/Ki
0.019	1.0	0
0.037	0.977	0.034
0.056	0.760	0.450
0.074	0.707	0.591
0.111	0.566	1.093
0.148	0.457	1.694
0.370	0.231	4.746

Ki = 0.120 mg/l, CV = 0.074

CONCLUSION

The ammonia biosensor using Nitrosomonas europaea (ATCC25978) was developed and verified that the sensor could be applied to monitor ammonia. Toxicity effects on the output response of the ammonia biosensor were experimentally studied by using ortho-chlorophenol, trichloroethylene and tetrachloethylene. The results of experiment showed that there was a substantial effect on the output response of the ammonia biosensor for every toxic compound tested. However, toxicity analysis showed that the relative magnitude of output could not directly applied to measure the toxicity. Expression of toxicity can be done by using kinetic parameters as well as the relative magnitude of output. Although use of the ammonia biosensor as a monitoring tool for toxicity requires further study on system construction, the experimental results indicated high potential of practical use of the ammonia biosensor in the field of toxic management.

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RENOVATION OF AN EXTENDED AERATION TREATMENT PLANT FOR BIOLOGICAL NUTRIENT REMOVAL PROCESS

by

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Cincinnati, Ohio

ABSTRACT

The effect of the oxidation reduction potential (i.e., ORP) control system on the simultaneous removal of nitrogen and phosphorus utilizing the oxic-anoxic-oxic process was investigated. A full-scale wastewater treatment plant which was designed as an extended aeration process was modified for this process to study. The optimum ORP at the first basin for the simultaneous removal was 125 mV in this experiment. In the first oxic basin the nitrification reaction and the denitrification reaction occurred simultaneously to maintain the optimum level of the ORP value. The averaged values of the effluent of this process were approximately 2 mg/l of total nitrogen (T-N), 0.5 mg/l of total phosphorus (T-P) and 5 mg/l of BOD respectively. It was also clearly shown by the statistic analysis that the ORP control was more effective in comparison with the control of dissolved oxygen (DO) concentration for the operating method of this process. Additionaly, it was supposed that the simultaneous nitrification and denitrification using the ORP control system was a reliable nitrogen removal system for renovating the conventional activated sludge process because the hydraulic retention time of the first basin used in this study was equal to that of the conventional process.

INTRODUCTION

Lakes and reservoirs in Japan have played an important role as resources for municipal water supply, recreation, fresh water fisheries, and other purposes. However, the Environmental Water Quality Standards achievement rate, which is defined as the ratio of number of water bodies meeting the standards to the total number of water bodies, was extremely low for lakes and other enclosed waters; moreover their quality has remained at the same level for several years. This is closely related to eutrophication processes in these water bodies. Eutrophication has brought about a variety of harmful effects, such as scenic damage due to decrease in the transparency or changes in the color of water, filter trouble with water service and unpleasant odor or taste of tap water, and changes in the kinds of fish caught by fishing operators.

Under these circumstances, in December 1982, the Environment Agency established the Environmental Water Quality Standards on nitrogen and phosphorus as a first step to control lake eutrophication. And in July 1985, the Water Pollution Control Law laid down uniform National Effluent Standards on nitrogen and phosphorus in order to control eutrophication at the designated lakes and reservoirs. The number of lakes and reservoirs to which the effluent standards were applied is 1,022 for phosphorus and 45 for nitrogen and phosphorus as of July 15, 1985. Also, sewage treatment plants which discharge treated wastewater to such bodies of water as Lake Biwa, are obliged to remove nitrogen and phosphorus.

Twelve sewage treatment plants are applying nitrogen and phosphorus removal processes at present. And the biological nitrogen removal process using nitrified liquor recirculation with the addition of alum to aeration tanks is mainly employed for the removal of phosphorus and nitrogen. This process can be accomplished by slight modification of a usual aeration tank where an anoxic zone is arranged by mixer installation and a mixed liquor recirculation line is added. The aeration tank of this process including the anoxic tank is designed with a total detention time of 12-16 hours. The

nitrogen removal efficiency of this process is in the range of 60-70% according to a recirculation ratio. However, for phosphorus removal, enhanced biological phosphorus removal will be a desirable alternative due to the increased sludge produced as a result of the addition of alum.

This presentation is about our experiment concerning a new process for nutrient removal on Lake Hamana which is located in Hamamatsu City, east of Nagoya. Lake Hamana is a brackish lake and the activity of inshore fishing is very high. Domestic wastewater originating from the residential area in the catchment basin of Lake Hamana is treated in several sewage treatment plants. The final effluent from both plants must satisfy the strict water quality standards which are 5 mg/l of BOD, 10 mg/l of COD, 3 mg/l of SS, 5 mg/l of T-N, and 1 mg/l of T-P.

All but one of the plants which was newly built, were originally designed and constructed with an extended aeration mode activated sludge process (without a primary sedimentation tank) followed by a coagulation tank, separate stage nitrification denitrification tank, rapid sand filter and activated carbon. However, a simpler treatment system is being sought to cut down the maintenance and operation costs.

To meet these strict standards and cut down the O&M cost, Hamamatsu City and the Public Works Research Institute, Ministry of Construction are conducting joint research for a new process of nutrient removal.

In this study a new biological process for simultaneous removal of nitrogen and phosphorus utilizing oxic-anoxic-oxic process was investigated and it proved to be an effective technology for improving the extended aeration process. The following items were examined in this study.

- o Effects of the ORP value set in this process on removal of nitrogen and phosphorus.
- o Mass balance of alkalinity in this process.
- o Relation between removal of nitrogen and that of phosphorus.
- o Relation between ORP control and DO concentration.
- o Response of this process to load variations.

PLANT DESCRIPTION AND EXPERIMENTAL PROCEDURE

A full-scale wastewater treatment plant, the Hitomigaoka plant, which was designed as an extended aeration process of 16 hours aeration time was modified for this process to study. This plant treats approximately 1000 m^3/d of domestic wastewater. The collection system is a separate sewer system. The effluent of this plant was discharged into Lake Hamana, and the discharge standards are 5 mg/l of BOD, 10 mg/l of COD, 3 mg/l of SS, 5 mg/l of T-N, and 1 mg/l of T-P.

A schematic flow of the plant is shown in Fig. 1. The reactor configuration of the process was composed of three sequential basins. The first oxic basin of 8 hours aeration time, that was completely mixed by two submerged mixers with the diffuser lines, was a simultaneous reaction zone of nitrification and denitrification. In this zone, the nitrification reaction and the denitrification reaction occur simultaneously to maintain the optimum level of the ORP using aeration control coupled with ORP measurement. This

denitrification reaction was assumed as "aerobic denitrification". The second basin of 6 hours hydraulic retention time was post-denitrification zone for nitrate nitrogen leaked out from basin A. The last oxic basin of 2 hours aeration time provided the assimilation of phosphorus and aerobic effluent employing the coarse bubble aeration at a fixed rate. Since in this method nitrification and denitrification take place simultaneously in the first aerobic basin, no recycle of nitrified mixed liquor is necessary unlike the anoxic-oxic process.

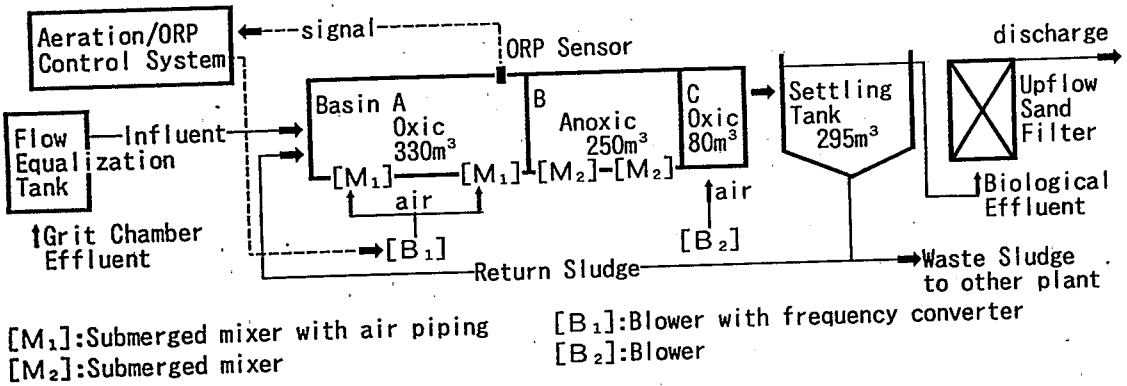


Fig. 1 Flow chart of the Hitomigaoka plant

The experiment was commenced in Oct. 1987, and is going on as of Aug. 1989. This paper analyzed the data obtained from Jan. to Aug. 1988. The operation condition is summarized in Table 1. The ORP values are measured values when a saturated silver-chloride electrode is adopted as a reference electrode. The ORP value in the first oxic basin was studied within the range between 100 mV and 180 mV. The plant was operated for approximately one month for one ORP value. Eventually, 125 mV was identified as the optimum value, and the long-term operation was conducted at this ORP value.

TABLE 1 Operation Conditions

Influent flow	900-1100 m /day
Retention time in the biological reactor (Basin A; 8 hours, Basin B; 6 hours, Basin C; 2 hours)	15-17 hours
Return sludge ratio	60-90%
MLSS	2800-3200 mg/l
BOD-SS loading	0.05-0.08 kg•BOD/kg•SS/day
TN-SS loading	0.01-0.02 kg•TN/kg•SS/day
Sludge retention time	17-25 days
Sludge aeration time	10-16 days
ORP value set in Basin A	100→125→150→180→115→125 mV (changed stepwise)

RESULTS AND DISCUSSION

Nitrogen Removal

Outline of nitrogen removal Fig. 2 and Fig. 3 show variations in the influent and effluent nitrogen concentrations and variations in the nitrogen concentration in Basin A, respectively. ORP values are also indicated in these figures. The T-N concentration of the influent varies between 30 mg/l and 45 mg/l. On the other hand, the T-N concentration of the effluent was kept in the best condition of less than 2 mg/l at 125 mV of ORP. The T-N concentration of the effluent was 5-10 mg/l at ORP other than 125 mV. The most of the effluent T-N concentration at 100 mV and 115 mV consisted of ammonia nitrogen (NH_4-N), while at 150 mV and 180 mV nitrate nitrogen (NO_3-N). Fig. 3 indicates that the dissolved total nitrogen (D-T-N) concentration in Basin A was 2-4 mg/l at the optimum ORP of 125 mV, and that ORP control in Basin A resulted in simultaneous progress of nitrification reaction and denitrification reaction in Basin A. This suggests that nitrogen removal can be achieved using a completely mixed flow single reactor with a hydraulic retention time of 8 hours.

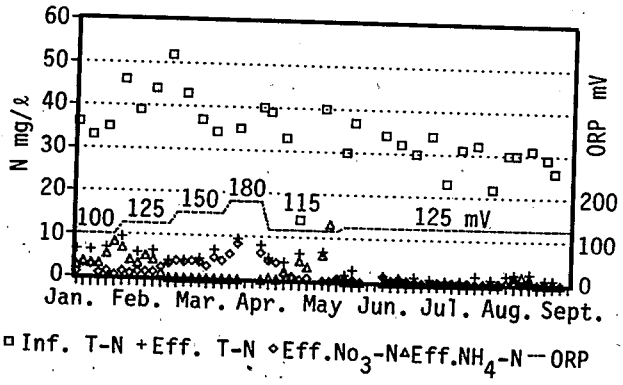


Fig. 2. Variations of Nitrogen concentrations in the influent and the effluent

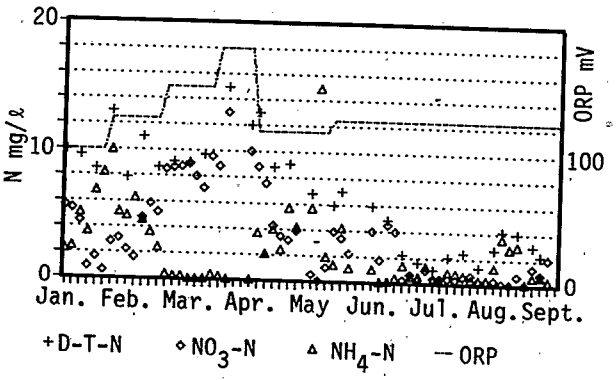


Fig. 3. Variations of nitrogen concentrations in Basin A

The denitrification in Basin A is a reaction taking place in an aerobic condition with the DO concentration of 0.2-0.5 mg/l and the ORP value of +125 mV (approximately +330 mV when the value is converted to the absolute value assuming a hydrogen electrode as the reference electrode). Therefore, it is assumed that this reaction is an "aerobic denitrification" which is different from denitrification in a usual anoxic condition. The evidence concerning aerobic denitrification are reviewed by Robertson and Kuenen (1, 2). The observed DO concentration distribution in the basin has confirmed that no anoxic zone existed in Basin A.

Fig. 4 shows a relationship between the ORP values and the nitrogen concentration in Basin A. It indicates that the nitrogen concentration and form depend upon the ORP set in Basin A. The range where nitrification and denitrification can be achieved simultaneously and where it has the highest efficiency is limited to a narrow zone of 120 to 125 mV, but as shown in Fig. 2 or Fig. 3, ORP control can be easily achieved.

Fig. 5 gives the typical behavior of nitrogen in the process, when the optimum ORP value of 125 mV is adopted. As mentioned before, it has been proved that the nitrogen concentration as low as that of the effluent is already achieved in Basin A. The behavior of orthophosphate (PO_4-P) was also shown in Fig. 5.

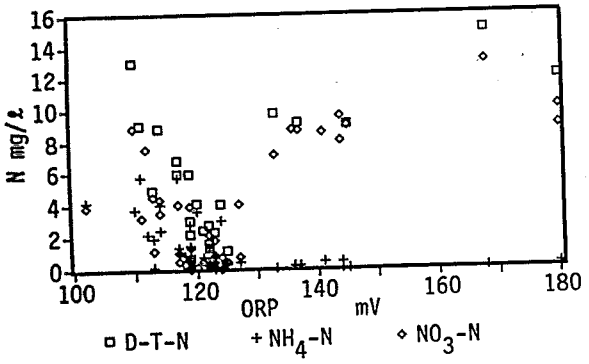


Fig. 4. Relationships between the ORP value and nitrogen concentrations in Basin A

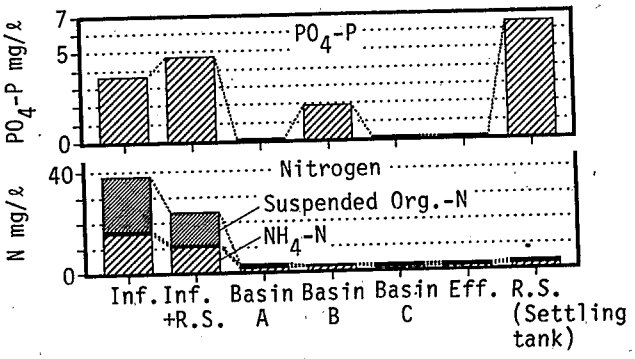


Fig. 5. Nitrogen and phosphorus variations in the process

Alkalinity balance Fig. 6 gives a change in the alkalinity. When the alkalinity in Basin A at the optimum ORP is adopted as a standard, then a ORP value lower than 125 mV produces high alkalinity because of imperfect nitrification. While a ORP value higher than 125 mV yields low alkalinity because of low efficiency of denitrification, although nitrification is perfect. The mass balance of alkalinity in the simultaneous nitrification and denitrification reactor is shown in Fig. 7. On the assumption that the denitrification efficiency, E, is 1, the apparent consumption of alkalinity in this simultaneous reactor is 3.57 times of the concentration of nitrified nitrogen. Then, Fig. 8 illustrates the relationship between the concentration of nitrified nitrogen at the optimum ORP value and consumed alkalinity. Measured values are consistent with the solid line. Although the alkalinity of the influent is not high enough to nitrify perfectly the nitrogen in the influent, this process has been proved effective, so that no shortage in alkalinity takes place.

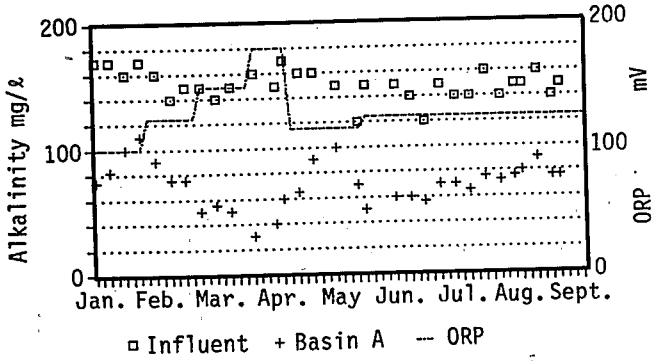


Fig. 6 Alkalinity variations in the influent and Basin A

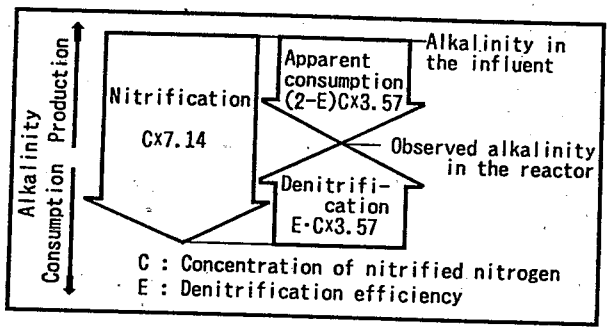


Fig. 7. Alkalinity balance in the simultaneous nitrification-denitrification reactor

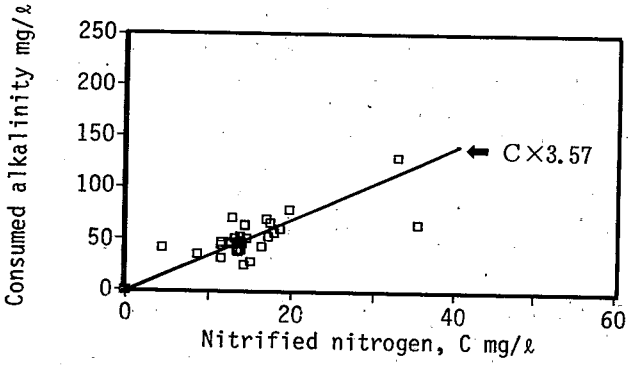
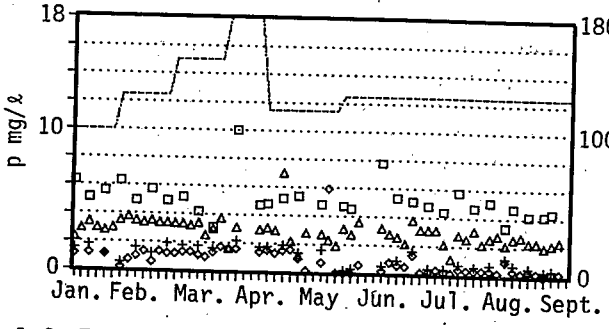


Fig. 8. Relationship between nitrified nitrogen and consumed alkalinity in Basin A

Phosphorus Removal

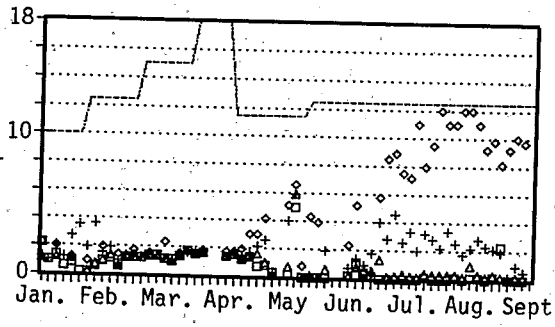
Outline of phosphorus removal Fig. 9 shows variations in the T-P concentration. The T-P concentration of the influent was 4-6 mg/l. The T-P concentration of the effluent was less than 1 mg/l (0.2-0.6 mg/l as PO_4-P concentration) at the ORP values of 100 mV and 125 mV. As for phosphorus, as the same with nitrogen, the T-P concentration of the effluent depends upon the ORP set in Basin A.

Next, the PO_4-P concentrations at different points in this process are shown in Fig. 10. Unlike nitrogen, phosphorus showed different behaviors at each point during two different periods of efficient phosphorus removal and inefficent phosphorus removal: During the poor phosphorus removal period (February to March) the PO_4-P level was nearly constant at 1.4-1.8 mg/l at all sampling points. On the other hand, during the efficient phosphorus removal period at the ORP of 125 mV (May to August) and the periods when the ORP value was 100 mV (January) and 115 mV (April to middle May), the PO_4-P concentrations in Basin A and the effluent were below 0.4 mg/l, and that in Basin B was 2-4 mg/l. The PO_4-P concentration of 4-12 mg/l was also detected in the return sludge line. The sludge blanket in the settling tank was functioning to phosphorus release zone.



□Inf. T-P +Eff. T-P ◇Eff.PO_4-P △Inf.PO_4-P—ORP

Fig. 9. Variations in the phosphorus concentrations



□Basin A +Basin B ◇R.S. △Effluent—ORP

Fig. 10. Variations in the PO_4-P concentration within the process

A comparison between PO_4-P and nitrogen in Fig. 5 suggests a phosphorus release taking place and an extremely low NO_3-N level in the settling tank. It has been known that in the AO process phosphorus release does not take place because of delivery of NO_3-N contained in the return sludge into the anaerobic basin when an anaerobic condition is not met and as a result that phosphorus removal decreases. In our method, too, a similar phenomenon seems to take place in the settling tank.

Relationship between nitrogen removal and phosphorus removal In the previous section, it has been suggested that the NO_3-N concentration relates to the released PO_4-P concentrations in the settling tank. Fig. 11 illustrates the result of an examination of this relationship. This figure indicates that it is required to make the NO_3-N concentration in the settling tank to be approximately below 0.5 mg/l in order to assure the PO_4-P concentration higher than 4 mg/l in the settling tank.

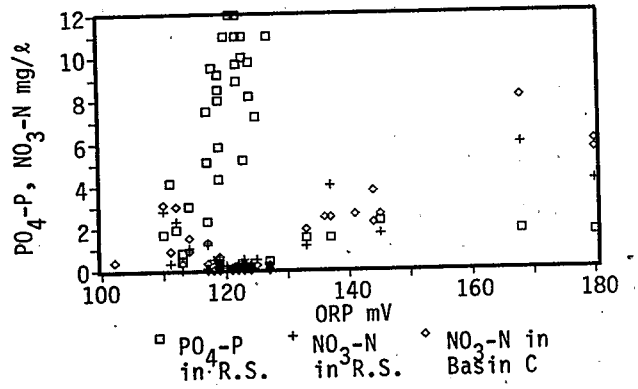


Fig. 11. Relationships between PO_4-P and NO_3-N concentrations

ORP Value and DO Concentration

Fig. 12 shows a relationship between the measured ORP values and the measured DO concentrations. The relationship shown with the solid line is recognized over the entire region of ORP control. In the range from 110 mV to 125 mV of ORP, however, the measured values clusters around one point in spite of a great gap in nitrogen removal, as shown in Fig. 3 and Fig. 4, showing no relationship which is recognized over the entire region of ORP control. Next, the relative frequencies of the ORP values and the measured DO concentrations at 115 mV and 125 mV of the ORP values set are given in Fig. 13 and Fig. 14, respectively. The relative frequency of ORP values show a clear difference at the different ORP values set, while there is nearly no difference in the relative frequency of DO concentrations, which are rather reversed against the ORP values. This proves that the ORP value is more suitable than the DO concentration as a control factor for Basin A in this process.

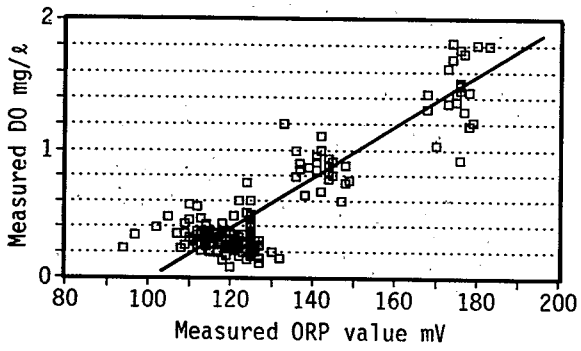


Fig. 12. Relationship between the measured ORP value and the measured DO in Basin A

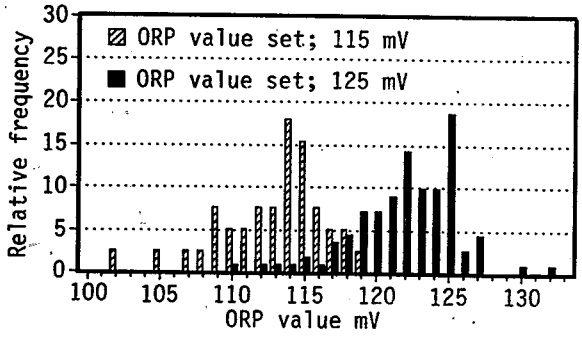


Fig. 13. Relative frequency of the ORP value in Basin A

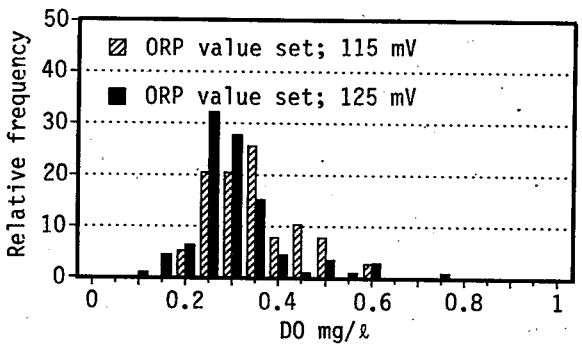


Fig. 14. Relative frequency of DO concentration in Basin A

Response to Diurnal Influent Variations

Fig. 15 shows diurnal variations in the influent and the response of aeration control coupled with ORP measurement to the variations. It proves that the supplied air flow is controlled in response to the influent variations that the ORP value is kept constant. Such control system of supplied air flow can maintain an environment where nitrification and denitrification take place simultaneously. This control system allows the optimum air flow according to a load and resultant energy saving. The air to flow ratio at the optimum ORP value is 8 on the average. Furthermore, this control system seems to be effective not only for nitrogen and phosphorous removal but also as a general control method according to loading variations in an activated sludge process.

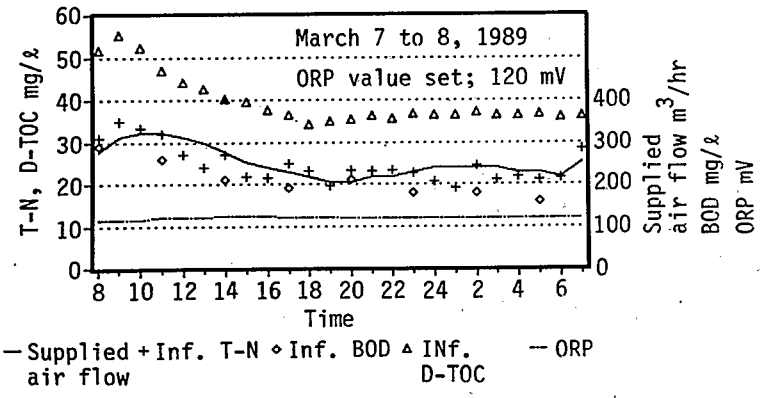


Fig. 15. Variations in the influent load and diurnal variations in the supplied air flow

CONCLUSIONS

The following conclusions were obtained in this study:

1) It has been proved that in this method, the optimum ORP value for simultaneous removal of nitrogen and phosphorus exists, and that a completely mixed flow single reactor, when its ORP value is kept at this proper value, can provide a simultaneous nitrification and denitrification reactor, where simultaneous removal of nitrogen and phosphorus can be achieved.

2) Concretely speaking, it was required to put the NO_3-N concentration in Basin C below 0.5 mg/l by controlling ORP. The necessary anaerobic condition in the sludge blanket of the settling tank was assured in this manner, and phosphorus removal was thus achieved.

3) Examinations of the relationship between measured ORP values and measured DO concentrations showed that ORP values are more suitable than DO concentrations, as a control factor in this method.

4) In the conditions of the optimum ORP value and 16 hours of the retention time in the biological reactor during a high water temperature period, the nitrogen removal rate of 90-95%, the effluent T-N of less than 2 mg/l, the effluent T-P of 0.5-1.0 mg/l and the effluent PO_4-P of 0.5 mg/l were achieved.

5) When nitrogen removal is aimed at, then a completely mixed flow single reactor with an approximately 8 hour retention time and an ORP control system can achieve the effluent T-N of approximately 10 mg/l.

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Development of a High Efficiency Nitrogen Removal System Using Immobilized Nitrifiers in Synthetic Resin

by

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ABSTRACT

This technology is a biological nitrogen removal process which uses immobilized nitrifiers (the IN Process) to reduce the reactor volume required for nitrogen removal. Nitrifiers are immobilized in 2 to 3 mm diameter polyethylene-glycol resin. The immobilized nitrifiers in the resin (Biopellets) are mixed with activated sludge in the nitrification tanks to promote quick nitrification. Since nitrification is completed at shorter sludge retention times using Biopellets, the reactor volume required for denitrification is also reduced. As a result, sufficient nitrogen removal within the conventional reactor-volume (6 - 8 HRTs) has become feasible with this technology. The IN Process was developed through a joint research project between the Hitachi Plant Construction, Co., Ltd. and the Japan Sewage Works Agency (JSWA) under a research contract between the Ministry of Construction and JSWA. This paper introduces selection of sythetic polymer, polymerization methods, nitrifying activity of Biopellets, separation and fluidization of Biopellets, and applicability of Biopellets to nitrogen removal processes.

Keywords; biological nitrogen removal, nitrified liquor recycling activated sludge process, immobilized nitrifiers, synthetic polymer, polyethylene-glycol

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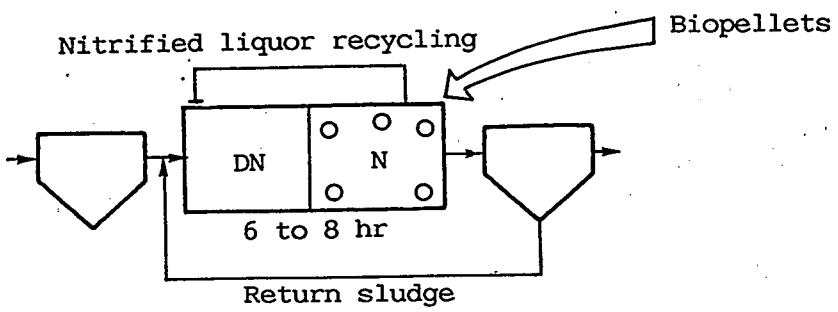
Introduction

Nitrogen removal from municipal wastewater is biologically accomplished using nitrification and denitrification mechanisms. Nitrifiers oxidize ammonium nitrogen to nitrite nitrogen, then to nitrate nitrogen. Denitrifiers reduce nitrate nitrogen to nitrogen gas. Two biological nitrogen removal processes have been applied in full-scale plants in Japan. Those processes are the nitrified liquor recycling activated sludge system, and the endogenous denitrifying activated sludge system. Although these processes result in effective nitrogen removal, the applications have been limited due to the larger reactor volume required. To maintain a sufficient amount of nitrifiers in an activated sludge reactor during periods of low temperature, the reactor volume has to be twice that of a conventional system. A hydraulic retention time (HRT) of 16 hours is usually provided for activated sludge reactors removing 70% of the nitrogen at 13°C.

This technology is a biological nitrogen removal process which uses immobilized nitrifiers (the IN Process) to reduce the reactor volume required for nitrogen removal. Nitrifiers are immobilized in 2 to 3mm diameter polyethylene-glycol resin. The immobilized nitrifiers in the resin (Biopellets) are mixed with activated sludge in the latter half of the aeration tank to promote quick nitrification. Since nitrification is completed at shorter sludge retention times using Biopellets, the reactor volume required for denitrification is also reduced without any carbon sources, excepting that in the influent wastewater. As a result, sufficient nitrogen removal within the conventional reactor-volume (6 - 8 HRTs) has become feasible with this technology.

The differences between the IN Process and the conventional nitrified liquor recycling system are shown in Figure 1. The IN Process was developed through a joint research project between the Hitachi Plant Construction, Co., Ltd. and the Japan Sewage Works Agency (JSWA) under a research contract between the Ministry of Construction and JSWA. Research was conducted on the following topics to develop the IN Process:

IN Process



Conventional nitrified liquor recycling system

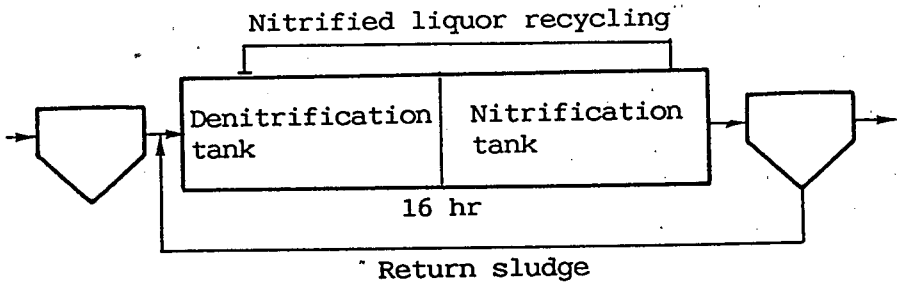


Figure 1. Difference between the IN Process and the conventional nitrified liquor recycling system.

a. The selection of polymer resin suitable for immobilizing nitrifiers,
b. Methods for the immobilization of nitrifiers,
c. The separation and fluidization of immobilized nitrifiers,
d. Secondary effluent nitrification using immobilized nitrifiers, and
e. Biological nitrogen removal using immobilized nitrifiers in activated sludge reactors.

In 1988, a pilot plant study was initiated to evaluate the process performance of the IN Process.

Selection of the polymer resin, and immobilization method

Polymer resin

The immobilization of microorganisms in polymer resin is a widely applied technique in drug manufacturing and food processing. The resins used in those fields are natural polymers (such as agar and carrageenan), and synthetic polymers (such as acryl-amide and epoxy). Generally, the characteristics required of the resin for immobilizing microorganisms are:

a. to be water soluble, and able to gel at ambient temperatures,
b. to have a low toxicity to the immobilized microorganisms during and after gelling,
c. to have a high dispersion coefficient for the substrate to be treated in the resin,
d. to have low biodegradability and to be durable,
e. to be physically strong, and
f. to be inexpensive.

Among the resins used for microorganism immobilization, the natural resins are less toxic and less expensive compared to synthetic ones. However, they are physically weak and easily biodegraded within short time periods. In contrast, synthetic resins have the opposite characteristics.

When applying microorganism immobilization to wastewater treatment, the resin should have good durability to reduce operation and maintenance costs. This must be the most important point when selecting a desirable resin. Therefore, it was decided that synthetic resins would be used as the immobilizing material early in that selection process. The physical characteristics of polyethylene-glycol (PEG), acryl-amide (AAm), epoxy (Epx), urethane, and melamine were examined. The results showed that melamine resin did not gel well at ambient temperatures, and that it was difficult for urethane resin to take on a spherical shape, even though it gelled.

The physical characteristics and nitrifying activity of PEG, AAm, and Epx pellets as immobilizing nitrifiers (Biopellets) are compared in Table 1. Nitrifying activity was measured as the respiration rate per unit volume of Biopellets in aerated ammonium nitrogen solution. The composition of the solution is shown in Table 2. The initial and final respiration rates were measured just after gelling, and after 80 or more days of acclimation with ammonium nitrogen solution (Table 2), repectively. Those physical characteristics did not differ very much. However, the Biological characteristics were significantly different. Both the intitial and final respiration rates of the PEG pellets were the highest among all three resins tested.

Table 1. Physical characteristics and nitrifying activity of pellets immobilizing nitrifiers.

Evaluation item	Polyacryl-amide	Epoxy	Polyethylene-glycol
Compressive strength (kg/cm^2)	2.1	2.1 - 5.2	2.8 - 3.6
Deformation ratio (%)	49	30 - 69	33 - 52
Degree of swelling (%)	101	81 - 107	93 - 113
Durability	Good	Good	Good
Initial respiration rate (%)	1 or less	1.8 - 3.3	5 - 15
Final respiration rate (mgO_2/ℓ-pellets/hr)	500	180 - 500	1100

Note: The initial respiration rate is the ratio of respiration rates before and after immobilization. The final respiration rate is measured after 80, or more, days of acclimation with ammonium nitrogen solution.

Table 2. Composition of ammonium nitrogen solution.

Chemicals	Concentration
$NH_4C\ell$	76.1 mg/ℓ (20 mg-N/ℓ)
$Na_2HPO_4 \cdot 12H_2O$	23.1 mg/ℓ
$NaC\ell$	10.1 mg/ℓ
$KC\ell$	4.7 mg/ℓ
$Cac\ell_2 \cdot 2H_2O$	4.7 mg/ℓ
$MgSO_4 \cdot 7N_2O$	16.7 mg/ℓ
$NaHCO_3$	243.3 mg/ℓ

Note: Sodium bicarbonate is added as a pH buffer.

The change in respiration rate with time for PEG and AAm pellets are shown in Figure 2. The respiration rate of the PEG pellets increased quickly as compared to the AAm pellets. This means that PEG provides a more suitable environment for nitrifying bacteria than AAm. As a result of these physical and biological considerations, PEG was chosen as the resin for immobilizing nitrifiers.

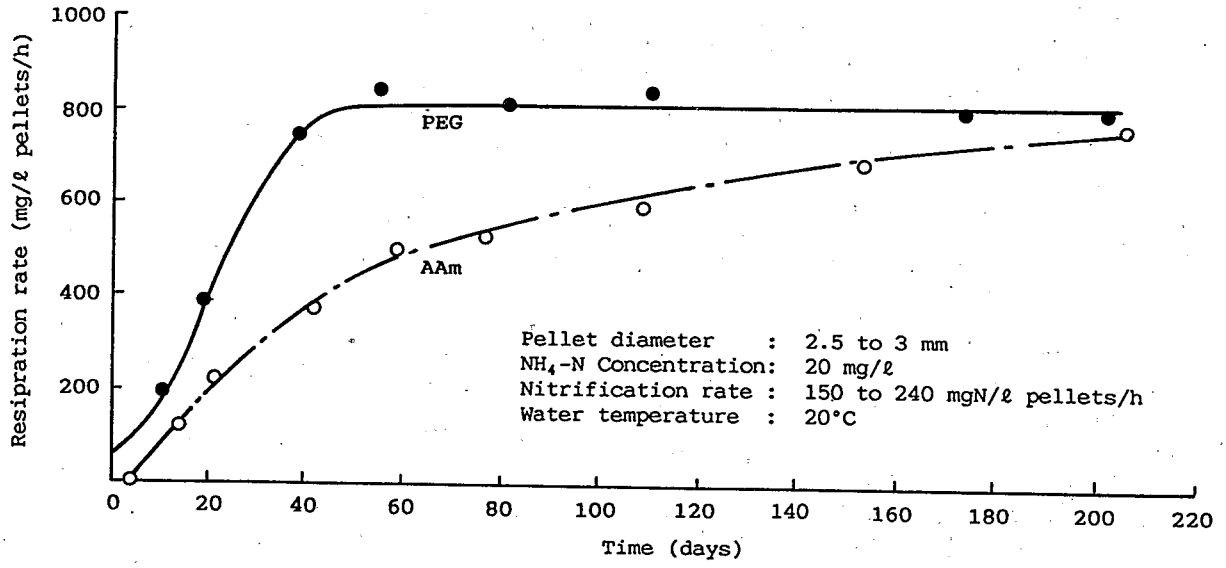


Figure 2. Change of respiration rates with time for PEG and AAm.

Immobilization method

Biopelletes must be produced in large amounts for use in full-scale treatment plants. From this point of view, four immobilization methods were evaluated. The methods tested were:

a. intra-tube polymerization,
b. sheet polymerization,
c. drop polymerization, and
d. suspension polymerization.

The applicability of each method is summarized as follows:

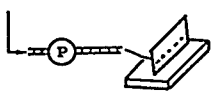
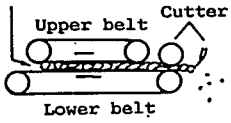
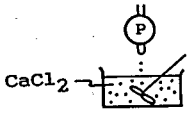
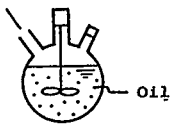
a	b	c	d
Intra-tube polymerization	Sheet polymerization	Drop polymerization	Suspension polymerization
Sludge Immobilizing material/Gelling catalyst 	Sludge Immobilizing material/Gelling catalyst + 	Sludge Immobilizing material/Gelling catalyst + Sodium alginate 	Sludge Immobilizing material/Gelling catalyst 
Cylindrical shape	Cubic shape	Spherical shape	Spherical shape

Figure 3. Immobilizing methods.

a. Intra-tube polymerization (Figure 3 a)

A mixture of activated sludge (nitrifier enriched), PEG prepolymer, and gelling catalysts (accelerator and starter) is pumped into a tube. In the tube gelation of the mixture occurs. The gelled mixture is extruded from the tube, and then cut into small, cylindrically shaped pellets. Several tubes must be prepared for mass production. However, there are some mechanical constraints involved in cutting several bars of gelled mixture at the same time into a uniform shape.

b. Sheet polymerization (Figure 3 b)

The mixture is cast between two sheets which are separated by a fixed distance. The mixture gels there, and then the gelled sheet mixture is cut horizontally and vertically to produce cubic pellets. Since this process seems to be easily automated, the sheet polymerization method should be applicable to mass production.

c. Drop polymerization (Figure 3 c)

A mixture of activated sludge, PEG prepolymer, sodium alginate, and gelling starter is dropped into a calcium chloride solution. The alginate and calcium ions react and produce a thin membrane. The mixture is covered with the membrane and gelation proceeds inside the membrane. An appropriate sized nozzle producing spherical pellets with a set diameter was determined through experiments.

d. Suspension polymerization (Figure 3 d)

A mixture of activated sludge, PEG prepolymer, and gelling catalysts is added to oil. The mixture is dispersed and suspended in the oil, and with agitation forms small drops. In such a suspended condition, gelation proceeds. This process does not appear to be suitable because the drops stick to each other and become large clusters during suspension.

It was considered from the above evaluations that only the methods of sheet polymerization and drop polymerization were applicable to immobilization. The latter method was chosen temporarily for producing Biopellets to be used in experiments on treatment performance. This was because of its mechanical simplicity. The procedures for producing Biopellets by drop polymerization are shown in somewhat more detail in Figure 4.

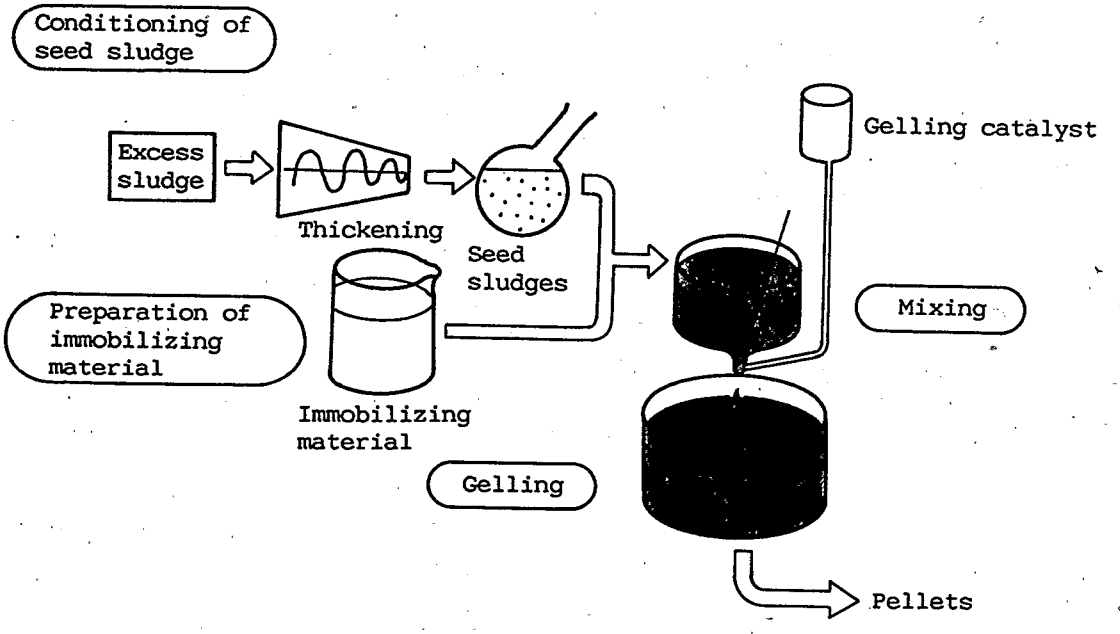


Figure 4. Procedures to produce Biopellets by drop polymerization.

Acclimation of Biopellets

The respiration rates of nitrifiers decrease to 5 to 15% of their initial rates just after immobilization in PEG. Within two months (or less) acclimation with wastewater, the rates increase, and then plateau. To determine the period required for acclimation, the effects of temperature on nitrifying activity, and in which part of a PEG pellet the nitrifiers reside, a laboratory study was conducted.

The effects of activated sludge types on acclimation are shown in Figure 5. Two different types of activated sludge were seeded in PEG pellets. One was sludge from a nitrified liquor recycling activated sludge plant, the other was the same sludge but pre-acclimated for 80 days with ammonium nitrogen solution. The two kinds of pellets were applied to wastewater treatment at 20 and 30°C. As shown in Figure 5, the respiration rate of pellets seeded with acclimated sludge increased rapidly at a temperature of 20°C. By comparison, pellets seeded with the original sludge needed a longer period to reach the plateau of respiration at the same temperature. However, at 30°C the pellets did not require such a long period, and were more rapidly acclimated than nitrifier-enriched pellets at 20°C. The respiration rates after 100 days were 600 to 800 mgO_2/ℓ-pellets/hr, and there were no differences among the three cases. These results mean:

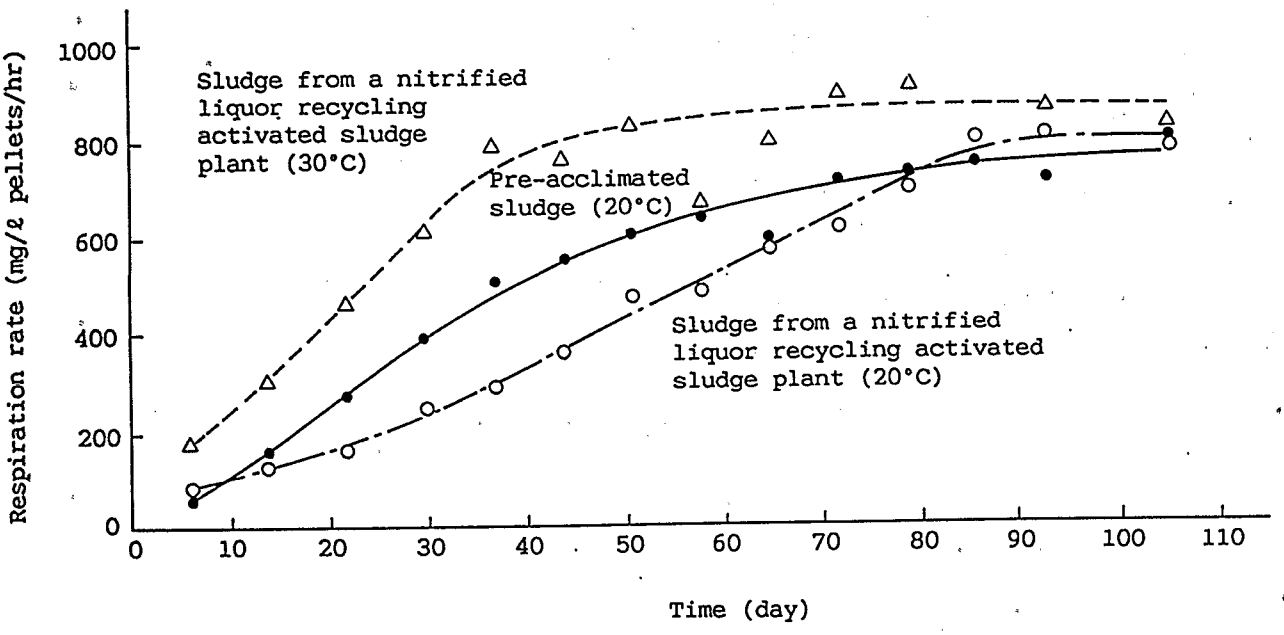


Figure 5. Effects of activated sludge types on acclimation.

a. the acclimation period required depends upon the seed sludges and temperatures,
b. the final nitrifying activity is not related to the seed sludge or temperature,
c. at higher temperatures, the sludge type seeded with pellets does not have a significant effect on starting up reactors, and
d. at lower temperatures, the nitrifier-enriched sludge should be used as seed sludge for immediate starting up.

Microorganism distribution in a pellet was observed using an electron microscope. The observed results over time are shown in Figure 6. The growth of microorganisms in the pellets proceeded as follows:

a. Microorganisms exit dispersively in a pellet just after immobilization.
b. Microorganisms form colonies in a pellet during increasing nitrifying activity. The growth of colonies is significantly high near the pellet's surface (until the 60th day).
c. The density of the colonies increases near the pellet's surface, where the biological environment is better than that inside, colonies inside the pellets tend to fade out (from the 60th to the 200th day).
d. The microbial layer near and/or on the surface grows to a thickness of 60 μm, and there seem to be no mircoorganisms inside the pellets (after the 200th day).

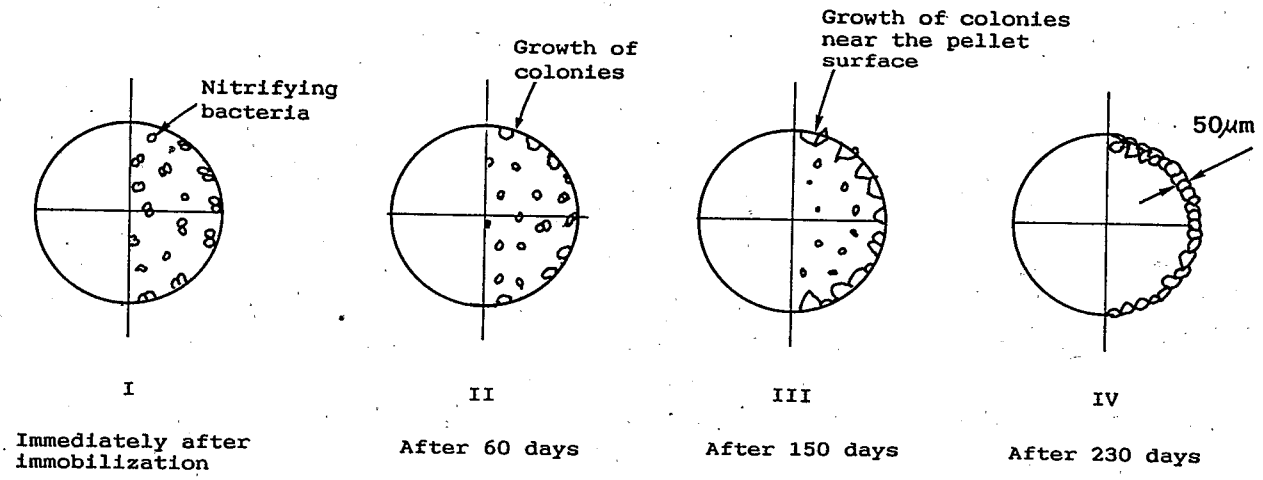


Figure 6. Results of electron microscopic observation over time.

From the above observations, it was shown that nitrifying bacteria exist only near and/or on the pellet's surface after sufficient acclimation, and that the inside of a pellet works merely as a support medium. This indicates that smaller pellets produce better nitrification performance when only a small amount of pellet material is available. The relationships between pellet diameters and final respiration rates are shown in Table 3. Smaller pellets showed higher respiration rates per unit pellet volume. However, respiration rates per unit pellet surface area were approximately 400 mgO_2/m^2/hour, and almost the same among pellets with different diameters. This result clearly shows that the pellet diameter must be as small as possible. The separability of pellets from activated sludge will determine the suitable diameter for Biopellets.

Table 3. Relationships between pellet diameters and final respiration rates.

Diameter	1.00 - 1.41	1.41 - 1.68	1.68 - 2.00	2.00 - 2.38	2.38 - 2.83
Respiration rate per unit pellet volume (mgO_2/ℓ/hr)	1,730	1,600	1,480	1,260	1,110
Number of pellets per unit pellet volume (1000/ℓ)	1,105	523	307	179	109
Surface area per unit pellet volume (m^2/ℓ)	5.00	3.90	3.26	2.73	2.31
Respiration rate per unit pellet surface area (mgO_2/m^2/hr)	346	411	454	462	481

Separation and fluidization of Biopellets

Separation of pellets

Biopellets have to be separated from treated wastewater or activated sludge to maintain them in a reactor. Two separation methods, wedge-wire separation (Figure 7) and gravity separation, were tested. It was thought before the experiments that pellets would be separated from activated sludge by a slight difference in the setting velocities of the pellets and the sludge. But the apparent density of the pellets decreased from 1.06 to 1.03, or less, due to the attached sludge so that stable pellet separation was difficult. On the other hand, wedge-wire worked well. It was found that wedge-wire should be installed at an angle of 30° to 60° from the water's surface.

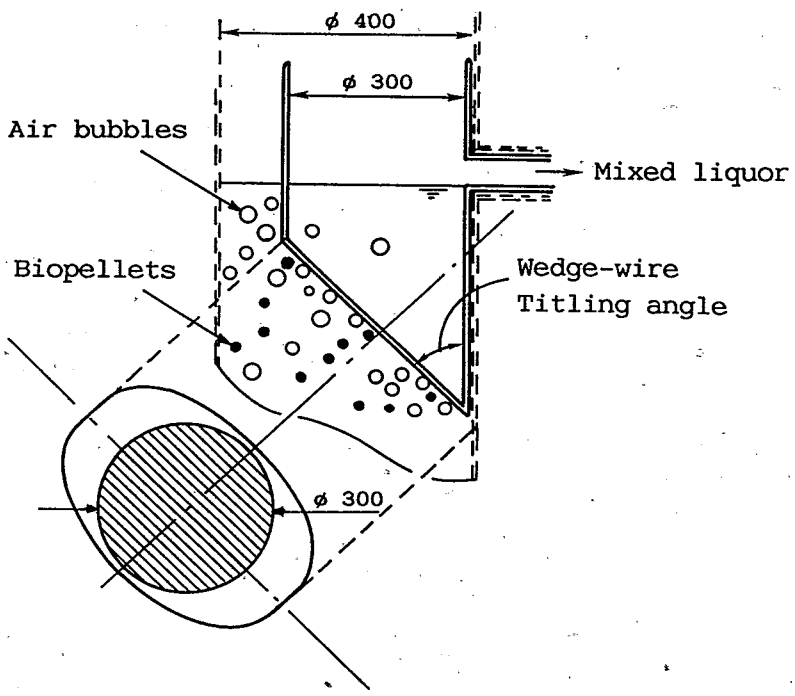


Figure 7. Biopellet separation with wedge-wire.

To determine the conditions for wedge-wire separation of pellets, several studies were carried out using a 200 ℓ reactor. The effects of wedge-wire opening on the separation of pelltes with different diameters are shown in Table 4. Each separation test was continued for one hour. Both pellet loss ratios and wedge-wire clogging ratios were measured after one hour operation. For the combination of a wedge-wire with a 1mm opening and pellets with 2.00 to 2.83mm diameters, pellet loss was less than 0.1%, and the clogging ratio was less than 5%. When a wedge-wire with a 1.5mm openning and pellets with diameters of more than 2.83mm were used, the ratios were the same as with the previous combination. These ratios, 0.1% loss and 5% clogging, did not cause any operational problems. This result indicates that the pellet diameter should be at least two times as large as the wedge-wire opening.

Table 4. Effects of wedge-wire opening on separation of pellets with different diameters.

Wedge-wire opening (mm)	0.5			1.0			1.5		
Pellet diamter (mm)	1.68-2.00	2.00-2.83	2.83 <	1.68-2.00	2.00-2.83	2.83 <	1.68-2.00	2.00-2.83	2.83 <
Pellet loss ratio (%)	0	0	0	0.4	0.1	0	9.1	1.4	<0.1
Wedge-wire clogging ratio (%)	0	0	0	20	5	0	40	20	<5

Note: The pellet loss ratio is the ratio of the pellet loss measured after one hour operation to the pellet volume in the nitrification tank. The wedge-wire clogging ratio is the ratio of the clogged area to the entire wedge wire area.

Operating conditons:
Wedge-wire separation rate = 1.25$m^3/m^2/min$.
Air flow rate = 0.08m^3/m^3 - tank volume/min.

The effects of separation rates and air flow rates on pressure loss at the wedge-wire are shown in Table 5. Approximately 2m^3/hour of air flow to unit volume of the aeration tank with 3-hour HRT is required for biological treatment with nitrification. This is based on 25 mg/ℓ influent ammonium nitrogen and 15% oxygen transfer efficiency. The air flow rate is equivalent to an aeration intensity of 1.28m^3/m^2-floor area/min at a water depth of 4m. Unit area of wedge-wire separated 1.28m^3/min of mixed liquor from a aeration tank with 10% pellets by volume at aeration intensities between 0.08 and 0.16$m^3/m^2/min$. When the pellet volume was increased to 20%, the pressure loss increased, and the separating capacity of unit area wedge-wire decreased considerably. This means that a larger wedge-wire area is required for effective separation of mixed liquor at higher pellet volume ratios. The mixed liquor flow to be separated is equal to the sum of the inflow rate and recycling rate when a nitrified liquor recycling system is employed. A IN Process treating 10,000m^3/day of wastewater at a 10% pellet volume ratio and 300% recycling rate requires a wedge-wire separating system with a 20m^2 area.

Table 5. Effects of separation rates and air flow rates on pressure loss at wedge-wire

Air flow rate	Inflow rate			
	0.96 $m^3/m^2/min.$	1.28 $m^3/m^2/min.$	1.60 $m^3/m^2/min.$	1.92 $m^3/m^2/min.$
No aeration		x/x		
0.08 $m^3/m^2/min.$	o/Δ	o/x	x/x	x/x
0.16 $m^3/m^2/min.$	o/Δ	o/x	o/x	
0.24 $m^3/m^2/min.$	o/o	o/Δ		x/x
0.40 $m^3/m^2/min.$		o/Δ	o/Δ	x/x

Note: Pellet volume ratio of 10% / Pellet volume ratio of 20%

o: No operation problem

Δ: Pellets continue attaching and detaching to the wedge-wire randomly. This causes a sudden increase/release of pressure loss. Due to the fluctuation, the separation is not be maintained at a constant rate.

x: Great pressure loss. Continuous operation is not possible.

Operating conditions:
Pellet diamter > 1.68 mm.
Wedge-wire opening = 0.5 mm

Fluidization of pellets

The fluidization test on the pellets was conducted under whole-floor aeration in a pilot-scale nitrification tank (effective volume = $4.2m^3$). The pellet volume ratio to the nitrification tank was set at 15 and 20%. Air supplied at $1.25m^3$/hour to unit cubic meter of tank volume fluidized pellets uniformly under both conditions. The air requirement for fluidizing pellets uniformly increased to $2.5m^3/m^3$-tank/hour when the pellet volume ratio was 30%. The boilogical air requirement for both BOD removal from and nitrification of ordinary municipal wastewater is about $2m^3/m^3$/hour at 3 hour HRT in a whole-floor aeration tank. This means that fluidization will be achieved with air supply to meet the biological need when the pellet volume ratio is less than 20%.

Application of Biopellets to secondary effluent nitrification

Temperature largely influences the nitrification of wastewater. In ordinary activated sludge plants, nitrification proceeds to some extent in summer, but not in winter. Since partial nitrification often causes effluent BOD_5 to increase, nitrification must be carefully monitored even in plants targeting only organic removal. Considering such problems in this section, the application of Biopellets to secondary effluent nitirification is discussed.

A pilot study was carried out using a 200 L reactor nitrifying secondary effluent. The pilot plant was as shown in Figure 8. The treated secondary effluent was pumped from a secondary settling tank in an activated sludge plant to the pilot plant. Nitrified wastewater in the reactor containing Biopellets was separated by wedge-wire.

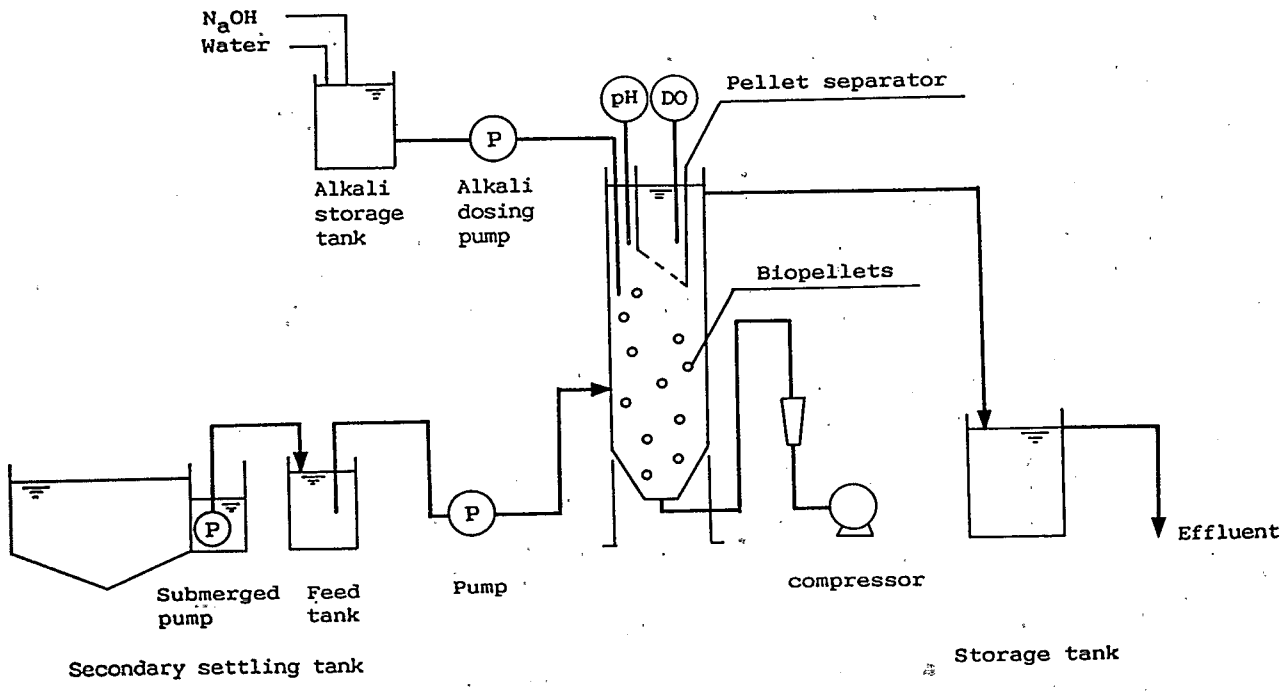


Figure 8. Schematic flow diagram of the pilot plant nitrifying secondary effluent.

The operating conditions were as shown in Table 6. The pellet volume ratio was 20% at 2 hour HRT, and 30% at 1 hour HRT. Water temperatures were between 10 and 12°C. Influent ammonium nitrogen concentrations were 20 to 30mg/ℓ. The experimental results are shown in Figure 9. Complete nitrification was achieved when the ammonia volumetric loadings were 0.2 to 0.4kg-N/m^3/day at 2 hour HRT. Increasing ammonia loadings to a range between 0.5 and 0.7kg-N/m^3/day caused nitrification to lightly deteriorate. However, effluent ammonium nitrogen concentrations were still less than 2mg/ℓ, and more than 95% of the ammonium nitrogen was removed. When the dissolved oxygen concentrations in the reactor decreased, the nitrifying efficiency dropped. This might have been caused by oxygen deficiency.

Table 6. Operating conditions for secondary effluent nitrification.

	Run 1	Run 2
Water temperature (°C)	10 - 12	10 -12
Pellet volume ratio(%)	20	30
Retention time (hour)	2	1
Ammonium nitrogen volumetric loading (kg/m^3/day)	0.2 - 0.4	0.5 - 0.7

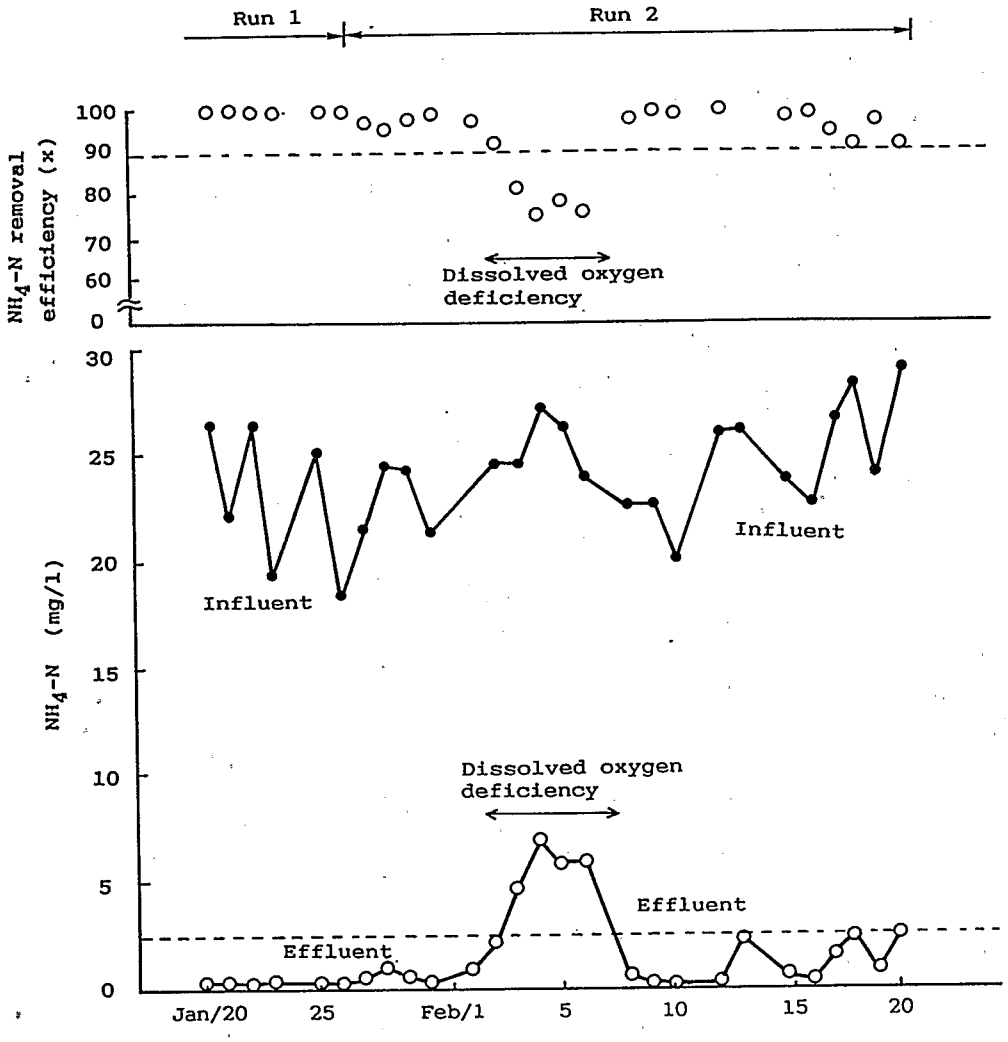


Figure 9. Results of secondary effluent nitrification.

The nitrifying rates of the pellets used in the pilot experiment were measured in a small batch reactor. The pellet volume ratio was increased by steps, from 10%, to 20%, to 40%. The results of the batch study are as shown in Figure 10. The nitrifying rates were 147mg-N/L-pellets/hour at 10% pellet volume ratio, 140mg-N/L/hour at 20%, and 83mg-N/L/hour at 40%. Increasing the pellet volume ratio means decreasing the nitrifying rate per unit volume of pellets. Oxygen transfer to liquid may have controlled the nitrifying rates of the pellets. In addition, a larger area of wedge-wire was required for pellet separation at higher pellet volume ratios. Therefore, the feasible pellet volume ratio should be limited to 20%.

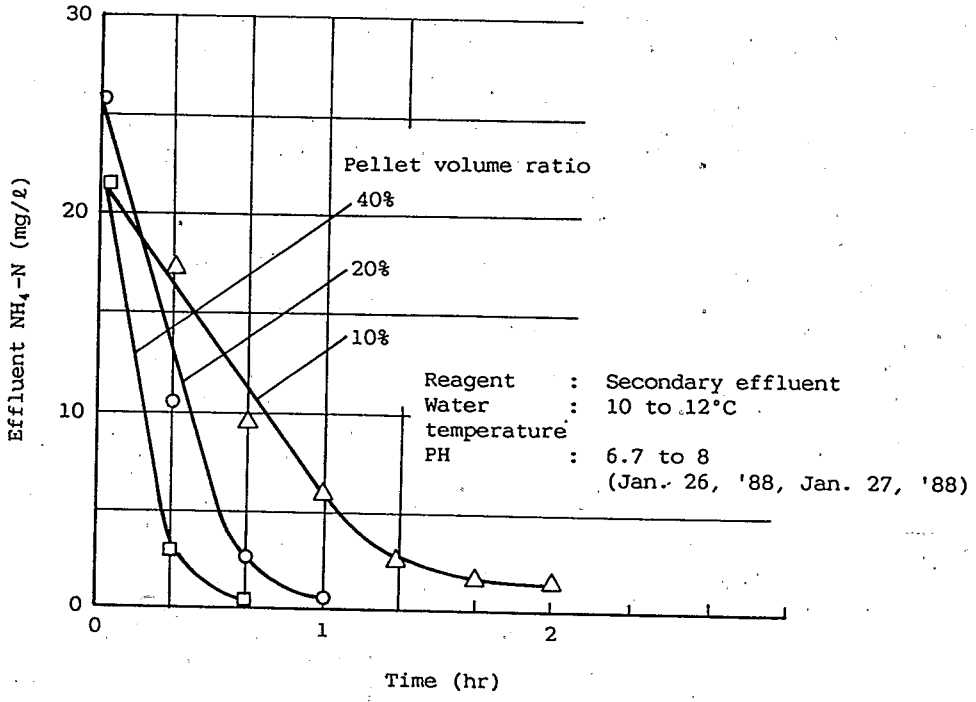


Figure 10. Results of batch experiment.

The detention time required for complete nitrification at 20% pellet volume ratio was only one hour, based on the batch study. However, it is difficult to maintain an oxygen concetration in a full-scale reactor as high as that in a small batch reactor. Moreover, maintaining high oxygen concentrations is not ecomonical in terms of the power requirement. The practical HRT of a reactor nitrifying ordinary secondary effluent should be approximately two hours at 20% pellet volume ratio. Under these conditions, complete nitrification was achieved in the pilot study.

Application of Biopellets to nitrogen removal (the IN Process)

Preliminary study

The application of Biopellets to biological nitrogen removal processes enables a significant reduction in the required reactor volume. To confirm the treatment performance and to determine operation conditions, a preliminary study was conducted. Biopellets were applied to the nitrified liquor recycling activated sludge process.

The reactor used was as shown in Figure 11. The reactor consisted of a denitrification tank, a nitrification tank, and a final settling tank. In the nitrification tank, wastewater was mixed with return sludge, and the nitrified liquor recycled. Denitrification occurred in the tank, and, simultaneously, some of the organics were consumed as carbon sources for denitrification. The mixed liquor then flowed into the nitrification tank, containing Biopellets. Organic removal by the activated sludge, and nitrification by both the activated sludge and Biopellets proceeded at this time. The wedge-wire installed near the top of the tank separated nitrified liquor from the pellets. The separated nitrified liquor was partly recycled to the denitrification tank, and the rest was introduced into a final settling tank.

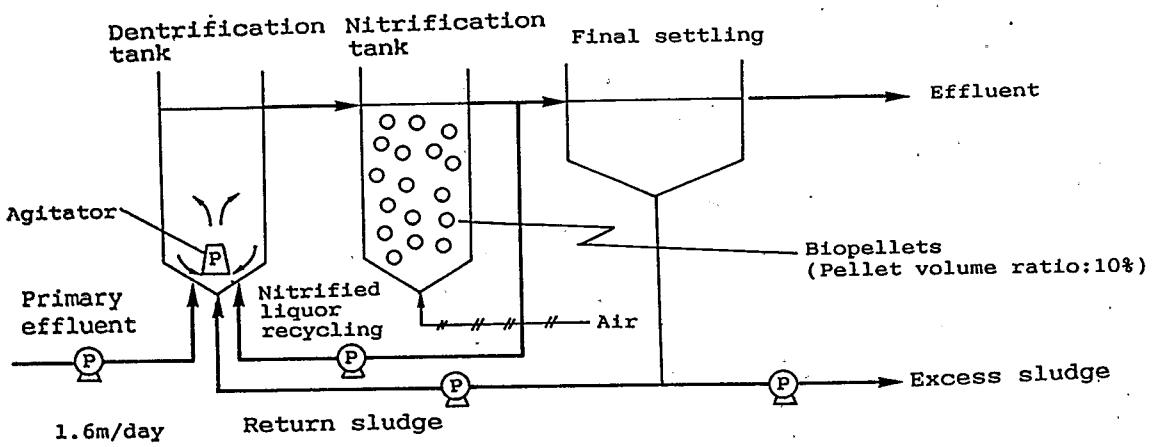


Figure 11. Flow diagram of IN Process for preliminary study.

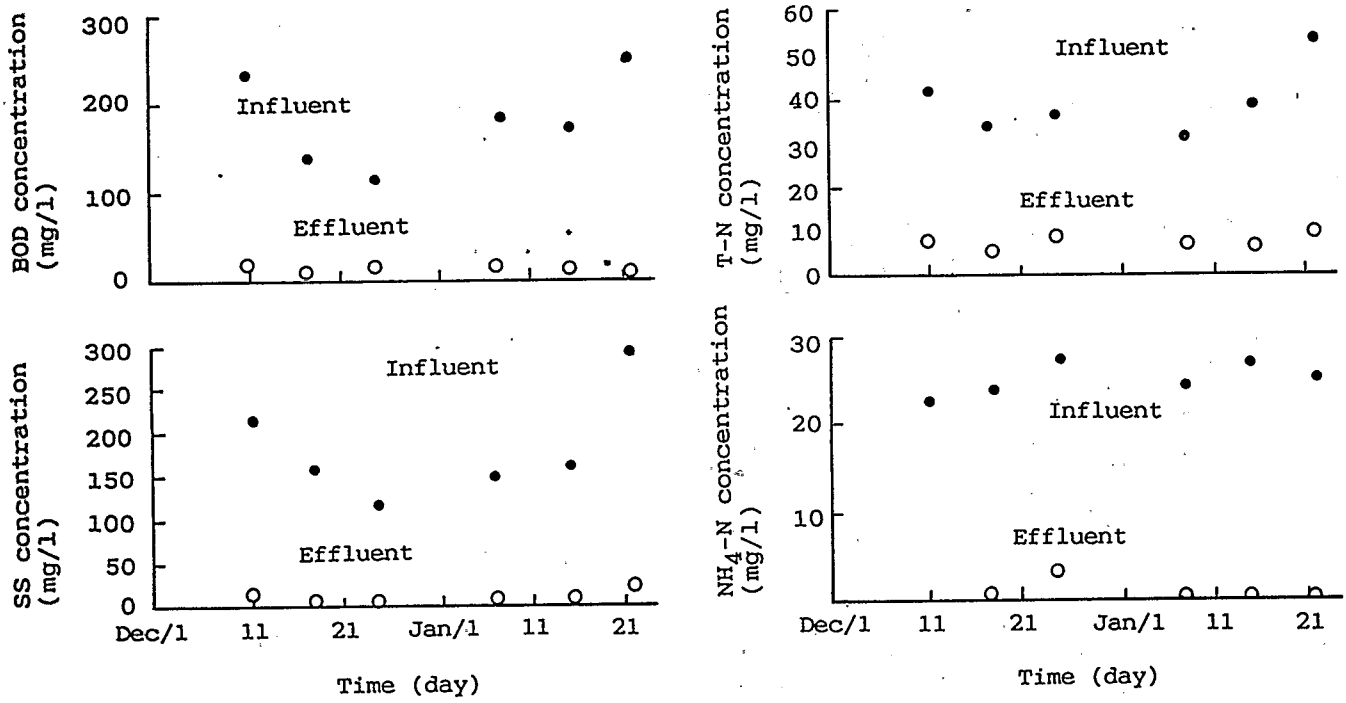
The operating conditions are shown in Table 7. The treated wastewater was the primary effluent of the municipal wastewater. The retention times in the denitrification, nitrification, and final settling tanks were 3 hours, 3 hours, and 1.8 hours, respectively. The reactor volume requirement was almost the same as that for the conventional activated sludge process. The Sludge return ratio was 100%, and the nitrified recycling ratio was 300%. Twenty liters of pellets were added to the nitrification tank. The pellet volume ratio was equivalent to 10%.

Table 7. Operating conditions for preliminary study of the IN Process.

Nitrification tank	Effective volume: 200ℓ, Retention time: 3 hours
Denitrification tank	Effective volume: 200ℓ, Retention time: 3 hours
Final settling tank	Effective volume: 120ℓ, Retention time: 1.8 hours
Pellet volume ratio	10% of the effective volume of the nitrification tank
Sludge return ratio	100%
MLSS	2,000 to 3,000 mg/ℓ
Nitrified recycling ratio	300%
BOD volumetric loading	1.46 kg/m³/day to the denitrification tank
T-N volumetric loading	0.31 kg/m³/day to the nitrification tank

Note: Operating condition during winter: Water temperature is 15 to 16°C.

The experimental results are shown in Figure 12. The influent BOD_5, suspended solids, and total nitrogen were 120 to 300mg/ℓ, 120 to 300mg/ℓ and 30 to 60mg/ℓ, respectively. The average volumetric loadings during the experimental period were 1.46kg/m^3/day for BOD_5, and 0.31kg/m^3/day for total nitrogen. During the experiment, both the BOD and suspended solids were removed to 20mg/ℓ, or less, and the total nitrogen was reduced to less than 10mg/ℓ. The ammonium nitrogen concentrations remaining in the effluent were less than 2mg/ℓ. These results demonstrated that Biopellets work well, even in activated sludge, and that the IN Process is feasible as a biological nitrogen removal process.



Average BOD load : 1.46 kg/m^3/day (denitrification tank)
Average T-N load : 0.31 kg/m^3/day (nitrification tank)
MLSS : 2,000 to 3,000 mg/ℓ
Water temperature: 15 to 16°C

Figure 12. Results of preliminary study on IN Process.

The sharing of ammonium nitrogen between the Biopellets and the activated sludge in summer and in winter is shown in Figure 13. The nitrification rates of the Biopellets and activated sludge were nearly equal in summer. The rate of the activated sludge decreased in winter, while the rate of the Biopellets increased and supplemented activated sludge. The nitrification rate of the Biopellets in winter was 100mg-N/L-pellets/hour. This rate is not much lower than the figure obtained in secondary effluent nitrification (140 mg-N/L-pellets/hour).

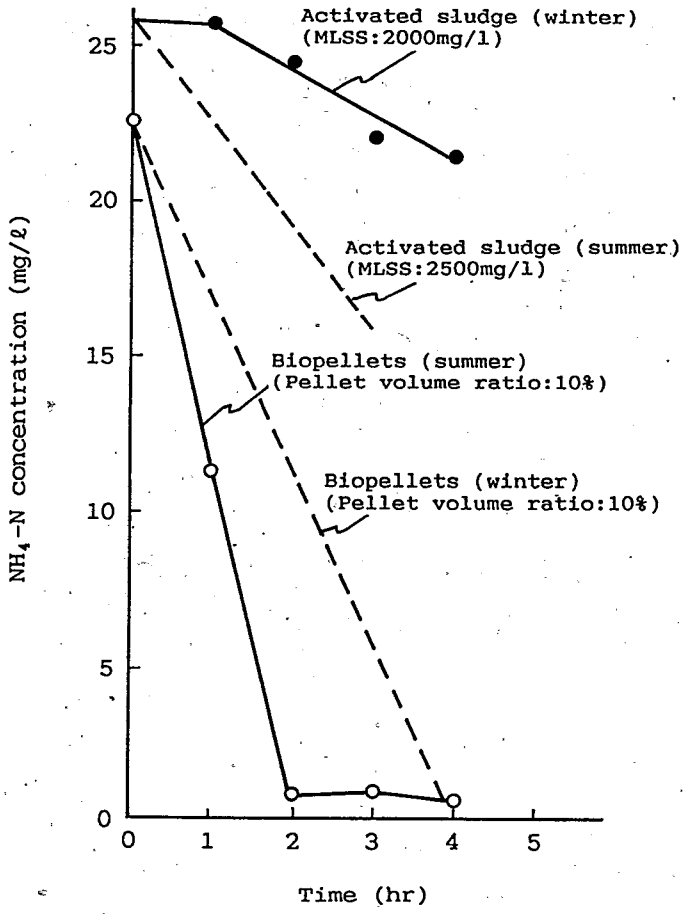


Figure 13. Sharing of ammonium nitrogen between Biopellets and activated sludge.

Pilot plant study

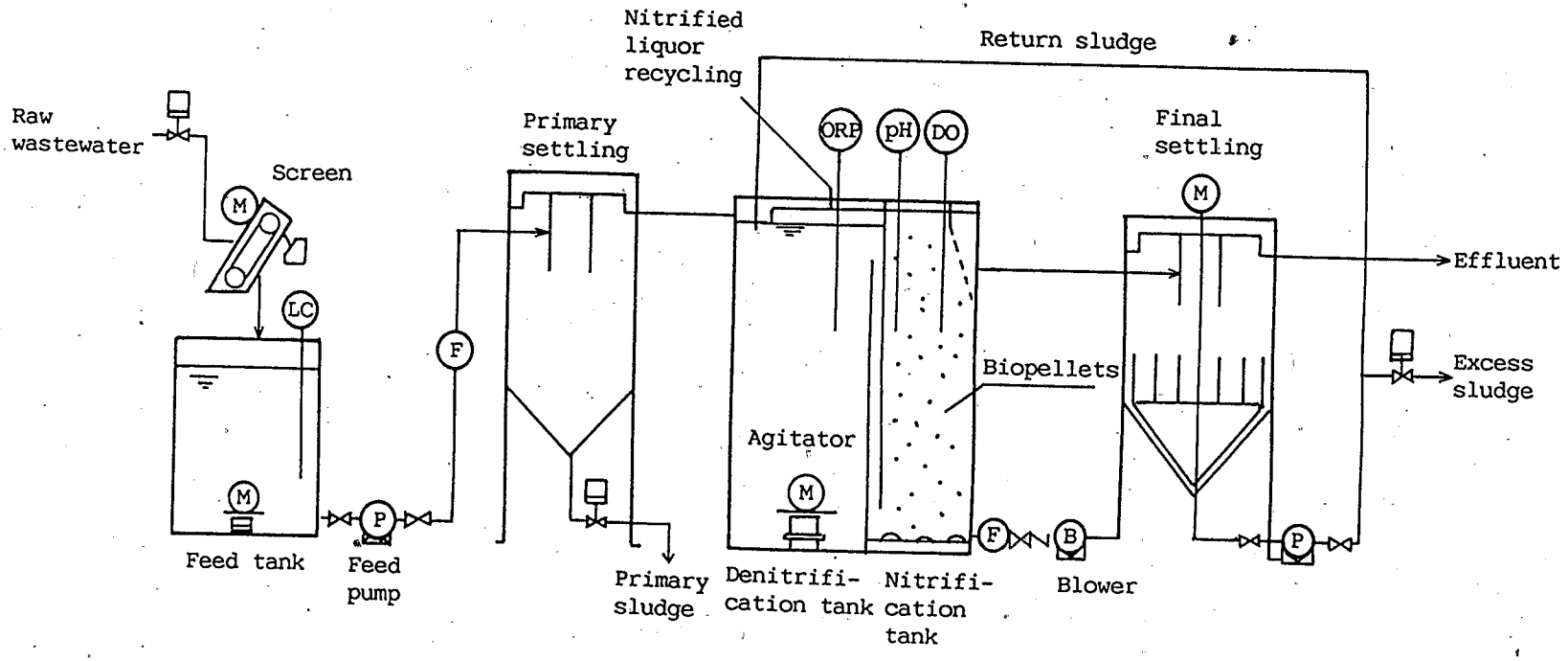
A pilot plant study was initiated in November 1988. The plant capacity is 50m^3/day. The primary objectives of this study were:

a. to confirm the treatment performance throughout a year,
b. to clarify operational characteristics, such as fluidization and separation of pellets and nitrified liquor recycling, and
c. to esatablish design and operation/maintenance procedures.

The pilot plant experiment will last to the end of this year. This section introduces interim results obtained last winter.

A flow diagram and the specifications of the pilot plant are shown in Figure 14. Raw wastewater is pumped up from a municipal wastewater treatment plant to the experimental yard in the same plant. After temporary storage in a feed tank, the raw wastewater is settled in a primary sedimentation tank. The primary effluent is treated by a scaled-down pilot IN Process. The process configuration is alomst the same as that for the preliminary study, but excludes the nitrified recycling system. The nitrification tank is aerated, making the contents overflow into the denitrification tank by air-lift effect, to thus recycle the nitrified liquid. The volume of Biopellets added into the nitrification tank is 10% of the tank's effective volume.

The operating conditions last winter are shown in Table 8. The influent BOD_5 and total nitrogen concentrations were 100 to 200mg/ℓ and 20 to 30mg/ℓ, respectively. The detention times were 4.8 hours for denitrification, and 3.2 hours for nitrification at the constant inflow rate of 31.5m^3/day. The air flow rate, sludge return ratio, and nitirified liquor recycling ratio were controlled at 140L-air/min, 40%, and 300%, respectively. The MLSS concentration fluctuated in the range between 2,000 and 3,000mg/ℓ, and the water temperatures were between 10 and 12°C.



Flow rate	:	48 m^3/day x 2 trains
Primary settling tank	:	1.2 mW x 2.4 mℓ x 5.0 mH x 6 m^3 x 2 tanks
Reactor	:	1.2 mW x 3.4 mℓ x 3.8 mH x 12 m^3 x 2 tanks
Final settling tank	:	2.2 mø x 3.8 mH x 6 m^3 x 2 tanks

Figure 14. Flow diagram and specifications of the pilot plant for IN Process.

Table 8. Operating conditions for pilot plant study of the IN Process.

Primary settling tank	Water surface area: 2.9 m^2, Overflow loading : 10.9m^3/m^2/day
Nitrification tank	Effective volume: 4.2 m^3, Retention time : 3.2 hours
Denitrification tank	Effective volume: 6.3 m^3, Retention time : 4.8 hours
Final settling tank	Water surface area: 3.8 m^2, Overflow loading : 8.3 m^3/m^2/day
Pellet volume ratio	7.5% of the effective volume of the nitrification tank.
Nitrified recycling ratio	40%
MLSS	2,000 to 3,000 mg/ℓ
Nitrified recycling ratio	300%
BOD volumetric loading	0.56 kg/m^3/day to the denitrification tank
T-N volumetric loading	0.22 kg/m^3/day to the nitrification tank

The treatment performance is shown in Figure 15. The average volumetric loadings during the winter were 0.56kg-BOD/m^3/day to the denitrification tank, and 0.22kg-TN/m^3/day to the nitrification tank. Even at lower temperatures, the pilot plant indicated excellent performance. The BOD_5 and total nitrogen concentrations were removed to less than 20mg/ℓ, and to less than 10mg/ℓ, respectively. The ammoniun nitrogen that remained in the final effluent was 2mg/ℓ or less. This means that nitrification was almost completed. The transparency of the final effluent was not as good as that from the conventional system. More consideration should be given to this point.

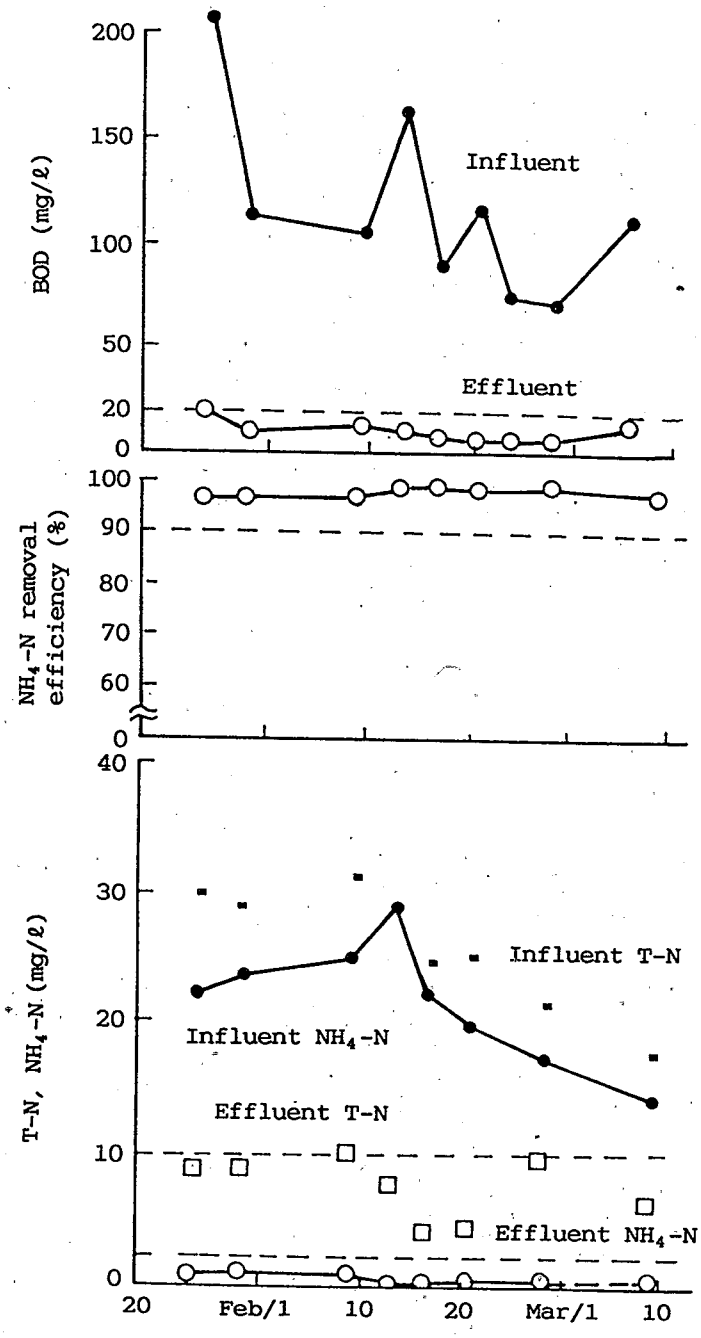


Figure 15. Results of pilot study on IN Process in Winter.

Initial and O/M costs for the In Process

The initial cost and the operation/maintenace costs for the IN Process were estimated by the interim results, and were compared with those for the conventional biological nitrogen removal process. Conditions for the estimation are shown in Table 9, and the results are shown in Table 10. Since the reactor volume requirement was reduced to half that for the conventional system, the initial cost was reduced greatly. The estimate did not include land cost. This means that the saving should be greater in larger cities, where land costs are extremely expensive. Furthermore, one feature on the IN Process is that it can save land area, which is the biggest advantage for existing treatment plants. In fact, there is no space for retrofitting advance treatment in most of the treatment plants in large cities, nor surround them.

Table 9. Conditions for estimation of initial cost, and operation/maintenance costs.

	IN Process	Conventional biological nitrogen removal process
Flow rate	10,000 m^3/day	
Influent BOD	120 mg/ℓ	
Effluent BOD	20 mg/ℓ or less	
Influent NH_4-N	25 mg/ℓ	
Effluent T-N	10 mg/ℓ or less	
Lowest water temperature	13°C	
Nitrification rate	96 mgN/ℓ-pellets/hour	0.67 mgN/gSS/hour
Denitrification rate	1.04 mgN/gSS/hour	0.71 mgN/gSS/hour
MLSS	3,000 mg/ℓ	3,000 mg/ℓ

Table 10. Estimates of initial and O/M costs for the IN Process.

	IN Process	Conventional biological nitrogen removal process
Reactor volume (m^3)	3,330	6,670
Retention time (hour)	8	16
Nitrification tank (m^3)	1,330	4,170
Retention time (hour)	3.2	10
Denitrification tank (m^3)	2,000	2,500
Retention time (hour)	4.8	6
Pellet volume (m^3)	100 (7.5%)	-
Initial cost (yen)	540,000,000	790,000,000
Civil works (yen)	250,000,000	480,000,000
Installations (yen)	290,000,000	310,000,000
Operation and maintenance cost (yen/year)	35,000,000	46,000,000
Power requirement (yen/year)	15,000,000	31,000,000
Pellet cost (yen/year)	6,000,000	-
Repairs (yen/year)	14,000,000	15,000,000
Treatment unit cost (yen/m^3)	10	13

Note: It is supposed that the air-lift recycling system of nitrified liquor is employed for the IN Process, while a pump recycling system for the conventional biological nitrogen removal process. Owing to the higher aeration strength of the nitrification tank for the IN process (2.02 m^3/m^3 tank/hour, versus 0.50 m^3/m^3 tank/hour for the conventional process), a greater gas hold up in the nitrification tank is achieved. This simplifies recycling of nitrified liquor by the air-lift method. As a result, the IN process requires no driving power for recycling. The estimation is made in terms of the reactor only. Therefore, the above costs do not include those for the primary and final settling tanks.

The estimate of O/M costs did not include the cost for personnel. Therefore, the costs for agitating denitrification tanks, aerating nitrification tanks, and maintaining mechanical and electrical devices are dominant in the estimate. Additionally in the IN Process, the cost for pellets had to be included because Biopellets would be damaged after a long-term operation. The duration of Biopellts was assumed to be 5 years. The results indicated that the O/M costs for the IN Process is less expensive, (by 3yen/m^3-treated) than that for the conventional process. This is due mainly to the air lift recycling system of nitrified liquor, which was applied for the IN Process. If pumps are employed for recycling, the costs will be increase by 4yen/m^3, and will be slightly more expensive compared to the conventional system. From this viewpoint, the establishment of the air-lift recycling system is important to promote the application of the IN Process as well as prolonging the duration of the Biopellets.

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CURRENT STATUS AND FUTURE DIRECTION OF ORGANIC TOXICS CONTROL IN THE WATER ENVIRONMENT - DISCHARGE REGULATIONS AND STREAM STANDARDS

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I. INTRODUCTION

Annually, more than 550 million pounds of toxic substances are discharged to surface waters in the United States. Some of these substances, such as pesticides, have been developed specifically for their toxicological effects and therefore may have serious environmental consequences. Others, such as surface active agents, are formulated for different purposes but can ultimately be toxic to aquatic life and result in unforeseen environmental impacts. Still others have certain toxicological properties, such as persistence and low solubility, which can lead to long-term degradation of aquatic ecosystems.

Until recently, many problems associated with the discharge of such pollutants had gone largely unrecognized. While we know, for example, that many organic compounds have a high capacity for accumulating in edible fish and shellfish tissue, we often do not know whether the consumption of a contaminated tissue is hazardous to human health. The biological consequences of certain properties of organic chemicals add a new dimension for potential risk to public health and the environment and create a profound challenge for environmental managers at all levels of government.

Over the past decade, technological advances have resulted in the development of advanced monitoring devices, such as gas chromatography and mass spectrography (GC/MS), that produce wastewater scans which indicate the presence of thousands of organic chemicals. Frequently, these chemical mixtures cannot be fully identified. For those chemicals that are identified, there is often a lack of data on their toxicological properties. Determining the environmental effects of releases of these compounds is yet another obstacle. Potential difficulties in determining the effects of releases include: the great number of toxic substances that may be discharged to receiving waters; the difficulty of obtaining positive identification of these chemicals; the changes in the toxic effects of a chemical due to reactions with the matrix of constituents in which it exists, and; the inability to predict the effects of exposure to an individual organic toxicant or to combinations of such chemicals.

To overcome some of these problems, the Environmental Protection Agency (EPA) and many State governments have initiated the use of aquatic toxicity tests and various human health assessment techniques to complement chemical analyses of effluent and receiving water samples. It is EPA's policy to investigate the bioaccumulative and toxicological properties of municipal and industrial effluents and implement effective controls through an integrated strategy using both biological and chemical assessment methods.

II. LEGISLATIVE FRAMEWORK

o Congress passed the Federal Water Pollution Control Act (FWPCA) amendments in 1972. This marked a turning point in water pollution control efforts in the United States. Unlike previous attempts to regulate water pollution through water quality standards which were to be developed and enforced at the State level, the Federal government assumed the dominant role in directing and defining water pollution control efforts across the country.

- The emphasis changed from strictly water quality-based standards (biological endpoint-based) to technology-based standards (wastewater engineering-based). The introduction of technology-based effluent guidelines shifted the burden to the discharger to meet the discharge limits achievable through already existing, defined control technologies.

- The NPDES permit program was established limiting the levels of pollutants allowable in effluent discharges. The program is under EPAs' authority and applies to all direct dischargers. To date, NPDES permits have been issued to over 65,000 dischargers. This includes more than 50,000 industrial dischargers and over 15,000 municipal dischargers which release wastewater directly to surface waters. There are as many as an additional 100,000 "indirect" industrial dischargers releasing wastewater to the sewer systems of publicly owned treatment works (POTWs). These indirect dischargers are regulated by the POTW through the national pretreatment program.

- Enforcement authority was succinctly defined by making illegal all discharges to U.S. waters without a permit, and allowed for civil and criminal penalties in cases of noncompliance.

- A national policy on toxics was established which prohibited the discharge of toxic pollutants in toxic amounts.

o In 1977 Congress passed the Clean Water Act (CWA) to refine some of the requirements originally set out in the 1972 law in order to emphasize the control of toxic pollutants. Among the most important modifications made to the 1972 requirements were the revisions to the Effluent Guidelines and Pretreatment programs:

- The National Effluent Limitation Guidelines program sets forth uniform national standards for direct dischargers, developed on an industry-by-industry basis utilizing existing control technologies. In an effort to stimulate advances in toxic control mechanisms the CWA established a list of 126 "priority pollutants", produced by 21 categories of industries, which must be specifically addressed in the development of effluent limitations guidelines. This list includes 111 organic compounds, 13 heavy metals, asbestos and cyanide.

- The National Pretreatment Program regulates the introduction of pollutants from nondomestic sources into sewer systems of POTWs.

o In 1987, amendments were made to the CWA, known as the Water Quality Act (WQA), in order to further encourage control of and research on toxic pollutants and their effects.

- Because Congress recognized that progress on chemical specific controls was slow, a dual approach was developed to facilitate water quality improvement. Section 303(c)(2)(A) of the WQA requires that States develop water quality criteria based on biological monitoring or assessment methods.

- Under section 304(l) of the WQA each State must prepare lists of its waterways which do not meet water quality standards. The States must develop individual control strategies (NPDES permits), to reduce the amount of toxic substances being released by point sources.

- Increased Congressional concern for toxics is also reflected in amendments to section 405 which requires EPA to identify toxic substances in sewage sludge and develop numerical limits for maximum allowable concentrations. A total of 16 organic chemicals are to be controlled via numeric limitations.

III. CURRENT STATUS OF ORGANIC TOXICS CONTROL EFFORTS

Technology-Based Standards

The CWA requires the development of national standards for treatment technologies for direct dischargers, which are implemented through the NPDES permit program. Technology-based effluent limits define a minimum level of control and are imposed at the point of discharge or "end-of-pipe". Organic toxics are controlled through Best Available Technology Economically Achievable (BAT).

The most significant effluent limitation guideline to be developed to date for the control of organic toxics is the organic chemicals, plastics, and synthetic fibers (OCPSF) effluent limitations guideline issued in November of 1988. Typically, organic chemical manufacturing facilities discharge more than 60 toxic substances in their wastestreams. With at least 400 indirect dischargers and 300 direct dischargers nationally, organic chemicals facilities affect a large number of receiving waters.

Water Quality-Based Standards

Existing technology-based permits require limits on a specific set of known pollutants which are treated to a specific level by BAT. This approach may not address complex effluents where combinations of chemicals may have additive effects, or effluents containing toxicants not found on EPA's priority pollutant list. Water quality-based permitting is used in cases where limits more stringent than technology-based are necessary to protect the designated use (such as a fishable river usage) of the receiving waters.

Water Quality Criteria and Advisories. Using experience gained by States and EPA Regions in toxic impact analysis, EPA is developing new water quality criteria for high priority toxicants as well as advisories for toxicants for which more limited information is

available. For example, EPA is issuing a criterion for organotin compounds because these compounds have been found to cause toxic impacts in harbors where organotin paints are used to protect ship hulls. Other criteria are being developed for pesticides found to pass through typical treatment systems.

Whole Effluent Toxicity Testing is often a preliminary step in the process of setting water quality-based permit limits for toxic pollutants. Effluent toxicity is measured by exposing test organisms to varying concentrations of an effluent/water mix to measure the aggregate toxic effect on living organisms.

Toxicity Reduction Evaluation (TRE). A TRE is initiated if the whole effluent toxicity test indicates an unacceptable level of toxicity. A TRE is a site-specific study conducted in a step-wise process to narrow the search for effective control measures for effluent toxicity. These studies are designed to identify the causative agents of effluent toxicity, isolate the sources of effluent toxicity, evaluate the effectiveness of toxicity control options, and then confirm the reduction in effluent toxicity.

Whole effluent toxicity tests have been used in detecting toxicity in the effluents of POTWs across the nation. TREs have then aided in identifying the causative agents of this effluent toxicity. For example, in a number of cases organophosphate pesticides have been identified as the cause of toxicity in POTW effluents. One such compound, diazinon, has been repeatedly identified as either the primary toxicant or a substantial contributor to the effluent toxicity.

Due to the unexpected presence of toxic concentrations of diazinon in diverse effluents, the National Effluent Toxicity Assessment Center (NETAC) has conducted a national survey of 28 POTW effluents for toxicity caused by diazinon. The results of this study indicate that diazinon is present in 61 percent of the POTWs tested, though toxic levels of diazinon tended to be restricted to the southern region of the nation. The relatively widespread occurrence of this compound in POTW effluents would likely have gone undetected were it not for the use of toxicity testing and TREs.

Pretreatment

Pretreatment programs require that industrial dischargers to POTWs reduce the level of pollutants in their process wastewaters. Approximately 1500 of the 15,000 POTWs in the U.S. are required to implement pretreatment programs as part of their NPDES permit requirements. There are three categories of controls utilized under the pretreatment program:

Prohibited Discharge Standards include both general and specific prohibitions. General prohibitions include prohibitions against pollutant discharges from any nondomestic user which cause pass through or interference. This includes prohibitions against effluents which may cause: 1) fire or explosion hazards; 2) corrosive structural damage; 3) interference due to flow obstruction; 4)

interference due to flow rate or concentration; and 5) interference due to heat.

Categorical Pretreatment Standards are incorporated into local permit requirements to limit potentially hazardous discharges. Categorical standards are developed from effluent limitations guidelines and identify what level of a specific pollutant will cause interference with treatment systems and what amount of the pollutant is removed by typical treatment processes.

Local limits are utilized by the POTW because categorical standards are minimum requirements which apply to broad classes of industries and do not necessarily address all industrial discharge problems that might occur at a given POTW. To prevent site-specific problems, each POTW must assess all of its industrial discharges and employ sound technical procedures to develop defensible local limits that will adequately protect the POTW, its personnel and the environment.

In developing local limits a POTW must conduct an industrial waste survey to locate and identify all industrial users as well as the character and volume of their pollutant discharges. The POTW then determines which pollutants have the potential for interference, pass through or sludge contamination. Then for each pollutant of concern the POTW must determine the maximum loading which can be accepted by the treatment facility without problems. The POTW must then implement a system of local permit limits to ensure that the maximum loadings will not be exceeded.

V. Future Directions

As environmental monitoring and detection capabilities become more sophisticated, the pervasiveness and consequences of toxic pollution become increasingly evident. In response to our expanding awareness of the impacts associated with toxic contamination, control efforts have been refined to greater levels of efficacy and comprehensiveness. In order to continue to progress in this area EPA is both building on established programs and developing new methodologies.

Technology-based effluent limitations guidelines are continuously being developed for additional industries. Two effluent guidelines currently under development will have considerable impact on organic toxics discharges. These guidelines are for the pharmaceutical and pesticide industries. Both of these industries discharge wide varieties of organic toxics and few have met BAT control standards.

Water quality-based limits will continue to be a major focus of pollution control. Point sources and other sources of pollution such as storm water discharges will be given limits to protect local water quality.

Sewage sludge is produced at the rate of 7.6 million tons per year in the U.S., and this rate is expected to double by the year 2000. Sludge may contain unacceptable levels of toxic substances,

particularly heavy metals and organics. When properly treated, sludge can be beneficially used as a soil conditioner and fertilizer. EPA encourages beneficial reuse of sludge whenever environmentally feasible. To this end, EPA is developing regulations and procedures for identifying, managing and reducing the amount of toxic pollutants in municipal sludge based on sludge use and disposal practices.

Bioconcentratable Contaminants are compounds that have the ability, due to their chemical properties, to accumulate in fish and shellfish tissues in concentrations hazardous to human consumers. Water quality criteria and effluent guidelines designed to prevent acute and short-term toxicity to aquatic life may not be sufficient to protect human health. EPA is exploring methods that will allow environmental managers to identify and quantitate bioconcentratable pollutants in effluents, and develop acceptable ambient criteria, wasteload allocations and effluent limitations for these pollutants.

Sediment Assessment is utilized to evaluate toxic contaminants in the bottom sediments of lakes, rivers and coastal waters for potential adverse impacts to biota and water quality. Historically, water quality-based standards have not been developed to directly protect sediment quality or the beneficial uses associated with uncontaminated sediments. Sediment assessment is being analyzed as a means to target dischargers and develop more stringent effluent limits.

Pollution Prevention reflects EPA's policy to reduce the amount and toxicity of pollutants being discharged, either directly or indirectly through POTWs, to waters of the U.S. Increasingly stringent controls and better detection methods encourages dischargers to begin in-plant recycling programs or to substitute less toxic compounds in their production processes to reduce the levels or toxicity of their discharges.

PERFORMANCE OF FINE PORE AERATION SYSTEMS IN PROCESS WATER

by

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INTRODUCTION

The supply of oxygen for aeration is the single largest energy consumer at activated sludge wastewater treatment plants, representing 50-90 percent of total plant energy requirements (1). Replacement of less efficient aeration systems with fine pore aeration devices can save up to 50 percent in aeration energy costs and has resulted in typical simple pay back periods of two to six years. As a result of these very impressive cost savings, more than 1,300 municipal and industrial wastewater treatment facilities in the United States and Canada now use fine pore aeration. In 1985, the U.S. Environmental Protection Agency (EPA) funded a Cooperative Research Agreement with the American Society of Civil Engineers (ASCE) to evaluate the existing data base on fine pore diffused aeration systems in both clean and process waters, conduct field studies at a number of municipal wastewater treatment facilities employing fine pore aeration, and prepare a comprehensive manual addressing the design, operation, maintenance, control, and economics of fine pore aeration systems.

A Design Manual on Fine Pore Aeration was produced from this effort by U.S. EPA (2). This presentation summarizes a portion of this manual which discusses the performance of fine pore aeration systems under process conditions.

The term "fine bubble" diffused aeration is elusive and difficult to define. The term "fine pore" was adopted to more nearly reflect the porous characteristics of the high efficiency diffusers marketed today. Typically, fine pore diffusers will produce a headloss due to surface tension in clean water of greater than about 5 cm water gauge (w.g.). For the purposes of this paper, fine pore diffusers are defined as including the following devices:

- Porous ceramic plates, discs, domes, and tubes
- Rigid porous plastic plates, discs, and tubes
- Nonrigid porous plastic tubes
- Perforated membrane tubes and discs

FACTORS AFFECTING PERFORMANCE

The performance of diffused aeration systems under process conditions is affected by a myriad of factors, some of the more important of which are:

- Changes in performance due to fouling, aging, fatigue, etc.
- Wastewater characteristics
- Loading conditions
- Process type and flow regime
- Basin geometry and diffuser placement
- Diffuser performance and characteristics
- Mixed liquor DO control and air supply flexibility
- Mechanical integrity of the system
- Quality of preventive maintenance

Manual of Practice FD-13 (3) is a good general reference on the importance of the above factors. To minimize life-cycle costs of an aeration system, all of these factors must be considered during design and some must be controlled during operation.

The areas of greatest concern in process water oxygen transfer performance of fine pore aeration systems are changes in performance due to fouling, aging, etc.; wastewater characteristics; process type and flow regime; and loading conditions. In a given case, any combination of these these factors can have a significant effect on the alpha (α) profile of a system, DO control, and changes in aeration system performance with time due to diffuser fouling. Some of these factors are discussed in greater detail later in this paper.

CALCULATION OF PROCESS PERFORMANCE

A substantial data base exists on the performance of fine pore diffusers in clean water. In designing aeration systems to operate under process conditions, clean water data are corrected to account for the influences of process water characteristics, temperature, and pressure. The corrections are made using the following equations:

$$OTF_f = K_La_{20}\ \alpha F\ \Theta^{(T-20)}\ (\tau\ \beta\ \Omega\ C^*_{\infty 20} - C)\ V \qquad (1)$$

where,

OTR_f	=	oxygen transfer rate under process conditions, mass/time
K_La	=	apparent volumetric mass transfer coefficient in clean water, 1/time
K_La_{20}	=	K_La at 20°C, 1/time
V	=	volume of water, length^3
$C^*_{\infty 20}$	=	steady-state dissolved oxygen (DO) saturation concentration attained at infinite time for a given diffuser at 20°C and 1 atm, mass/length^3
C	=	process water DO concentration, mass/length^3
α	=	(process water K_La of new diffuser)/(clean water K_La of new diffuser)
β	=	(process water C^*_∞/(clean water C^*_∞)
$\Theta^{(T-20)}$	=	K_La_T/K_La_{20}
T	=	temperature of process water being treated
Ω	=	$(P_b + \text{gamma}\ d_e - P_{VT})/(P_S + \text{gamma}\ d_e - P_{VT})$
Ω approx.	=	P_b/P_S (for basin depths <6.1 m or for elevations <600 m)
P_b	=	atmospheric pressure at aeration system location, mass/(length x time^2)
gamma	=	weight density of water at the test water temperature, force/length
P_S	=	standard atmospheric pressure (1.0 atm of air at 100% relative humidity), mass/(length x time^2)
P_{VT}	=	saturated vapor pressure of water at temperature T, mass/(length x time^2)
d_e	=	effective depth at infinite time, length
τ	=	$C^*_{\infty T}/C^*_{\infty 20} = C^*_{ST}/C^*_{S20}$
C^*_{ST}	=	tabular value of DO surface saturation concentration at temperature T and standard total pressure (1 atm at 100% relative humidity), mass/length^3
F	=	(process water K_La of a diffuser after a given time in service)/(K_La of a new diffuser in the same process water)

Since standard oxygen transfer rate in clean water, SOTR, is:

$$SOTR = K_La_{20}\ C^*_{\infty 20}\ V \qquad (2)$$

Equations 1 and 2 may be combined to calculate the process water oxygen transfer rate, OTR_f:

$$OTR_f = \alpha F(SOTR)\Theta^{T-20} (\tau \beta \Omega C^*_{\infty 20} - C)/C^*_{\infty 20} \quad (3)$$

This equation can be rearranged as follows:

$$\alpha F(SOTR) = (OTR_f C^*_{\infty 20} \Theta^{20-T})/(\tau \beta \Omega C^*_{\infty 20} - C) \quad (4)$$

The term $\alpha F(SOTR)$ is often used to express oxygen transfer rate under field conditions corrected to standard temperature and pressure and a driving force of $C^*_{\infty 20}$ (i.e., $C = 0$).

Further, since standard oxygen transfer efficiency, SOTE, may be defined as:

$$SOTE = SOTR/W_{O2} \quad (5)$$

where,

W_{O2} = mass flow rate of oxygen, mass/time

Equation 3 may also be used to describe process water oxygen transfer efficiency, OTE_f, by direct substitution of $\alpha F(SOTE)$, oxygen transfer efficiency under field conditions corrected to standard temperature and pressure, and a driving force of $C^*_{\infty 20}$, for $\alpha F(SOTR)$.

Although employing clean water SOTR values to estimate oxygen transfer rates in process water is conceptually straightforward, the estimate of OTR_f is subject to considerable doubt because of the uncertainties contained in α and F. These uncertainties are magnified when the process water application is based on a basin geometry and process temperature that differ from those of the clean water test.

Table 1 is a guide for applying Equation 3 and indicates the source of information for the parameters needed to estimate OTR_f.

DIFFUSER FOULING

All fine pore diffusers are susceptible to waterside buildup of biofilms and/or deposition of inorganic precipitates that can alter the operating characteristics of the diffusers. Porous diffuser media may also be susceptible to air-side clogging of pores due to particles in the supply air. Air diffusion media may also suffer changes in properties, and the integrity of the installation may change with time. These changes, which include alteration of diffuser media properties over time and leaks in the air piping or around diffuser gaskets, may be caused by type of materials selected, poor equipment design, improper installation, or inadequate inspection and maintenance.

The rate of diffuser fouling has historically been gauged by the change in backpressure while in service. Since significant levels of fouling can take place with little or no increase in backpressure but with substantial reductions in OTE, this provided a crude and qualitative measure at best.

Better methods of measuring the degree of diffuser fouling and the effectiveness of diffuser cleaning became available in the early 1980s. These methods include dynamic wet pressure (DWP), bubble release vacuum (BRV), the ratio of one to the other, and chemical and microbiological analysis (4). The practice of employing pilot diffusers that could be removed from the basin and individually analyzed also came into use during this period. DWP is defined as the

pressure differential (headloss) in cm w.g. across the diffusion element when operating in a submerged condition at some specified airflow rate. BRV is a measure of the negative pressure in cm w.g. required to form and release bubbles in tap water from a localized point on the surface of a thoroughly-wetted porous element. Normally, BRV denotes the average of the point BRV values for a specific diffuser.

Concurrently, better methods were being developed to measure the oxygen transfer performance of operating aeration systems. These methods permitted better appraisal of the effects of fouling and facilitated improved preventive maintenance scheduling. These methods include inert gas tracers, off-gas analysis, a nonsteady-state technique that uses hydrogen peroxide, and DO and respiration rate profiles (5-8). Off-gas analysis equipment has been effectively used to evaluate the adverse effects of fouling on both full-scale diffused air systems and individual diffusers (5). For the U.S. EPA/ASCE Cooperative Agreement on fine pore aeration (2), off-gas analyses were used in all field studies. All investigators conducting field studies on this project were trained in the use of this equipment, all off-gas analyzers were of the same design, and calculation methodology was standardized.

TYPES OF FOULING

Fouling can be classified as one of two general types, Type I or Type II. The two types have distinct characteristics and can occur alone or in combination with variable dominance from treatment plant to treatment plant and within the same treatment plant from time to time.

Type I Fouling

Type I fouling is characterized by clogging of the diffuser pores, either on the air side by airborne particulates or on the liquid side by precipitates such as metal hydroxides and carbonates. In the process of fouling, the areas of the diffusers with the highest local air flux foul more rapidly. This serves to reduce the flux in the high flow areas and increase it in low flow areas. The combined effect is to improve uniformity of air distribution. As fouling progresses, the BRV coefficient of variation decreases (9).

Eventually, the accumulation of foulant in the pores reduces the pore size and DWP rises correspondingly. The increase in DWP can exceed the capabilities of the air supply system, and process air delivery may fall short of requirements. Also, the reduced effective pore diameters produce smaller bubbles such that OTE does not decline and can actually increase slightly. A schematic representation of Type I fouling is illustrated in Figure 1. An idealized plot of how OTE and DWP change with time is shown in Figure 2.

Type II Fouling

Type II fouling is characterized by the formation and accretion of a biofilm layer on the surface of the diffuser.

Figure 3 is a model of Type II fouling proposed by Costerton (10) based on scanning electron microscopy data collected on the biofilms that had thicknesses >1 mm. The microscopic work showed that the biofilms were composed of bacterial cells enmeshed in a matrix of their own amorphous exopolysaccharides. Inorganic particles were trapped within the bacterial matrix. This composition has been observed in other biofilms found in natural and industrial aquatic systems (11). When a biofilm of this structure develops on a fine pore diffuser, the bubble release surface changes dramatically. Preliminary studies (10) indicate that bubble sizes increase, which may significantly reduce OTE.

Thin biofilm layers are usually not a problem, but as the biofilm thickness increases, BRV and its coefficient of variation will increase, DWP will increase slightly or not at all, and OTE will decrease substantially. The net result of nonuniform bubble release and an increase in bubble size is a substantial reduction in OTE. Figure 4 is an idealized representation of OTE and DWP changes with time under Type II fouling conditions.

EFFECTS OF FOULING ON PERFORMANCE

As indicated above, fouling processes may affect BRV, DWP, and OTE. Based on fouling studies at nine municipal plants (12), the effects of diffuser fouling were divided into three categories as shown in Table 2. The diffusers from the same plant could fit into more than one category depending on the time in service and operating conditions of the plant.

Equation (1) introduced a new term, F, that describes the impairment of diffuser performance caused by diffuser fouling or diffuser media deterioration. It has been introduced to separate the effects of wastewater characteristics (α effects) and diffuser fouling (F effects) on the estimate of process water oxygen transfer. F generally decreases from a value 1.0 with time in service. The characteristics of fouling dynamics are site and diffuser specific as hypothesized in Figure 5. At this time, a linear model describing fouling dynamics has been introduced in the design manual (2) and that conceptual model is shown in Figure 6. The slope of the F vs. time curve represents the fouling rate factor, f_F, expressed in terms of a unit decrease per month. Therefore:

$$f_F = (F - 1.0)/t = [\alpha F(SOTE)_0 - \alpha F(SOTE)_t]/\alpha F(SOTE)_0\, t \qquad (6)$$

This model assumes that there is a minimum value of F, F_{min}, that occurs after some critical service period, t_c. It may be reasonably assumed that F_{min} approaches the ratio of the oxygen transfer performance of a coarse bubble diffuser to that of an unfouled fine pore diffuser in question, both diffusers being in the same physical basin configuration and operating under the same process loadings.

F and its functional relationship with time can be estimated by:

- Conducting full-scale OTE tests in process water over a period of time
- Monitoring aeration system efficiency using operational data
- Conducting OTE tests on fouled and new diffusers, ex situ, in an aeration column (2)

The results of preliminary tests conducted at municipal plants under full-scale monitoring of OTE appear in Table 3. It should be emphasized that these data are very preliminary and based on a simplified assumption of linear reduction of F with time in service to some minimum value.

MEDIA EFFECTS ON PERFORMANCE

Fine pore aeration systems have sometimes exhibited reductions in OTE that could be attributed to mechanical failures or material changes rather than α or fouling effects. A number of examples have been cited where oxygen transfer performance decreased due to gasket failures, broken support bolts, and cracked plenums and air piping (2).

Changes in diffuser media properties can also affect DWP and OTE. Plasticized PVC (a thermoplastic elastomer) and EPDM (a thermoset elastomer), the two principal perforated membrane materials in use, can experience various physical property changes with time when used as wastewater aeration devices. These changes, however, may differ significantly in nature and degree between the two materials. Specific, comprehensive descriptions of the materials employed in media production are therefore essential for prediction of perforated membrane diffuser

performance. Conditions that can substantially affect perforated membrane performance and life include loss of plasticizer, hardening or softening of the material, loss of dimensional stability through creep, absorptive and/or extractive exchange with wastewater, and chemical changes resulting from environmental exposure. Plasticizer migration can cause hardening and reduction in membrane volume, resulting in dimensional changes. Absorption by the membrane of various constituents, including oils, can result in softening of the membrane with volumetric increases and consequent dimensional changes. Membrane creep, which may be influenced by the above factors, will reduce OTE in some cases. It may also be accompanied by a reduction in DWP after cleaning to lower than the original value. This reduction is not recoverable by known maintenance procedures.

Reported data on performance and changes in characteristics of perforated membranes under service conditions, although limited, are becoming increasingly available. Table 4 presents the results of some of these data (2).

EFFECT OF WASTEWATER ON PERFORMANCE

The literature is replete with studies on the effects of wastewater characteristics on oxygen transfer. Both α and the nature and dynamics of fouling are attributed to wastewater properties. Surfactants (surface active agents) are believed to play an instrumental role in the depression (and occasionally the increase) of α in wastewaters. The rise in α as treatment progresses down the length of plug flow aeration basins has been attributed to surfactant removal from the wastewater. Other wastewater properties may also have an impact on α, including total dissolved solids and transition elements such as iron and manganese (2).

It is instructive to note that αF values obtained from a significant number of in-process studies (Table 5) are lower than first anticipated for fine pore aeration systems in municipal wastewater. Although αF values presented in this table are dependent on several process and design variables for the specific treatment plants tested, it is apparent that the average mean weighted αF is usually <0.5.

The variability of αF is site specific. Examples of typical variations over a 24-hr period are presented in Table 6. A review of αF variations at these sites revealed a maximum αF:average αF of about 1.2, with a range of 1.08-1.47. The average value of minimum αF:average αF was approximately 0.86, with a range of 0.77-0.96.

PROCESS LOADING EFFECTS

A review of the dynamics of αF(SOTE) in a variety of activated sludge systems suggests that several process variables affecting oxygen transfer are not clearly identifiable based on our current knowledge of the process. For example, αF(SOTE) data collected at Madison, WI (13) over an 800-day period in the first pass of a three-pass plug flow system demonstrate a significant variability in αF(SOTE) with time. Some of this variability may be attributed to the properties of the wastewater (i.e., composition and strength), but cannot account for all of it. Multiple linear regressions of the data including sludge retention times (SRTs), food-to-microorganism (F/M) loadings, volumetric organic loadings, mixed liquor volatile solids (MLVSS) concentrations, and air-flow rates could account for up to about 60-70 percent of the variability. Similar findings were described for the Whittier Narrows treatment plant (14), where 30-74 percent of the variability in αF(SOTE) could be accounted for by F/M loadings, airflow rates, and time-in-service.

These studies suggest that elements of process loading may have some impact on diffuser performance. There are theoretical reasons why αF should be a function of SRT, F/M loading, or MLVSS concentration. Current biological treatment models developed for activated sludge predict

substrate concentration as a function of these parameters. Since the substrates are partially comprised of surfactants, lower substrate concentrations imply lower surfactant levels and higher α values. The dynamics of biomass accumulation and depletion on diffuser surfaces will very likely influence F. The impact of these process loading factors on aeration system performance may be short or long term depending on their combined effects on both α and F.

Data from Table 5 were analyzed to determine whether a relationship between process loading and αF(SOTE) may exist. A plot for the ceramic diffuser installations in this table, using SRT as the loading parameter, is given in Figure 7. Although wide variations in system design and operation, as well as wastewater characteristics, are evident at these sites, it appears a trend does exist between process loading and αF(SOTE). Nitrifying treatment plants have been highlighted in this figure to indicate their relative importance to the relationship.

FLOW REGIME EFFECTS

Aeration basin flow regime affects the mixing pattern of the basin and, therefore, the residence time distribution of the influent wastewater. Because wastewater components may have an impact on αF(SOTE), it is reasonable to expect that mixing patterns will also affect αF(SOTE). A study conducted on the Madison, Wisconsin, ceramic dome diffuser system (13) illustrates this concept (Figure 8). Single-day αF(SOTE) off-gas profiles are compared as a function of grid position for one aeration basin when it was operated first in a step feed mode vs. two months later when it was operated in a plug flow pattern of three passes in series. The switch from step feed to plug flow was made approximately half way between the two off-gas test days. In both cases, SRT was approximately 2.2 days.

The αF profile along the length of any aeration basin will depend on the degree of mixing that exists in that basin. Typical results for a variety of basin geometries are presented in Table 7. The site key given in the first column refers to the treatment plants described in Table 5. In general, it has been found that treatment plants with high length-to-width ratios, operating as plug flow basins, generate significant αF gradients. Conversely, treatment plants that are short and wide or that employ step feed configurations exhibit much less variability in αF along the basin length.

DIFFUSER FLOW RATE EFFECTS

The OTE of a fine pore diffuser in clean water decreases as the airflow rate increases. Similar results are reported for process waters (2). A plot of αF(SOTE) versus airflow rate per unit media surface area for selected municipal plants listed in Table 5 shows this trend (Figure 9). This curve should not be used for design purposes, since many other variables, as described above, affect oxygen transfer performance.

CONCLUSIONS

Field research over the last three years has significantly improved our understanding of fine pore aeration systems. Based on the information available at this time, the following conclusions can be formulated.

- Fine pore aeration system performance is significantly influenced by diffuser fouling as well as by changes in media properties.
- Attempts to quantify the effects of diffuser fouling and media changes have been initiated using the fouling factor, F, and fouling rate, f_F.

- Diffuser fouling has been classified as Type I or Type II, each having distinct characteristics that can occur alone or in combination with variable dominance from site to site and time to time.
- A number of factors effect the performance of a fine pore aeration system under process conditions including the wastewater, process loading, flow regime, and airflow rate.
- The SRT (F/M) of an operating plant appears to affect fine pore aeration system performance. An increase in SRT (decrease in F/M) generally will result in an increase in αF SOTE.

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TABLE 1

Guide to Application of Equation 3

Parameter	Source of Information
$C^*_{\infty 20}$	Clean water test results.
SOTR	Clean water test results.
d_e	Clean water test results.
C	Process water conditions.
T	Process water conditions.
τ	Calculated based on tabulated DO surface saturation values.
Ω	
α	Calculated based on site barometric pressure and effective depth data.
β	Estimated based on experience or on measured values of K_La in clean and process waters using a clean diffuser.
	Calculated based on total dissolved solids measurements.
Θ	Taken as 1.024 unless experimentally proven to differ.
F	Estimated based on field experience, field measurements, or laboratory analysis of diffusers taken from the field.

TABLE 2

Fine Pore Diffuser Fouling Categories (2)

Condition	BRV (cm, w.g.)	DWP/BRV	F (After 1 year in service)
Severe	>100	<0.3	<0.7
Moderate	40-100	0.3-0.6	0.7-0.9
Light	<40	>0.6	>0.9

TABLE 3

Fouling Rates (f_F) and Fouling Factors (F) for Selected Treatment Plants (2)

City - Plant	Time in Service,[a] years	Test Period, months	No. Tests	f_F, month^{-1}	Correlation, r	F (@ 1 month)	Comments
Frankenmuth, MI	3.3	13.0	17	0.029	-0.52[b]	0.97	Ceramic disc diffusers; Basin 5
		11.7	16	0[c]		1.0	Ceramic disc diffusers; Basin 6
Green Bay, WI (2nd operating period)	3.0	6.5	9	0.064	-0.43	0.94	Ceramic disc diffusers: contact basin
		6.5	9	0.067	-0.66[b]	0.93	Ceramic disc diffusers; reaeration basin
Green Bay, WI (1st 6 months only)	3.3	6.0	5	0.028	-0.56	0.97	Perforated membrane tube diffusers; contact basin
		6.0	5	0.046	-0.65[b]	0.95	Perforated membrane tube diffusers; reaeration basin
Milwaukee, WI; Jones Island East	6.0	30.0	20	0.002	-0.22	1.0	Ceramic plate diffusers
Milwaukee, WI; South Shore	15.0	56.7	20	0[c]	-	1.0	Ceramic plate diffusers; various basins
Madison, WI; Nine Springs	3.5	26.7	37	0[c]	-	1.0	Ceramic disc diffusers
Monroe, WI	3.0	4.5	5	0.043	-0.81	0.96	Ceramic disc diffusers during bypass of aerated flow from equalization basin
Los Angeles County, CA; Whittier Narrows	9.3	24.0	23	0.027	-0.82[b]	0.97	Ceramic disc diffusers injected with HCl gas 5 times about every 3 months
Ridgewood, NJ	6.0	5.0	48	0.074	-0.15	0.93	Ceramic dome diffusers cleaned often by hosing or brushing with HCL solution

[a] As of 4/1/89.
[b] Significant at 95 percent confidence level.
[c] Zero fouling rate based on visual inspection.

TABLE 4

Characteristics of Perforated Membrane Diffusers in Service

Site	Diffuser Type	Membrane Type	Wastewater Type	Time in Service, months	Diffuser Condition[a]	Diffuser Physical Characteristics: Weight, g	Thickness, in	Durometer Shore A	Tensile Modulus[b], psi	Type of Test Specimen	DWP (@ 1 scfm/diffuser), in	SOTE[c] (@ 1 scfm/diffuser), %
A-1[d]	Tube	PVC (U.S.)	Municipal/ Industrial	7	New	120	0.031	63	630		7.3	18.0
					AS[e]						6.1	11.9
					AC	114	0.028	63	900	Pilot	3.2	13.2
A-2[d]	Disc	EPDM (U.S.)	Municipal/ Industrial	14	New	174	0.105	60	612[f]		12.0	19.5
					AS					Pilot	12.7	18.2
					AC	174	0.106	68	728[f]	Pilot	10.7	18.0
B	Tube	PVC (U.S.)	Municipal	48	New	118	0.028	63	653		5.5	18.0
					AS[e]						6.2	13.5
					AC[e]	108	0.028	75	3,273	Field	4.3	14.2
C	Tube	PVC (U.S.)	Industrial	13	New	117	0.030	60 -	600		8.8	18.0
					AC	87.5	0.024	81	17,000	Field	3.8	14.8
D	Tube	EPDM (Eur.)	Industrial	24	New	280	0.080		375		15.3	23.0
					AC	222	0.062		1,450	Field	6.3	17.5
E	Tube	EPDM (Eur.)	Industrial	8	New		0.083		360		11.7[g]	21.8
					AS						14.5[g]	19.8
					AC		0.078		487	Pilot	9.65[g]	21.8
F	Tube	EPDM (Eur.)	Municipal	13	New		0.078		375		11.0	23.0
					AC		0.075		560	Field	11.3	19.8
G	Disc	EPDM (Eur.)	Municipal	12	New	263	0.086	50.8	334[f]		8.4	17.8[h]
					AS					Field	15.5	16.7[h]
					AC	243	0.078	51.1	318[f]	Field	7.7	18.0[h]
H	Disc	EPDM (U.S.)	Municipal	11.5	New	174	0.105	60			12.0	19.5
					AS					Field	16.3	18.5
					AC	178	0.106	62		Field	11.6	20.0
I	Disc	EPDM (U.S.)	Municipal	7	New	174	0.105	60			12.0	19.5
					AS					Pilot	19.4	19.5
					AC	177	0.103	61		Pilot	11.1	19.4
J	Disc	EPDM (U.S.)	Municipal	3	New	175	0.105	54			10.0	19.5
					AS					Pilot	15.9	17.9
					AC	179	0.099	59		Pilot	11.3	20.5
K	Disc	EPDM (U.S.)	Municipal	7	New	174	0.105	60			12.0	19.5
					AS					Pilot	12.5	18.9
					AC	176	0.105	64		Pilot	11.0	19.5

[a] AS - after indicated period of service; AC - after cleaning (performed after indicated period of service).
[b] Radial for discs; circumferential for tubes.
[c] Clean water test in laboratory column at 10-ft diffuser submergence.
[d] Not concurrent in time.
[e] Mean of 6 samples.
[f] @ 30 psi stress.
[g] DWP @ 0.75 scfm/diffuser.
[h] @ 3 scfm/diffuser.

TABLE 5

Oxygen Transfer and Plant Process Data

Site No.	City - Plant	αF(SOTE)/ft, percent			αF			scfm/sq ft[a]			lb BOD_5/d/ 1,000 cu ft	F/M[b], day^{-1}	SRT[2], days	scfm/BOD_5[c]	No. Samples
		Mean	Min	Max	Mean	Min	Max	Mean	Min	Max					
1	Frankenmuth, MI	0.73	0.58	0.95	0.37	0.30	0.48	0.80	0.63	0.97	95	0.52	11.0	0.56	9
2	Green Bay, WI	0.71	0.57	0.91	0.43	0.35	0.54	0.73	0.46	1.02	149	0.59	3.0	0.36	9
3	Green Bay, WI	0.86	0.54	1.19	0.49	0.36	0.64	0.62	0.39	0.93	149	0.59	3.0	0.30	9
4	Hartford, CT	0.64	0.46	0.84	0.36	0.24	0.48	0.36	0.18	0.52	52	0.16	7.6	0.45	12
5	Milwaukee, WI; Jones Island East	1.09	0.81	1.36	-	-	-	0.20	0.12	0.35	72	0.77	3.8	0.20	30
6	Milwaukee, WI; Jones Island West	0.77	0.44	1.04	-	-	-	0.17	0.08	0.28	71	0.82	3.3	0.16	21
7	Madison, WI; East	0.74	0.54	1.04	0.43	0.31	0.57	0.28	0.15	0.38	36	0.63	2.2	0.49	21
8	Madison, W; East	1.14	0.97	1.36	0.66	0.56	0.79	0.23	0.19	0.27	12	0.12	14.0	1.25	8
9	Madison, WI; West	0.92	0.87	0.98	0.48	0.44	0.51	0.20	0.19	0.22	16	0.15	9.6	0.74	5
10	Ridgewood, NJ	0.67	0.38	0.94	0.41	0.23	0.58	0.39	0.11	0.67	20	0.24	3.1	1.32	48
11	Milwaukee, WI ; South Shore	1.25	0.99	1.48	-	-	-	0.14	0.08	0.28	26	0.37	7.5	0.36	20
12	Los Angeles County, CA; Whittier Narrows	0.72	0.47	0.94	0.31	0.21	0.40	0.25	0.15	0.38	30	0.61	2.6	0.56	35
13	Los Angeles County, CA; Whittier Narrows	0.55	0.25	0.89	0.24	0.11	0.39	0.32	0.20	0.55	30	0.61	2.6	0.71	34
14	Los Angeles, CA; Terminal Island	0.48	0.27	0.67	-	-	-	0.68	0.39	1.32	-	0.27	6.6	-	8
15	Los Angeles, CA; Terminal Island	0.80	0.62	1.19	0.45	0.30	0.74	0.47	0.23	0.94	-	0.27	6.6	-	7
16	Los Angeles County, CA; Valencia	0.56	0.40	0.62	0.28	0.26	0.29	0.43	0.27	0.47	-	-	-	-	7
17	Monroe, WI	0.63	0.54	0.80	0.35	0.28	0.54	0.36	0.29	0.42	33	0.40	8.1	0.73	12
18	Phoenix, AZ; 23rd Ave.	0.47	0.44	0.51	0.27	0.24	0.31	0.53	0.31	0.66	38	0.76	1.0	0.75	6
19	Phoenix, AZ; 91st St.	0.58	-	-	0.29	-	-	0.37	-	-	34	-	2.0	0.70	1
20	Phoenix, AZ; 91st St.	0.59	-	-	0.28	-	-	0.33	-	-	33	-	1.6	0.70	1
21	Minneapolis, MN; Metro	0.70	0.56	0.84	0.37	0.29	0.45	0.43	-	-	62	-	5	0.45	6
22	Minneapolis, MN; Metro	0.75	0.64	0.86	0.40	0.34	0.46	0.43	-	-	28	-	10	0.99	18
23	Minneapolis, MN; Metro	1.02	0.92	1.11	0.52	0.45	0.59	0.39	-	-	24	-	9	1.04	44
24	U.K.; Rye Meads	1.00	0.90	1.10	-	-	-	0.21	0.19	0.24	36	-	11.7	1.32	-
25	U.K.; Rye Meads	0.67	0.57	0.89	-	-	-	0.21	0.19	0.22	66	-	2.7	0.73	-
26	Glastonbury, CT	0.55	0.41	0.70	0.56	0.42	0.67	0.19	0.14	0.29	27	0.19	-	0.46	6

[a] scfm of airflow/sq ft basin floor area.
[b] Based on MLSS under aeration.
[c] Average scfm of airflow based on measured off-gas flux rate per unit area of basin floor/average lb BOD_5 applied per day.

TABLE 6

24-Hr αF and αF(SOTE) Variations at Selected Municipal Treatment Plants

Site No.	City - Plant	αF			αF(SOTE)			Position in Basin
		Average	Minimum	Maximum	Average	Minimum	Maximum	
4	Hartford, CT	0.30	0.23	0.44	8.3	6.4	11.2	Influent pass
9	Madison, WI; West	0.24	0.22	0.29	8.7	7.7	10.4	Inlet end
10	Ridgewood, NJ	0.46	0.44	0.59	10.7	9.5	13.1	Entire basin weighted
12	Los Angeles* County, CA; Whittier Narrows	0.25	0.21	0.27	7.8	6.4	8.7	Influent grid
12	Los Angeles* County, CA; Whittier Narrows	0.26	0.20	0.30	8.7	6.6	9.9	Middle grid
13	Los Angeles County, CA; Whittier Narrows	0.45	0.41	0.50	12.2	11.1	13.5	Effluent grid
17	Monroe, WI	0.23	0.19	0.28	-	-	-	Influent pass
17	Monroe, WI	0.39	0.33	0.45	-	-	-	Effluent pass

* Data for 6-hr period.

TABLE 7

αF Profiles for Various Aeration Systems[a,b,c]

Site No.	City - Plant	αF Zone 1 Mean	Zone 1 Min	Zone 1 Max	Zone 2 Mean	Zone 2 Min	Zone 2 Max	Zone 3 Mean	Zone 3 Min	Zone 3 Max	Total Basin Mean	Total Basin Min	Total Basin Max	Diffuser Type	Flow Regime
2	Green Bay, WI	0.45	0.35	0.55	0.43	0.35	0.59	0.40	0.31	0.54	0.43	0.36	0.53	PVC Perf. Memb. Tubes	Plug Flow
3	Green Bay, WI	0.49	0.41	0.68	0.50	0.34	0.67	0.46	0.30	0.64	0.49	0.36	0.64	Ceramic Discs	Plug Flow
4	Hartford, CT	0.37	0.18	0.49	0.37	0.28	0.49	0.35	0.24	0.45	0.36	0.24	0.48	Ceramic Domes	Step Feed
5	Milwaukee, WI; Jones Island East	0.45	0.32	0.60	0.58	0.44	0.79	0.60	0.47	0.77	0.54	0.44	0.68	Ceramic Plates	Plug Flow
6	Milwaukee, WI; Jones Island West	0.34	0.18	0.46	0.40	0.27	0.49	0.43	0.25	0.60	0.39	0.23	0.52	Ceramic Plates	Plug Flow
7	Madison, WI; East	0.32	0.24	0.44	0.44	0.29	0.62	0.52	0.36	0.76	0.43	0.31	0.57	Ceramic Domes	Plug Flow
8	Madison, WI; East	0.40	0.33	0.47	0.64	0.54	0.78	0.92	0.77	1.00	0.66	0.56	0.79	Ceramic Domes	Plug Flow
9	Madison, WI; West	0.33	0.26	0.40	0.54	0.52	0.56	0.55	0.52	0.58	0.48	0.44	0.51	Ceramic Discs	Plug Flow
11	Milwaukee, WI; South Shore	0.64	0.50	0.92	0.62	0.47	0.83	0.64	0.51	0.83	0.63	0.51	0.75	Ceramic Plates	Step Feed
12	Los Angeles County, CA; Whittier Narrows	0.25	0.15	0.42	0.30	0.15	0.40	0.38	0.22	0.51	0.31	0.21	0.40	Ceramic Discs	Plug Flow
13	Los Angeles County, CA; Whittier Narrows	0.16	0.09	0.27	0.23	0.08	0.40	0.31	0.17	0.49	0.24	0.11	0.39	Ceramic Domes	Plug Flow
18	Phoenix, AZ; 23rd Ave.	0.29	0.25	0.34	0.27	0.23	0.31	0.25	0.21	0.30	0.27	0.24	0.31	Ceramic Domes	Step Feed
21	Minneapolis, MN; Metro	0.36	0.32	0.40	0.36	0.23	0.42	0.37	0.24	0.45	0.37	0.29	0.45	Ceramic Domes	Step Feed
22	Minneapolis, MN; Metro	-	-	-	-	-	-	-	-	-	0.40	0.34	0.46	Ceramic Domes	Step Feed
23	Minneapolis, MN; Metro	0.50	-	-	-	-	-	-	-	-	0.52	0.45	0.59	Ceramic Domes	Step Feed
26	Glastonbury, CT	0.59	0.43	0.69	0.54	0.56	0.77	0.56	0.37	0.65	0.56	0.42	0.67	Rigid Porous Plastic Tubes	Plug Flow

[a] Each zone represents 1/3 of the aeration volume.
[b] Reaeration volume not included for contact stabilization systems.
[c] SOTE for plate diffusers was assumed to be 2 percent/ft submergence.

FIGURE 1

Schematic Structure of Type I Fouling

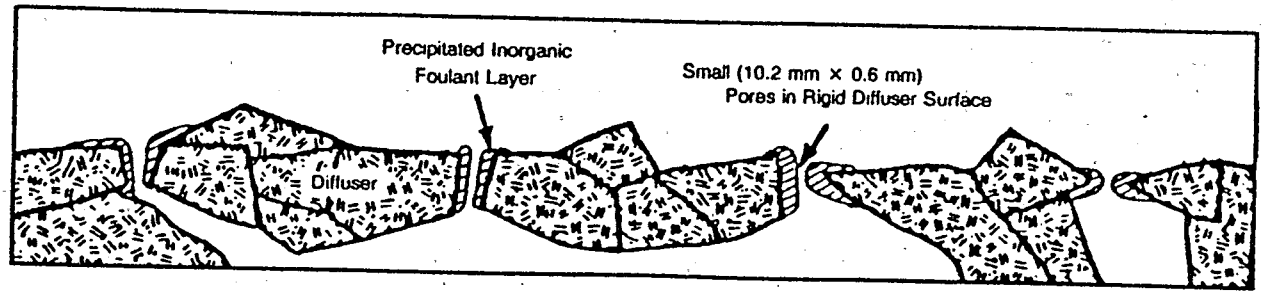


FIGURE 2

Idealized Plot Showing Effects of Type I Fouling on DWP and OTE

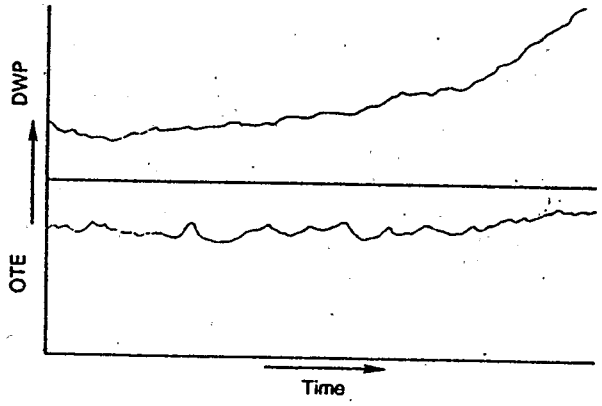


FIGURE 3

Schematic Structure of Type II Fouling

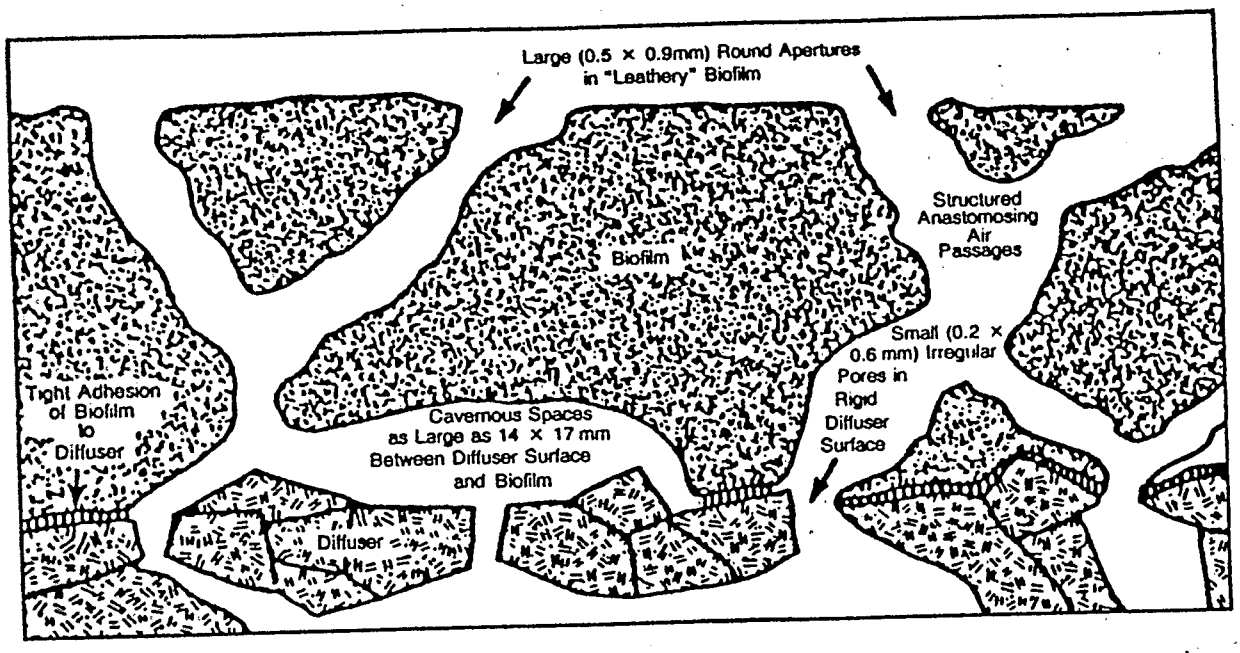


FIGURE 4

Idealized Plot Showing Effects of Type II Fouling on DWP and OTE

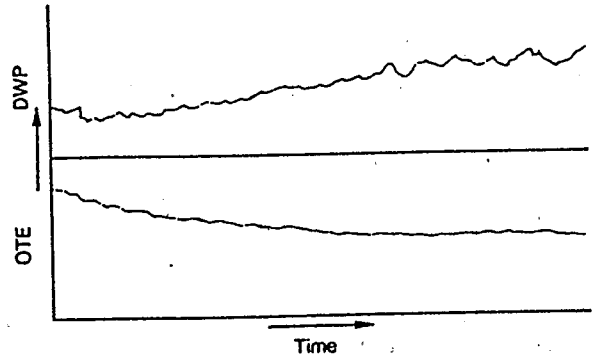


FIGURE 5

Hypothetical Fouling Patterns

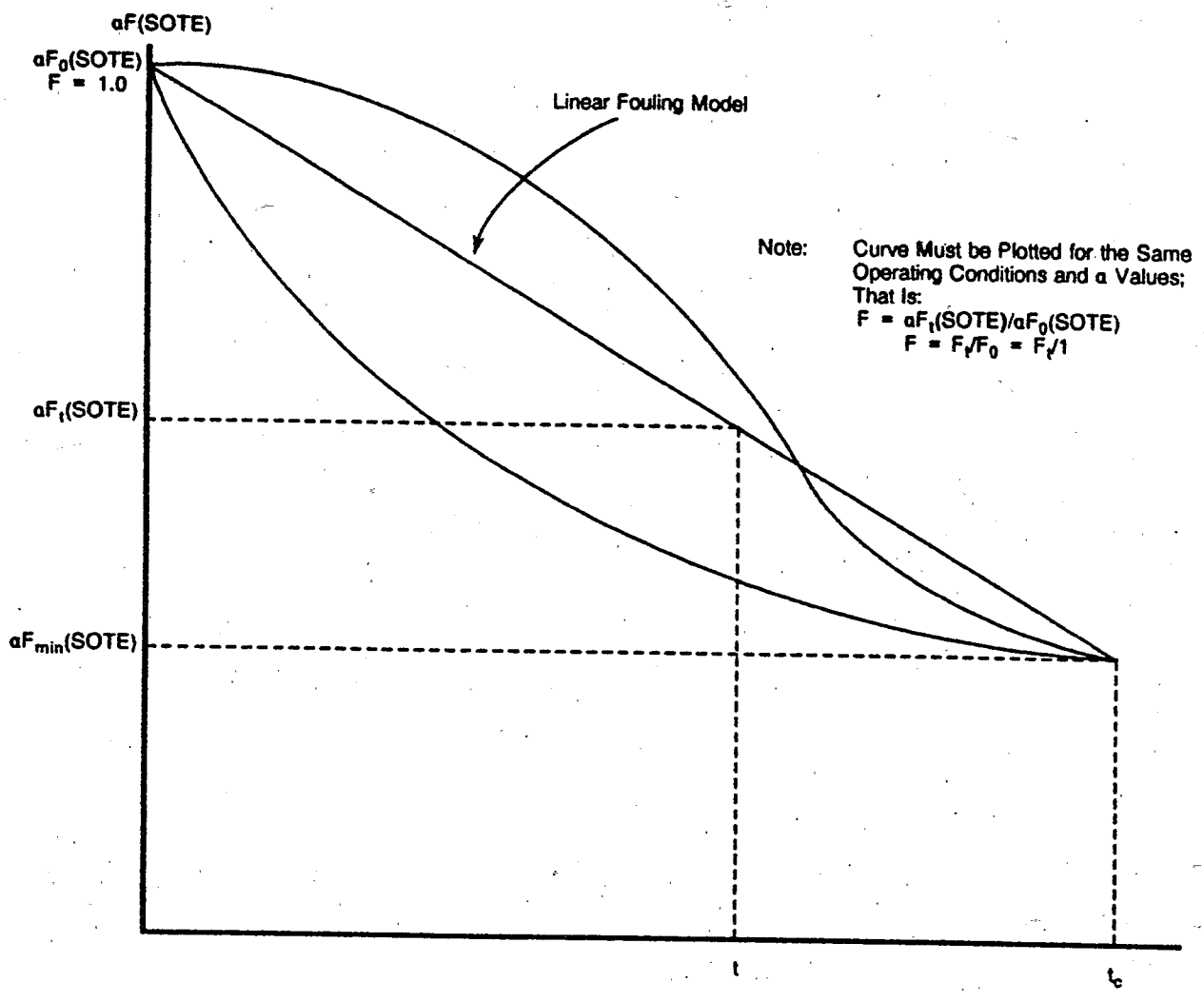


FIGURE 6

Linear Fouling Factor Model

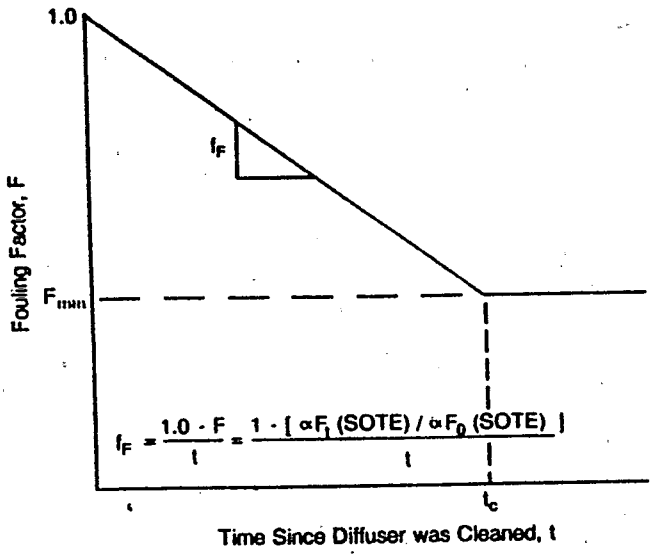


FIGURE 7

αF(SOTE) vs. SRT for Ceramic Diffuser Facilities

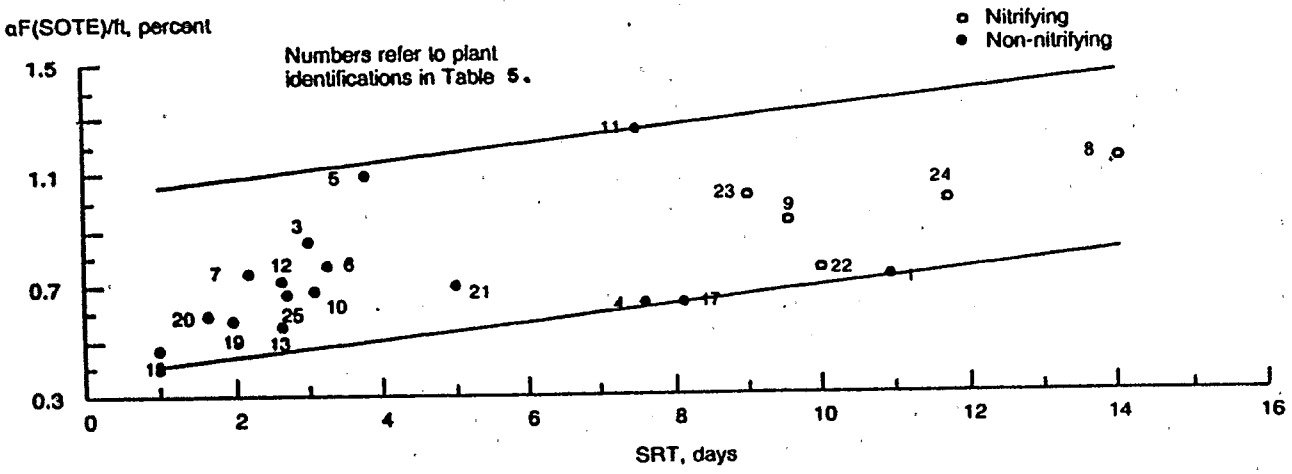


FIGURE 8

αF(SOTE) vs. Grid Position for Plug Flow and Step Feed Aeration Systems - Madison, WI

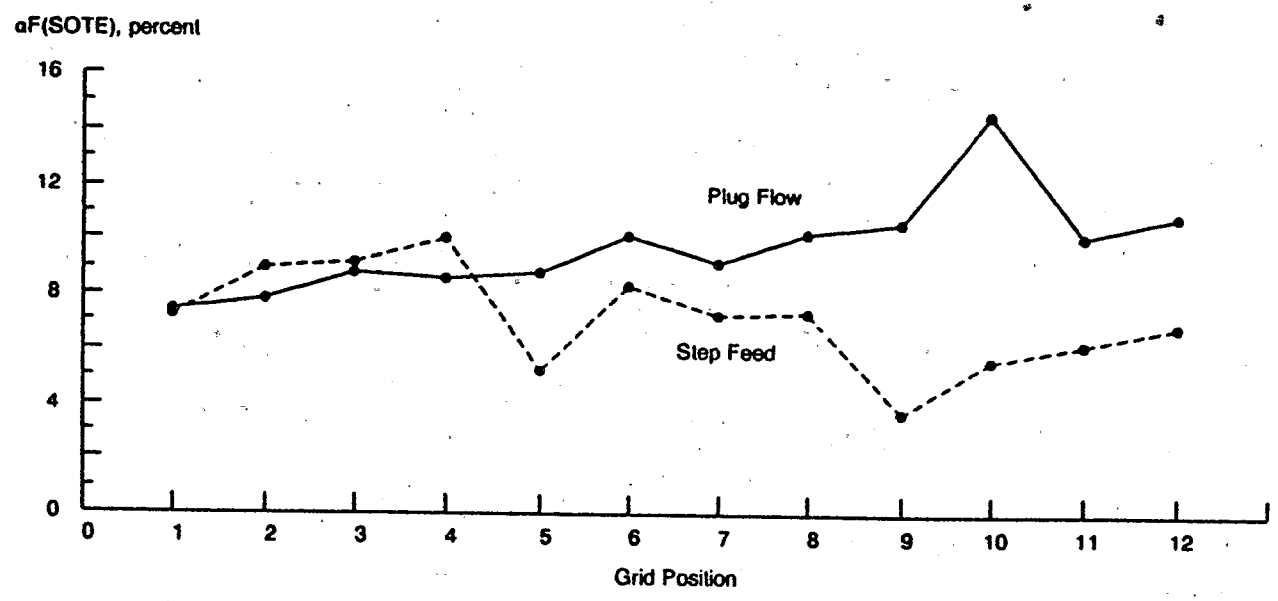
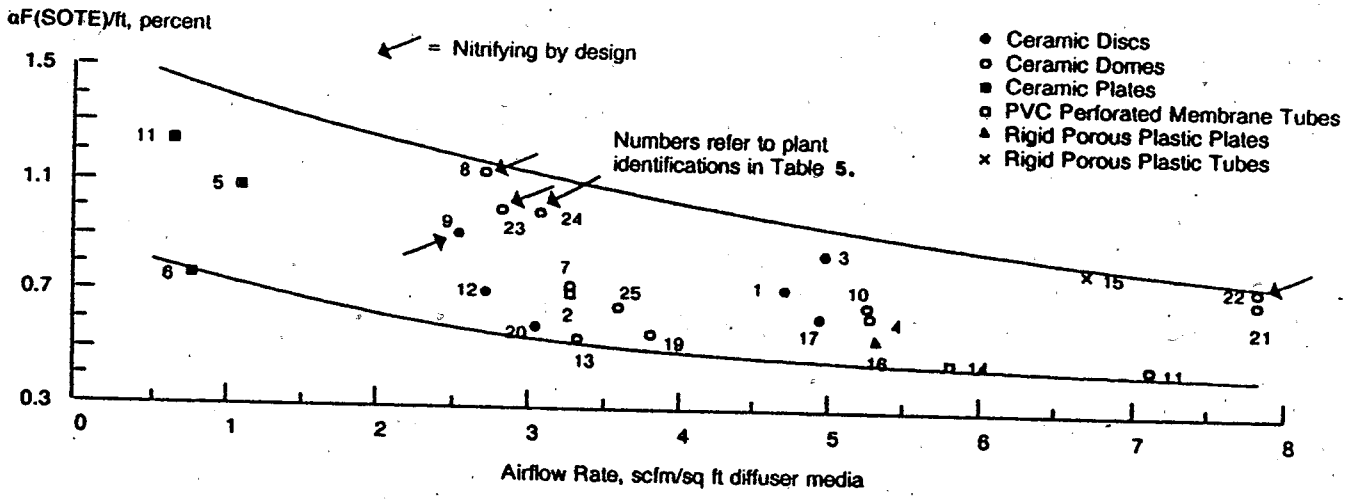


FIGURE 9

Mean αF(SOTE) vs. Airflow Rate Per Unit Area of Diffuser Media



BIODEGRADABILITY AND BIODEGRADATION KINETICS ASSESSMENT FOR TOXIC ORGANIC COMPOUNDS USING OXYGEN UPTAKE MEASUREMENTS

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BIODEGRADABILITY AND BIODEGRADATION KINETICS ASSESSMENT FOR TOXIC ORGANIC COMPOUNDS USING OXYGEN UPTAKE MEASUREMENTS

ABSTRACT

Electrolytic respirometry involving natural sewage, sludge and soil microbiota is being applied to the fate studies of priority pollutant and RCRA toxic organics to generate data on their biodegradability and on biodegradation/inhibition kinetics. This paper discusses the experimental design and procedural steps for the respirometry biodegradation and toxicity testing approach for individual organics or specific industrial wastes. The discussion also includes a review of the electrolysis BOD measuring system inherent in electrolytic respirometry and the factors effecting respirometric determination and measurement of respiration rate.

A developed multi-level protocol is presented for determination of the biodegradability, microbial acclimation to toxic substrates and first order kinetic parameters of biodegradation (μ and μ') and for estimation of the Monod kinetic parameter (μ_m, K_s and Y_g) of toxic organic compounds, in order to correlate the extent and rate of biodegradation of these organics with a predictive model based on chemical properties and structure of these compounds.

Respirometric biodegradability/inhibition and biodegradation kinetic data are provided for representative RCRA alkyl benzenes, phenolic compounds, phthalate esters, and ketones. Data on the effects of the source of sludge biomass, temperature, and concentrations of microbial inoculum and of toxic substrate on the kinetics of biodegradation, are also included.

INTRODUCTION

Electrolytic respirometry is attaining prominence in biodegradation studies and is becoming one of the more suitable experimental methods for measuring the biodegradability and the kinetics of biodegradation of toxic organic compounds by the sewage, sludge and soil microbiota and for determining substrate inhibitory effects to microorganisms in wastewater treatment systems.

Biodegradation of toxic and hazardous organic compounds holds a great promise as an important fate mechanism in wastewater treatment and in soil detoxification. Information about the extent and rate of biodegradation is a prerequisite for informed decision making on the applicability of the biodegradation approach. Unfortunately, relatively little quantitative data are available from which engineering judgement can be made, because of the large effort required to assess biodegradation kinetics.

Current research in our laboratories has shown that it is possible to assess biodegradation kinetic parameters from oxygen uptake data, obtained through the use of electrolytic respirometry. This method greatly reduces the work and expense involved in evaluation of biodegradation kinetics. The ongoing biodegradation studies are concerned with the generation of biokinetic database so that it can be ultimately used to establish a possible correlation between molecular substrate configuration (chemical/physical characteristics) and biomass activity (kinetic parameters) as an index of biodegradation. The experimental respirometric testing is also providing data on the concentration levels of toxic organics inhibitory to microbial activity.

The ongoing biodegradation studies are concerned with the prediction of the biological fate of toxic organic compounds using electrolytic respirometry as an approach to measure the biodegradation of selected organic compounds and generate biokinetic data so that these can be ultimately used to establish a possible correlation between molecular substrate configuration (physical/chemical characteristics) and biomass activity (kinetic parameters) as an index of biodegradation. The experimental respirometry testing is also providing data on the concentration levels of toxic organics inhibitory to microbial activity.

Initially, the inter-laboratory, ring test, Organization of Economic Cooperation and Development (OECD) studies at the EPA laboratory, Cincinnati, Ohio, were undertaken to develop confirmatory respirometric biodegradability testing procedure. Respirometric biodegradability and biokinetic data were provided for the selected non-inhibitory and non-adsorbing compounds, tetrahydrofuran, hexamine, pentaerythritol, 1-napthol, sodium benzene sulphinate, thioglycolic acid and the biodegradable reference compound, aniline.

Subsequently, similar electrolytic respirometry studies were initiated to determine biodegradation kinetic parameters for selected representative toxic compounds of varied classes of organics included in the Priority Pollutant, RCRA and superfund CERCLA lists, and to demonstrate presence of

any inhibitory effects of these organics of specified concentration levels on the sludge biomass and on the metabolism of biogenic compounds.

The objectives of the present study were to utilize the electrolytic respirometry oxygen uptake data to: (1) determine the biodegradability of selected RCRA alkyl, chloro, and nitrobenzenes, phenols, phthalates, and ketones and representative CERCLA leachate toxic organics; (2) generate information on their acclimated times (t_0) and the initiation and termination time values for the declining growth phase (t_1 and t_2); (3) determine their first order kinetic parameters of biodegradation (specific growth rate constants for the exponential growth phase (μ) and for the declining growth phase (μ'); (4) estimate the Monod kinetic parameters (μ_m, K_s and Y) of these compounds without initial growth or growth yield assumptions; (5) demonstrate presence of any inhibitory effects of these compounds on the metabolism of the biodegradable reference compound, aniline; and (6) to correlate the extent and rate of biodegradation of these compounds with a predictive model based on chemical properties and structure of these compounds.

The purpose of this study was to obtain information on biological treatability of the benzene, phenol, phthalate and ketone organics and of the Superfund CERCLA organics bearing wastes in wastewater treatment systems which will support development of an EPA technical guidance document on the discharge of the above organics to POTWs. The study was to generate basic information on the fate of CERCLA leachate organics during on-site treatment and biodegradation and inhibition data for pollutants found in Superfund site wastewater that could be discharged to POTWs. Respirometric biodegradability, biokinetic and inhibition data were generated for the selected RCRA benzene, phenolic, phthalate and ketone compounds.

BACKGROUND

Measurement of Oxygen Consumption

Measurement of oxygen consumption is one of the oldest means of assessing biodegradability. Time consuming manual measurement of oxygen uptake (dilution BOD measurements) was replaced gradually by a more direct and continuous respirometric method for measurement of oxygen consumption in biochemical reactions, for use in routine examination of sewage and in control of sewage treatment process.

A rather comprehensive review of the use of respirometers for the study of sewage and industrial waste and their application to water pollution problems was published by Jenkins in 1960 (1). Montgomery's (2) review of respirometric methods summarized the design and application of respirometers for determination of BOD.

The application of respirometry was gradually directed to research studies to assess the toxicity and biodegradation of specific wastes or compounds, to evaluate factors affecting biological growth and to provide an insight into nitrification reaction. Of the commercial respirometers which have been developed for respirometric studies, the electrolytic respirometers were shown to be most applicable for measurement and

quantitation of biodegradation activity because they automatically produce oxygen as needed, thereby eliminating some of the limitations of other techniques and allowing output data to be collected automatically for direct recording and processing (3-9). A recent detailed review of respirometric techniques and their application to assess biodegradability and toxicity of organic pollutants was published by King and Dutka (10).

Respirometric Biodegradability Testing

Most uses of electrolytic respirometry in biodegradability testing have been for screening purposes to measure the extent of biodegradation as a percentage of the theoretical oxygen demand exerted in some time period (9, 11-17). A more recent study by Painter and King (18) concluded that a procedure based on electrolytic respirometry was reliable for assessing biodegradability, and could serve as an adequate Level I screening test for biodegradability (19).

A considerable number of studies using electrolytic respirometry to determine the biodegradability of wastes and specific organics is available in published literature and significant data on biodegradation of pollutants based on oxygen uptake have been generated (2, 4, 5, 20-34).

There are many techniques that have been used to evaluate biodegradation kinetics and these were reviewed in detail by Howard et al. (35, 36) and Grady (37). These techniques utilize continuous, fed-batch and batch type reactors for providing data from which kinetic parameters can be evaluated. The use of batch systems in biotechnology and biological wastewater treatment represents a less labor intensive, less expensive and much faster way to model biokinetics.

Hopefully, the kinetic parameters obtained by the above techniques would be intrinsic, that is, dependent only on the nature of the compound and the degrading microbial community and not on reactor system used for data collection. If this condition is satisfied, then the parameters obtained can be used for any reactor configuration and can be used in mathematical models to estimate the fate of toxic organics.

Batch techniques have been shown to be successful in obtaining intrinsic kinetic parameters by applying non-linear curve fitting techniques to single batch substrate removal curves, provided initial conditions are selected with proper care (Simkins and Alexander (38, 39), Robinson and Tiedje (40), Cech et al. (41), and Braha and Hafner (42). Batch systems can be used with either acclimated or unacclimated biomass for providing kinetic data and require that samples be taken at discrete time intervals during the course of biodegradation [Tabak et al. (43), Larson and Perry (24) and Paris and Rogers (44)].

Use of Electrolytic Respirometry to Generate Biokinetic Data

Dojlido (11), Larson and Perry (24), Tabak et al. (45), Oshima et al. (46), Gaudy et al. (47, 48), Grady et al. (49-51) investigated the use of respirometry to generate biodegradation kinetic data. Larson and Perry (24) showed that the electrolytic respirometer can be used to measure biodegradation of complex organics in natural waters when specific analytical methods or radiolabeled materials are unavailable. However, they used empirical kinetic expressions which were system specific.

Dojlido (11) divided the oxygen uptake curve into seven different phases and then proposed an empirical model for each phase and evaluated the biodegradability and toxicity of a test compound by measuring empirical rate constants and time interval associated with each phase. Various phases were distinguished by identifying inflection points in the curve through the plots of the logarithm of the slope versus time.

Tabak et al. (45) and Oshima et al. (46) sought to capitalize on Dojlido's method of identifying inflection points in order to quantify more fundamental rate coefficients. In order to identify more fundamental (and therefore intrinsic) kinetic coefficients they used the generalized concept of oxygen uptake by Gaudy and Gaudy (52). Substrate removal was divided into exponential and declining phases separated by an inflection point, and the endogenous phase. They coupled substrate removal and cell growth to oxygen consumption by imposition of an electron balance and consequently were able to evaluate μ_m from oxygen consumption data up to the inflection point. They proposed the use of the lag time as an indicator of how difficult it is to achieve acclimation to a test compound and their respirometric studies were carried out with an unacclimated biomass.

The justification for using respirometry to obtain intrinsic kinetic lies in the concept of oxygen consumption as an energy balance [Busch et al. (53); Gaudy and Gaudy (52)]. This concept states that all of the electrons available in a substrate undergoing biodegradation must either be transferred to the terminal acceptor or be incorporated into new biomass or soluble microbial products. If the concentrations of the substrate, the products and the biomass are all expressed in units of chemical demand (COD), then the oxygen uptake can be calculated in a batch reactor from an equation relating oxygen uptake to substrate, biomass and soluble products [Grady et al. (49)]. Furthermore, since biomass growth and product formation are proportional to substrate removal, this suggests that an oxygen uptake curve can provide the same information as either a substrate removal curve or a biomass growth curve. This latter concept has recently been used by Gaudy et al. (47, 48) to calculate biodegradation kinetics. Specific growth rates obtained from growth studies as slopes of plots of Ln (biomass concentration) or Ln(X) versus time at different substrate concentrations, compared favorably with those obtained from exponential phase of respirometric oxygen uptake curves as slopes of plots of Ln (d oxygen uptake)/d time or Ln (dO_u/dt) versus time.

Studies of Grady et al. (49, 50) have demonstrated that it is possible to determine intrinsic kinetics of single organic compounds by using only measurements of oxygen consumption in respirometric batch reactors. With the use of computer simulation techniques and non-linear curve fitting

methods, intrinsic kinetic parameters were obtained from oxygen consumption data and were shown to be in agreement with those obtained from traditional measurement of substrate removal (DOC, SCOD, ^{14}C) or cell growth.

MATERIALS AND METHODS

Experimental Approach

The electrolytic respirometry approach to determine the biodegradability of the organic test compounds in this study was chosen because of the specific advantages of the respirometric methods over that of manometric procedures in tracking oxygen utilization during the exertion of biochemical oxygen demand (BOD). These advantages are listed in Table 1. General classification of respirometers based on principle of operation and on techniques and applications is presented in Tables 2 and 3.

The electrolytic respirometry studies were conducted using an automated continuous oxygen uptake and BOD measuring Voith Sapromat B-12 (12 unit system) electrolytic respirometer-analyzer. The instrument consists of a temperature controlled waterbath, containing measuring units, a recorder for digital indication and direct plotting of the decomposition velocity curves of organic compounds; and a cooling unit for the conditioning and continuous recirculation of waterbath volume. The recorder shows the digital indication of oxygen uptake and constructs a graph for these values of each measuring unit. The cooling unit constantly recirculates water to maintain constant temperature in the waterbath. Each measuring unit as shown in Figure 1 is comprised of a reaction vessel with a carbon dioxide absorber mounted in a glass joint flask stopper, an oxygen generator and a pressure indicator. This measuring unit is interconnected by hoses, forming an air sealed system, so that the atmospheric pressure fluctuations do not adversely affect the results.

The activity of the microorganisms in the sample creates a vacuum which is recorded by the pressure indicator, which triggers the oxygen generator. The pressure conditions are balanced by electrolytic oxygen generation. The quantity of the sample, the amperage for the electrolysis and the speed of the synchronous motor are so adjusted that, with a sample of 250 mL, the digital counter indicates the oxygen uptake directly in mg/L. The CO_2 generated is absorbed by soda lime. The nitrogen/oxygen ratio in the gas phase above the sample is maintained throughout the experiment and there is no depletion of oxygen. The recorder-plotter concomitantly constructs an oxygen uptake graph for the selected values. The oxygen generators of the individual measuring units are electrolytic cells which supply the required amount of oxygen by electrolytic dissociation of a copper sulfate solution combined with sulfuric acid.

The nutrient solution used in these studies was an OECD synthetic medium (19, 56) consisting of measured amounts per liter of deionized distilled water of (1) mineral salts solution; (2) trace salts solution, and (3) a solution (150 mg/L) of yeast extract as a substitute for vitamin solution.

The microbial inoculum was an activated sludge from The Little Miami wastewater treatment plant in Cincinnati, Ohio, receiving municipal wastewater. The activated sludge sample was aerated for 24 hours before use to bring it to an endogenous phase. The sludge biomass was added to the medium at a concentration of 30 mg/L total solids. Total volumes of the synthetic medium in the 500 mL capacity reactor vessels were brought up to a final volume of 250 mL.

The test and control compound concentrations in the media were 100 mg/L. Aniline was used as the biodegradable reference compound, at a concentration of 100 mg/L.

The typical experimental system consisted of duplicate flasks for the reference substance, aniline, and the test compounds, a single flask for the physical/chemical test (compound control), a single flask for toxicity control (test compound plus aniline at 100 mg/L each) and an inoculum control. The contents of the reaction vessels were preliminarily stirred for an hour to ensure endogenous respiration state at the initiation of oxygen uptake measurements. Then the test compounds and aniline were added to it. The reaction vessels were then incubated at 25°C in the dark (enclosed in the temperature controlled waterbath) and stirred continuously throughout the run. The microbiota of the activated sludge used as an inoculum were not pre-acclimated to the substrates. The incubation period of the experimental run was between 28 to 50 days. A more comprehensive description of the procedural steps of the respirometric tests is presented elsewhere (45, 55 and 56).

For fully automatic data acquisition, frequent recording and storage of large numbers of oxygen uptake data, the Sapromat B-12 recorders are interfaced to an IBM-AT computer via the Metrabyte interface system. The use of Laboratory Handbook software package allows the collection of data at 15 minute intervals.

The oxygen utilization by the biomass based on the oxygen uptake velocity (BOD) curves, consisting of the exponential and declining phases of microbial growth was the basis for measurement of substrate utilization and growth rate. Figure 2 provides a generalized plot of the substrate concentration, biological solids concentration (biomass), and oxygen utilization during exertion of BOD, versus time in a respirometric biodegradability treating vessel system. A typical relationship between the BOD and substrate degradation curves as well as between biomass and substrate concentrations and growth yield is illustrated in Figure 3.

Possible curves of oxygen uptake attributed to the organic test substrate that can be generated in a respirometric run and which are dependent on the acclimation time, the extent and rate of bio-oxidation of the substrate, the presence of cometabolite(s) and the presence of biocatalytic additive in the nutrient solution are demonstrated in Figure 4.

Factors affecting respirometric BOD determination have been taken into consideration in the study, and are listed in Table 4.

Chemical Analysis

Indirect analysis of culture samples from respirometric vessels during the study included determination of chemical oxygen demand as soluble and total COD and of dissolved oxidizable carbon (DOC) with the use of a Beckman Model 915 B system.

Specific substrate analysis of culture samples for the residual parent compound and the possible intermediate and final products of metabolism was performed with the use of a Gas Chromatograph Varian 3700, equipped with FID and EC detectors and with different detector columns (depending on the substrate to be analyzed) and the Finnigan automated GC/EI-CI Mass Spectrophotometer system. Samples were analyzed as dichloromethane extracts of cultures (continuous liquid-liquid extraction) or by direct aqueous injection.

Suspended solids and biomass were measured gravimetrically (solids retained on a 0.45 μm filter) and by optical density determined spectrophotometrically (percent transmission read at 540 nm).

Determination of Substrate Biodegradability from Oxygen Uptake Data

In this study, biodegradation was measured by three approaches: the first, as the ratio of the measured BOD values in mg/L (oxygen uptake values of test compound minus inoculum control - endogenous oxygen uptake values) to the theoretical oxygen demand (ThOD) of substrate as a percent; the second as a percentage of the test compound as measured by dissolved organic carbon (DOC) changes [OECD Guidelines for Testing of Chemicals (Method DGXI 283/82, Revision 5) (56)]; and the third, as a percentage of the test compound as measured by specific substrate analysis.

Graphical representation of percent biodegradation based on the BOD/ThOD ratio were developed against time for each test compound. The experimental DOC data for the initial samples and samples for reaction flasks collected at the end of experimental run were used to calculate the percent biodegradation based on the percent of DOC removal in the culture system.

Determination of Kinetic Parameters of Biodegradation

The Monod equation, relating cell growth to biomass and substrate concentration and the linear law, relating cell growth to substrate removal are the most popular kinetic expressions which can provide adequate description of growth behavior during biodegradation of substrate. The Monod relation states that cell growth is first order with respect to biomass concentration (X) and mixed order with respect to substrate concentration (S) by tbe equation

$$dX/dt = (Su_mX)/(K_s + S) \quad (1)$$

Cell growth is related to substrate removal by the linear law by the equation

$$dX/dt = -Y(dS/dt) \quad (2)$$

The kinetics of biodegradation were evaluated by quantifying μ_m, K_s and Y kinetic parameters expressed in the equation for the rate of substrate removal, r_s:

$$r_s = -\mu_m X/Y \quad (3)$$

where X is the concentration of biomass capable of utilizing the organic substrate and Y is the biomass yield coefficient for the compound and in the Monod equation:

$$\mu = \mu_m S_s/K_s + S_s \quad (4)$$

(if the compound is not inhibitory to its own biodegradation) or by the Haldane equation:

$$\mu = \mu_m S_s/K_s + S_s + (S_s^2/K_I) \quad (5)$$

if the compound is inhibitory. In these equations, μ_m is the maximum specific growth rate, K_s is the half saturation coefficient, K_I is the inhibition coefficient and S is the concentration of substrate.

A graphical presentation of the Monod substrate utilization equation is illustrated in Figure 5 which provides a relationship between rate of substrate utilization and substrate concentration.

Determination of Rates of Exponential and Declining Growth

The first order kinetic rate constants (specific growth rate parameters) were determined by the linearization of the BOD curves or transforming the typical BOD curve to the linear function of time t, by the relationship of log dO_u/dt to t, which gives straight lines expressing the exponential and declining endogenous phases of the BOD curve as shown in Figures 6 and 7. The slope of the Ln(d oxygen uptake/dt) versus t give specific rate constants of the exponential growth phase (μ values) and the declining growth phase (μ' values) of the BOD curve as described by Dojlido (11), Tabak et al. (45), Oshima et al. (46), and Tabak et al. (54, 55).

Acclimation time values (t_0) and the time values for the initiation and termination of the declining growth phase (t_1 and t_2) for each test compound were determined from linearized expressions of BOD curves.

Estimation of Monod Kinetic Parameters

The estimations of the Monod Kinetic parameters, maximum specific growth rate constant, μ_m, half saturation constant, K_s and growth yield constant, Y were determined directly from experimental oxygen uptake curves without the consideration of initial growth and growth yield assumption [Jobbagy, Grady and Tabak (57) and Tabak et al. (54)].

If the concentrations of the substrate, the products and the biomass are all expressed in BOD units, then the oxygen uptake (O_u at any time in the batch reactor may be calculated from

$$O_u = (S_{so} - S_s) - (X-X_o) - (S_p-S_{po}) \quad (6)$$

where S_{so}, S_{po} and X_o are the concentrations of substrate, products and cells, respectively, at time zero, and S_s and S_p are the concentrations of substrate and product, respectively, at time t.

To apply equation (4) for the determination of kinetic coefficients, equations must be available which express the concentrations of soluble substrate (S_s), soluble product (S_p) and biomass (X) as functions of time. For batch reactors, those equations are:

$$dS_s/dt = -(\mu_m/Y)S_sX/K_s + S_s) \quad (7)$$

$$dS_p/dt = (Y_p\mu_m/Y)S_sX/(K_s + S_s) \quad (8)$$

$$dX/dt = \mu_mS_sX/(K_s + S_s) - K_sbX/(K_s + S_s) \quad (9)$$

where S_s = soluble substrate concentration; S_p = soluble product concentration; Y_p = product yield; and b = decay coefficient.

To calculate oxygen uptake in a batch reactor, equations (7), (8) and (9) must be solved simultaneously, and the resulting values of S_s, S_p and X over time are substituted into equation (6).

Determination of Y Constant

Y - the true yield parameter or the ratio of growth of biomass to substrate utilization, can be obtained from the experimental oxygen uptake curve at the initiation of the plateau of the curve as shown in Figure 8, with the use of equation:

$$Y = (1-O_{upt}/S_o)-Y_p \quad (10)$$

Where O_{upt} is the cumulative oxygen update value at the initiation of plateau, S_o is the concentration of substrate at time zero, and Y_p is the product yield.

A vertical line is drawn at the point of intersection of the tangents of the exponential and plateau phases of the curve. The oxygen uptake value obtained at the point of intersection of the vertical line (drawn through intersection of tangents) and oxygen uptake curve is the O_{upt} value - [cumulative oxygen uptake value at the initiation of plateau].

The O_{upt} value is then substituted into equation (10) , for Y determination. Y_p is soluble product concentration formed, divided by initial substrate concentration. In this study, product yield (Y_p) was negligible.

Determination of μ_m Constant

The initial estimate of μ_m is obtained by the technique of Gaudy et al. (47, 48). If Y_p is assumed to be zero, equation (8) is eliminated and if b (decay constant) is assumed to be zero, equation (9) is simplified to

$$dX/dt = (\mu_m S_s X)/(K_s + S_s) \qquad (11)$$

Combining equation (7) and (11) and integrating from S_{so} to S_s and from X_o to X gives:

$$X = X_o + Y(S_{so} - S_s) \qquad (12)$$

If the assumption that $S >> K_s$, the term $S_s/K_s + S_s$ in equations (7) and (11) or (1) approaches one, and these systems can be simplified through the use of equation (12) to give equation

$$dX/dt = \mu_m X \qquad (13)$$

Integrating equation (13) and combining with equation (2) and than substituting in equation (6) with P and P_o both equal to zero gives

$$Ln[X_o + O_u/(1/Y)-1] = Ln(X_o) + \mu_{m(est)} \qquad (14)$$

The plot of $Ln[(X_o + O_u)/((1/Y) - 1)]$ versus time will give a straight line with slope μ_m. The accuracy of μ_m(est) will depend upon the size of K_s, but this value is good enough as an estimate.

μ_m - the maximum specific growth rate can be determined from experimental oxygen uptake curve plot in the following manner:

(1) Values of the change of O_u with time (dO_u/dt) or slopes are determined along the entire experimental oxygen uptake curve as shown in Figure 9.

(2) These dO_u/dt (slope) values are then plotted against the cumulative O_u values for each time interval, as shown in Figure 10.

(3) The slope of the developed linearized form of oxygen uptake curve is the estimated μ_m value.

Determination of K_s Constant

K_s - the half saturation constant or the substrate concentration at which the specific growth rate is 1/2 the maximum specific growth rate can be obtained from the experimental oxygen uptake curve in the following manner:

(1) Value of O_{ut} can be calculated from the plot of (dO_u/dt) versus O_u provided the K_s value is 1 or less (insignificant in comparison to S_o value) and the plot contains a linear section with the slope μ_m, as shown in Figure 11.

(2) Other (dO_u/dt) versus O_u plot in which the slope deviates from μ_m because of larger K_s values (more significant in comparison to S_o) is illustrated in Figure 12.

(3) The value of dO_u/dt is determined at the intercept of the straight line developed from the plot of dO_u/dt versus O_u (Figure 11) which contains a linear section with slope μ_m.

(4) Beginning with the value of 1/2 the intercept value, another straight line (b) is constructed with the slope 1/2 that of the slope of original line (a) whose slope is μ_m.

(5) At the point where line (b) intercepts the declining experimental curve of the plot, a vertical line from that point of interception can provide the value of O_{ut} on the x axis.

(6) This O_{ut} value is then used in the determination of K_s with the use of the equation

$$S_t = S_o - O_{ut}/(1-Y_p-Y) = K_s$$

where S_o = initial substrate concentration and S_t = substrate concentration at time t.

(7) When the O_{ut}, Y, Y_p, and S_o values are plugged into the equation, the value of S_t can be calculated - which is the value of K_s (in systems where K_s value is 1 or less).

Thus the oxygen uptake value O_{ut} associated with 1/2 of the estimate of u_m is used in $K_s = S_o - O_{ut}/(1-Y_p-Y)$ to get the estimate of K_s. The major impact of K_s is upon the shape of the oxygen uptake curve in the region of the plateau. Comparison of the experimental curves to a family of standardized curves as an initial estimate provides an initial estimate of K_s that is sufficient for non-linear curve fitting techniques for quantitation of the kinetic parameters.

Quantitation of Monod Kinetic Parameters

The methodology for quantitation of the Monod kinetic parameters requires the use of the above specific methods for estimating them initially and subsequently followed by computer simulation methods coupled with non-linear curve fitting techniques and is based on the use of measured values of initial growth and growth yield. The method requires the use of the kinetic equation relating growth rate of biomass in presence of substrate, the substrate utilization rate, product formation rate and rate of oxygen consumption from O_2 uptake (BOD) curves to calculate and use the theoretical oxygen consumption data to quantitate the biokinetic parameters.

The determination of the kinetic parameters associated with biodegradation requires a series of steps. The initial substrate (S_{so}) and biomass (X_o) concentration must be carefully measured in COD units. The ratio of the two values must lie in a certain range in order to allow independent evaluation of μ_m, K_s and Y [Simkins and Alexander (38, 39)]. Grady's studies (49, 50) have shown that a S_{so}/X_o ratio of around 20 works

well. The value of Y_p may be estimated by determining the residual stable SCOD concentration after substrate depletion (plateau area). It is numerically equal to the residual SCOD divided by the initial SCOD. The value of the decay coefficient, b, may be determined by fitting to the oxygen consumption curve after the plateau when the only activity contributing to oxygen consumption is endogenous metabolism and cell decay. Once X_o, S_{so}, Y_p and b are known, μ_m, K_s and Y may be determined by non-linear curve fitting techniques [Grady (49, 50)].

The technique involves the calculation of a theoretical oxygen consumption using oxygen uptake equation (4) and equations for substrate, product and biomass concentrations (5) - (7) with assumed Monod parameters. The residual sum of squared errors (RSSE) associated with the difference in calculated and experimental oxygen uptake values is used to obtain new estimates. The above procedure is repeated until a minimum RSSE is found.

The Grid Search technique was selected as a most suitable non-linear curve fitting technique for application in the determination of the kinetic parameters from oxygen uptake data, because it can allow easy discrimination between local minima and the global minimum RSSE. This technique enables a comparison between the calculated and experimental oxygen uptake data. The value of Y is fixed. For this value of Y, a pair of μ_m and K_s which give RSSE is found on a μ_m:K_s plane. The above procedure is repeated with other values of Y. Values of μ_m, K_s and Y which give minimum RSSE associated with the difference in calculated and experimental oxygen uptake data constitute the best values of the kinetic parameters.

The values of μ_m, K_s and Y developed from grid search technique, which when substituted into equations 5-7, will provide X and S values, which (when substituted into oxygen uptake equation 4) will in turn provide calculated oxygen uptake values at the region of the plateau, closest to the experimental oxygen uptake values, with a minimum RSSE, will constitute the best quantitative kinetic parameter values.

Development of Multi-Level Respirometric Biodegradation Testing Protocol

The oxygen consumption data generated with the use of electrolytic respirometry have been adequately utilized for assessing the biodegradative activity of sludge microbiota, the biodegradability/toxicity of toxic organic compounds, as well as for the determination of the intrinisic kinetic parameters of biodegradation. Methodologies have been developed for quantitating biodegradability and biodegradation kinetics of representative classes of RCRA toxic organics, with the resultant development of a comprehensive multi-level respirometric biodegradation testing protocol based on oxygen consumption data.

The methods of each successive testing level of the respirometric protocol are characterized by increased complexity and a consequent higher cost for performing the tests pertaining to each testing level. Accordingly, the testing levels can be selected as appropriate to the research needs.

Studies to assess the biodegradability and biodegradation rates of toxic organics by sludge microbiota with the use of respirometric oxygen uptake data can involve the use of one of more levels of the protocol, depending on the amount of information needed for assessing the biodegradability of the organic toxic pollutant or the toxicant bearing waste for determination of its fate and rate of biotreatment in the municipal or industrial waste treatment systems. A logic flow diagram of the respirometric biodegradation testing protocol, providing a brief description of each of the testing levels, is shown in Figure 13.

RESULTS AND DISCUSSION

Respirometric biodegradability, biokinetic and Monod kinetic data for selected RCRA alkyl benzenes, phenols, phthalates and ketones are reported in this paper. The electrolytic respirometry oxygen uptake data for the test compounds, the control reference compound aniline, the inhibition and endogenous control systems were generated revealing the lag phase (acclimation phase), the biodegradation (exponential) phase, the different bio-reaction rate slopes (characteristic of the test compound) as well as the plateau region at which the biooxidation rate reaches that of the endogenous rate of microbial activity. Figure 14 illustrates a representative oxygen uptake curve for aniline and the endogenous controls. Figure 15 shows the replicate pentaerythritol oxygen uptake curves and the toxicity control (pentaerythritol plus aniline) curve, Figure 16 illustrates a representative graphical treatment of the percent biodegradation of pentaerythritol with time, which was developed for each test compound (OECD studies).

Based on the biokinetic equations relating growth rate of microbiota in presence of above compounds, the substrate utilization rate, and rate of oxygen uptake (BOD) curves, specific growth rate kinetic parameters (biodegradation rate constants) were derived as slope values of the linearized plots (plots of the log of DO_u/dt) of exponential and declining growth phases of the BOD curve. The acclimation time values (t_0), and time values for the initiation and the termination of the declining growth phases (t_1 and t_2) for the test compounds and aniline were also generated.

The estimations of the Monod kinetic parameters for benzene, phenol, phthalate, and ketone compounds reported here, were determined directly from experimental oxygen uptake curves without the consideration of initial growth and growth yield assumption.

Respirometric Studies with Selected RCRA Alkyl Benzene Compounds

The biodegradation of benzene, toluene, ethyl benzene, m- and p-xylenes, tert-butyl benzene, sec-butyl benzene, butyl benzene, cumene, 1-phenyl benzene and the reference compound, aniline at 100 mg/L concentration by 30 mg/L sludge biomass (as measured by oxygen consumption by sludge microbiota in mg O_2/L) was followed over a period of 20 days. The electrolytic respirometry oxygen uptake and BOD curves were generated and graphical treatment of the percent biodegradation was established for each

compound. Figure 17 demonstrates typical oxygen uptake and BOD curves for p-xylene and p-xylene + aniline and Figure 18 illustrates graphically the % biodegradation of p-xylene with time.

The percent biodegradation data based on the BOD/ThOD ratios for benzene, toluene, ethyl benzene, m- and p-xylene and the reference compound, aniline, are summarized in Table 5. All of the above alkyl benzene compounds were shown to be biodegradable substrates at concentration levels of 100 mg/L when exposed to 30 mg/L of activated sludge biomass under the environmental conditions of the respirometric testing procedure, and within the period of 20 days of incubation.

The toxicity test control flask respirometric data revealed no inhibitory effects by these test compounds at the 100 mg/L concentration levels on the bio-oxidation of aniline by sludge microbiota.

Table 6 summarizes the bio-kinetic data for the benzenes studied, showing the specific growth rate constants for the exponential growth phase (μ values) and for the declining growth phase (μ' values) of the linearized form of the BOD curves of these compounds, as well as the t_0, t_1 and t_2 kinetic parameters. Figure 19 shows a typical plot of $Ln(dO_u/dt)$ vs. time for toluene, from which the kinetic parameters were determined.

Table 7 summarizes the Monod kinetic parameter (μ_m, K_s, Y_g) data for these benzene compounds.

Respirometric Studies with Selected RCRA Phenolic Compounds

The biodegradation of phenol, resorcinol, o-, m- and p-cresols, catechol, 2,4-dimethyl phenol and the reference compound aniline at 100 mg/L concentration levels and exposed to 30 mg/L biomass was followed over a period of 20 days.

All of the phenols were shown to be biodegradable substrates under the conditions of the respirometric testing procedure. The toxicity test control flask respirometric data revealed no inhibitory effects by these compounds at the 100 mg/L levels on the biodegradation of aniline by the sludge biomass.

Table 8 summarizes the bio-kinetic data for the phenols studied, showing the specific growth rate constants as well as the t_0, t_1, and t_2 kinetic parameters. Table 9 provides the Monod kinetic parameter data for these phenolic compounds.

Respirometric Studies with Selected RCRA Phthalate Ester Compounds

Evaluation of the biodegradability and determining of bio-kinetics of degradation of phthalate compounds, dimethyl phthalate, diethyl phthalate, dipropyl phthalate and butyl benzyl phthalate was achieved with use of respirometric oxygen uptake data.

All of the above phthalates were shown to biodegradable under the conditions of the respirometric tests and were shown not to exhibit any inhibitory effects at the 100 mg/L levels on aniline biodegradation by the sludge microbiota.

Tables 10 and 11 summarize respectively the biokinetic (first order) and Monod kinetic parameter data for the selected phthalate esters under study.

Respirometric Studies with Selected RCRA Ketone Compounds

Respirometric oxygen uptake data from the studies with the selected ketone compounds, acetone, 2-butanone, 4-methyl-3-pentanone and a cyclic ketone, isophorone were utilized to determine their biodegradability and biodegradation kinetic parameters.

All the ketones were shown to be biodegradable at 100 mg/L concentration levels in media containing 30 mg/L biomass and did not exhibit any toxicity to aniline biodegradation at these concentrations.

Tables 12 and 13 summarize respectively the first order and Monod kinetic parameter data for these ketones.

CONCLUSIONS

The experimental data of respirometric studies with several classes of organic compounds definitely demonstrate that it is possible to measure the biodegradability (percent biodegradation - as a ratio of BOD to ThOD) and to determine the kinetics of degradation of single organic compounds by using only measurements of oxygen consumption in respirometric batch reactors. The values of the kinetic parameters determined from oxygen consumption data were demonstrated to be similar to those based on the measurements of substrate removal and those made with cell growth data.

The generated data on biodegradation, biodegradation rates and substrate inhibition kinetics through the use of electrolytic respirometry, will enable the classification of biodegradability of toxic priority pollutant and RCRA toxic organic compounds and ultimate projection of the fate of organic compounds of similar molecular structure to those experimentally studied by way of the established predictive treatability models based on structure-activity relationships.

With the electrolytic respirometry approach, a data base on the removal of the above compounds by biodegradation fate mechanism can be adequately generated to support the development of predictive models on fate and removal of toxics in industrial and municipal waste treatment systems. A possible relationship between the kinetic parameters and the effect of different factors on these parameters, as determined through electrolytic respirometry and the structural properties of the organic pollutant, can eventually facilitate prediction of the extent and the rate of biodegradation of organic chemicals in the field of wastewater treatment systems from the knowledge of the structural properties of the pollutant organics.

A preliminary predictive biodegradation - structure/activity model based on the group contribution approach was developed from the generated biodegradation kinetic data (first order kinetic parameters) with the use of electrolytic respirometry. It is expected that the model will closely predict the results found experimentally. In this way, the fate of other organic compounds may be anticipated without the time and expense of experimental work.

The electrolytic respirometry biodegradation studies will provide basic pilot scale treatability information and data which will be used to confirm methods to predict treatability and the need for pretreatment of structurally related pollutants (e.g., by structure, anticipated treatability properties, etc.). This study will thus provide a more extensive list of pollutants than was covered by experimental data, for consideration in guiding the Agency to predict the fate of such compounds without costly experimental testing.

Studies are currently in progress to determine the effect of temperature, and different sources of sludge biomass (domestic and industrial wastewater treatment) on the biodegradation kinetics derived from electrolytic respirometry.

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TABLE 1

ADVANTAGES OF RESPIROMETRIC METHODS

1. Oxygen uptake can be monitored continuously and constantly and more precisely.

2. Automation provides numerical or binary output data for direct recording or processing (electrolytic).

3. Sample may not require dilution and therefore its oxygen uptake characteristics are measured in a more natural state.

4. Larger sample volumes can be used so that more representative samples are obtained and sampling errors are minimized.

5. The samples are mixed continuously to provide uniform contact of microorganisms, substrate and oxygen.

6. No chemical titrations are required.

7. A continuous record of O_2 uptake is provided with some units having automatic recording devices.

8. Permit treatment plants and in-stream conditions to be simulated more closely than in dilution test.

9. Convenient for measuring the effect of various factors on oxygen uptake, such as dilution, substrate type and concentration, temperature and presence of toxics, effect of ph, nutrient addition, volume and source of sewage seed and seed adaptation.

10. Can be used to determine bacterial growth and substrate removal coefficients.

11. Much lower coefficient of variability can be obtained with respirometry than with the dilution test because of the larger and more representative samples used and the lack of dilution factor.

12. Conditions in nature and in a treatment plant may be simulated more closely in respirometer than in a BOD bottle.

13. Usefull information is often available in very much less than in 5 days.

14. It is possible to stop the test at a recognizable point on the oxygen curve (such as the beginning of the "plateau" which corresponds to the exhaustion of readily oxidized substrate).

15. Recommended for use in: A) routine examination of sewage and trade wastes and in B) control of sewage treatment processes, as well as for C) research studies because of the ease and precision with which variables can be controlled.

TABLE 2 GENERAL CLASSIFICATION OF RESPIROMETERS

Types	Basic Principle of Operation
1. Manometric	Determination of O_2 weight changes in a closed system by measuring or responding to O_2 pressure changes at constant temperature and volume or volume changes at constant pressure.
2. Electrolytic	Same.
3. Dissolved O_2 depletion	Use of dissolved oxygen probe to make direct measurements of depletion of dissolved O_2 from solution.

TABLE 3 TYPES OF RESPIROMETERS BASED ON TECHNIQUES AND APPLICATIONS

I. Respirometers measuring gas exchange

A. Small constant pressure respirometers –
Measurements are made by observing the change in volume of gas-phase in contact with the respiring liquid, as gaseous O_2 is absorbed.

B. Small constant volume respirometers –
The changes in pressure due to O_2 uptake is observed. Calibration is necessary.

C. Large respirometers measuring gas exchange –
Recommended especially for studies of treatability.

D. Electrolytic respirometers –
Oxygen pressure is automatically maintained at a constant value by an electrolysis cell.

II. Respirometers in which gas exchange is not measured
Oxygen uptake (BOD) is measured by means of a solid O_2 electrode.

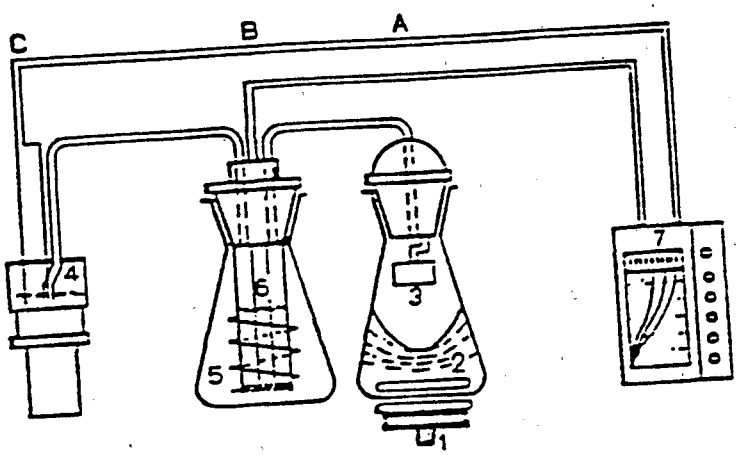
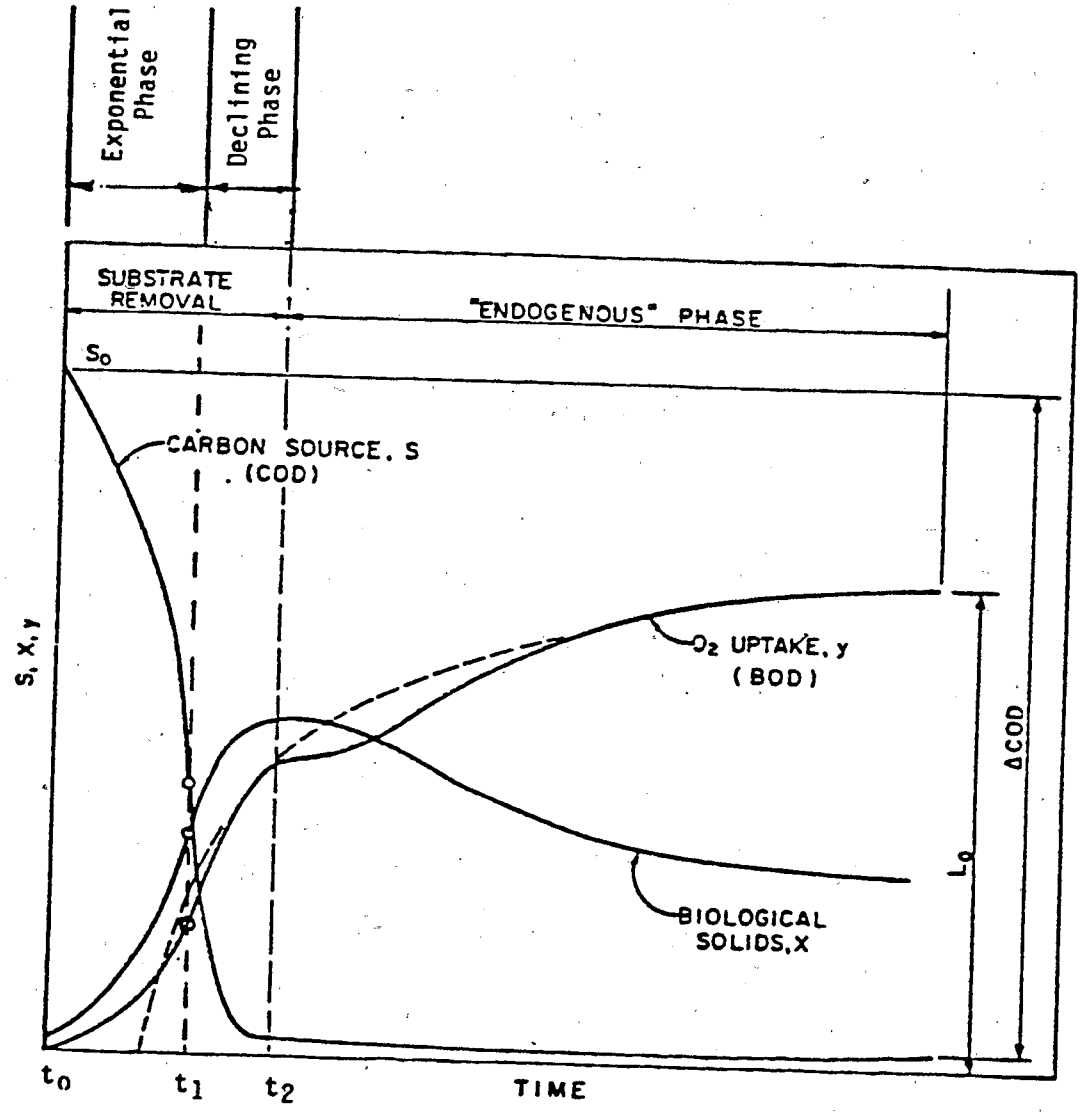


Diagram of a measuring unit
A reaction vessel
B oxygen generator
C pressure indicator

1 magnetic stirrer
2 sample (250 ml)
3 CO_2 absorber
4 pressure indicator
5 electrolyte
6 electrodes
7 recorder

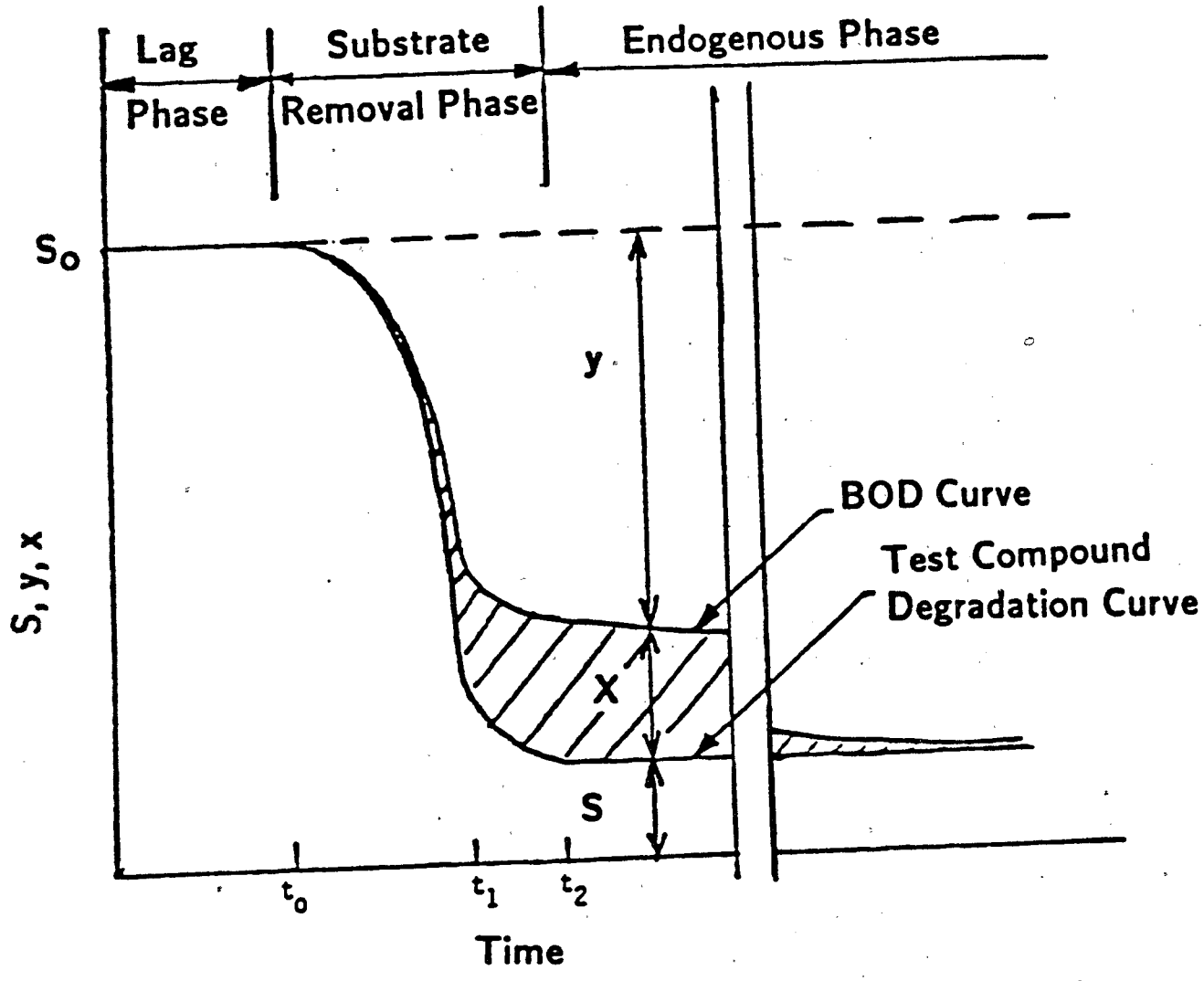
SCHEMATIC DIAGRAM OF A MEASURING UNIT

FIGURE 1



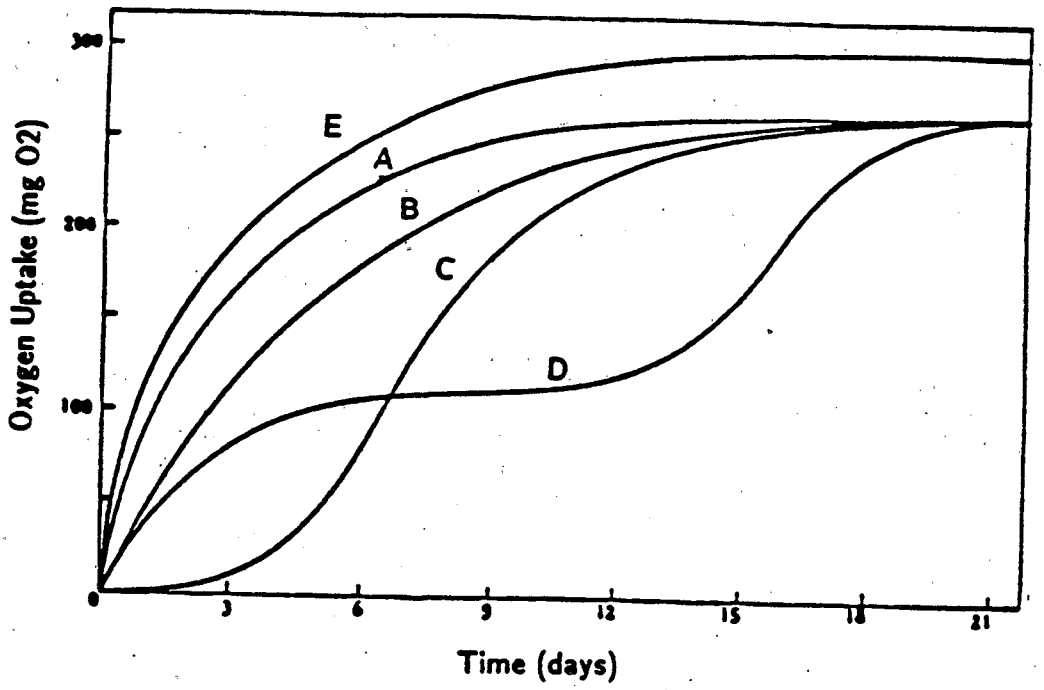
Generalized plot of substrate concentration, biological solids concentration, and oxygen utilization during exertion of biochemical oxygen demand. Circles mark inflection points.

FIGURE 2



The schematic diagram for the relationship of S, y and x.

FIGURE 3



A – O_2 uptake curve of a rapidly oxidizing substrate.

B – O_2 uptake curve of a rapidly oxidizing substrate with lower rate of O_2 uptake than in (A).

C – O_2 uptake curve with an initial lag period, demonstrating oxidation of substrate by microbiota only after undergoing a period of acclimation.

D – O_2 uptake curve of a non–homogenous substrate showing immediate oxidation of part of the substrate followed by acclimation period before oxidation of the remaining part of substrate. Illustration of catabolite repression.

E – O_2 uptake curve of biodegradable substrate showing higher rate of O_2 uptake (higher oxidation rate) and an increase of total cumulative O_2 uptake attributed to biocatalytic additive than in (A) and (B) curves for a substrate in presence of only the control seed organisms.

POSSIBLE CURVES OF THE OXYGEN UPTAKE ATTRIBUTED TO THE PROBLEM SUBSTRATE

FIGURE 4

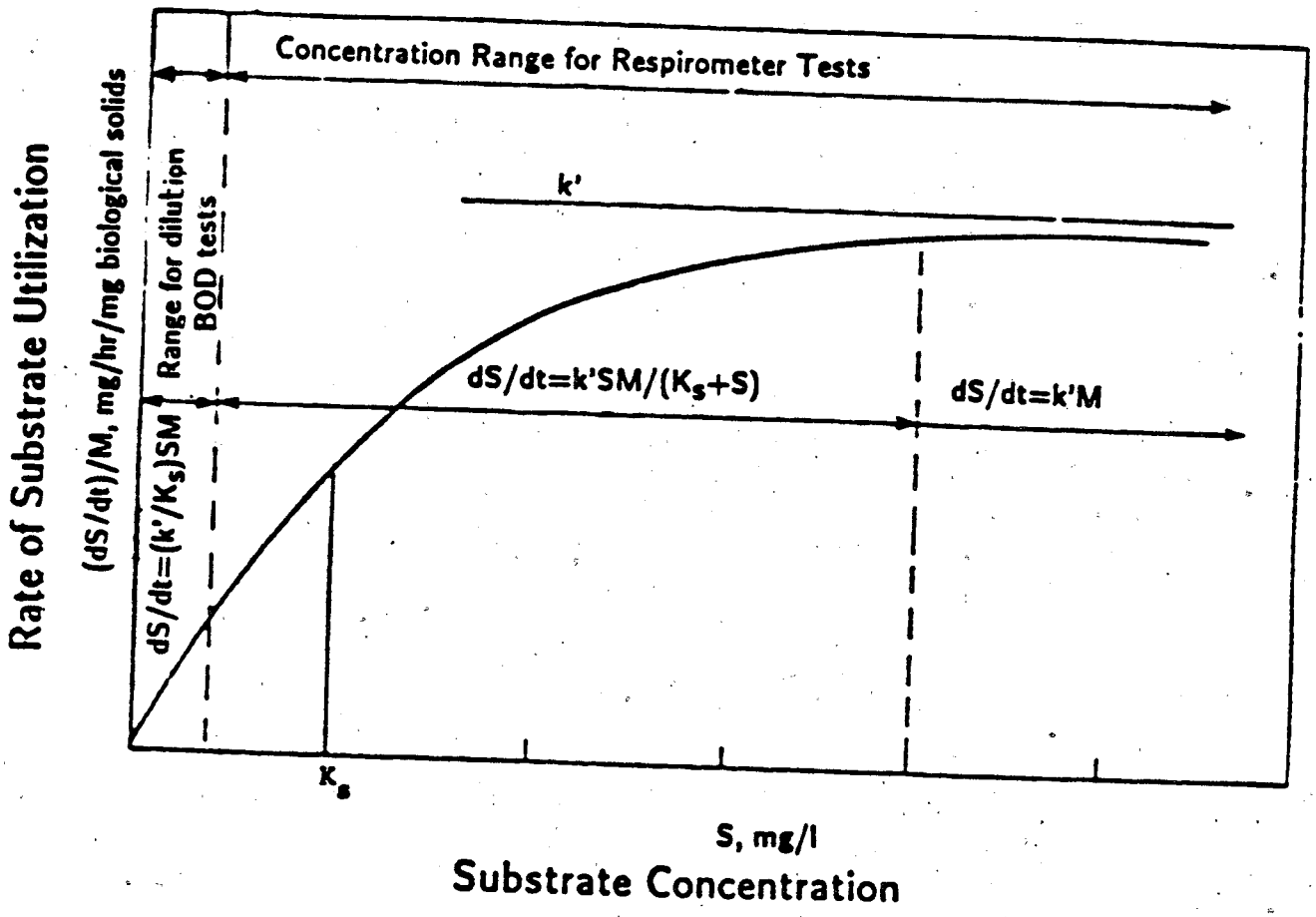
TABLE 4 FACTORS AFFECTING RESPIROMETRIC BOD DETERMINATION

Affect both microorganism growth and substrate utilization rates

- Population dynamics – growth characteristics
- Substrate concentration – nutrient deficiency
- Rate of oxygen transfer or CO_2 evolution
- Type of inoculum
  - A. General – Indigenous microbial populations
  - B. Addition of specific microbiota to sample
  - C. Protozoa
- Concentration of inoculum
- Dissolved oxygen concentration
- Toxicity
- Nitrification
- Inhibition of protein synthesis

Physico – chemical factors

- Temperature
- pH value and buffering capacity
- Effect of light
- Air bubble effect
- Mixing
- Turbulence
- Nutrient concentration Mineral nutrients
- Storage of sample before analysis



Graphical presentation of the "Monod" substrate utilization equation. The reaction approaches first order at low substrate concentration and a zero—order at high substrate concentrations.

FIGURE 5

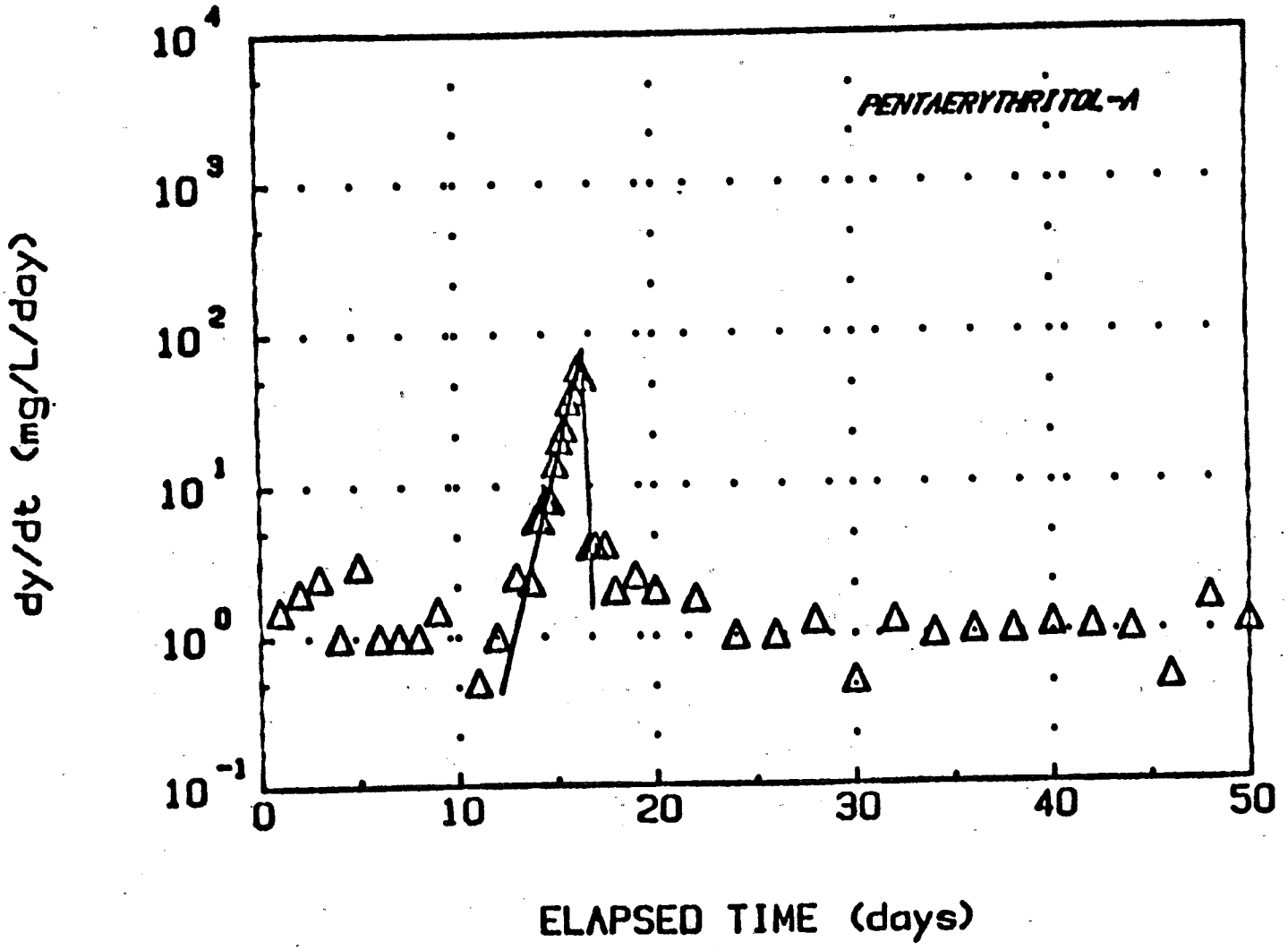


FIGURE 6 Log (dy/dt) vs t Plot
(PENTAERYTHRITOL)

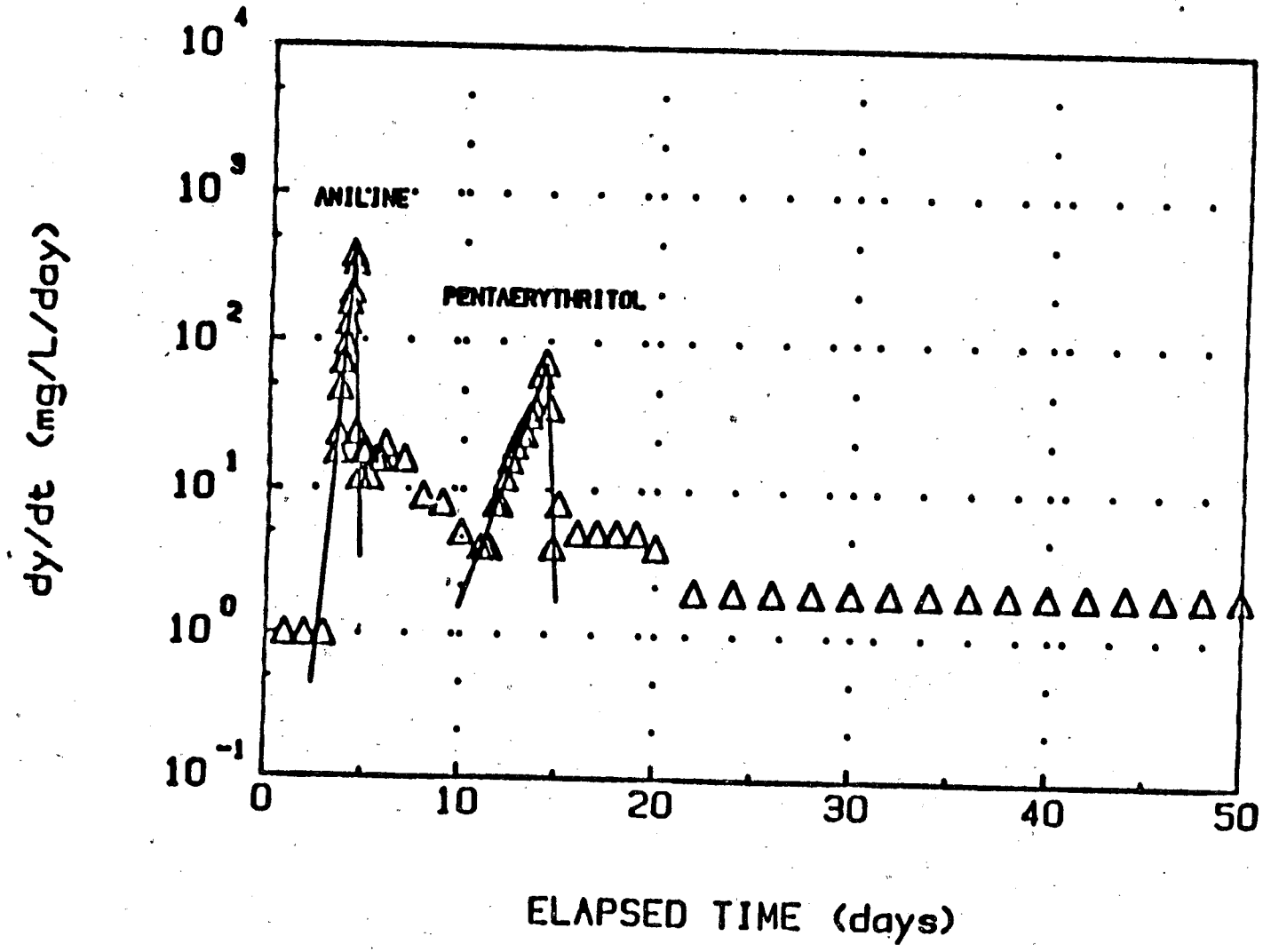


FIGURE 7 Log (dy/dt) vs t Plot
(PENTAERYTHRITOL & ANILINE)

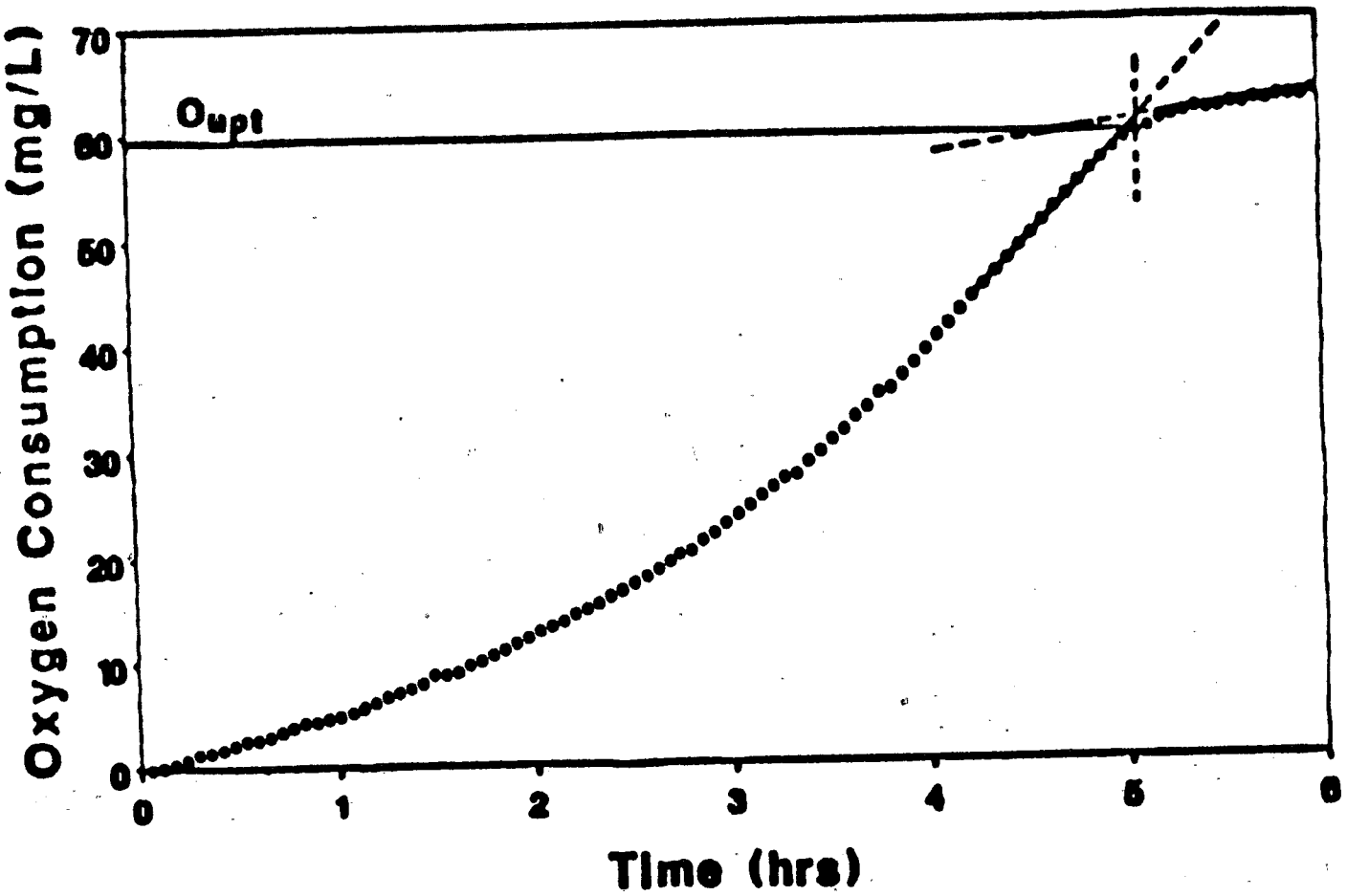


FIGURE 8 DETERMINATION OF THE START OF THE PLATEAU IN OXYGEN CONSUMPTION FOR USE IN ESTIMATING Y_g

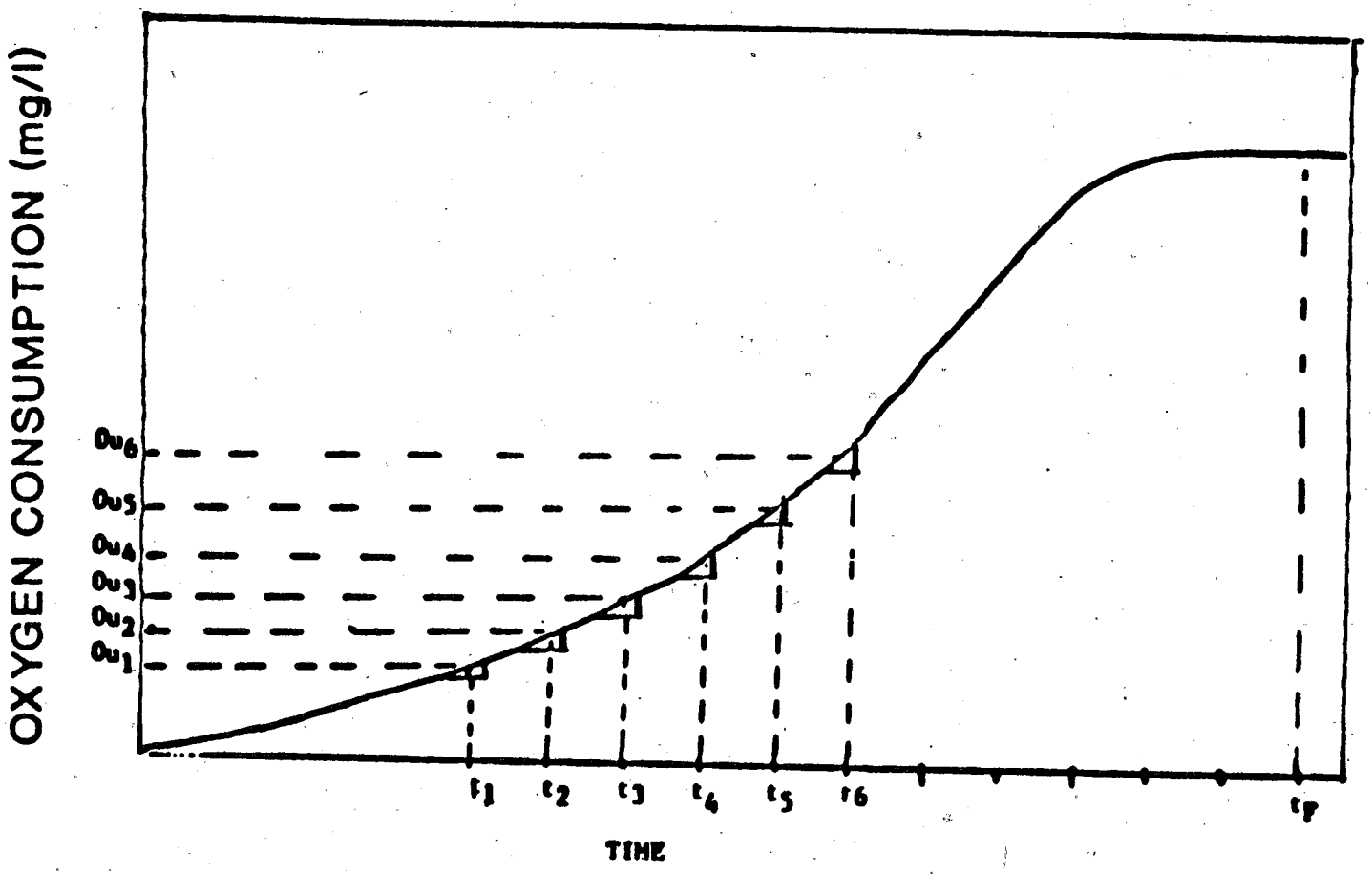


FIGURE 9 PLOT OF OXYGEN UPTAKE RATE
(Ou VERSUS TIME)

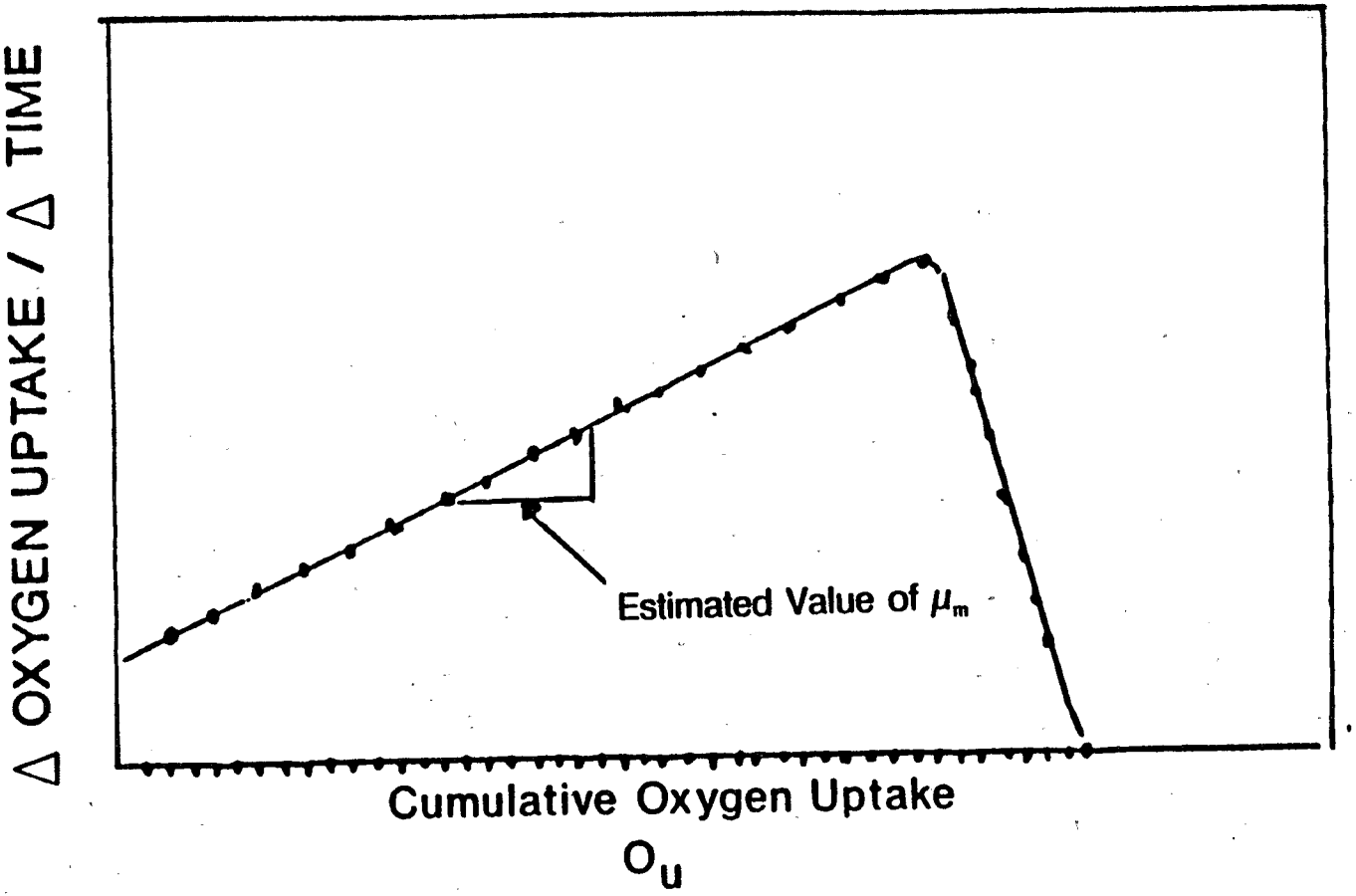
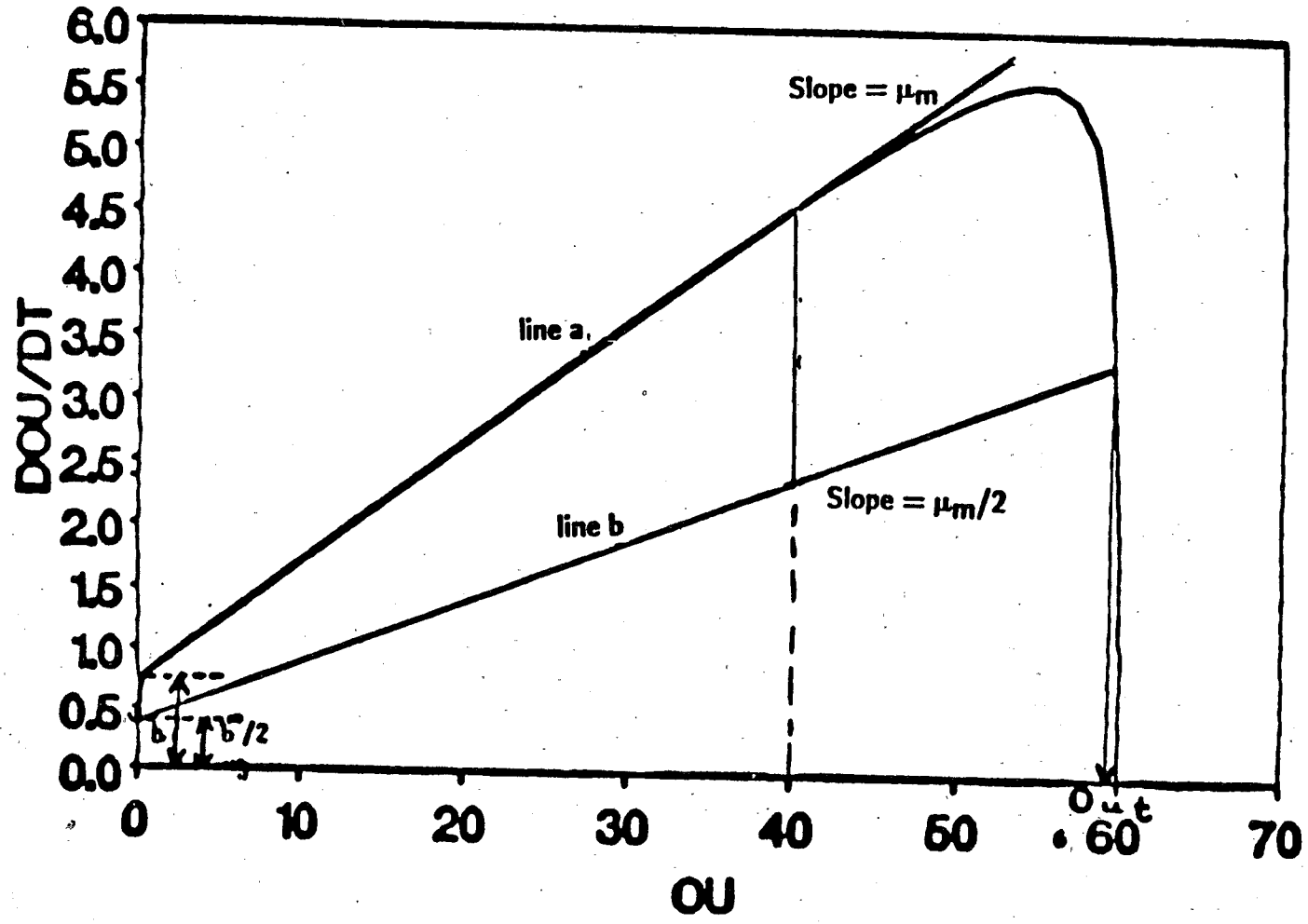


FIGURE 10 PLOT OF $\Delta Ou / \Delta t$ VALUES OR SLOPES VERSUS THE CUMULATIVE Ou VALUES



UM=0.1, KS=1, YG=0.4, X0=5, S0=100

FIGURE 11 Plot showing ($\Delta Ou/\Delta t$) versus Ou_t curve containing a linear section with a slope μ_m to be used in estimating the values of K_s in cases where K_s values are negligible in comparison to S_0.

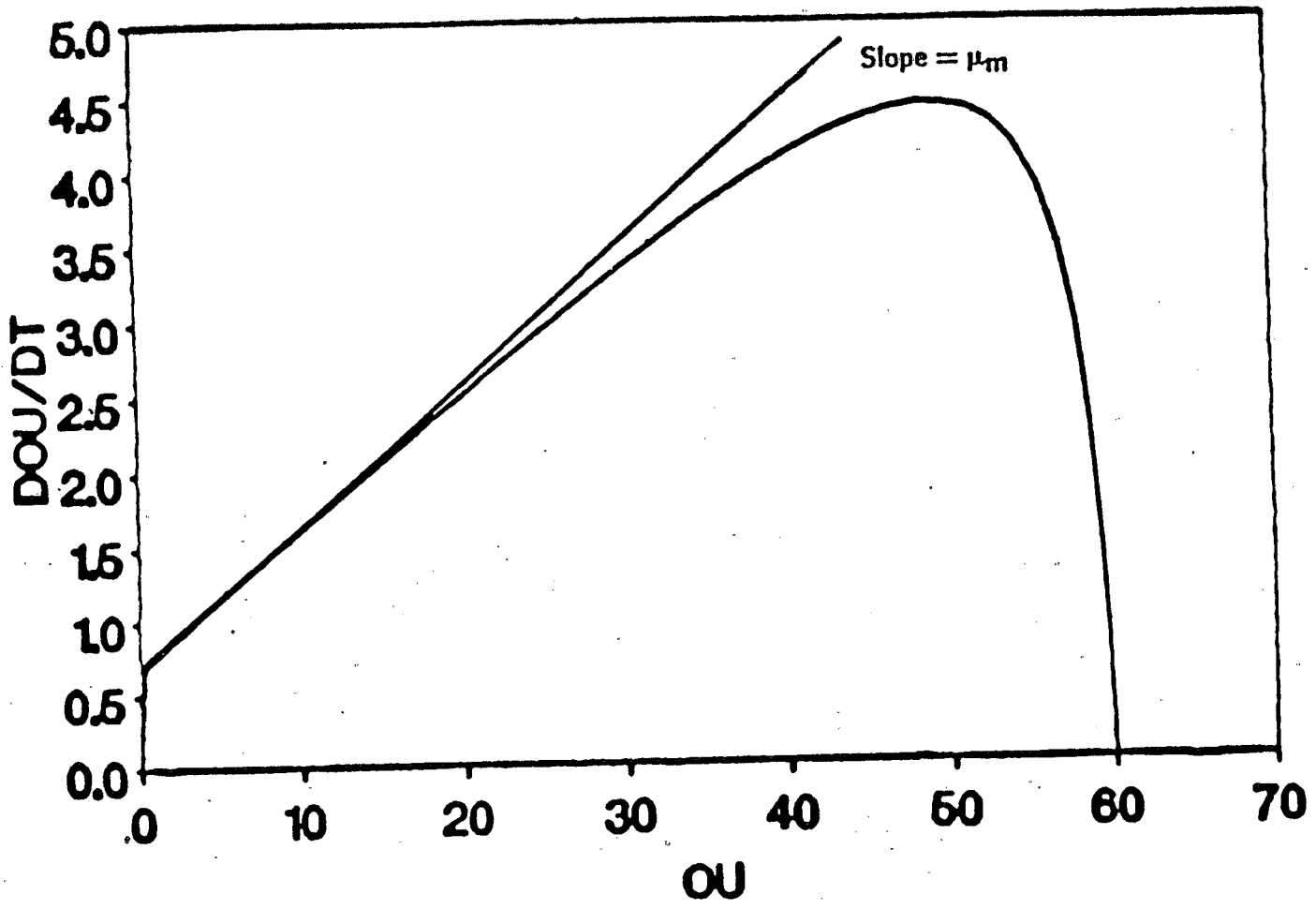


FIGURE 12 Plot of ($\Delta Ou/\Delta t$) versus Ou_t in which the slope deviates from μ_m because K_s values are more significant in comparison to S_0. ($K_s = 5$)

FIGURE 13. LOGIC FLOW DIAGRAM FOR THE RESPIROMETRIC BIODEGRADATION TESTING PROTOCOL

Level I -- Biodegradation Screening Method

Determination of Primary Biodegradation from Oxygen Uptake Data

Determination of Acclimation Time Acclimation Time Studies	Determination of Percent Biodegradation Biodegradability Testing Studies	Determination of Microbial Inhibition/Toxicity Substrate Inhibition/Toxicity Studies
Acclimation + or -	Biodegradation + or -	Inhibition + or -
(1) Determination of acclimation (lag) time of microbial biomass to toxic substrate (t_0 values)	(1) Measurement of the ratio of BOD to ThOD in mg O_2/L of the toxic substrate as a percent.	(1) Determination of the inhibitory effects of toxic substrate or the toxicant bearing waste on microbial biomass as a function of concentration.
(2) Determination of the initiation and termination time values for declining growth phase of BOD curve (t_1 and t_2 values)	(2) Measurement of the percentage of the residual toxic substrate by DOC value changes.	(2) Determination of the inhibitory effects of toxic substrate or toxicant bearing waste on the biodegradative activity of biomass on biogenic substrate(s).
	(3) Measurement of the percentage of the residual toxic substrate by the specific substrate analysis.	(3) Determination of substrate inhibitory levels to biomass and to normal rate of biodegradation.

FIGURE 13 (continued)

Level II -- Studies of Biokinetics of Biodegradation Based on Respirometric Oxygen Uptake Data

Determination of Specific Growth Rates and the Initial Estimates of Monod Kinetics of Biodegradation

Determination of the Rates of Exponential and Declining Growth Phase of BOD Curve Determination of μ and μ' Constants

Application of O_2 uptake data and BOD values to the determination of BOD kinetics. Utilizaiton of biokinetic data (exponential and declining growth rates) for biodegradation efficiency determination.

First order kinetic rate contstants (specific growth rate parameters) determined by linearization of BOD curves through a relationship of log (dOu/dt) to t which gives straight lines expressing exponential and declining phases of BOD curves. Slopes of these lines represent μ and μ' parameter values respectively.

Determination of the Monod Kinetic Parameters from the Experimental Oxygen Uptake Curves Determination of μ_m, K_s and Y Parameters

Application of O_2 uptake data and BOD values for the determination of the initial estimates of Monod kinetic parameters directly from experimental O_2 consumption curves without the consideration of initial growth and growth yield assumption.

These initial kinetic parameters estimate data can be successfully applied for biodegradation efficiency determination in wastewater treatment systems.

Specific methodologies have been developed for obtaining the initial estimates of the Monod kinetic parameters.

Use of pre-accclimated biomass or biomass acclimated in respirometric vessels.

FIGURE 13 (continued)

Level III -- Studies to Quantitate Monod Kinetic Parameters from Respirometric Oxygen Uptake Data

Use of Initial Estimates of Kinetic Parameters, u_m, K_s and Y, to Quantitate these by Means of Computer Simulation Methods Coupled with Non-Linear Curve Fitting Techniques

Methodology based on the kinetic relationship between growth rate, oxygen uptake rate and toxic substrate concentration in a mass balance equation to develop intrinsic kinetic parameter data for prediction of the fate of the substrate in full scale treatment systems

Methodology supported by Proof-of-Concept Data which show an agreement between the values of the kinetic parameters obtained from O_2 consumption data and those obtained from traditional measurement of substrate removal (DOC, SCOD, ^{14}C) or cell growth.

Methodology requires the use of kinetic equations relating growth rate in presence of toxic substrate, substrate utilization rate, soluble product formation rate and rate of oxygen uptake in order to calculate and use the theoretical O_2 consumption data (O_2 uptake curves) to quantitate the biokinetic parameters.

Methodology is based on the use of measured values of initial growth (X_0) (biomass concentration), substrate concentration (S_{so}) and growth yield (Y), product yield (Y_p) (determined by fitting O_2 uptake curve after the plateau (O_2 consumption due to endogenous metabolism and cell decay).

Once X_0, S_{so}, and Y_p and b (cell decay) values are known, the kinetic parametes, u_m, K_s and Y are quantitated by the non-linear curve fitting techniques. Initial kinetic parameter estimates are used in the method.

The non-linear curve fitting technique (Grid Search Method) was shown to be most applicable, since it allows easy discrimination between the calculated and experimental O_2 uptake data. RSSE associated with differences in calculated and experimental O_2 uptake values are used to obtain new esimates of kinetic parameters. This procedure is repeated till minimum RSSE is formed.

Grid Search Technique - Value of Y is fixed. For this value of Y, a pair of u_m and K_s values which give RSSE is found on a u_m:K_s plane. The above procedure is repeated with other values of Y. Values of u_m, K_s and Y which give minimum RSSE associated with the difference in calculated and experimental O_2 uptake data constitute the best values for the kinetic parameters.

Level III uses either pre-acclimated biomass or biomass acclimated in respirometric vessels.

FIGURE 13 (continued

Level IV -- Studies to Determine the Extent of Biodegradation

Tests for Assessment of Ultimate Biodegradability during Respirometric Experiments

Radioisotope Respirometry Electrolytic Respirometry Coupled with Radioisotope Technology	Respirometry Coupled with Specific Substrate Analysis
Use of ^{14}C labeled toxic substrates. Trap $^{14}CO_2$ and measure ^{14}C intermediate metabolites during incubation. Measure radioactivity of the liquid and gaseous phases of the contents in the respirometric vessel system by scintillation counter techniques. Liquid and gaseous samples for the radioactivity analyses can be taken intermittently during the respirometric run by means of specially adapted syringes from side arms of the respirometric vessel.	Analysis of the residual parent toxic substrate, intermediate metabolites and end products of metabolism by GC, GC/MS and HPLC technology. Liquid and gaseous samples of the culture system can be manually taken from side arms of the respirometric vessel with specially adapted syringes during the respirometric run. Automated analysis of the culture system can be performed by way of specially designed system coupling the respirometric vessels to the analytical instruments.

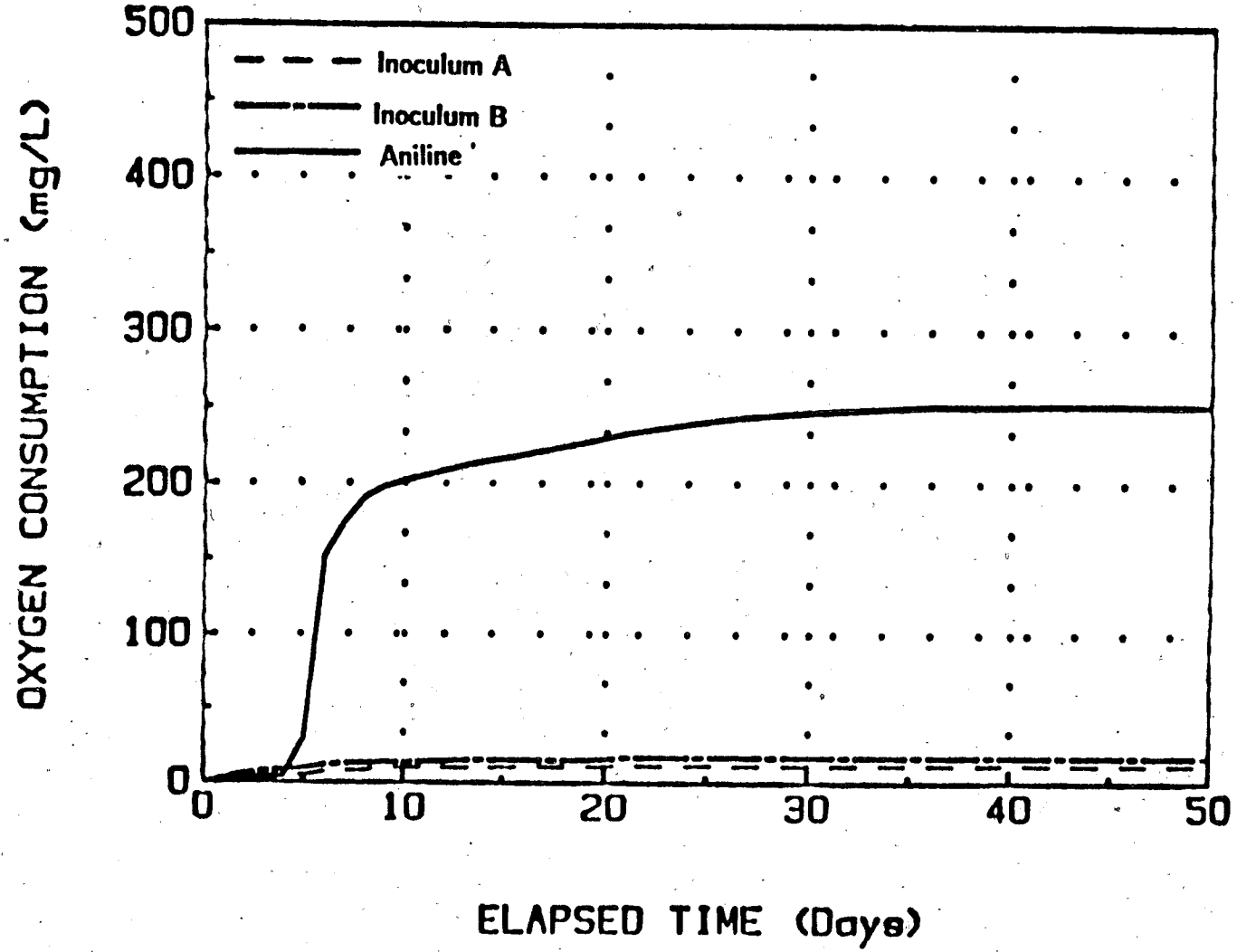


FIGURE 14 OXYGEN UPTAKE DATA ON ANILINE

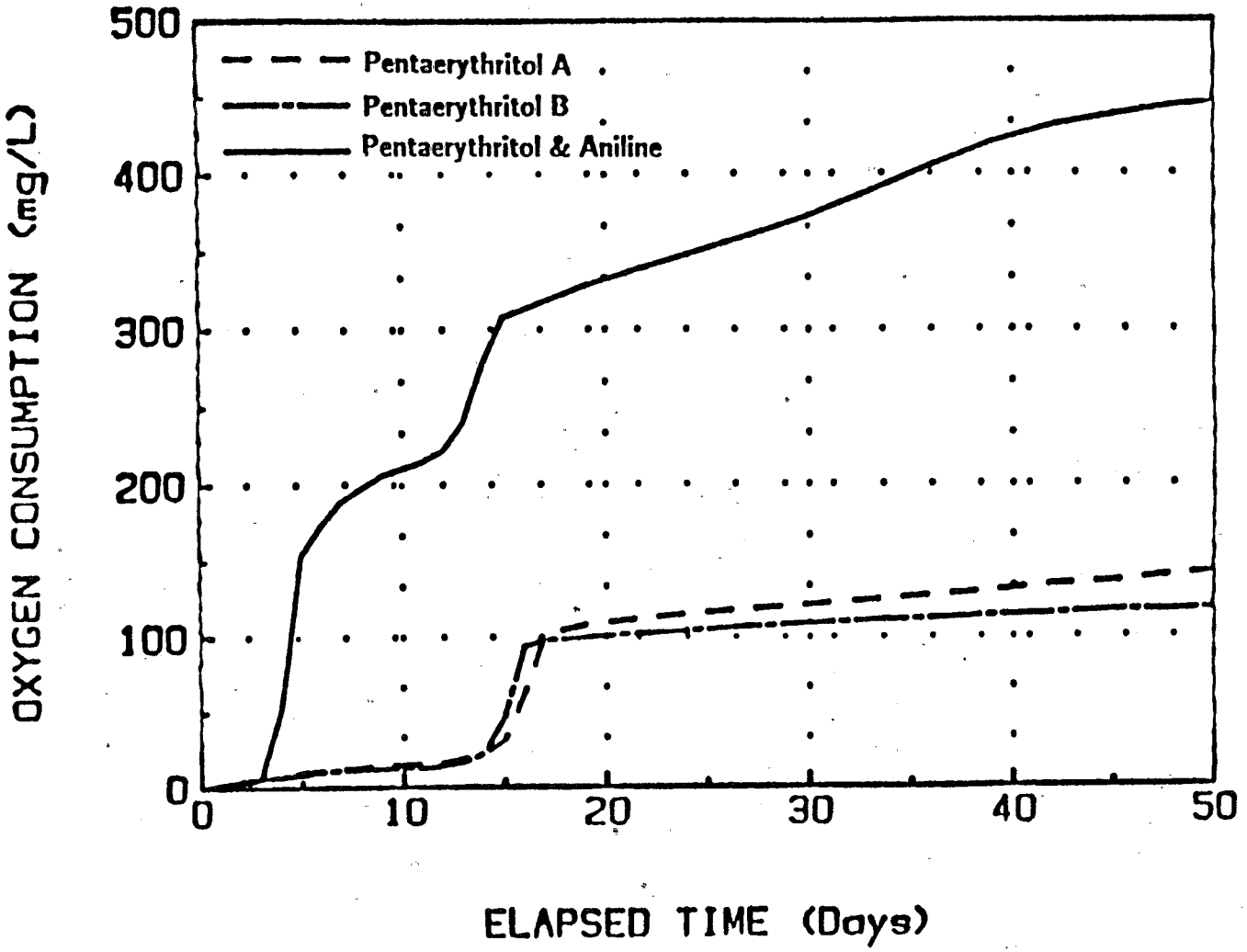


FIGURE 15 BIOLOGICAL OXYGEN UPTAKE CURVE
(Run 1 Sample No. 10-12)

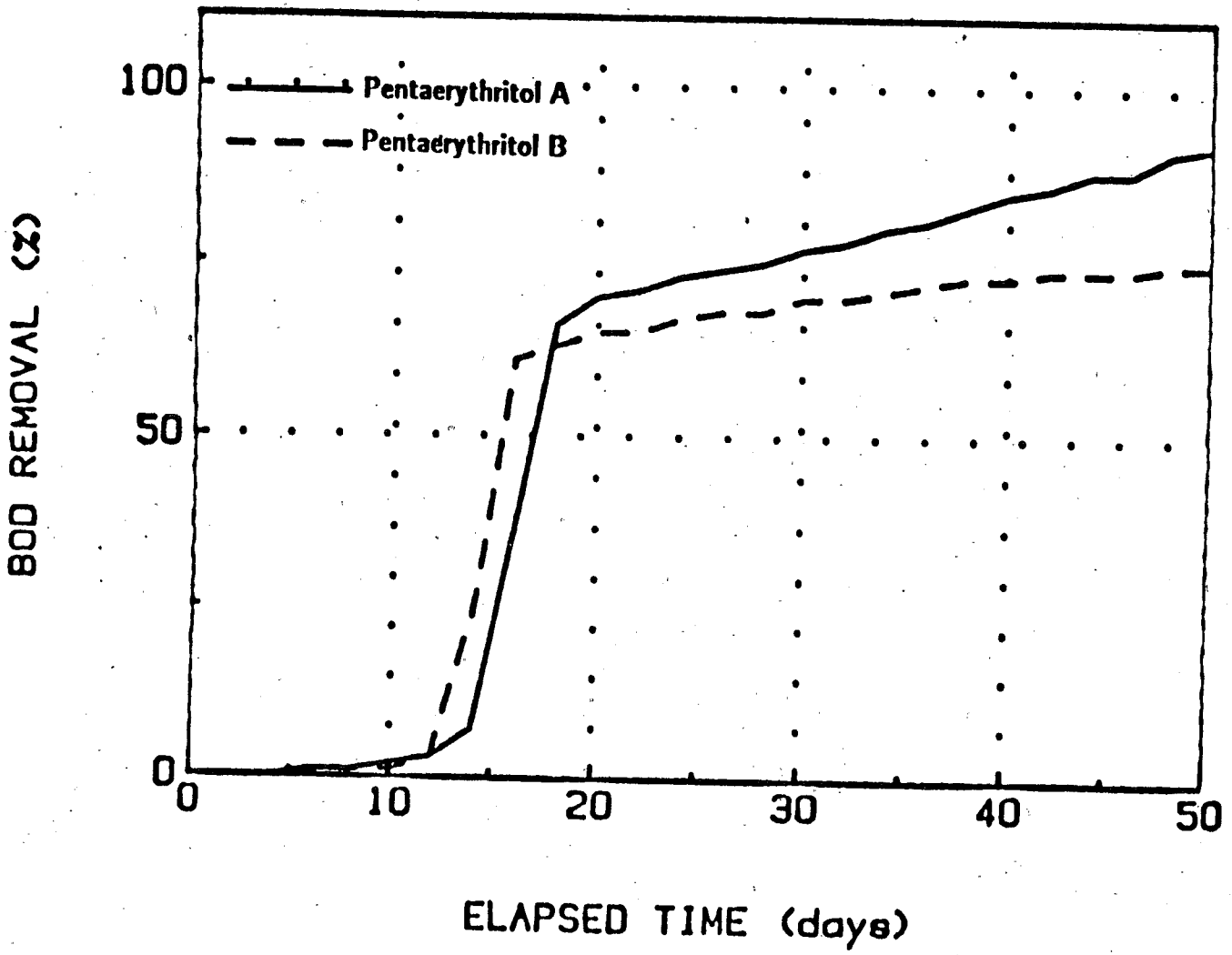


FIGURE 16 BIODEGRATION (% BOD REMOVAL) CURVE
(PENTAERYTHRITOL)

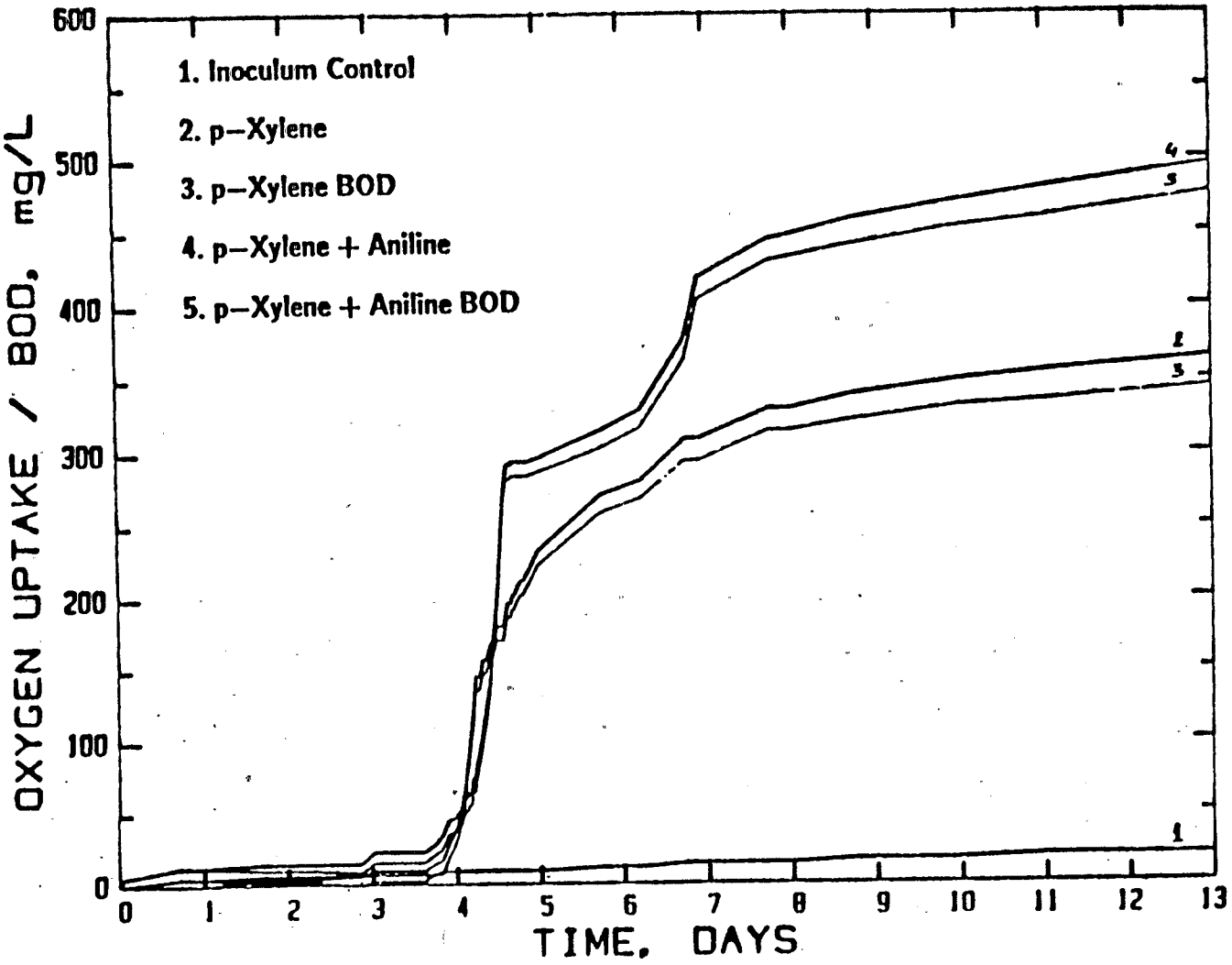


FIGURE 17 BIOCHEMICAL OXYGEN UPTAKE AND BOD DATA FOR P-XYLENE AND P-XYLENE + ANILINE

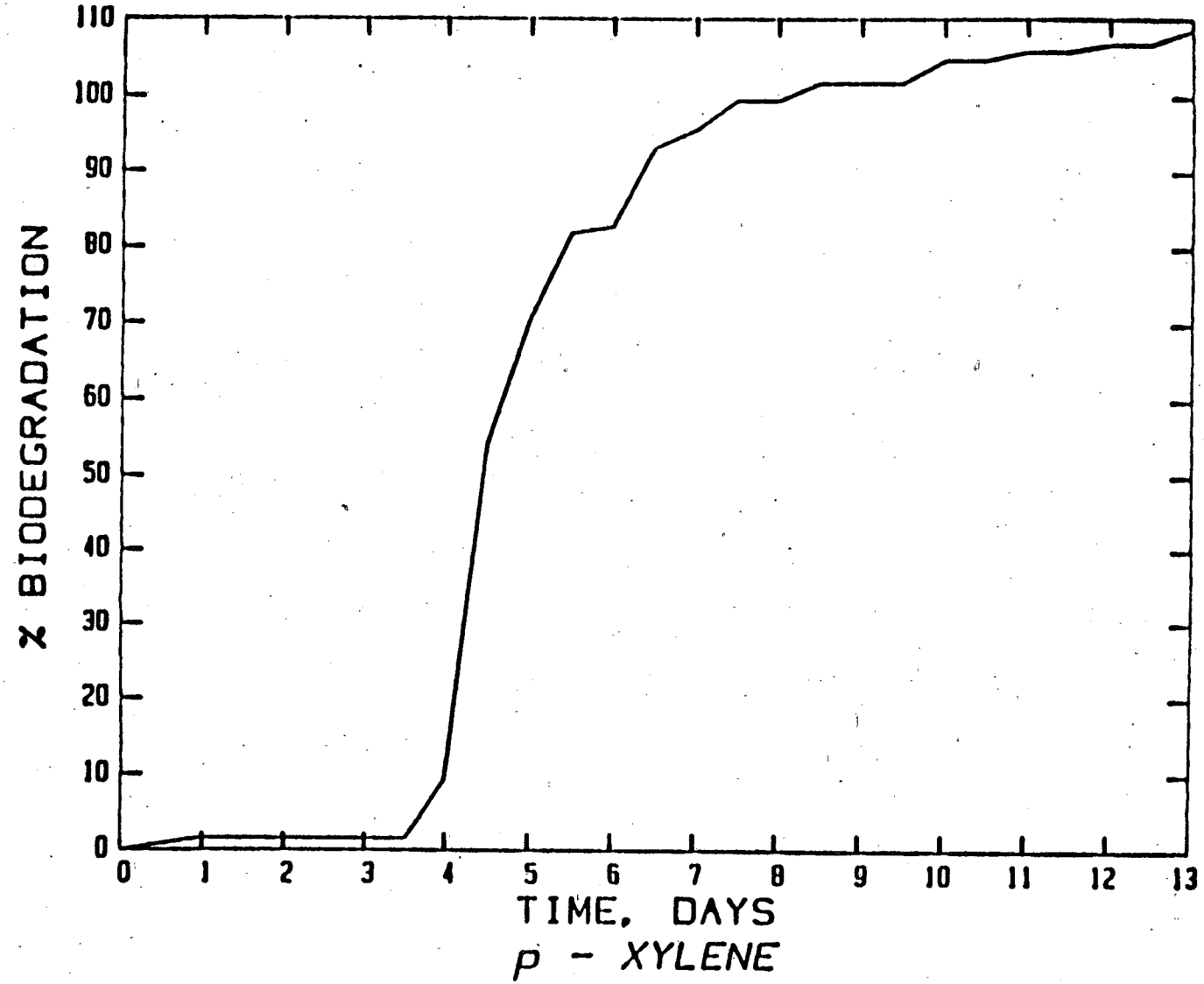


FIGURE 18 BIODEGRADATION DATA FOR P-XYLENE

TABLE 5. SUMMARY OF RESPIROMETRIC BIODEGRADATION DATA FOR SELECTED BENZENES

PERCENT BIODEGRADATION (BASED ON % BOD REMOVAL)

Time (days)	Aniline	Benzene	Toluene	Ethylbenzene	m-Xylene	p-xylene
0.0	0.0	0.0	0.0	0.0	0.0	0.0
1.0	2.58	2.11	2.78	2.27	1.35	1.58
2.0	3.43	2.11	8.81	2.27	2.14	1.58
3.0	3.94	4.15	85.11	2.99	69.68	1.58
4.0	14.95	4.97	89.04	3.78	69.68	9.4
5.0	102.0	71.5	94.34	4.29	76.68	70.38
6.0	102.0	74.22	97.92	40.25	82.58	82.68
7.0	102.0	81.85	100.0	73.91	84.67	95.58
8.0	102.0	93.37	101.18	73.78	87.03	99.49
9.0	102.0	93.89	103.48	83.94	87.79	101.76
10.0	102.0	95.45	103.48	88.86	90.12	104.7
11.0	102.0	95.45	103.48	88.86	90.15	105.78
12.0	102.0	96.13	103.48	89.65	90.97	106.7
13.0	102.0	97.46	103.48	91.00	91.80	108.6

TABLE 6. SUMMARY OF BIO-KINETIC DATA FOR SELECTED BENZENE COMPOUNDS

COMPOUNDS	ThOD for 100 mg	t_0 (days)	t_1 (days)	t_2 (days)	μ (day-1)	μ' (day-1)
Aniline						
(Experiment 1)	310	4.00	4.70	4.83	2.78	3.29
(Experiment 2)	310	4.00	4.65	4.79	3.80	8.60
Benzene	308	4.50	4.87	5.00	8.57	25.30
Ethyl benzene	317	4.00	4.21	4.83	8.33	44.80
Toluene	313	2.00	2.20	2.42	8.75	14.93
p-Xylene	317	3.90	4.22	4.83	9.94	4.50
m-Xylene	317	2.00	2.35	2.50	6.60	29.60
tert-Butyl benzene	322	4.40	5.12	5.70	1.21	1.69
sec-Butyl benzene	322	3.50	4.00	5.70	0.78	0.73
Cumene	320	2.40	2.79	3.00	2.31	2.24
Butyl benzene	322	3.30	3.92	4.56	2.42	2.42
1-Phenyl hexane	326	4.00	4.55	5.15	1.85	1.90

μ = specific growth rate constant for exponential growth phase of BOD curve.

μ' = specific growth rate constant for declining growth phase of BOD curve.

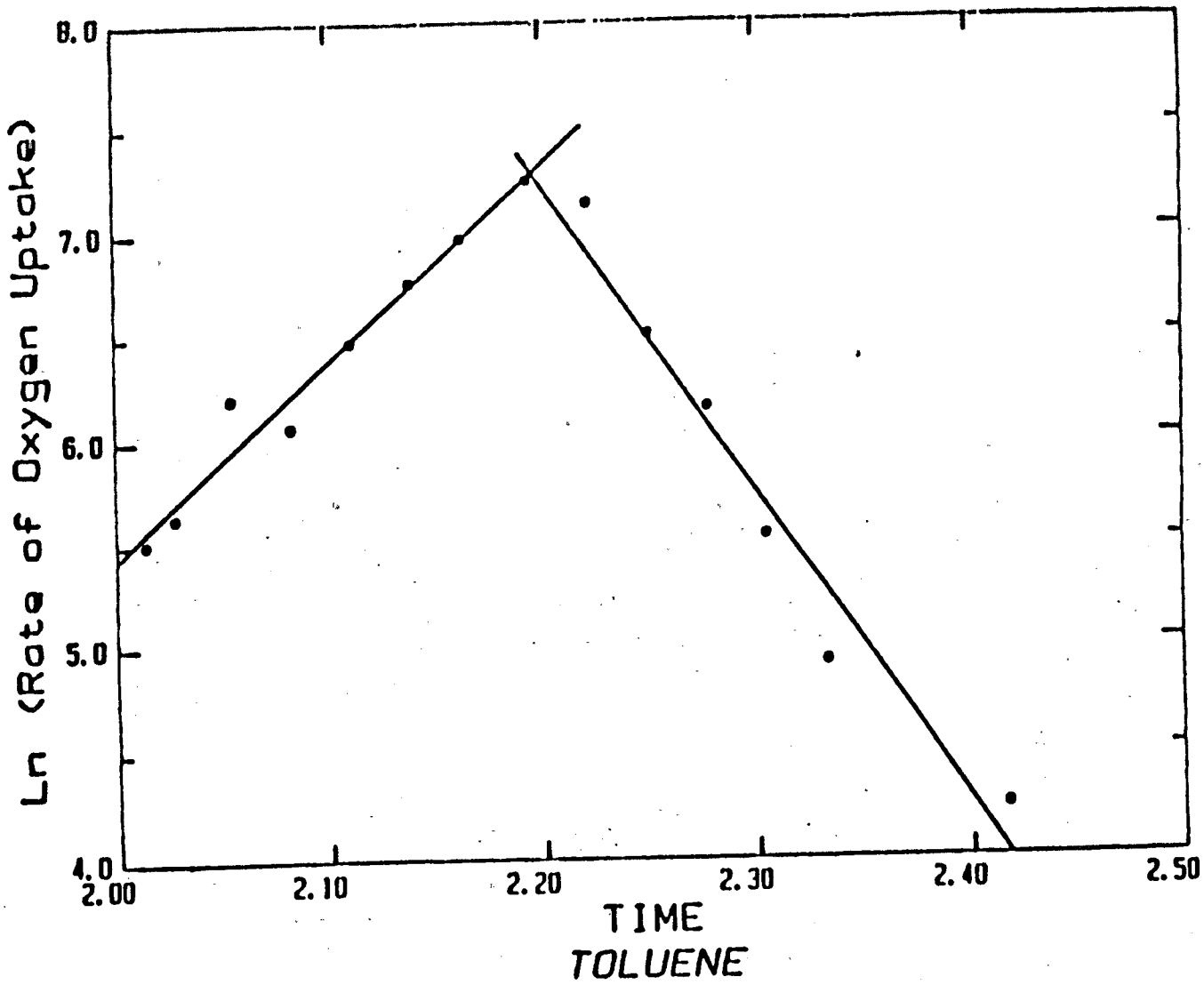


FIGURE 19 PLOT OF LOG ($\Delta Ou/\Delta t$) FOR TOLUENE

TABLE 7. SUMMARY OF MONOD KINETIC PARAMETER DATA FOR SELECTED BENZENE COMPOUNDS

COMPOUNDS	Lag Time (t_0) days	Y_g mg biomass / mg substrate	μ_m (day-1)	K_s mg/l
Aniline	4.00	0.38	6.15	6.10
Benzene	4.50	0.23	9.30	22.16
Ethyl benzene	4.00	0.23	9.00	11.81
Tolouene	2.00	0.20	9.09	56.74
p-Xylene	3.90	0.26	8.33	67.00
m-Xylene	2.00	0.14	11.50	35.61
tert-Butyl benzene	4.40	0.66	1.80	69.04
sec-Butyl benzene	3.50	0.64	2.39	22.00
Cumene	2.44	0.66	2.44	87.64
Butyl benzene	3.30	0.80	5.17	34.50
1-Phenyl hexane	4.00	0.67	2.15	35.09

μ = maximum specific growth rate.

K_s = half saturation constant; concentration of substrate at $\mu_m/2$.

Y_g = growth yield, mg biomass formed/mg substrate consumed.

TABLE 8. SUMMARY OF BIO-KINETIC DATA FOR SELECTED PHENOLIC COMPOUNDS

COMPOUNDS	ThOD for 100 mg	t_0 (days)	t_1 (days)	t_2 (days)	μ (day-1)	μ' (day-1)
Aniline	310	4.00	4.70	4.83	2.78	3.29
Phenol	238	1.00	1.56	1.67	2.75	8.26
Resorcinol	189	1.50	1.83	2.14	5.42	6.53
p-Cresol	252	1.00	1.34	1.52	4.70	5.78
o-Cresol	252	1.20	1.78	1.90	3.76	3.92
m-Cresol	252	1.44	2.04	2.40	5.17	6.15
Catechol	189	0.85	0.94	1.12	11.80	11.85
2,4-Dimethyl phenol	262	2.00	2.64	2.84	3.21	5.02

μ = specific growth rate constant for exponential growth phase of BOD curve.

μ' = specific growth rate constant for declining growth phase of BOD curve.

TABLE 9. SUMMARY OF MONOD KINETIC PARAMETER DATA FOR SELECTED PHENOLIC COMPOUNDS

COMPOUNDS	Lag Time (t_0) days	Y_g mg biomass / mg substrate	μ_m (day-1)	K_s mg/l
Aniline	4.00	0.38	6.15	6.10
Phenol	1.00	0.58	9.82	9.43
Resorcinol	1.50	0.48	12.22	6.31
p-Cresol	1.00	0.33	6.11	27.78
o-Cresol	1.20	0.41	4.10	16.41
m-Cresol	1.44	0.46	7.97	17.62
Catechol	0.85	0.49	12.80	43.87
2,4-dimethyl phenol	2.00	0.39	5.62	14.07

μ = maximum specific growth rate.

K_s = half saturation constant; concentration of substrate at $\mu_m/2$.

Y_g = growth yield, mg biomass formed/mg substrate consumed.

TABLE 10. SUMMARY OF BIO-KINETIC DATA FOR SELECTED PHTHALATE ESTER COMPOUNDS

COMPOUNDS	ThOD for 100 mg	t_0 (days)	t_1 (days)	t_2 (days)	μ (day-1)	μ' (day-1)
Aniline	310	4.00	4.70	4.83	2.78	3.29
Dimethyl phthalate	168	3.46	3.98	4.25	2.76	4.71
Diethyl phthalate	195	2.00	2.97	3.30	2.16	2.92
Dipropyl phthalate	211	2.40	2.87	3.40	2.04	2.00
Butyl benzyl phthalate	226	2.00	2.28	2.80	4.12	2.33

μ = specific growth rate constant for exponential growth phase of BOD curve.

μ' = specific growth rate constant for declining growth phase of BOD curve.

TABLE 11. SUMMARY OF MONOD KINETIC PARAMETER DATA FOR SELECTED PHTHALATE ESTER COMPOUNDS

COMPOUNDS	Lag Time (t_0) days	Y_g mg biomass / mg substrate	μ_m (day-1)	K_s mg/l
Aniline	4.00	0.38	6.15	6.10
Dimethyl phthalate	3.46	0.43	7.07	41.68
Diethyl phthalate	2.00	0.46	3.00	11.67
Dipropyl phthalate	2.40	0.48	5.78	15.81
Butyl benzyl phthalate	2.00	0.61	7.80	36.25

μ = maximum specific growth rate.

K_s = half saturation constant; concentration of substrate at $\mu_m/2$.

Y_g = growth yield, mg biomass formed/mg substrate consumed.

TABLE 12. SUMMARY OF BIO-KINETIC DATA FOR SELECTED KETONE COMPOUNDS

COMPOUNDS	ThOD for 100 mg	t_0 (days)	t_1 (days)	t_2 (days)	μ (day-1)	μ' (day-1)
Aniline	310	4.00	4.70	4.83	2.78	3.29
Acetone	221	3.70	3.99	4.18	2.45	3.98
2-Butanone	244	2.00	2.20	2.35	2.41	4.98
4-Methyl-2-pentanone	272	1.85	2.24	2.35	2.31	4.80
Isophorone	278	22.30	23.70	25.40	0.73	0.38

μ = specific growth rate constant for exponential growth phase of BOD curve.

μ' = specific growth rate constant for declining growth phase of BOD curve.

TABLE 13. SUMMARY OF MONOD KINETIC PARAMETER DATA FOR SELECTED KETONE COMPOUNDS

COMPOUNDS	Lag Time (t_0) days	Y_g mg biomass / mg substrate	μ_m (day-1)	K_s mg/l
Aniline	4.00	0.38	6.15	6.10
Acetone	3.70	0.36	4.86	9.76
2-Butanone	2.00	0.39	5.11	10.79
4-Methyl-2-pentanone	1.85	0.45	6.40	24.70
Isophorone	22.30	0.43	1.57	27.42

μ = maximum specific growth rate.

K_s = half saturation constant; concentration of substrate at $\mu_m/2$.

Y_g = growth yield, mg biomass formed/mg substrate consumed.

PREDICTION OF BIODEGRADATION KINETICS OF TOXIC ORGANIC COMPOUNDS

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PREDICTION OF BIODEGRADATION KINETICS OF TOXIC ORGANIC COMPUNDS

ABSTRACT

Biodegradation is the most important mechanism in controlling the concentration of chemicals in an aquatic system because it can mineralize toxic pollutants to harmless forms. So the fate of organic chemicals in an aquatic environment is dependent on the susceptibility to biodegradation. Because of the large number of chemicals, it would be expensive and labor intensive to gather this information in a reasonable amount of time. Hence, there is need for a prediction method to obtain these data.

The paper describes a predictive biodegradation—structure/activity model based on the group contribution approach using biodegradation kinetic data obtained from the literature and through electrolytic respirometry studies. The model, incorporating chemical properties and structure of organic compounds, predicts biodegradation kinetics for related toxic compounds. Biodegradation kinetic data for representative RCRA toxic organics presented in this paper, were used to establish a pattern of predictability for biodegradation of related compounds with the group contribution predictive model.

Results are presented on the development of the structure—activity relationship for biodegradation using the group contribution approach. With the use of this approach, reported results of the kinetic rate constants agree within 20% with the predicted values. Additional compound studies are essential to further extend the methodology.

INTRODUCTION

It has been estimated that 50,000 organic chemicals are commercially produced in the United States and a large number of new organic chemicals are added to the production each year (1). The presence of many of these chemicals in the environment is a serious public health problem. Their presence could be attributed to inadequate disposal techniques.

Since many of these hazardous chemicals can be detected in wastewater, their fate in wastewater treatment system is of great interest. Of the many factors that affect the fate of these compounds, microbial degradation is probably the most important (2). The high diversity of species and the metabolic efficiency of microorganisms suggest that they play a major role in the ultimate degradation of these chemicals (3). Biodegradation can eliminate hazardous compounds by biotransforming them into innocuous forms, degrading them by mineralization to carbon dioxide, water and other simple and harmless molecules. So the information regarding the extent and the rate of biodegradation of these chemicals is very important for regulation of their manufacture and use. Due to the large number of these chemicals, gathering of this information in a reasonable amount of time would be both expensive and labor intensive. Thus, the only practical way out is to develop correlations and predictive techniques to assess biodegradability (4). Lack of an adequate database on biodegradation kinetics has hindered the development of such techniques.

EXPERIMENTAL TECHNIQUES

There are many techniques for collection of biodegradation data and these are reviewed in great detail by Howard, et al. (5) and Grady (6). Experimental techniques for biodegradation data collection fall into three broad categories : continuous, fed–batch and batch reactor systems.

Continuous culture reactors require an acclimated biomass. They also require long transitional time intervals for reaching quasi steady state condition (7). So it is time consuming, tedious and expensive. However, the analysis of data obtained is simpler because the equations for continuous reactors, operating at steady state, reduce to algebraic equations that are easily solved. This technique is used more for evaluating parameters for the design of treatment systems rather than for biodegradation kinetics (8,9).

Fed–batch reactors have also been used to estimate the kinetic parameters of biodegradation. In these reactor configurations, a quantity of biomass is added into a reactor and a substrate stream is continuously added in negligibly small amounts with respect to the reactor volume. Because of small amounts of both substrate and microorganisms, a pseudo steady state is achieved. So this technique reduces the time required and also alleviates the problems associated with changes in the composition of the microbial community. This configuration requires an acclimated biomass because the microbes must be capable of responding instantaneously to the input of substrate. This technique cannot be used to determine the Monod biokinetic parameters, but it is an excellent procedure to determine the parameters for treatment plant design or operation.

The use of batch cultures in biotechnology and biological wastewater treatment represents a less expensive and much faster way to model biokinetics in fermenters

and in activated sludge tanks (7). The batch method commonly used for a large number of compounds (10) is one in which the substrate of interest, at different concentrations, is inoculated with the small amount of biomass. Then the increase in biomass concentration in each reactor is measured as a function of time. Another technique focuses on the substrate removal rather than the microbial growth. This is commonly used in engineering studies. The batch reactors are inoculated with large quantities of biomass and the substrate removal is measured as a function of time. Both of these techniques have been widely used to measure parameters such as five day BOD and COD. Batch reactor can be used with either acclimated or unacclimated biomass. It requires that samples be taken at discrete time intervals during the course of biodegradation. If unacclimated biomass is used, the number of samples required may be large depending on the acclimation time. Tabak, et al. (11) have collected degradability and acclimation data on 96 compounds by static culture screening procedure and culture enrichment process. Larson and Perry (12) and Paris, et al. (13) have done biodegradation studies with unacclimated biomass in the batch reactors and have evaluated the kinetic parameters.

In the past, the number of data points obtained was small because of manual sampling. This can be avoided by monitoring oxygen consumption as an indirect measure of biodegradation using an electrolytic respirometer. The automatic data collection and recording allows sufficient accumulation of data, that the reliability of the kinetic parameters evaluated is maximized.

In the respirometric methods of BOD measurement, wastewater samples are kept in contact with the gas phase source of oxygen. Oxygen uptake by the microogranisms over a period of time is measured by the changes in volume or pressure of the gas phase. An alkali is included in the apparatus to absorb carbon dioxide produced during biodegradation. Samples are usually incubated at constant temperature and are kept away from light. The latest development in respirometric techniques has been the advent of an electrolytic respirometer. It supplies oxygen, produced from electrolysis of water, to the air space above the sample in a completely sealed reaction vessel. The production of oxygen is triggered due to the pressure changes in the reaction vessel. Studies by Larson and Perry (12), Young and Baumann (14), Tabak, et al. (15) and Dosanjh and Wase (16) have shown that the electrolytic respirometer eliminates most of the technical difficulties associated with other methods for determining BOD. It is particularly useful for the rate studies because it provides both, a continuous record of oxygen uptake and it maintains an unchanging atmosphere over the sample regardless of the length of the test.

EXPERIMENTAL SET–UP

The electrolytic respirometer Sapromat B–12, consists of a temperature controlled waterbath, which contains the measuring units, a recorder for digital indication and direct plotting of the oxygen uptake curves and a cooling unit. The waterbath has 12 reaction flasks, each connected to the recorder. Each unit as shown in figure 1 consists of a reaction vessel C, with a carbon dioxide absorber (sodalime) 3 mounted in a stopper, an oxygen generator B, and a pressure indicator A. The vessels interconnected by hoses, form a sealed measuring system which assures that the barometric pressure fluctuations do not adversely affect the result. The magnetic stirrer 1 in the sample 2 to be analyzed provides vigorous agitation, thus ensuring an effective exchange of gases. The activity of the microorganisms in the sample creates a reduction in the pressure which is recorded by the pressure indicator. It controls both the electrolytic oxygen generation and the indication and the plotting of the measured values.

The consumption of oxygen by microorganisms creates a reduction in pressure in the reaction vessel. As a result, the level of 0.5% sulphuric acid in the pressure indicator rises and comes in contact with a platinum electrode. This completes the circuit and triggers the generation of oxygen by the electrolytic cell. The oxygen gas is provided to the reaction vessel, alleviating the negative pressure. The level of electrolyte in the pressure indicator drops, breaking contact with the electrode. This switches off the electrolytic cell. The amount of oxygen supplied to the sample is recorded directly in milligrams per litre by the recorder. The recorder is connected to an IBM AT which records data from the measuring units every 15 minutes.

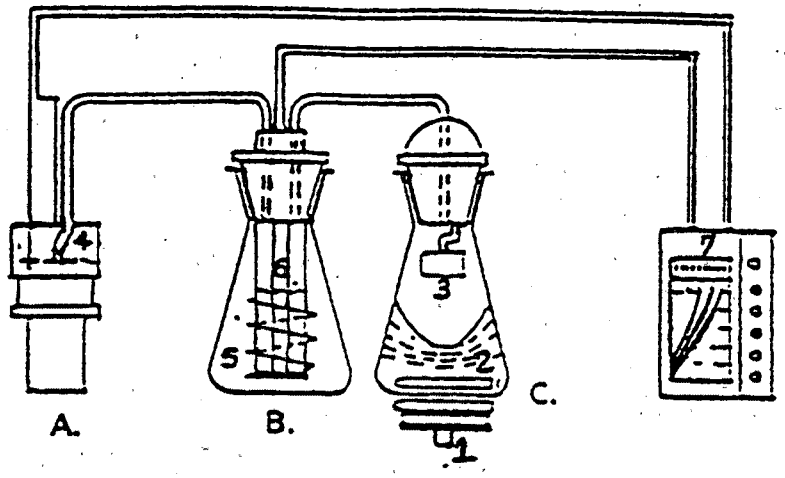

Figure 1. Diagram of a Measuring Unit

A. Pressure indicator
B. Oxygen generator
C. Reaction vessel
1. Magnetic stirrer
2. Sample (250ml)
3. CO_2 absorber
4. Pressure indicator
5. Electrolyte
6. Electrodes
7. Recorder

BIODEGRADATION KINETICS

A considerable amount of information concerning biodegradation kinetics is available in the published literature. Early literature shows widely differing kinetic rates in different studies. The evaluation and prediction of the extent and rate of biooxidation is affected by methodological and experimental factors. Regardless of the different assumptions involved in the measurement of biodegradation rates, it is generally considered associated with microbial cell growth and so most of the models

for it are the same as those used to model growth and substrate removal.

Although many models have been proposed for microbial growth, the Monod relation is the most popular kinetic expression (17). The Monod model, in combination with the linear law for substrate removal can provide an adequate description of microbial growth behavior. It states that the cell growth is first order with respect to the biomass concentration (X) and mixed order with respect to the substrate concentration (S)

$$dX/dt = (S\,\mu_m\,X) / (K_s + S) \quad \ldots\ldots \quad (1)$$

Cell growth is related to the substrate removal by the linear law

$$dX/dt = -\,Y_g\,(dS/dt) \quad \ldots\ldots\ldots \quad (2)$$

The kinetic parameters of interest are maximum specific growth rate μ_m, half saturation constant K_s, (it is the concentration of substrate when $\mu=0.5\mu_m$), and the yield coefficient Y_g. The Monod equation has two limiting cases. When the substrate concentration is much greater than the saturation constant the term (S/K_s+S) approaches 1.0 and cell growth and substrate removal are zero order with respect to substrate concentration. When the substrate concentration is much smaller than the saturation constant, the term (S/K_s+S) approaches (S/K_s) and cell growth and substrate removal are first order with respect to the substrate concentration. Many researchers have used either of the above two approaches. The characteristics of these kinetic expressions have been discussed by Simkins and Alexander (18).

The electrolytic respirometer has been mostly used to measure the extent of biodegradation as a percentage of the theoretical oxygen demand over a certain period of time. Several researchers have tried to extract kinetic parameters from oxygen uptake data. Larson and Perry (12) used empirical kinetic expressions which were system specific. Dojlido (19) divided the oxygen uptake curve into seven different phases and then proposed an empirical model for each phase and evaluated the biodegradability and toxicity of a test compound by measuring empirical rate constants and time intervals associated with each phase. Tabak, et al. (15) divided the substrate removal region of oxygen uptake curve into two regions, separated by an inflection point. The first period was called exponential phase where it was assumed that the substrate was not limiting and cell growth was occuring, while the second period was called declining phase and here it was assumed that the substrate was limiting. So work done to obtain kinetic parameters from oxygen uptake curve has been empirical.

STRUCTURE–ACTIVITY RELATIONSHIP

The structure–activity relationships (SAR) have been widely used in pharmocology and medicinal chemistry. The different methods and models used are free energy models, Free–Wilson mathematical model, discriminant analysis, cluster analysis, pattern recognition, topological methods, and quantum mechanical methods.

The free energy model of Hansch, et al. (20) is widely used. They incorporated octanol–water coefficient, log P, in the linear free energy relationship as a measure of lipophilicity. This provided a general SAR model for biological activity. The success of this model has led many researchers to include additional physicochemical parameters and properties, structural, topological and molecular indices. Using similar principles other researchers have proposed models to include more complex relationships between the bioactivity and the chemical structure or the properties. Martin (21) has dicussed

these models.

Free and Wilson proposed a mathematical model to assess the additive effects of substituents and quantitatively estimate their magnitude (22). According to their model the structure of a compound is composed of different groups or a core which is substituted in various positions, resulting in a series of linear equations of the form

$$BA = \Sigma a_j X_j + \beta$$

where BA is biological activity, X_j is the j^{th} group, a_j is the contribution of the j^{th} group and β is the overall average activity. These linear equations are solved by the least square method for a_j and β. This method requires a large number of compounds for a meaningful analysis and it will breakdown if there are interactions between different groups. Fujita and Ban (23) suggested that BA should be expressed as log(1/C), where C is the concentration of the compound that produces a constant biological response. This modified Free–Wilson model is in common use in medicinal chemistry.

Discriminant analysis is used where only semiquantitative or qualitative data have to be evaluated. In this method, a linear combination of parameters called a linear discriminant function is formed, which classifies the observations. Martin (21) has discussed the background of this method and has given examples. The Principal component method is used as a preliminary step in multiple regression analysis of the Hansch type (24). Factor analysis is used to gain insight into the structure of multidimensional data set and involves manipulation of the eigenvectors of variance–covariance matrix of the dataset (25). Cluster analysis is used to group similar substituents when various combination of parameters are considered (26).

Pattern recognition is used to examine structural features and chemical properties for the patterns associated with different biological activities (27). In this method, a set of descriptors is generated for each compound and then a suitable algorithm is applied to find some combinations and weight of the descriptors which give a perfect classification for a set of compounds. This classification is then applied to another set of compounds of known classification and performance is judged by the percentage of correct prediction.

Various methods have been proposed to relate topology of the molecule with its biological activity. Verloop (28) has proposed a method to treat directionality of steric effects. A computer representation of a compound is created and then measured by tangential planes which results in five STERIMOL parameters. Kier and Hall (29) have used molecular connectivity index, a number calculated from graph theoretical principles for SAR correlation. Blankley (30) has reviewed all these and other methods used in SAR.

In the field of biodegradation there are several studies which have attempted to correlate some physical, chemical or structural property of a chemical with its biodegradation. Based on the type and the location of the substituent groups, Geating (31) developed an algorithm to predict biodegradation. Qualitative relationships for different compounds have been investigated by others, but quantification is required for regulatory purposes.

In the literature, first order rate of biodegradation or five day BOD of chemicals have been correlated with physical or chemical properties. Paris, et al. (32,33,34) established a correlation between second order biodegradation rate constant and the

van der Waal's radius of substituent group for substituted anilines and for a series of para–substituted phenols. Wolfe, et al. (35) correlated second order alkaline hydrolysis rate constant and biodegradation rate constants for selected pesticides and phthalate esters. Several workers have observed a correlation between biodegradability and lipophilicity, specifically octanol/water partition coefficients (log P). Paris, et al. (36) found a good correlation between biodegradation rate constant and log P, for a series of esters of 2,4–dichlorophenoxy acetic acid. Banerjee, et al. (37) obtained a similar relationship for chlorophenols. Vaishnav, et al. (38) correlated biodegradation of 17 alcohols and 11 ketones with octanol–water coefficients using 5–day BOD data. Pitter(39) has found a dependence of biodegradation rate on electronic factors, like Hammett substituent constant, for a series of anilines and phenols. Dearden and Nicholson (40) have correlated 5–day BOD with modulus difference of atomic charge across a selected bond in a molecule for amines, phenols, aldehydes, carboxylic acids, halogenated hydrocarbons and amino acids. A direct correlation between the biodegradability rate constant and the molecular structure of the chemical has been used by Govind (41) to relate the first order biodegradation rate constant with the first order molecular connectivity index and by Boethling (42) to correlate the biodegradation rate constants with the molecular connectivities for dialkyl esters, carbamates, dialkyl ethers, dialkyl phthalate esters and aliphatic acids.

GROUP CONTRIBUTION APPROACH

Using a group contribution approach, a very large number of chemicals of interest can be constituted from perhaps a few hundered functional groups. The prediction of the property is based on the structure of the compound. According to this method the molecules of a compound are structurally decomposed into functional groups or their fragments, each having unique contribution towards the compound property. This method is similar to the Free–Wilson method used in medicinal chemistry.

The biodegradability rate constant k, is expressed as a series function of contribution α_j, of each group of the compound. The first order approximation of this series function representing biodegradation rate constant can be expressed as

$$\text{Ln}\,(k) = \sum_{j=1}^{L} N_j \alpha_j \quad \ldots\ldots\ldots\ldots\ldots \quad (3)$$

where N_j is the number of groups of type j in the compound, α_j is the contribution of group of type j and L is the total number of groups in the compound.

CALCULATION OF GROUP CONTRIBUTION PARAMETER, α

Using equation 3 for each compound, a linear equation in α's is constructed. For a given dataset this generates a series of linear equations which are solved for α's using regression analysis.

If there are n α's and m compounds ($n<m$) then there will be m equations of the form of equation 3, each representing a different compound. The sum of squares of deviation S, for these equations is given by

$$S = \sum_{i=1}^{m} \{ Ln(k_i) - \sum_{j=1}^{n} N_{ji}\alpha_j \}^2 \quad \ldots\ldots\ldots \quad (4)$$

The estimates of α's should be such that it produces the least value of S. These estimates are obtained by differentiating equation 4 with respect to α and setting it equal to zero

$$dS/d\alpha_k = -2 \sum_{i=1}^{m} \{[Ln(k_i) - \sum_{j=1}^{n} N_{ji}\alpha_j] N_{ki}\} = 0 \quad \ldots\ldots \quad (5)$$

This will generate a series of n linear equations which are solved for α's. If **N** is the matrix of coefficients of α's and **Y** is the vector of Ln(k) values then the solution vector α is given by

$$\alpha = (N'N)^{-1} (N'Y)$$

where ' denotes the transpose of matrix.

MODEL VERIFICATION

The contribution of several different groups was calculated from the data obtained from the literature (43,44,45). The data used to calculate these contribution parameters is given in Table 1. The groups and their contribution parameters are given in Table 2.

The biodegradation experiments were carried out for cresols, phenol, 2,4–dimethyl phenol, 2–butanone, acetone, 1–phenyl hexane and butyl benzene using an electrolytic respirometer Sapromat B–12 (Voith–Morden, Milwaukee, WI). The chemicals were from Aldrich chemical company with 99+% purity. These compounds were used to validate the model, but were not used in the calculation of the group contribution parameters. The experimental conditions were : temperature 25°C, biomass concentration 30 mg/L and compound concentration 100mg/L. The biomass was obtained from Little Miami wastewater treatment plant in Cincinnati, which receives 95% domestic waste. The biomass was aerated for 24 hours and then used for the experiment. The biodegradation rate constants for these compounds were determined using the following simplified Monod form, obtained by assuming that $K_s >> S$ and change in biomass concentration is negligible.

$$dS/dt = -KX_oS. \quad \ldots\ldots\ldots\ldots \quad (6)$$

where $K = (\mu_m/Y_g K_s)$ and X_o is initial biomass concentration. Integrating equation 6 we get

$$S = S_o \exp(-KX_ot). \quad \ldots\ldots\ldots \quad (7)$$

In terms of BOD the above equation transforms to

$$BOD = ThOD (1 - \exp\{-KX_ot\}). \quad \ldots\ldots \quad (8)$$

The above equation was used for only the rising part of the BOD versus time curve, and initial data and data for endogenous phase were neglected. Note that the selection of the rising part of the BOD curve was arbitrary and was based on visual inspection of the BOD curve. The biodegradation rate constants were also predicted using the group contribution parameters of Table 1. The comparison of these values is given in

Table 3 and Figure 2.

RESULTS AND DISCUSSION

The prediction values of Ln(K) are within 10% of the experimental values. It is important to emphasize that there are several reasons for the rather large discrepancy between the predicted and experimental values. These reasons are as follows :

1. The data used for calculating the group contribution parameters were obtained from the literature (refer Table 1). The experimental BOD curves, obtained from electrolytic respirometry, were approximated by an exponential relation (refer equations 6, 7 and 8). The range of curve to which this equation was applied was selected arbitrarily. This resulted in some error between the experimental data and the calculated value of the kinetic constant, K.

2. The group contribution method described here is the first order approximation. It assumes that the contributions of the groups are independent of each other. So the rates predicted are only expected to be within an order of magnitude.

3. Biomass and experimental methods used in the different studies were different. The studies, used for determination of group contribution values had acclimated their biomass to the compound before using it in the experiment, while our biomass was from domestic wastewater treatment plant and was not acclimated to the compounds used in the study.

Inspite of these errors, the prediction is within an order of magnitude. With availability of more data and using a detailed kinetic model rather than a first order exponential equation, the prediction error may be reduced considerably.

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TABLE 1. DATA USED TO CALCULATED GROUP CONTRIBUTION PARAMETERS

No.	COMPOUNDS	Molecular formula	K(L/Mg.Hr)*E4	REF.
1	Ethyl alcohol	C2H5OH	5.749	43
2	Butyl alcohol	C4H9OH	11.10	43
3	Ethylene glycol	C2H4(OH)2	40.00	43
4	Acetic acid	CH3COOH	22.22	43
5	Propionic acid	C2H5COOH	23.14	43
6	n—Butyric acid	C3H7COOH	10.50	43
7	n—Valeric acid	C4H9COOH	6.516	43
8	Adipic acid	C4H8(COOH)2	2.781	43
9	Methyl ethyl ketone	CH3COC2H5	3.071	43
10	Hexamethylenediamine	C6H12(NH2)OH	2.994	43
11	n—Hexylamine	C6H13NH2	4.425	43
12	Mono ethanol amine	C2H4(NH2)OH	3.720	43
13	Acetamide	CH3CONH2	4.576	43
14	Benzene	C6H6	288.3	44
15	Benzyl alcohol	C6H5CH2OH	16.34	43
16	Toluene	C6H5CH3	8.446	43
17	Acetophenone	C6H5COCH3	6.408	43
18	Hydroxybenzoic acid	C6H4(OH)COOH	10.50	43
19	Aminobenzoic acid	C6H4(NH2)COOH	6.332	43
20	Aminophenol	C6H4(NH2)OH	4.179	43
21	2—Cl phenol	ClC6H4OH	1000.	45
22	1,2—Dichlorobenzene	C6H4Cl2	678.4	45
23	1,2,3—Tri—Cl—benzene	C6H3Cl3	446.7	45

No.	COMPOUNDS	Molecular formula	K(L/Mg.Hr)*E4	REF.
24	l—Glutamic acid	C2H4(COOH)CH(NH2)COOH	24.17	43
25	Glycerol	CH(CH2OH)2OH	44.78	43
26	iso—Valeric acid	CH(CH3)2COOH	15.43	43
27	Propionaldehyde	C2H5CHO	4.010	43
28	n—Butylaldehyde	C3H7CHO	5.106	43
29	Benzaldehyde	C6H5CHO	25.13	43
30	Acetonitrile	CH3CN	1.455	43
31	Benzonitrile	C6H5CN	6.326	43
32	Chloroform	CHCl3	178.1	44
33	2,4—Dinitrophenol	C6H3OH(NO2)2	93.00	44
34	Nitrobenzene	C6H5NO2	147.3	44
35	1,2—Dichlorobenzene	C6H4Cl2	233.5	44
36	Tri—Br—methane	CHBr3	1.000	45
37	Br—di—Cl—methane	CHBrCl2	1.000	45
38	Di—Br—Cl—methane	CHBr2Cl	1.000	45

TABLE 2. GROUP CONTRIBUTION PARAMETERS

No.	Group		α_j
1	Methyl	CH3	−3.460
2	Methylene	CH2	−0.1354
3	Ketone	CO	−0.9650
4	Amine	NH2	−4.061
5	Acid	COOH	−3.201
6	Hydroxy	OH	−2.983
7	Aromatic CH	ACH	−0.8340
8	Aromatic C	AC	1.973
9	Chlorine	Cl	−2.352
10	Methylidyne	CH	3.223
11	Nitro	NO2	−2.448
12	Aldehyde	CHO	−3.909
13	Nitrile	CN	−5.272
14	Bromine	Br	−4.657

TABLE 3. COMPARISON OF ACTUAL AND PREDICTED Ln(K) VALUES

Compound	Actual Ln(K)	Predicted Ln(K)	Error % based on Ln
o—Cresol	−6.0890	−5.8330	4.20
m—Cresol	−5.7706	−5.8330	−1.08
p—Cresol	−5.8659	−5.8330	0.56
2,4—Dimethylphenol	−6.2472	−6.4860	−3.82
Butylbenzene	−6.5297	−6.0632	7.14
1—Phenylhexane	−6.7983	−6.3340	6.83

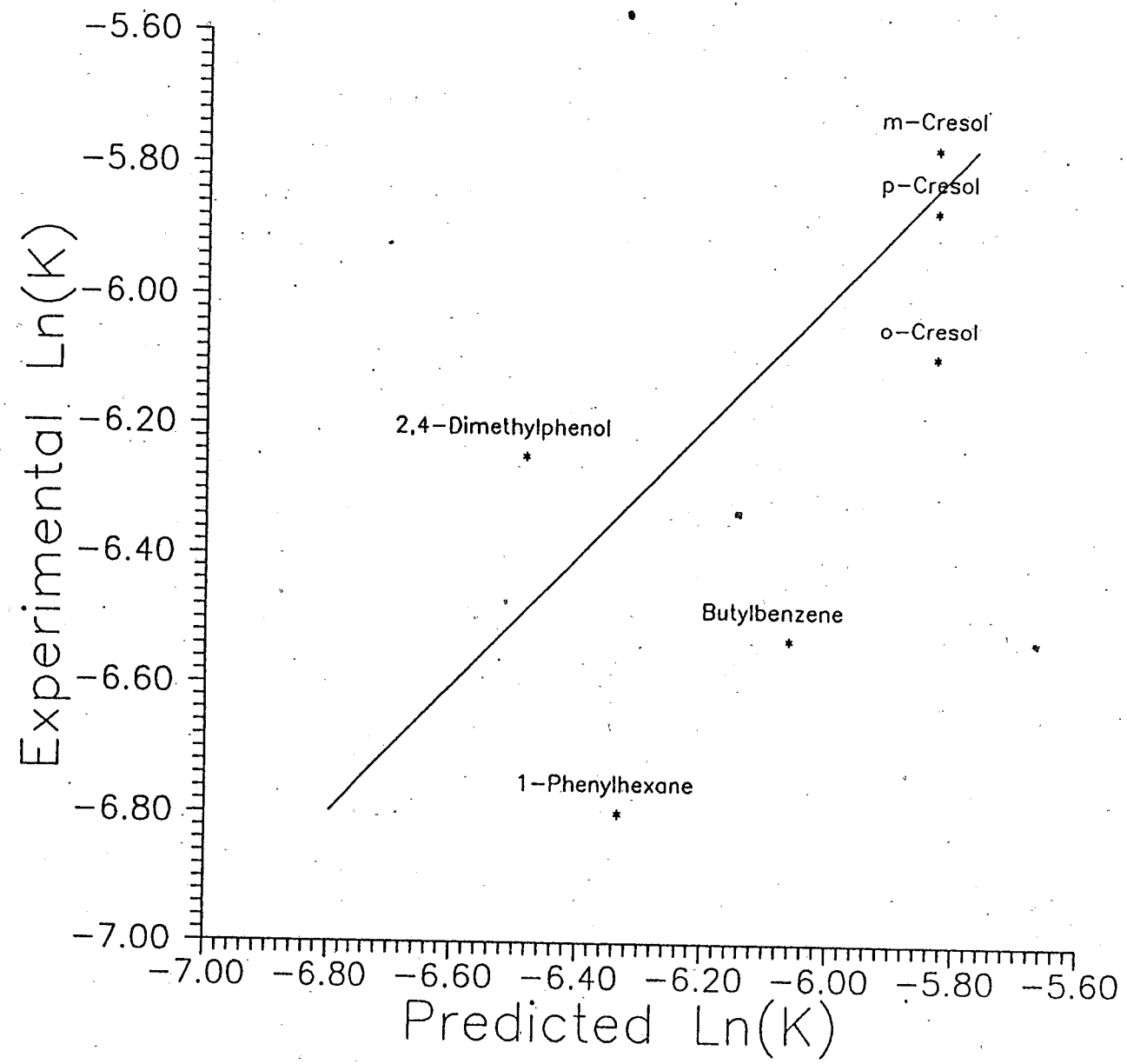

Figure 2. Comparison of experimental and predicted Ln(K)

PARTITIONING OF TOXIC ORGANIC COMPOUNDS BETWEEN WASTEWATER AND WASTEWATER SOLIDS

by

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PARTITIONING OF TOXIC ORGANIC COMPOUNDS BETWEEN WASTEWATER AND WASTEWATER SOLIDS

ABSTRACT

An experimental protocol which eliminates the effect of biodegradation and volatilization losses on the sorption measurement has been developed and applied to uptake of toxic organics by wastewater solids. Sorption capacities obtained for primary, mixed-liquor and digested solids from municipal wastewater treatment plants have been correlated with octanol/water partition coefficients and with modified Randić indexes. It has been shown that the correlations are the same for all three types of wastewater solids if partition coefficients are calculated on the basis of organic content of the solids. The correlations developed are useful for assessing the role of sorption in the treatment of toxic or hazardous compounds in conventional biological wastewater treatment plants. Correlations developed provide a basis for predicting the concentration of toxic compounds associated with wastewater solids at a given equilibrium concentration in the aqueous phase. Organics in sludge can impact anaerobic digestion, land spreading, incineration and ocean dumping of sludges. Estimates of the concentration of toxics in sludge allows assessment of the impact of toxics on sludge disposal options.

INTRODUCTION

Sorption on solids is one of the fundamental processes controlling removal of toxic organic compounds in biological wastewater treatment plants. The fate of hydrophobic organic pollutants in wastewater is highly dependent upon their sorptive behavior. The degree of sorption not only affects a chemical's mobility, but also is a dominant factor in fate processes such as volatilization and biodegradation. Organics may accumulate in sludges at concentrations several orders of magnitude greater than the raw wastewater concentrations. Concentrations factors of 50-39,000 have been reported (1, 2).

In dilute solutions which are typical of most environmental situations the partition coefficient K_p for sorption on wastewater solids can be defined as follows:

$$K_p = C_s/C_L \qquad (1)$$

where: C_s = concentration of pollutant in the solid phase and C_L = concentration in the liquid phase. Proper choice of concentration units results in a dimensionless value for K_p.

Partition coefficients are usually determined from a sorption isotherm. Wastewater solids are contacted with different initial concentrations of the toxic organic compound being studied. After the system reaches equilibrium, the concentration of toxic organics in both the solution and solid phases are measured using a suitable analytical method. Solid phase concentrations can be calculated based on the difference between the initial concentration added and the solution concentration at equilibrium. Data are fitted to the Freundlich equation which can be written:

$$X/M = KC_e^{\,n} \qquad (2)$$

where: $X = (C_o-C_e)V$ which is the amount of toxic organic sorbed from a volume of solution equal to V; C_o = initial concentration of toxic organic added; C_e = concentration of toxic organic in the solution phase at equilibrium; M = weight of wastewater solids added to the solution; K and n = empirical constants. Data are fitted to the logarithmic form of equation (2) which can be written as follows:

$$\log X/M = \log K + n \log C_e \qquad (3)$$

In dilute solutions this equation yields a straight line with a slope of n and an intercept equal to the value of K (X/M at $C_e = 1.0$) when X/M is plotted as a function of C_e on logarithmic paper. K is an indicator of sorption capacity and the slope of sorption intensity. K is also referred to as the sorption or adsorption coefficient. If X/M is in mg/gm and C_e is in mg/l, or X/M is in μg/gm and C_e is in μg/l Kp is obtained from the following equation:

$$K_p = \frac{X/M}{C_e/10^3} \quad (4)$$

The Freundlich equation is a useful method for treatment of sorption data. Since $X = C_o - C_e$, for a given volume of solution, substitution of $C_o - C_e$ for X in Equation (2) yields:

$$C_o - C_e/M = KC_e^n \quad (5)$$

which gives removal of toxic organic as a function of wastewater solids concentration. However, to relate the logarithmic form of the Freundlich equation (Eq. 3) to the equation for the partition coefficient (Eq. 4), one must assume that the slope (n) is unity. In most cases where the concentration of toxic organics in the aqueous solution are low, the slopes are generally close to unity. Deviations can result from difficulties in measuring the sorption isotherm caused by biodegradation, volatilization losses, or analytical problems.

In the present study, partition coefficients were measured for selected toxic organic compounds on raw, mixed-liquor, and digested wastewater solids. Sorption data were correlated with fundamental properties of the compounds studied.

MATERIALS AND METHODS

REAGENTS

Organic compounds were obtained from Eastman Kodak Co., Rochester, NY; Aldrich Chemical Company, Inc., Milwaukee, WI; and Chem Service Inc., West Chester, PA.*

ANALYTICAL METHODS

Analytical methods for analysis of the selected organic compounds used in the validation study are summarized in Table 1.

Volatile chlorinated organics were analyzed using a Tracor LSC-2 sample concentrator and a Tracor 222 gas chromatograph equipped with a Tracor 700 Hall electrolytic conductivity detector. Pesticides and the cresol were analyzed using a Varian Model 3700 gas chromatograph equipped with a Varian CDS-111 chromatographic data system. Dyes were determined with a Beckman Model 25 spectrophotometer equipped with a clinical sipper system.

COLLECTION OF WASTEWATER SOLIDS

Wastewater solids were collected from municipal wastewater treatment plants within a fifty mile radius of the Risk Reduction Engineering Laboratory, Cincinnati, Ohio. Wastewater plants were selected to include a wide range of domestic/industrial contribution to the total plant flow. Samples of primary sludge, mixed-liquor solids, and digested sludge were

*Mention of trade names or commercial products does not constitute endorsement or recommendation for use.

collected in a series of five gallon carboys. Solids were removed by centrifugation in a perforate bowl centrifuge. Analysis of the sludge or mixed liquor and the centrate showed solids capture of 95 ± 3% by centrifugation. The collected solids (approximately 10% on a dry weight basis) were stored at 4°C until used. No visible change was observed during the cold storage period required for completion of experimental testing for a given wastewater solid system.

TABLE 1. SUMMARY OF ANALYTICAL METHODS USED FOR ANALYSIS OF TOXIC ORGANIC COMPOUNDS

Compound	Analytical method	Conditions
Methylene chloride Chloroform 1,1-dichloroethylene Carbon tetrachloride Tetrachloroethylene Chlorobenzene	Purge and trap Gas chromatography	Tenax trap; 3% Carbowax 1500 on Chromosorb column Hall detector
Lindane Heptachlor epoxide Dieldrin	Hexane extraction Gas chromatography	1.5% SP-2250/1.9% SP-2401 column Electron capture detector
2,4-dinitro-o-cresol	Freon extraction Gas chromatography	1% SP-1240 DA on Sepelcoport column flame ionization detector
Methylene blue dye Azo dye 88 Azo dye 151	Visible spectroscopy	665 nm 505 nm 512 nm

EXPERIMENTAL PROCEDURE

Experimental Protocol Development

Key issues to be resolved in the development of a protocol for measuring partition coefficient include: (1) effect of solids-to-water ratio on the partition measurement; (2) rate of attainment of equilibrium; and (3) methods to eliminate biodegradation during the equilibration period.

Several reports in the literature state that the equilibrium partition coefficient (K_p) decreases as the ratio of solids-to-water increases (3, 4, 5). These observations challenge the classical view of the sorption process. The two procedures available for sorption capacity measurements involve the use of a constant concentration of compound with varied amounts of solids or the use of a varied concentration of compound with a constant amount of solids. If solids-to-water ratio affects the measurement of partition coefficients, the first method described should not be used. In order to resolve this issue, a preliminary set of isotherms were measured using both methods. The test involved determination of the sorption or partition coefficient for methylene blue dye using mixed-liquor solids.

In the first of the two procedures, a constant methylene blue dye concentration of 10.0 mg/L was used with mixed-liquor solids concentrations of 0, 0.05, 0.10, 0.25, 0.30, and 0.50 g/L on a dry weight basis. These solids doses selected gave a 10-fold variation in solids-to-liquid ratio and are typical of the conditions used to measure partition coefficients. In the second procedure, a constant mixed-liquor solids concentration of 0.1 g/L dry weight was used in conjunction with methylene blue dye concentrations of 0, 2.5, 5.0, 7.5, 10.0, and 20.0 mg/L. Methylene blue dye was chosen for the preliminary test because it is non-biodegradable, non-volatile, and easily determined by visible spectroscopy at 665 nm. Test samples containing the specified solids and dye concentration were prepared and equilibrated for 8 hours. Solids were separated by centrifugation and the concentration of dye measured. The isotherms obtained by both methods are shown in Figure 1 where X/M is plotted as a function of C_e according to the logarithmic form of the Freundlich equation (Eq. 3).

The sorption coefficients (K values) calculated from the least squares treatment of the separate isotherm data were the same for both methods (22.2 mg/gm). The pooled data yielded a K value of 22.6 mg/gm in excellent agreement with the two individual isotherm values. These data indicate that the sorption coefficient and the partition coefficient remain constant over the range of solid-to-water ratios used in this study. It has been emphasized that if precautions are taken to eliminate or account for nonsettling (or non-filtrable) microparticles which remain in the aqueous phase during sorption tests, the partition coefficient does remain constant (6) in agreement with the present study. Even though either method can be used to experimentally determine sorption and partition coefficients, the second method discussed was selected for use in the present study. The choice was based on the desire to maintain a constant solids level in the samples during the solids' separation step by centrifugation or filtration prior to analysis. A second reason for the use of constant solids in the sorption test is that frequently measurable levels of a given test compound

will be found in the wastewater solids as a natural background. If this occurs, it is easier to correct analytical values of the equilibrium concentrations in the samples by a constant amount.

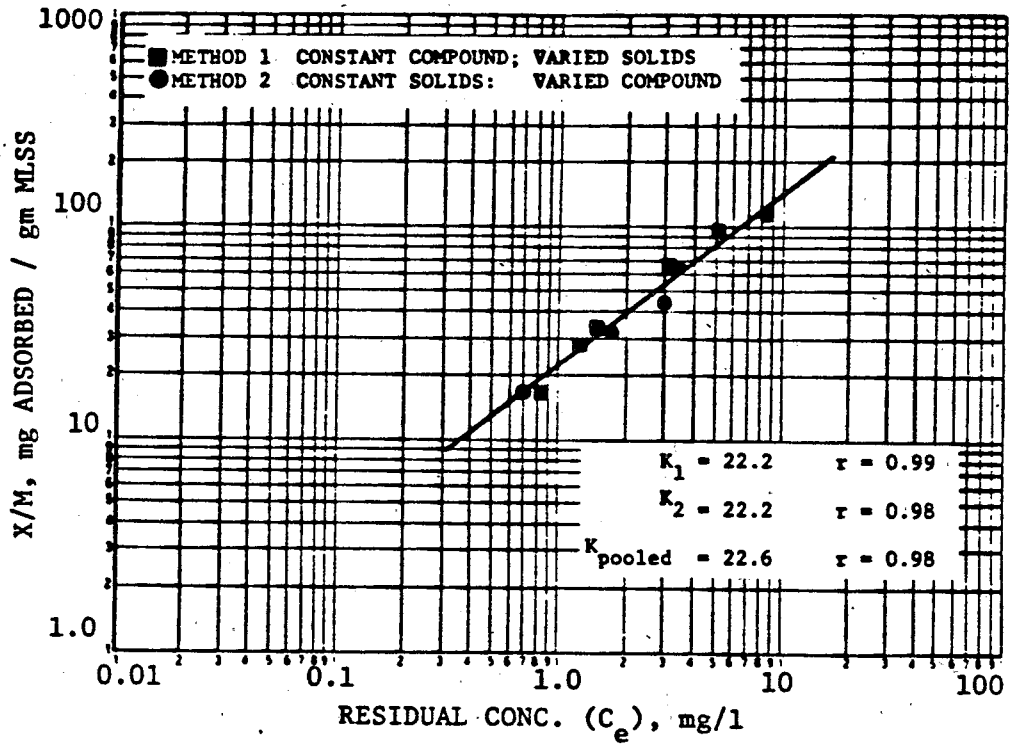


Figure 1. Comparison of two isotherm methods for sorption capacity measurements.

A great deal of confusion exists in the literature regarding the rate of attainment of sorption equilibrium. Sorption kinetics for phenol and 1,4-dichlorobenzene were reported to be exceedingly fast (equilibrium in < 4 minutes) (7). In the case of polynuclear aromatic hydrocarbons on natural sediments 35 to 60% of the equilibrium sorption capacity was achieved in minutes (mixing limited) with apparent equilibrium after a few hours (8).

Based on the conflicting reports in the literature on the rate of attainment of equilibrium, kinetic measurements on the sorption of methylene blue dye on mixed-liquor solids were made to determine the proper contact time for experimental purposes. Mixed-liquor solids from three different municipal wastewater treatment plants were collected. Rate of sorption of methylene blue dye from a 10 mg/L solution was measured at 0.1, 0.2, and 0.5 g/L dry weight solids. All three mixed-liquor solids concentrations from the same wastewater treatment plant used in the isotherm runs are presented in Figure 2.

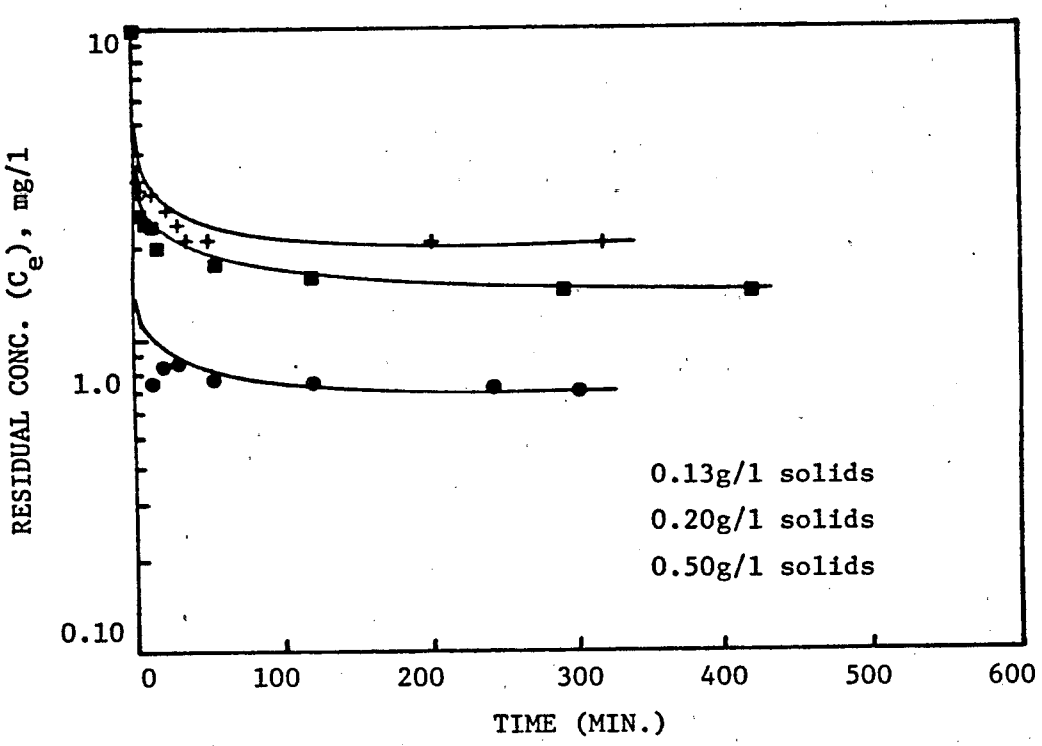


Figure 2. Kinetics of sorption of methylene blue dye on mixed-liquor solids.

The kinetic data for the sorption of methylene blue dye are in agreement with the two box model proposed for hydrophobic pollutants in water (9). The two box model described sorption dynamics in terms of a rapid or "labile" exchange and a highly retarded or "non-labile" sorption requiring days to weeks to occur. In the case of methylene blue dye and 0.5 g/L mixed-liquor solids dose, 92% of the compound was sorbed in approximately 30 minutes, while the remainder required several hours. In order to minimize biodegradation during the sorption test, an equilibration period of 6 hours is recommended. The remaining organic compounds were equilibrated for 6 hours. This avoids overnight contact and permits samples to be separated from the solids the same day.

The third issue to be resolved in the protocol development studies involved methods to eliminate biological degradation of test compounds during the partitioning process. It is obvious that other removal mechanisms such as biodegradation and volatilization would have to be controlled if accurate sorption data are to be obtained. Otherwise, measurements of changing aqueous phase compound concentrations would include the effects of all removal mechanisms resulting in apparent sorption capacities greater than actually achieved. Volatilization can be eliminated from sorption measurements by using completely filled containers with zero headspace. The impact of biodegradation on the sorption measurements was more difficult to control. Lyophilization followed by dry heat treatment has been reported to produce biological solids which had flocculation and settling properties which were indistinguishable from live biomass (7).

These authors claimed the batch experimental technique with nonviable biomass appeared to give realistic results compared to tests with viable biomass. Based on the reported successful use of lyophilization for control of biological activity, this technique was evaluated. A batch of mixed-liquor solids was collected and separated from the supernatant in a perforate bowl centrifuge. The moist solids were divided into two portions. One portion was freeze-dried and the other was stored at 4°C. Sorption isotherms were measured using 2,4-dinitro-o-cresol. The isotherms were conducted at pH 3.0 for two reasons. First, the low pH would inhibit or eliminate biological activity in the test with viable biomass. Secondly, the decrease in pH would convert the 2,4-dinitro-o-cresol to the undissociated (sorbable) form (pka = 4.35). Results of the isotherm comparison are shown in Figure 3.

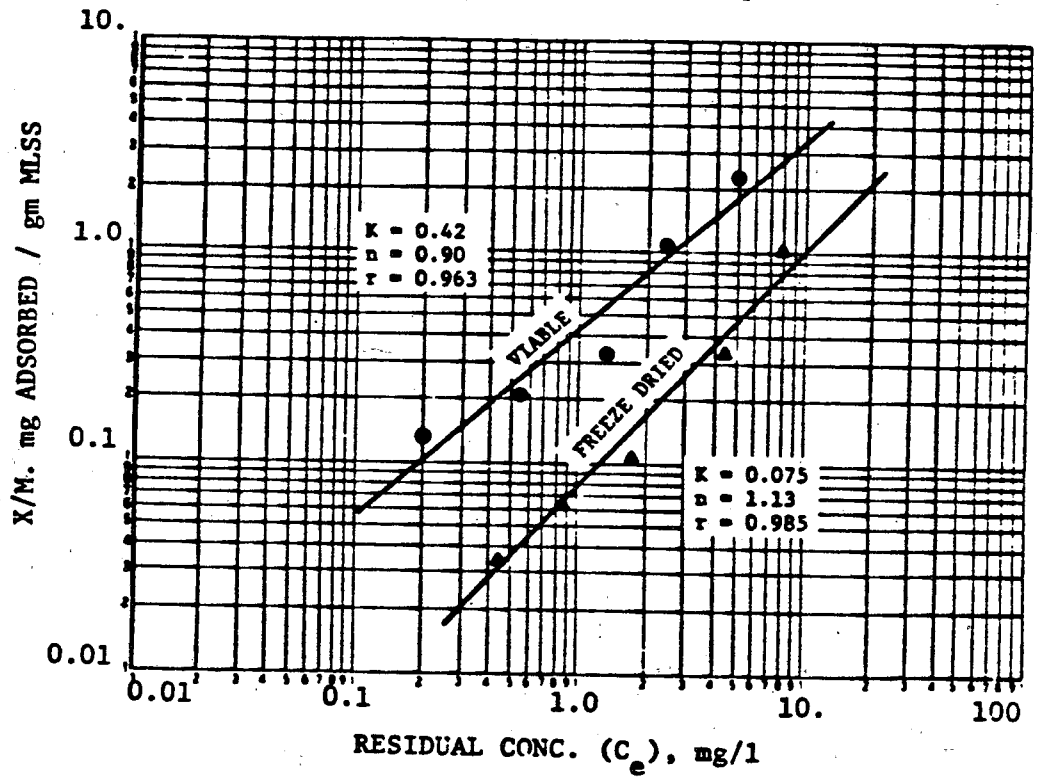


Figure 3. Comparison of freeze-dried and viable biomass for sorption of 2,4-dinitro-o-cresol at pH 3.0.

The use of freeze-dried solids in the isotherm procedure was not acceptable based on the results shown in Figure 3. The viable biomass clearly has a significantly greater sorption capacity than the freeze-dried material. A similar result has been reported for sorption of azo and triphenylmethane dyes on microbial populations (10).

Alternate techniques to control biodegradation were evaluated. Wastewater samples are often preserved prior to analysis by adjusting pH to 3 or less. Preliminary experiments indicated that this approach would minimize biological activity for primary and mixed-liquor solids but did not completely inhibit digested sludge. In addition, the technique would be limited to neutral organics, since pH change would alter charge and dissociation of acidic and basic organic compounds which would impact partitioning.

Two additional methods were evaluated as potential treatments to eliminate biodegradation during the equilibration period in the isotherm test. Pretreatment of the wastewater solids with cyanide was attempted based on the reported potent inhibition of methanogensis and carbon metabolism by this anion (11, 12, 13). Carbon tetrachloride was selected as a model compound to evaluate the required contact time to inactivate the biomass and the ratio of cyanide to dry weight solids required to minimize biodegradation over the 6-hour period recommended for equilibration. Results indicated that treatment of the wastewater solids for 24 hours using a cyanide to dry weight solids ratio of 0.1-0.2 was optimum to prevent biodegradation of the easily degradable carbon tetrachloride. In the second method wastewater solids were pasteurized at 85°C for thirty minutes prior to being used in the isotherm procedure. The effectiveness of the two methods for eliminating biodegradation is compared in Table 2 where whole samples (solid and liquid phases) were analyzed for five chlorinated solvents initially and after 6 hours of mixing.

TABLE 2. EFFECT OF CYANIDE ADDITION AND PASTEURIZATION ON BIODEGRADATION DURING ISOTHERM EQUILIBRATION

	Concentration of toxic, µg/L					
	Control		Cyanide addition		Pasteurization	
Compound	0 hr	6 hr	0 hr	6 hr	0 hr	6 hr
1,1-dichloroethane	106	91	113	100	103	99
Chloroform	100	103	97	94	101	100
Carbon Tetrachloride	100	49	100	101	99	95
Tetrachloroethylene	105	107	102	100	100	96
Chlorobenzene	100	109	93	97	97	106

Results in Table 2 demonstrate that both cyanide addition and pasteurization were effective methods to eliminate biodegradation during the isotherm test. Only carbon tetrachloride showed biodegradation losses in the control samples after the 6 hour equilibration period. Pretreatment of wastewater sludges and solids by either method will eliminate the effect of biodegradation on the sorption isotherm. The recommended procedure is to pretreat the solids by the method of choice, separate from the supernatant liquid, and conduct the isotherm test as soon as possible. One should not pretreat solids and store for later use.

With the completion of the preliminary experiments a general overall protocol for determination of the partition coefficient (sorption-isotherm) is outlined in Figure 4.

MEASUREMENT OF SORPTION ISOTHERMS

In the present study, the following general procedure was used. The moisture content of the solids from the perforate bowl centrifuge was determined by drying a portion to constant weight at 110°C. A stock slurry was prepared to give the desired concentration of solids for the isotherm test. Equal aliquots of the stock slurry were added to a series of centrifuge bottles. A different concentration of the organic compound was added to each centrifuge bottle from a stock solution of the toxic in methanol. The bottles were completely filled and capped to avoid loss of volatiles to the headspace or atmosphere. Isotherm samples were mixed on a multiple magnetic stirrer at approximately 100 rpm for six hours. After equilibration, the solid and liquid phases were separated by high-speed centrifugation and analyzed. Blanks, standards, and whole samples (both aqueous and solid phases) were also processed through the procedure and analyzed for quality control purposes.

Sorption isotherms were measured using the general procedure described for a selected organic compound on the three types of wastewater solids. Chemical structures for the toxic organic compounds investigated are shown in Figure 5.

Experimental Freundlich parameters for sorption on mixed-liquor solids are summarized in Table 3.

Sorption coefficients on mixed-liquor solids ranged from 0.056 mg/gm for methylene chloride to an extrapolated value of 38.9 mg/gm for dieldrin at equilibrium concentrations of 1.0 mg/L. Slopes (n values) were generally close to 1.0 and ranged from 0.57 to 1.28 for sorption on mixed-liquor solids. Slopes for sorption on soils have been reported to range from a low of 0.3 to a high of 1.7 (14). Compiled data on 26 chemicals (mostly pesticides) had a mean value of n of 0.87 with a coefficient of variation of ± 15% (15). If a measured value is not available, it is frequently assumed to be equal to 1.0. Excellent correlation coefficients (r values) were obtained for all the isotherms measured.

Isotherms for the chlorinated solvents listed in Table 3 were measured using primary sludge and anaerobically digested sludge to complete the testing of municipal wastewater treatment plant solids. Freundlich parameters obtained are presented in Table 4.

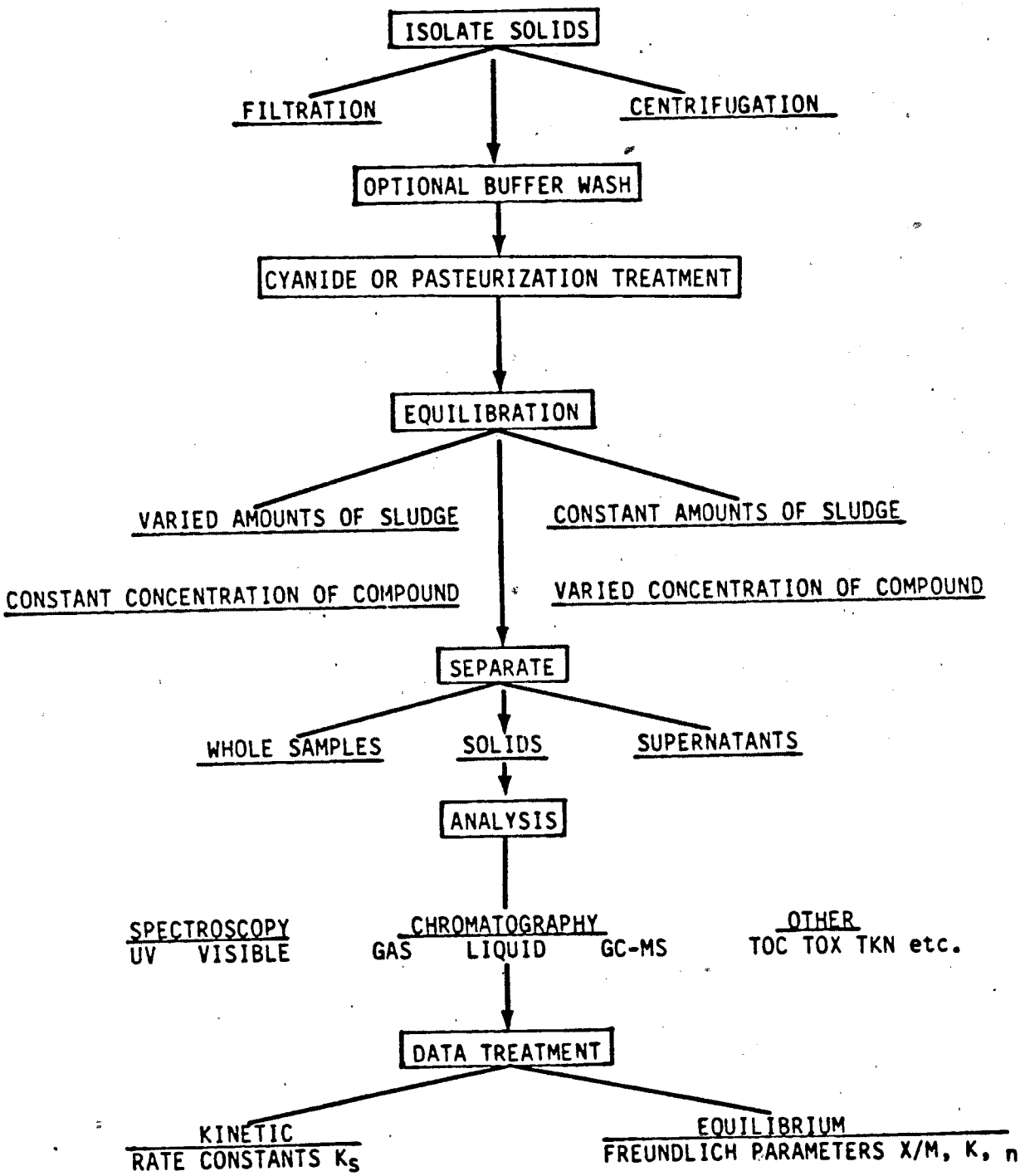
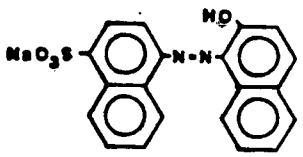
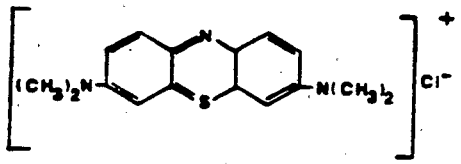


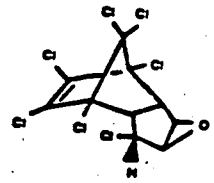
Figure 4. Protocol for sorption of toxics on wastewater solids.



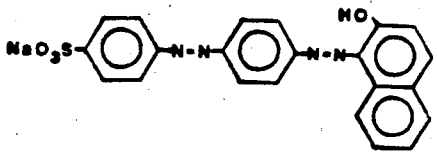
Azo Dye 88



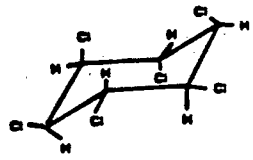
Methylene Blue



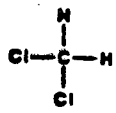
Heptachlor Epoxide



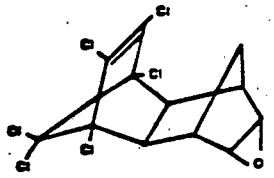
Azo Dye 151



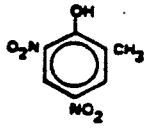
Lindane



Methylene Chloride

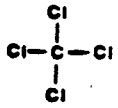


Dieldrin

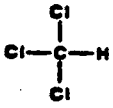


2,4-dinitro-o-cresol

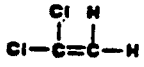
Chlorobenzene



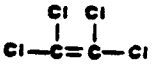
Carbon tetrachloride



Chlorofrom



1,1-dichloroethylene



Tetrachloroethylene

Figure 5. Structures of organic compounds used in sorption studies.

TABLE 3. SUMMARY OF FREUNDLICH PARAMETERS FOR SORPTION OF TOXIC ORGANICS ON MIXED-LIQUOR SOLIDS

Compound	K, (mg/gm)*	n+	r#
Methylene chloride	0.056	1.28	0.85
Chloroform	0.094	0.90	0.98
1,1-dichlorethylene	0.150	0.71	0.96
Carbon tetrachloride	0.497	0.57	0.95
Chlorobenzene	0.285	0.96	0.96
2,4-dinitro-o-cresol	0.420	0.90	0.96
Tetrachloroethylene	0.897	1.12	0.98
Lindane**	0.762	1.00	0.99
Heptachlor epoxide**	17.9	1.09	0.98
Dieldrin**	38.9	1.11	0.99
Methylene blue dye	22.6	0.82	0.98
Azo dye 151	8.30	1.00	0.99
Azo dye 88	1.35	0.65	0.99

*K = sorption coefficient = X/M at C_e = 1.0 mg/L

+n = slope of isotherm plot

#r = correlation coefficient

**K values extrapolated to C_e = 1.0 mg/L for comparison purposes

TABLE 4. SUMMARY OF FREUNDLICH PARAMETERS FOR SORPTION OF CHLORINATED SOLVENTS ON PRIMARY AND ANAEROBICALLY DIGESTED SLUDGES

Compound	Primary sludge			Digested sludge		
	K,(mg/gm)	n	r	K,(mg/gm)	n	r
Methylene chloride	0.188	1.08	0.75	0.075	0.80	0.94
Chloroform	0.267	0.81	0.99	0.054	0.90	0.61
1,1-dichloroethylene	0.142	0.54	0.93	0.130	0.82	0.96
Carbon tetrachloride	0.339	0.41	0.98	0.203	1.18	0.97
Chlorobenzene	0.483	1.05	1.00	0.285	0.76	0.98
Tetrachloroethylene	0.596	1.31	0.98	0.698	1.47	1.00

DISCUSSION OF RESULTS

In soil and sediment sorption, the extent to which an organic compound partitions itself between the solid and solution phases is expressed in terms of a parameter K_{oc}, which is largely independent of the properties of the soil or sediment (16, 17). Once K has been determined for a soil or sediment, it is divided by % organic carbon (OC) contained in the soil or sediment and multiplied by 100 to obtain K_{oc}. In the case of wastewater solids used in the present study, organic carbon content was based on weight loss on ignition at 600°C and is defined by the following equation:

$$K_{om} = K \cdot \frac{100}{\%VSS} \qquad (6)$$

where K_{om} = amount sorbed per unit weight of organic matter in solids and VSS = volatile suspended solids. Sorption data are usually correlated based on organic carbon or organic matter. Therefore, the sorption coefficient, K, must be converted to K_{om} for correlation purposes. Thus, we can define a corrected partition coefficient by the following equation:

$$K_p' = 1000\ K_{om} \qquad (7)$$

Emphasis on the environmental impact of toxic organic chemicals has resulted in increased reliance on the physical-chemical properties of these compounds to assess their environmental behavior. One such parameter, the octanol/water partition coefficient (K_{ow}), has proved useful as a means for predicting soil adsorption (18), biological uptake (19), lipophilic storage (20), and biomagnification (21-24). K_{ow} is defined by the following equation:

$$K_{ow} = C_o/C_w \quad (8)$$

where C_o = equilibrium concentration of organic compound in octanol layer and C_w = equilibrium concentration of organic compound in water layer.

Based on the soil-sediment literature, an attempt was made to correlate sorption on wastewater solids with the octanol/water partition coefficient for the compounds used in the present study. Sorption coefficients in Tables 3 and 4, were converted to K_{om} values using Equation 6. The K_{om} values obtained were used to calculate corrected partition coefficients (K_p' values) using Equation 7. Log of the average K_p' values for sorption of the toxic organic compounds on the three different types of wastewater solids are summarized in Table 5 along with the corresponding octanol/water partition coefficients. Log of the average K_p' values obtained for the three types of wastewater solids were used for each compound for correlation purposes. Justification for averaging the log K_p' values was based on replicate isotherms measured for methylene blue dye on the three kinds of solids studied. Log K_p' values measured for sorption of dye on separately collected replicate samples of wastewater solids averaged 4.35, 4.57, and 4.41 for primary solids, mixed-liquor solids, and digested solids, respectively. All individual replicates were within ± 0.1 of the average values. Thus, it was concluded that when sorption coefficients are based on

TABLE 5. AVERAGE LOG K_p' VALUES USED TO CORRELATE SORPTION ON WASTEWATER SOLIDS WITH OCTANOL/WATER PARTITION COEFFICIENTS

Compound	Log K_p'	Log K_{ow}*
Methylene chloride	2.06	1.26
Chloroform	2.15	1.97
1,1-dichloroethylene	2.23	2.13
Carbon tetrachloride	2.65	2.64
Chlorobenzene	2.64	2.84
Tetrachloroethylene	3.00	2.85
2,4-dinitro-o-cresol	2.77	2.88
Lindane	3.41	3.85
Heptachlor epoxide	4.19	5.40
Dieldrin	4.40	5.48

*K_{ow} values were taken from reference (25) or were calculated by the fragment method from reference (26).

organic matter in the sorbent, all three types of solids are equivalent. Methylene blue dye is an ideal model compound for sorption experiments because it in non-volatile, non-biodegradable and can be determined accurately in the complex matrices used in the isotherm procedure.

A simple linear regression of log K_p' on log K_{ow} yielded a correlation coefficient (r) of 0.963. A plot of log K_p' versus log K_{ow} is shown in Figure 6. The following equation was obtained for the relationship between partitioning on wastewater solids and octanol/water partition coefficient:

$$\log K_p' = 1.57 + 0.57 \log K_{ow} \quad (9)$$

If K_{ow} is given in numerical form, the equation can be written as follows:

$$K_p' = 37.2(K_{ow})^{0.57} \quad (10)$$

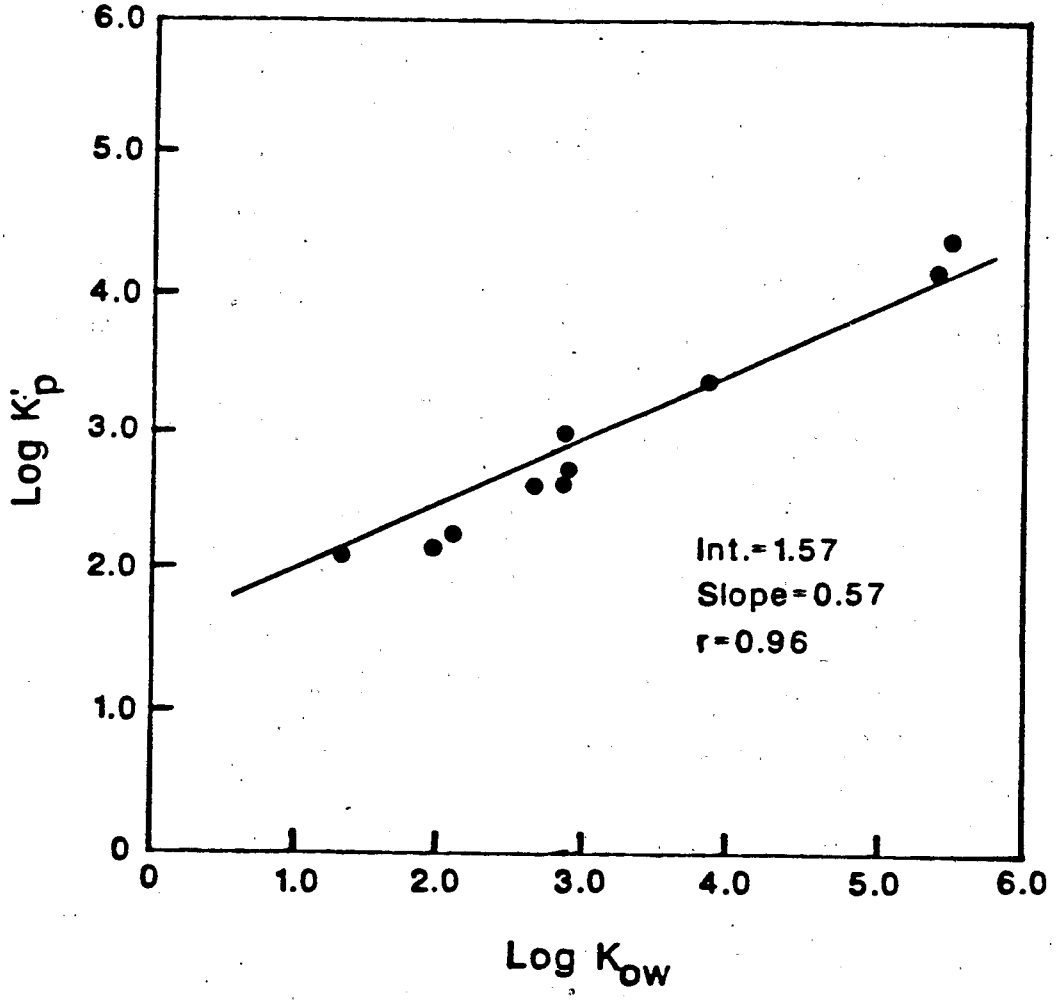


Figure 6. Correlation of partition coefficients with octanol-water partition coefficients.

Octanol/water partition coefficients are generally accepted as accurate estimates of soil sorption coefficients. However, at least two major difficulties have been described (27). First, the precision of available K_{ow} data is low; and secondly, a variety of quantitative linear models describing the relationship between K_{ow} and soil partition coefficients have been proposed, hence the choice of the model to be used for a given system becomes difficult. The same problems are encountered in correlating sorption on wastewater solids with K_{ow}.

Methods based on the topology of molecules have been found to predict a broad range of properties and have found applications beyond simple predictions of chemical behavior (28). The Randić molecular-connectivity index has a wider range of application than others yet devised. Significant correlations have been reported between the Randić index and log K_{ow} (29). The index was first described by Randić (30) and has since been applied to prediction of soil sorption coefficients for polycyclic aromatic hydrocarbons, alkylbenzenes, chlorobenzenes, chlorinated alkanes and alkenes, and halogenated phenols (27). Several extensive reviews of both the theory and calculation of molecular and connectivity indexes have been published (29-33). The conventional first-order molecular connectivity indexes are calculated from the non-hydrogen part of the molecule. Each non-hydrogen atom is described by its atomic δ value, which is equal to the number of adjacent non-hydrogen atoms. This index is then calculated from the atomic values by the following equation:

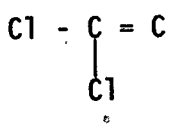
$$ {}^{1}\chi^{v} = \Sigma(\delta_i \delta_j)^{-0.5} \qquad (11) $$

where δ_i and δ_j correspond to the pairs of adjacent non-hydrogen atoms and summation is over all bonds between non-hydrogen atoms. In the present study heteroatoms such as oxygen and nitrogen are included in the analysis by using the appropriate valence assigned to the heteroatom. Thus,

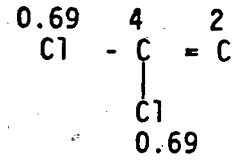
$$ \delta_i = Z^v - h_i \qquad (12) $$

where Z^v is the number of valence electrons in the site and h_i is the number of hydrogen atoms. Therefore, for oxygen $Z^v = 6$ and for nitrogen $Z^v = 5$. In the case of chlorine, a value of 0.690 has been assigned (30). The basic assumption is that a multiple bond in a molecule should be represented by the endpoint valencies derived from the formula for δ_i as described earlier. Double bonds are counted as double edges. To illustrate, the modified Randić index ($^{1}\chi^{v}$) for 1,1-dichloroethylene is calculated as follows:

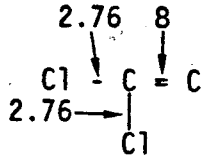
(1) Determine valence weighted graph (omit hydrogen atoms):



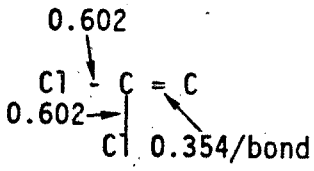
(2) Calculate valence at node or vertex (vertex valence = number of valence electrons = number of H atoms):



(3) Compute product of vertex valencies for each edge:



(4) Compute each edge term as the reciprocal square root product (e.g., $\frac{1}{\sqrt{8}} = 0.354$):



(5) Compute sum of the edge terms.

Therefore, $1_X^v = 2(0.602) + 2(0.354) = 1.912$.

Correlation of the modified Randić index with the corrected partition coefficients for sorption on wastewater solids was attempted using the average Log K_p' calculated from the Freundlich parameters from Table 5 and the corresponding values for log 1_X^v presented in Table 6.

A simple linear regression of log K_p' on 1_X^v yielded a correlation coefficient (r) of 0.972. A plot of log K_p' versus 1_X^v is shown in Figure 7. The following equation was obtained for the relationship between partitioning on wastewater solids and the modified Randić index:

$$\log K_p' = 1.79 + 0.29\ 1_X^v \qquad (13)$$

Methylene blue dye, and the two azo dyes used for sorption experiments were not included in the correlation with the modified Randić index because they were ionic species. Methods exist to treat polar and ionic compounds and these will be evaluated in future research work.

TABLE 6. MODIFIED RANDIC INDEX USED TO CORRELATE SORPTION ON WASTEWATER SOLIDS

Compound	$^1\chi^v$
Methylene chloride	1.70
Chloroform	2.08
1,1-dichloroethylene	1.91
Carbon tetrachloride	2.41
Chlorobenzene	2.51
Tetrachloroethylene	2.91
2,4-dinitro-o-cresol	3.46
Lindane	6.17
Heptachlor epoxide	8.52
Dieldrin	8.98

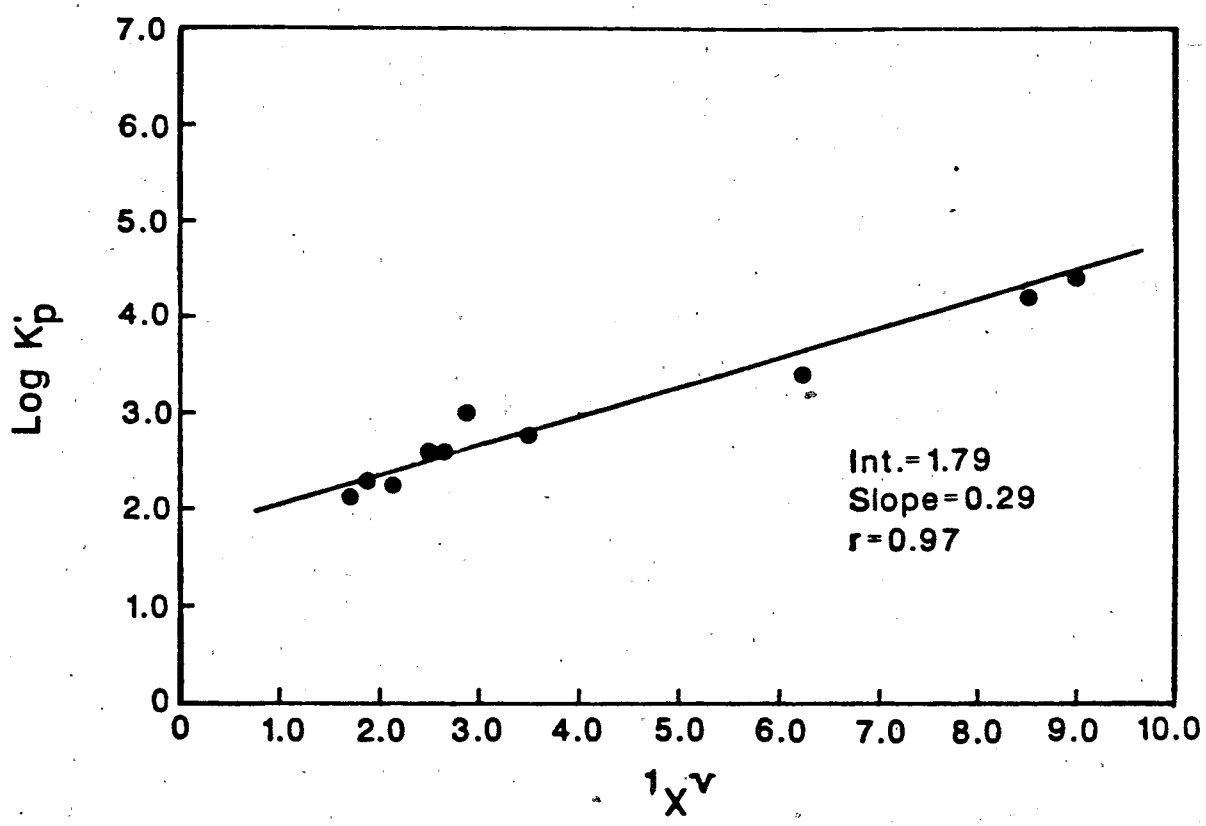


Figure 7. Correlation of partition coefficients with modified Randić indexes.

SUMMARY AND CONCLUSIONS

Sorption of toxic organics on primary sludge, mixed-liquor solids and digested sludge have been correlated with the octanol water partition coefficient and with modified Randić indexes. The same correlation is valid for all types of wastewater solids if results are expressed on the basis of the organic content of the solids as measured by weight loss on ignition at 550-600°C. Octanol/water partition coefficients are useful for estimating uptake of toxic organics on wastewater solids. However, the modified Randić indexes offer a more fundamental method for assessing sorption of toxic organic compounds based on structural features of the molecule. Correlations based on Randić indexes avoid the problems of low precision in the reported values for K_{ow} and provide at least a theoretical basis for predicting sorption of complex mixtures of organic compounds.

The relationships developed provides a basis for estimating the removal of toxic and hazardous organics from wastewater by the sorption process. The equations presented can be used to calculate removal of toxic organic compounds from wastewater treatment plants if the compounds are non-volatile and non-biodegradable. In addition, toxic organic compounds sorbed on

wastewater solids can impact treatment and disposal of sludges. Toxic and hazardous compounds can affect anaerobic digestion processes, land spreading, incineration, and ocean-dumping of sludges. The correlations developed provide a basis for assessing the impacts of sorbed organics on the sludge disposal options.

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LEVELS AND REMOVAL OF SELECTED INDICATOR AND PATHOGENIC MICROORGANISMS DURING CONVENTIONAL ANAEROBIC SLUDGE DIGESTION

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ABSTRACT

Samples of wastewater and wastewater sludge were collected on an approximate bi-weekly routine over a thirteen month period at three wastewater treatment plants in the Baltimore vicinity to characterize the reduction in the levels of selected indicator and pathogenic microorganisms through conventional anaerobic sludge digestion. The levels of selected microorganisms and human pathogens in the raw wastewater entering the three treatment plants were similar. The levels of human pathogenic bacteria, viruses and parasites were low in all the samples compared to the levels of indicators. Salmonella spp. were consistently found in raw wastewater, raw sludge, treated sludge and wastewater effluent. Parasites were found in most of the samples of raw wastewater. raw sludge and digested sludge but were rarely observed in the wastewater effluent. Enteric viruses were infrequently recovered from raw wastewater but found in many of the raw sludge samples and few of the digested sludge samples. Viruses were not recovered from the wastewater effluent. Correlations significant at the 95% and 99% level were found between the levels of indicator microorganisms and levels of pathogens. However, the correlation coefficients were low and had little predictive value. No indicator microorganism yielded particularly high or consistently higher correlation coefficients. Fewer significant correlations were found when the data was calculated as the #/gTS and #/gVSS. The anaerobic digestion process yielded consistent reductions in the densities of selected indicator and pathogenic microorganisms but varied at each of the three treatment plants. Approximately 1 to 3 log reductions were observed for the coliform indicators, while only 0.2 to 2 logs reduction were found for the streptococci. Salmonella were reduced 1.3 to 3.3 orders of magnitude. The log reductions for human virus and total parasites were difficult to determine because of low numbers but were less than those found for the bacteria. Only 0.4 to 1.4 logs reduction for virus and 0.5 to 1.2 logs reduction for parasites were observed. Correlations significant at the 95% and 99% level were found between the log reduction of indicator and pathogenic microorganisms. Correlations were generally higher (as high as 0.65) than that observed for the levels of microorganisms. Again no particular indicator consistently yielded higher correlations. When the correlations were calculated for each treatment plant, fewer significant correlations were found.

INTRODUCTION

Samples were collected at three wastewater treatment plants in the Baltimore vicinity to characterize the levels of reduction of selected indicator and pathogenic microorganisms during anaerobic sludge digestion to provide the necessary database to prepare and support regulations for the handling of domestic wastewater sludge including disposal to land. In addition, the study investigated the relationship between solids, microbial indicators and pathogens in wastewater and sludge during conventional sludge treatment processes to evaluate effective monitoring parameters that reflect the pathogen reductions.

METHODS

SAMPLING

Samples were collected after thorough flushing of the available sample lines to insure a representative sample. Sterile 2 to 5 gal tubs with sealable tight fitting covers were used for ease of handling and wide mouth to accommodate the thick samples. After collection, samples were transported immediately to the Water and Wastewater Microbiology Laboratories at The Johns Hopkins University in Baltimore and were processed immediately upon arrival. Typical travel times to the laboratory after sample collection were 20, 40 and 60 minutes for Back River, Sod Run and Blue Plains, respectively. The routine collection of samples was begun on August 26, 1987 and continued through September 1988. In addition, samples for the determination of human parasites were cooled and shipped by overnight express to Dr. Jan Sykora at the University of Pittsburgh. Samples for human enteric viruses were concentrated and the concentrates shipped frozen to Dr. Robert Safferman at the USEPA laboratory in Cincinnati, Ohio.

Each sample was mixed with a "lightning mixer" for 5 minutes at about 200 revolutions per minute prior to dispensing. Samples were dispensed with continuous mixing for analysis. Microbiological assays that required subsequent dilutions were mixed with an equal volume of 1 % peptone and blended at high speed in a Waring blender for 1.0 minute. Immediately after blending, the samples were diluted and plated.

CULTURE AND RECOVERY METHODS

A major effort during the preliminary phase was the comparative evaluation of culture methods for the selected indicator and pathogenic microorganisms. The intent of this portion of the study was not to develop new methods but to evaluate existing procedures and techniques to yield the most precise and accurate determination of the levels of microorganisms. Table 1 shows the selected microorganisms assayed and the method employed.

CHEMICAL AND PHYSICAL MEASUREMENTS

Total solids (TS) was performed at 103 C according to Standard Methods (APHA 1985). Total fixed solids (TFS) were determined at 550 C according to Standard Methods (APHA, 1985). Total volatile solids (TVS) were determined by subtraction (total solids-fixed solids). Suspended solids (SS) was performed at 103 C with Gooch crucibles containing Whatman 934-AH according to Standard

Methods (APHA, 1985). Fixed suspended solids (FSS) were determined at 550 C according to Standard Methods (APHA, 1985). Volatile suspended solids (VSS) were determined by subtraction (Suspended solids-fixed suspended solids). The pH was determined electrometrically according to Standard Methods (APHA, 1985).

RESULTS

LEVELS OF MICROORGANISMS

The logarithmic mean level per 100 ml for each of the microorganisms assayed is shown in Table 2a,b and c. The density of microorganisms for the raw sewage at the three treatment plants was remarkably similar. Higher logarithmic mean levels per 100 ml were found for the raw sewage and undigested sludge, while the plant effluents and digested sludges were lower. Similar trends were observed for the data presented per unit solids except that the logarithmic mean levels per unit solids for the plant effluents were not as reduced as that observed for unit volume. The most numerous indicators were the coliforms with total coliforms higher than fecal which were higher than E. coli. The fecal streptococci were about 2 orders of magnitude lower than the total coliform with fecal streptococci consistently higher than the enterococci. The logarithmic mean levels of Clostridium and Mycobacteria were about 3 and 4 orders of magnitude lower than the coliforms. The slow growing or late Mycobacteria were consistently higher than the fast growing Mycobacteria. The logarithmic mean levels of coliphage were generally 2 to 3 orders of magnitude lower than the levels of coliforms. The logarithmic mean levels observed for phage counted on E. coli C were similar to the levels observed for male specific phage counted with Havelaar's strain.

The pathogenic microorganisms were all observed at relatively low levels. In many of the samples, particularly the digested sludges, and effluents no pathogens were recovered. The logarithmic mean levels of Salmonella were about 5 orders of magnitude lower than the coliforms but about 2 orders of magnitude higher than the human enterovirus virus and parasites. Salmonella was consistently isolated from raw wastewater and sludges and less frequently found in digested sludges.

The level of human enteric viruses was very low in all samples. The levels of virus in the raw wastewater was particularly low and suggests that the level of virus was low in the populations served by the three plants. Only a few viruses were observed per 100 ml of the sludge samples.

Parasites were recovered from most of the samples collected from the early part of the treatment process. The average levels observed in raw sewage were less than 10/100 ml, but in primary, trickling filter sludge, and activated sludge, the levels were 29-170,45-150, 14-66 /100 ml, respectively. After anaerobic digestion, the frequency of the isolation of parasites decreased slightly. When found, the average level of parasites was 14-60/100 ml. The cyst concentrations in raw sewage ranged between 0 and 30 cysts per 100 ml, whereas the concentrations in final effluents ranged from 0 to 1 cyst per 100 ml.

Cysts of Giardia were the most frequently recovered parasite. Helminths were infrequently found and other human parasites were at best occasionally observed in the samples of wastewater and sludges.

CORRELATION OF LEVELS OF MICROORGANISMS

The correlation coefficients for the level of selected indicator

microorganisms and the level of selected human pathogens for all samples collected at the three wastewater treatment plants are shown in Table 3. The levels of microorganisms are presented in 3 different formats; #/100 ml, #/gTS and #/g VSS. The second number presented below each correlation coefficient was the number of sample pairs used to calculate the correlation coefficient. The significance was evaluated by determining the probability that there was no correlation between the levels of indicators and pathogens, the null hypothesis. The significance was indicated at the 95 % level by an asterisk (*) and at the 99% level by a double asterisk (**).

Significant correlations were observed at the 99% level for the #/ 100 ml of total coliforms, fecal coliforms, E. coli, fecal streptococci, phage C and phage H and the #/100 ml Salmonella. Similar significant correlations were found for human enteric viruses, except that the phage H correlation was not significant. Only the correlations between the #/100 ml of total coliforms, fecal streptococci and phage C and the #/100 ml of total parasites were significant. The correlation coefficients in each case were very low and were between 0.17 and 0.34 for Salmonella, 0.14 and 0.46 for human enteric virus and 0.18 and 0.19 for total parasites. No significant correlations were found between the #/100 ml of enterococci, Clostridium and Mycobacteria. It should be noted that the sample number for these calculations was between 230 and 427. Calculating the correlation coefficient with the microbial densities per gram of various solids categories yielded poorer correlations. Significant correlations at the 99% level were observed only for the levels of total coliforms, fecal coliforms, E. coli, fecal streptococci, phage C and phage H and the #/100 ml Salmonella. Few significant correlations were found for virus and parasites. Fewer significant correlations were found for correlations calculated with the microbial levels per various solids criteria.

LOG REDUCTION OF MICROORGANISMS THROUGH ANAEROBIC DIGESTION

The mean log reductions for the #/100 ml of the selected microorganisms are shown in Table 4. The mean log reduction was calculated from the log reductions observed for each set of samples collected at each of the three wastewater treatment plants as shown below. The relative log digester input was calculated as the log of the sum of the product of the #\100 ml x the flow (Q in MGD) for each input to the digester.

$$\text{MEAN LOG REDUCTION} = \frac{\text{SUM (LOG DIGESTER INPUT-LOG DIGESTER OUTPUT)}}{\text{NUMBER OF SAMPLES}}$$

WHERE:

LOG DIGESTER INPUT-LOG DIGESTER OUTPUT

FOR PLANT A=

$$\text{LOG } (\#/100\text{ ml}_{A2} \times Q_{A2}) + (\#/100\text{ ml}_{A3} \times Q_{A3}) + (\#/100\text{ml}_{A4} \times Q_{A4}) - \text{LOG } (\#/100\text{ ml}_{A5} \times Q_{A5})$$

FOR PLANT B=

$$\text{LOG } (\#/100\text{ ml}_{B2} \times Q_{B2}) + (\#/100\text{ml}_{B4} \times Q_{B4}) - \text{LOG } (\#/100\text{ ml}_{B5} \times Q_{B5})$$

FOR PLANT C=

$$\text{LOG } (\#/100\text{ ml}_{C2} \times Q_{C2}) + (\#/100\text{ml}_{C4} \times Q_{C4}) - \text{LOG } (\#/100\text{ ml}_{C5} \times Q_{C5})$$

AND:

Q=FLOW IN MGD

A2,A3,A4,A5,B2,B3,B4,B5,C2,C3, AND C5=SLUDGE SAMPLES AT PLANT A, B. AND C.

The values listed below the mean log reduction are the standard deviation for this value and the maximum and minimum reductions observed during the course of

the sampling period. It should be noted that considerable variability was observed for the mean log reduction values with almost 2 orders of magnitude range for the mean log reduction of the indicator bacteria. The ranges for the log reduction values for the coliform indicators were 1.27 to 1.45, 2.82 to 2.97 and 1.06 to 1.25 for plants A, B and C, respectively. Lower log reduction values were found for the streptococci and phage indicators. Relatively little reduction of Clostridium and Mycobacterium were found with no reductions observed at plant C.

Human pathogens were reduced by anaerobic digestion. The mean log reduction values for Salmonella were 1.92, 3.33 and 1.38 for plants A, B and C. The mean log reduction for human enteric virus was lower with 1.12, 1.37 and 0.45 for plants A, B and C. Total parasites mean log reduction was similar: 0.90, 1.27 and 0.55 for plants A, B and C. In each case plant B had the highest reduction in the level of indicator and pathogenic microorganisms.

CORRELATION OF LOG REDUCTION

The correlation coefficients for the linear regression for the log reductions of selected indicator and pathogenic microorganisms are shown in Table 5 for all samples. Significant correlations at the 99% level were found between the log reduction of the indicator microorganisms (except for the slow growing Mycobacteria which was significant at the 95% level) and Salmonella. The correlation coefficients ranged from 0.27 for the slow growing Mycobacteria which was significant at the 95% level to 0.65 for E.coli which was significant at the 99% level. The correlation coefficients for log reduction of the coliforms and streptococci indicators with log reduction of Salmonella were 0.58 to 0.65 and 0.47 to 0.62, respectively. Similar but lower correlations were observed for all samples with log reductions of human enteric virus. Lesser correlations were found for total parasites. However, when the log reductions for each treatment plant were considered separately, only the log reduction for E.coli at plant B was significant at the 95% level for the log reduction of Salmonella.

RELATIVE MICROBIAL BALANCE

The relative microbial mass balance through the three treatment plants is shown in table 6. The relative mean microbial mass for each selected indicator and pathogen was calculated from the product of the flow in MGD for the waste stream and the level of microorganism, #/100 ml. The data was reported as the relative microbial mass since no correction was made for the volume term. The units are # million gallons/100 ml day. The overwhelming majority of the microbial mass leaves the wastewater treatment plant after conventional secondary treatment in the liquid effluent. Without terminal disinfection, only 1 to 2 orders of magnitude reduction are accomplished by secondary treatment. The log reduction in the microbial mass from the raw wastewater after anaerobic sludge digestion was 2 to 3 orders of magnitude. The microbial mass leaving the plant by the solids route after anaerobic sludge digestion was about 10 fold lower than by the liquid route without disinfection.

LOG PROBABILITY

An alternative presentation for the levels of microorganisms in the raw sewage (open points) and the digested sludge (filled points) for plant A (circles), B (triangles) and C (squares) is shown in figures 1, 2, 3 and 4 for fecal coliform, fecal streptococci, coliphage C and Salmonella, respectively . For each indicator and pathogenic microorganism assayed, the log probability

plot approximates a straight line and indicates that the distribution for the levels of microorganisms follows a log-normal distribution. As seen in figure 1, 2, 3 and 4, the levels of each of the microorganism in raw wastewater at plant A, B, and C was similar. The differences found in the raw sewage was small. However, the levels of microorganisms in the digested sludge for each treatment plant varied. A visual inspection of each figure shows the approximate log #/gTS reduction from raw sewage thru anaerobic sludge digestion (open vs closed points) for each of the indicator and pathogenic microorganisms assayed. This visual log reduction differs from the mean log reduction calculation shown in table 4 and 6. Nevertheless, a 2 to 3 log reduction from raw sewage through sludge digestion is shown for the coliform (figure 1) and a 1.5 to 2 log reduction for fecal streptococci (figure 2). The levels of coliphage (figure 3) also yielded a straight line for the log probability plot and follows a log-normal distribution in raw sewage and digested sludge. About 2 logs reduction were observed for phage C (figure 3). Similar log probability plots are shown for Salmonella in figure 4. Like the indicator bacteria, the log of Salmonella levels also approxomated a straight line on the log probability plot and appeared to follow a log normal distribution. The levels of Salmonella for the raw sewage at each plant were a little more variable than the indicators and reflected the poorer precision of the multiple tube culture method and the most probable number calculation used for enumeration. About 2 to 3 logs reduction were observed from raw sewage through anaerobic digestion.

CONCLUSIONS

Considerable variation was observed for samples collected at the same sample station at a given treatment plant. The levels of selected microorganisms and human pathogens in the raw wastewater entering the three treatment plants were remarkably similar. The variability between treatment plants appeared to be no more than the variability between type of samples. The levels of the selected indicators and pathogens appeared to follow the order of densities for the samples of wastewater and wastewater sludge sampled in this study: total coliform > fecal coliform > E.coli > fecal streptococci > enterococci > phage C ~ phage H > Clostridium > Mycobacterium > Salmonella > enteric viruses > parasites.

The levels of human pathogenic bacteria, viruses and parasites were low in all the samples compared to the levels of indicators. Salmonella spp. were consistently found in raw wastewater, raw sludges, treated sludges and wastewater effluents. Parasites were found in most of the samples of raw wastewater, raw sludges, and digested sludges but rarely observed in the wastewater effluents. Enteric viruses were infrequently recovered from raw wastewater but found in many of the raw sludge samples and few of the digested sludge samples. Viruses were not recovered from the wastewater effluent. The low frequency of recovery in the liquid flows at the treatment plants was probably due to a combination of low virus densities and low volume (100 ml) assayed.

The anaerobic digestion process yielded reductions in the densities of selected indicator and pathogenic microorganisms at each of the three treatment plants. The mean log reduction varied for the three treatment plants for the same microorganism, but followed the same relative order. Plant B, Sod Run, had the highest mean log reductions, followed by plant A, Back River, and plant C, Blue Plains. One to three log reductions were observed for the coliform indicators thru the anaerobic digestion process, while only 0.2 to 2 logs

reduction for the streptococci were found at the three plants. Salmonella were reduced 1.3 to 3.3 orders of magnitude. While the levels of human virus and parasites were low and the log reductions difficult to determine, the log reduction observed were less than those found for the bacteria. Only 0.4 to 1.4 and 0.5 to 1.2 orders of magnitude reduction were observed for virus and parasites, respectively. Correlations significant at the 95% and 99% level were found between the log reduction of indicator and pathogenic microorganisms. Correlations were generally higher (as high as 0.65) than that observed for the levels of microorganisms. Again no particular indicator consistently yielded higher correlations.

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TABLE 1. METHODS EMPLOYED FOR THE RECOVERY OF MICROORGANISMS.

Microorganism	Methods	References
Total coliform	m-Endo Plate Count	Standard Methods (APHA 1985) Geldreich (1981)
Fecal Coliform	m-FC Plate Count	Standard Methods (APHA 1985) Geldreich (1981)
E. coli	m-Tech Plate Count	Dufour et al.(1984)
Fecal Streptococci	Plate Count	Standard Methods (APHA, 1985)
Enterococci	Membrane Filter	Dufour et al (1984)
Clostridium	Membrane Filter	Bisson et al (1979)
Mycobacteria	Membrane Filter	Engelbrecht & Haas (1977)
Salmonella	Multiple Tube Dilution	Standard Methods (APHA, 1985)
Phage	Agar Overlay, E. coli C S.typhimurium WG49	Adams (1959) Havelaar (1984)
Enteric Virus	Elution and organic flocculation	USEPA Manual of Methods for Virology (1984)
Protozoan cysts	Sedimentation with sucrose floatation*	Fox et al. (1981)
Helminth Ova	Sedimentation with zinc sulfate floatation	Fox et al. (1981)

TABLE 2a. LOGARITHMIC MEAN LEVELS PER 100 ML OF BACTERIAL INDICATORS AND PATHOGENS, HUMAN ENTERIC VIRUS AND HUMAN PARASITES FOR SAMPLES COLLECTED AT PLANT A

		TC	FC	EC	FS	ENT	CP	MY	LTMY	SAL	PHAGE C	HEV	PHAGE H	TP
RAW SEWAGE	LM	7.99	7.34	7.08	5.74	5.58	3.79	1.86	2.20	2.11	5.43	0.18	5.78	0.48
	STD	0.04	0.08	0.15	0.08	0.08	0.15	0.11	0.18	0.11	0.08	0.08	0.23	0.08
	N	24	24	24	24	24	23	23	18	23	23	23	13	23
PRIMARY SLUDGE	LM	8.20	7.34	7.00	4.92	4.52	4.45	3.36	3.58	1.99	4.62	0.79	4.54	1.32
	STD	0.15	0.15	0.18	0.15	0.20	0.20	0.11	0.20	0.18	0.15	0.15	0.23	0.11
	N	24	24	24	24	24	23	23	18	23	23	23	13	22
TFS	LM	8.62	7.61	7.34	6.18	5.97	4.79	3.18	3.75	2.66	5.11	0.34	5.36	0.82
	STD	0.15	0.18	0.18	0.20	0.15	0.28	0.18	0.23	0.08	0.08	0.11	0.26	0.18
	N	21	21	21	21	21	20	20	16	21	20	20	10	21
AS	LM	8.59	7.73	7.62	6.40	6.32	5.04	3.61	4.08	2.64	5.57	1.04	5.62	0.71
	STD	0.11	0.08	0.11	0.15	0.08	0.23	0.18	0.23	0.20	0.08	0.15	0.23	0.18
	N	23	23	23	23	23	22	22	17	24	22	22	12	23
DS	LM	8.15	7.20	6.99	5.72	5.64	5.32	4.60	4.95	1.68	5.11	0.40	5.26	1.34
	STD	0.08	0.11	0.11	0.11	0.08	0.23	0.15	0.20	0.15	0.15	0.11	0.18	0.18
	N	24	24	24	24	24	23	23	18	24	23	23	13	24
ME	LM	7.00	6.38	6.15	4.82	4.52	3.72	1.57	1.96	1.26	4.64	0.18	4.63	-0.03
	STD	0.15	0.18	0.20	0.15	0.15	0.11	0.18	0.18	0.18	0.11	0.08	0.32	0.04
	N	24	24	24	24	24	23	23	17	24	23	23	13	24

NOTE: LM = LOGARITHMIC MEAN. STD = LOGARITHMIC STANDARD DEVIATION, N = NUMBER OF OBSERVATIONS,
TFS = TRICKLING FILTER SLUDGE, AS = ACTIVATED SLUDGE,
DS = DIGESTED SLUDGE, ME = MIXED EFFLUENT

TABLE 2b. LOGARITHMIC MEAN LEVELS PER 100 ML OF BACTERIAL INDICATORS AND PATHOGENS, HUMAN ENTERIC VIRUS AND HUMAN PARASITES FOR SAMPLES COLLECTED AT PLANT B

		TC	FC	EC	FS	ENT	CP	MY	LTMY	SAL	PHAGE C	HEV	PHAGE H	TP
RAW SEWAGE	LM	8.04	7.00	6.93	5.90	5.53	3.83	2.08	2.40	1.60	4.93	0.40	5.30	0.57
	STD	0.08	0.11	0.11	0.08	0.04	0.15	0.15	0.20	0.15	0.04	0.11	0.15	0.08
	N	24	24	24	24	24	23	23	17	24	23	23	13	23
PRIMARY SLUDGE	LM	8.87	7.92	7.87	6.54	6.08	4.83	3.75	3.75	2.46	5.41	1.36	5.04	1.85
	STD	0.20	0.20	0.23	0.15	0.18	0.23	0.18	0.23	0.15	0.15	0.18	0.43	0.15
	N	22	22	22	22	22	21	21	16	21	21	21	12	22
TFS	LM	8.11	7.08	6.95	5.86	5.64	4.34	3.43	3.63	2.18	5.08	0.71	5.58	1.45
	STD	0.30	0.28	0.28	0.28	0.20	0.23	0.15	0.18	0.26	0.20	0.20	0.11	0.18
	N	22	22	22	22	22	21	21	17	21	22	21	13	20
AS	LM	8.58	7.40	7.20	6.30	6.11	4.73	3.48	3.79	2.53	5.59	1.15	5.78	0.96
	STD	0.08	0.15	0.15	0.08	0.08	0.23	0.18	0.18	0.15	0.08	0.15	0.11	0.15
	N	24	24	24	24	24	23	23	18	24	23	23	13	24
DS	LM	6.91	5.93	5.67	5.43	5.04	5.04	4.43	4.23	0.20	4.71	0.34	3.71	1.30
	STD	0.20	0.18	0.18	0.18	0.18	0.26	0.18	0.18	0.20	0.11	0.11	0.20	0.18
	N	24	24	24	24	24	23	23	18	24	23	23	13	24
DS (STORED)	LM	5.71	4.85	4.81	4.60	4.28	5.36	4.40	4.52	-0.36	4.75	0.26	3.23	0.51
	STD	0.18	0.18	0.18	0.18	0.15	0.28	0.15	0.20	0.20	0.08	0.11	0.20	0.18
	N	24	24	24	24	24	23	23	18	24	23	23	12	24
ME	LM	6.15	5.00	5.11	4.11	3.78	3.26	1.77	1.76	0.51	4.72	0.08	4.41	-0.02
	STD	0.08	0.15	0.15	0.11	0.11	0.08	0.23	0.15	0.20	0.08	0.04	0.18	0.00
	N	24	24	24	24	24	23	23	18	24	23	23	13	23
SEPTAGE	LM	7.15	6.34	6.00	5.28	5.20	4.66	4.04	3.64	1.00	4.15	0.74	3.45	0.00
	STD	0.34	0.28	0.38	0.28	0.26	0.28	0.20	0.38	0.26	0.34	0.32	0.34	0.00
	N	9	9	9	9	9	9	9	8	9	9	8	9	9

TABLE 2C. LOGARITHMIC MEAN LEVELS PER 100 ML OF BACTERIAL INDICATORS AND PATHOGENS, HUMAN ENTERIC VIRUS AND HUMAN PARASITES FOR SAMPLES COLLECTED AT PLANT C

		TC	FC	EC	FS	ENT	CP	MY	LTMY	SAL	PHAGE C	HEV	PHAGE H	TP
RAW SEWAGE	LM	7.89	6.90	7.00	5.88	5.59	3.78	1.83	2.60	1.94	5.08	0.34	5.56	0.52
	STD	0.04	0.08	0.11	0.04	0.04	0.18	0.15	0.15	0.15	0.11	0.11	0.15	0.08
	N	24	24	24	24	24	23	23	18	23	23	23	13	24
PRIMARY SLUDGE	LM	8.86	7.91	7.79	5.81	5.81	4.85	3.30	3.71	2.97	5.18	1.26	5.46	1.63
	STD	0.08	0.08	0.18	0.15	0.18	0.20	0.15	0.23	0.08	0.08	0.15	0.15	0.15
	N	24	24	24	24	24	23	23	18	24	23	23	13	23
AS	LM	7.36	6.53	6.54	5.59	5.28	4.64	3.34	3.97	2.08	4.67	0.15	4.43	0.48
	STD	0.11	0.11	0.15	0.11	0.11	0.23	0.20	0.18	0.11	0.08	0.08	0.23	0.11
	N	24	24	24	24	23	23	23	18	24	23	23	13	24
DS	LM	7.41	6.64	6.49	5.52	5.59	5.34	4.20	4.63	1.53	4.66	0.34	4.18	0.98
	STD	0.11	0.11	0.11	0.26	0.08	0.30	0.15	0.15	0.20	0.08	0.11	0.15	0.20
	N	23	23	23	23	23	22	22	17	23	22	22	13	22
EF	LM	5.04	4.30	4.04	3.20	3.30	3.30	1.49	1.75	0.32	2.84	0.00	2.70	0.00
	STD	0.18	0.11	0.11	0.08	0.23	0.11	0.26	0.20	0.18	0.08	0.00	0.00	0.00
	N	24	24	24	24	24	23	23	17	24	23	22	13	24

NOTE: LM = LOGARITHMIC MEAN, STD = LOGARITHMIC STANDARD DEVIATION, N = NUMBER OF OBSERVATIONS, EF = EFFLUENT, SEE TABLE 2a FOR OTHER ABBREVIATIONS

TABLE 3. CORRELATION COEFFICIENTS FOR THE LEVEL OF INDICATOR MICROORGANISMS AND THE LEVEL OF HUMAN PATHOGENS FOR ALL SAMPLES COLLECTED AT THREE WASTEWATER TREATMENT PLANTS

		correlation coefficient / number									
		TC	FC	EC	FS	ENT	CP	MY	LTMY	PHAGE C	PHAGE H
#/100 ML	SAL	0.32 **	0.28 **	0.17 **	0.34 **	0.02	-0.03	-0.06	0.06	0.33 **	0.26 **
		427	427	427	427	426	409	409	319	410	234
	HEV	0.28 **	0.15 **	0.18 **	0.14 **	0.00	-0.01	-0.03	-0.03	0.46 **	0.07
		412	412	412	412	411	412	412	320	395	235
	TP	0.20 **	0.09	0.05	0.19 **	0.00	0.08	-0.01	-0.01	0.18 **	-0.04
		423	423	423	423	422	405	405	313	406	230
#/gTS	SAL	0.10 *	0.19 **	0.16 **	0.24 **	-0.02	0.01	-0.05	-0.08	0.38 **	0.40 **
		427	427	427	427	426	409	409	319	410	234
	HEV	0.01	0.06	0.03	0.08	-0.01	0.00	-0.02	-0.03	0.15	0.02
		412	412	412	412	411	412	412	320	395	235
	TP	0.04	0.06	0.05	0.21 **	-0.02	0.04	-0.04	-0.05	0.04	0.08
		423	423	423	423	422	405	405	313	406	230
#/gVSS	SAL	0.09	0.27 **	0.06	0.27 **	0.00	0.07	-0.02	-0.04	0.36 **	0.29 **
		425	425	425	425	424	407	407	317	408	233
	HEV	0.03	0.07	0.53 **	0.20 **	-0.01	0.00	-0.01	-0.03	0.15 **	0.00
		410	410	410	410	409	410	410	318	393	234
	TP	0.04	0.05	0.01	0.24 **	-0.02	-0.03	-0.03	-0.07	0.03	0.12
		422	422	422	422	421	404	404	312	405	230

TABLE 4. MEAN LOG REDUCTION OF THE #/100 ML OF SELECTED MICROORGANISMS FROM RAW SEWAGE THRU CONVENTIONAL ANAEROBIC DIGESTION AT THREE WASTEWATER TREATMENT PLANTS

	TC	FC	EC	FS	ENT	CP	MY	LMY	SAL	PHC	PHH	HEV	TP
PLANT A													
MEAN LOG REDUCTION	1.25	1.39	1.35	1.17	1.10	0.60	0.12	0.93	1.87	1.16	1.88	1.14	1.01
STD	0.41	0.64	0.57	0.60	0.40	1.30	1.01	1.94	0.87	1.14	2.08	0.32	0.85
MAX	2.06	3.08	2.48	3.02	1.88	6.48	3.94	5.70	4.04	6.13	6.67	1.90	2.85
MIN	-0.13	-0.01	0.33	0.04	0.43	-0.37	-1.46	-1.43	0.51	0.07	0.21	0.31	-0.10
N	24.00	24.00	24.00	24.00	24.00	24.00	24.00	23.00	24.00	24.00	16.00	24.00	24.00
PLANT B													
MEAN LOG REDUCTION	2.97	2.82	2.91	2.04	2.13	0.81	0.20	0.65	3.33	1.81	3.03	1.37	1.22
STD	1.12	1.15	1.38	0.88	0.88	0.65	0.46	0.57	0.85	0.79	0.70	0.40	1.07
MAX	5.69	5.34	7.19	4.60	4.49	1.95	1.26	1.34	5.25	4.42	4.18	2.19	3.43
MIN	1.15	1.22	0.44	1.01	0.67	-0.21	-0.48	-0.65	2.10	0.77	1.91	0.74	-1.31
N	24.00	24.00	24.00	24.00	24.00	23.00	23.00	18.00	24.00	23.00	13.00	23.00	24.00
PLANT C													
MEAN LOG REDUCTION	1.25	1.06	1.18	0.43	0.26	-0.34	-0.54	-0.30	1.38	0.40	1.34	0.45	0.55
STD	0.73	0.55	0.79	1.35	0.58	0.74	0.64	0.83	1.04	0.40	0.86	0.39	0.80
MAX	2.85	2.36	3.03	6.35	1.53	1.02	0.92	1.57	2.88	1.27	2.90	1.22	2.03
MIN	-0.07	0.04	-0.60	-0.94	-1.16	-1.31	-2.16	-1.85	-1.32	-0.70	0.42	-0.23	-0.51
N	23.00	23.00	23.00	23.00	23.00	22.00	22.00	17.00	23.00	22.00	13.00	22.00	22.00

TABLE 5. CORRELATION COEFFICIENTS FOR THE LOG REDUCTION OF INDICATOR MICROORGANISMS AND THE LOG REDUCTION OF HUMAN PATHOGENS FOR ALL SAMPLES COLLECTED AT THREE WASTEWATER TREATMENT PLANTS*

CORRELATION COEFFICIENT
PROBABILITY THAT H=0
NUMBER

	TCR	FCR	ECR	FSR	ENTR	CPR	MYR	LTMYR	PHAGE CR	PHAGE HR
SALR	0.62816	0.58489	0.64799	0.47003	0.6199	0.37327	0.32343	0.27615	0.55113	0.56131
	0.0001 **	0.0001 **	0.0001 **	0.0001 **	0.0001 **	0.002 **	0.0081 **	0.0498 *	0.0001 **	0.0002 **
	69	69	69	69	69	66	66	51	66	38
HEVR	0.46712	0.48219	0.48995	0.451	0.64232	0.49419	0.33542	0.33927	0.61254	0.48638
	0.0001 **	0.0001 **	0.0001 **	0.0001 **	0.0001 **	0.0001 **	0.0059 **	0.0149 *	0.0001 **	0.002 **
	66	66	66	66	66	66	66	51	63	38
TPR	0.29146	0.23363	0.33649	0.26595	0.30236	0.02307	0.1196	0.13059	-0.2197	0.10912
	0.0159 *	0.0552 *	0.005 **	0.0284 *	0.0122 *	0.8553	0.3426	0.366	0.0787	0.5143
	68	68	68	68	68	65	65	50	65	38

* CALCULATION BASED ON THE MICROBIAL LEVELS AS #/ 100 ML

TABLE 6. RELATIVE MICROBIAL BALANCE FROM RAW WASTEWATER FOR THE SELECTED MICROORGANISMS AT PLANTS A,B AND C.*

	TC	FC	EC	FS	ENT	CP	MY	LMY	SAL	PHC	PHH	HEV	TP
PLANT A													
					RELATIVE MEAN LOG MICROBIAL MASS								
RAW	10.29	9.62	9.38	8.05	7.87	6.15	4.22	4.43	4.37	7.69	8.02	3.30	2.75
DIGESTER IN	9.15	8.33	8.08	6.64	6.48	5.46	4.27	4.61	3.30	5.82	5.98	1.85	2.08
DIGESTER OUT	7.90	6.95	6.73	5.47	5.38	5.06	4.34	4.46	1.42	4.86	5.04	0.75	1.08
EFFLUENT	9.27	8.64	8.40	7.08	6.77	5.97	3.82	4.24	3.52	6.95	6.88	3.22	2.22
					MEAN LOG REDUCTION FROM RAW SEWAGE								
DIGESTED	2.40	2.68	2.65	2.58	2.49	1.08	-0.12	-0.02	2.95	2.83	2.98	2.55	1.68
EFFLUENT	1.02	0.98	0.98	0.98	1.10	0.18	0.39	0.20	0.85	0.75	1.13	0.07	0.53
PLANT B													
					RELATIVE MEAN LOG MICROBIAL MASS								
RAW SEWAGE	8.88	7.83	7.77	6.73	6.36	4.66	2.92	3.26	2.44	5.82	6.16	1.80	1.40
DIGESTER IN	8.46	7.33	7.16	6.05	5.74	4.43	3.21	3.46	2.10	5.09	5.31	0.95	1.11
DIGESTER OUT	5.49	4.51	4.25	4.01	3.61	3.61	3.01	2.81	-1.23	3.28	2.29	-0.42	-0.11
EFFLUENT	6.97	5.84	5.92	4.93	4.60	4.07	2.60	2.61	1.34	5.59	5.26	1.81	0.81
					MEAN LOG REDUCTION FROM RAW SEWAGE								
DIGESTED	3.40	3.32	3.52	2.72	2.76	1.05	-0.09	0.44	3.67	2.55	3.87	2.23	1.51
EFFLUENT	1.92	1.99	1.84	1.81	1.77	0.59	0.32	0.65	1.10	0.24	0.90	-0.00	0.59
PLANT C													
					RELATIVE MEAN LOG MICROBIAL MASS								
RAW SEWAGE	10.37	9.39	9.50	8.36	8.07	6.25	4.31	5.09	4.42	7.56	8.04	3.51	3.00
DIGESTER IN	9.29	8.33	8.26	6.59	6.48	5.63	4.29	4.90	3.52	5.69	6.03	2.04	2.10
DIGESTER OUT	7.75	6.98	6.83	5.89	5.96	5.94	4.80	5.20	2.10	5.07	4.69	1.59	1.51
EFFLUENT	7.51	6.78	6.54	5.68	5.78	5.77	3.97	4.23	2.80	5.35	5.18	3.48	2.49
					MEAN LOG REDUCTION FROM RAW SEWAGE								
DIGESTED	2.62	2.40	2.66	2.47	2.11	0.31	-0.49	-0.10	2.32	2.49	3.35	1.92	1.48
EFFLUENT	2.87	2.60	2.95	2.68	2.29	0.48	0.34	0.86	1.62	2.21	2.85	0.03	0.50

* CALCULATION BASED ON THE MICROBIAL LEVELS AS #/ 100 ML

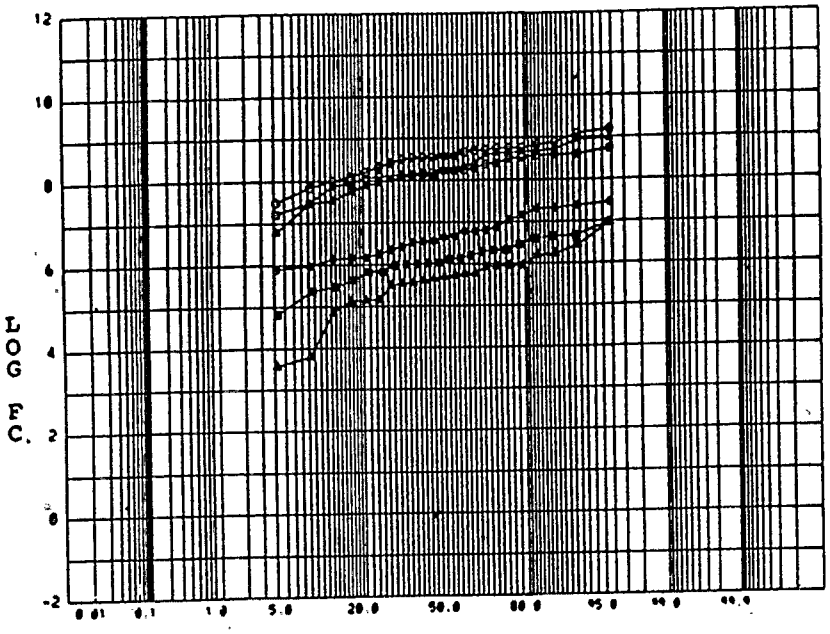

Figure 1. Log # fecal coliform/gts vs probability for samples raw sewage (open points) and digested sludge (closed points) at plant A (circles, plant B, (triangles) and C, (squares) from August 1987 thru September 1988.

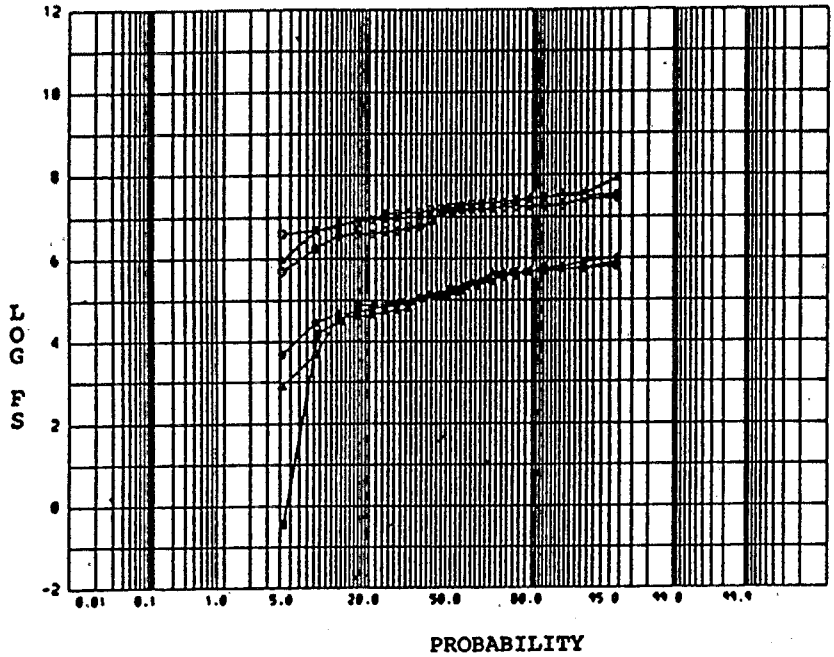

Figure 2. Log # fecal streptococci/gts vs probability for samples of raw sewage (open points) and digested sludge (close points) at plant A (circles), plant B, (triangles) and C, (squares) from August 1987 thru September 1988.

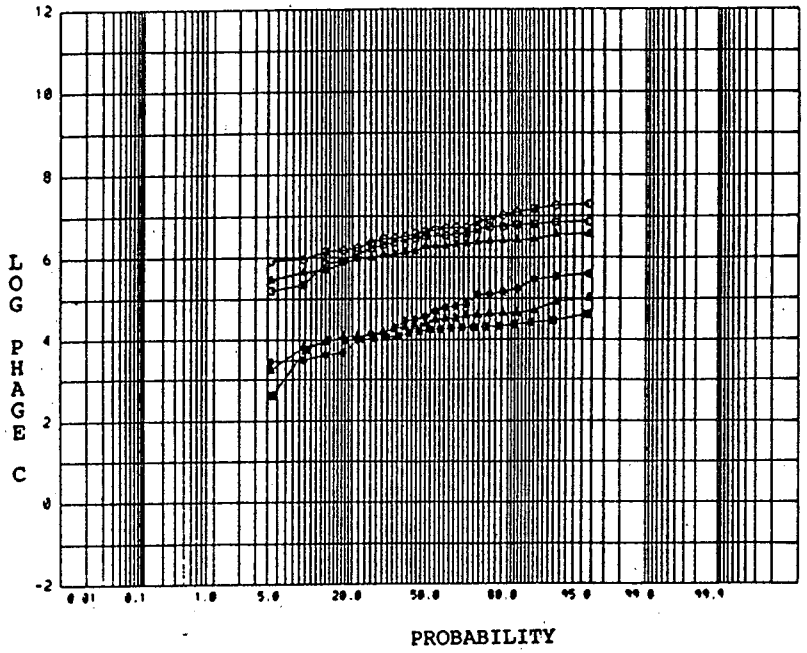

Figure 3. Log # coliphage C/gts vs probability for samples of raw sewage (open points) and digested sludge (closed points) at plant A (circles), plant B, (triangles) and C, (squares) from August 1987 thru September 1988.

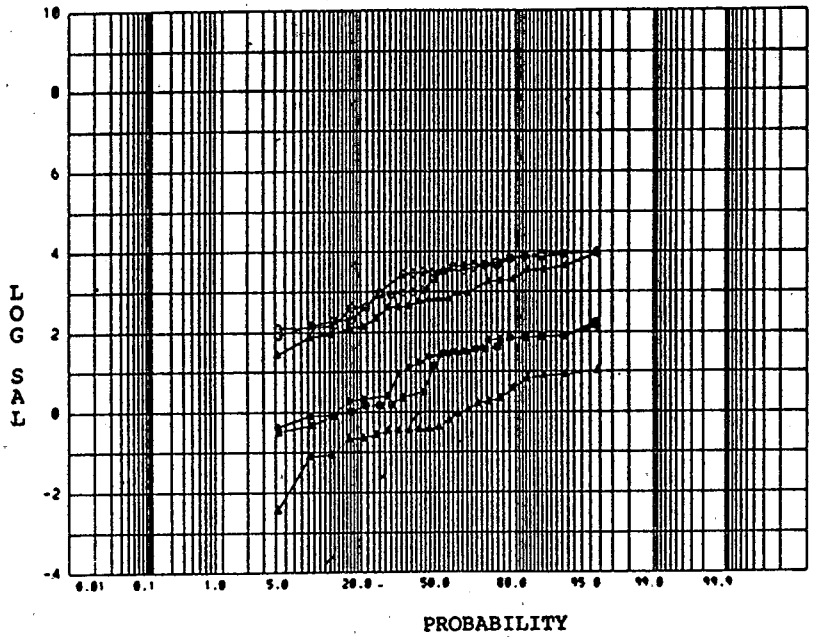

Figure 4. Log # Salmonella/gts vs probability for samples of raw sewage (open points) and digested sludge (closed points) at plant A (circles), plant B, (triangles) and C, (squares) from August 1987 thru September 1988.

IN-VESSEL COMPOSTING OF MUNICIPAL WASTEWATER SLUDGE

by

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INTRODUCTION

Since the late 1970s, composting has become a more frequently used method for stabilizing and processing municipal sewage sludge in the United States. The interest in this technology has been extremely rapid, from less than 10 facilities operating in 1975 to nearly 200 under design or in operation in 1989. Today much of the compost produced is being used as for soil improvement in accordance with the U.S. Environmental Protection Agency's (EPA's) policy that encourages the beneficial use of sewage sludge (49 FR 24358, June 12, 1984).

More recently, because of odor, labor, and materials-handling problems, a greater number of composting systems are being designed and built which contain the materials within a vessel. Other reasons for choosing an "in-vessel" system have been space limitations and adverse climates. In-vessel systems also differ from other composting systems because they use mechanical mixers to blend amendments with the sludge prior to composting, and they extensively use conveyors for materials handling. The evolution of these in-vessel systems has been very rapid, and municipalities continue to encounter problems in dealing with odors, removing moisture, and handling materials.

To learn from the experiences of these mostly first-generation in-vessel sludge composting systems and their operators, EPA decided in early 1988 to gather information from 8 of the 13 full-scale facilities operating in the United States.

This paper presents a brief discussion of the findings of this study. First, a description of the facilities surveyed and the reactor systems used will be presented. Then the major problem areas discovered by this survey -- materials handling, moisture removal, aeration of curing/storage piles, and odor control -- will be discussed. Finally, capital and operation and maintenance costs will be presented.

DESCRIPTION OF FACILITIES

The location of the facilities and the type of composting system used at each facility are shown in Figure 1. The eight facilities studied included all six types of in-vessel systems operating in early 1988. Akron, Cape May County, Newberg, Portland, Plattsburgh, and Schenectady were the first facilities of their kind to be built in this country. Only Cape May County and Portland were operating prior to 1986.

A survey team conducted a 1.5 to 2-day long site visit at each facility. The survey team interviewed plant supervisors, operators, and others familiar with each facility's procurement, construction, and operations. Contract documents and operating records were also examined when available. The information collected represents a "snapshot" of each facility taken at the

time of the site visit. The data presented are accurate only for the time of the site visit and may have subsequently changed.

Figure 2 gives a summary of the configuration of each of the facilities. Composting is often defined as a two stage process, a high-rate stage followed by a curing stage. The first stage is characterized by high oxygen uptake rates, high temperatures, rapid degradation of biodegradable volatile solids, and high potential for odor production. The second stage is characterized by lower temperatures, reduced oxygen uptake rates, and a lower, but significant, potential for odor production. No precise definition or distinction exists between these two stages, however. Moreover, although the first stage of composting is performed in reactors at all in-vessel facilities, the second stage can be performed in a reactor, an exterior pile, or both. Because of the lack of a clear definition, the term "first-step reactor" will be used to denote the first vessel that the compost enters. "Second-step reactor" will be used to denote the second vessel the compost enters in series. At several plants, there is no second reactor. The reactors will be discussed in more detail in a following section.

All of the facilites have exterior piles for curing or storage. Because no consistent measure of compost stability was used at the facilities, it was not possible to define whether material in outside piles was "curing" or just being "stored." Therefore, the term "exterior curing/storage piles" refers to any compost piles located outside the reactors, both aerated and unaerated.

AKRON, OHIO

Table 1 summarizes pertinent information for the Akron, Ohio facility, which uses four Paygro reactors. Each of the reactors is 6 meters wide, 3 meters deep, and 223 meters long. Akron, at 66 dry metric tons per day, had the largest design capacity of the facilities surveyed. Akron was operating at reduced capacity because of complaints about odors by neighbors.

Figure 3 is a schematic of a Paygro reactor. Compost material is placed in the reactors by a series of conveyors. The operation of the facility is similar to an extended aerated static pile facility in that sections of each reactor bin are treated as individual batches. The compost is then periodically mixed in place as desired by a machine called an "extractoveyor". The extractoveyor also removes the compost from the reactor. Air is supplied to the bottom of the reactor through a plenum covered by a steel grate. Air is exhausted from the top of the compost into the building which encloses the reactors, and the building air is exhausted by roof fans.

CAPE MAY COUNTY, NEW JERSEY

Table 2 summarizes pertinent information for the Cape May County, New Jersey facility, which uses two Purac reactors. This facility was originally designed with one first-step, and one second-step reactor. However, to increase the capacity of the facility, both reactors are now used as the first-step, and aerated piles located outside the building are used for the second-step. At the time of the survey, Cape May County was operating at reduced capacity because of complaints about odors by neighbors. Since the end of the summer of 1989, the facility has been operating at full capacity.

The reactors are rectangular in shape, 8 meters tall, and the walls slope slightly outward from top to bottom. One reactor has a volume of 850 cubic meters and the other reactor is 1050 cubic meters.

The reactors at Cape May County have had fires on several occasions, caused by both spontaneous combustion and welding accidents. Spontaneous combustion fires occurred when the composting materials were held in the reactor for too long because some other part of the system slowed or had stopped processing material. Fires can start at surprisingly low temperatures in dry compost. Cape May found that when the solids content of compost is greater than 65 percent and the temperature exceeds 80°C, it is possible for a series of exothermic chemical reactions to elevate the temperature to a point where true ignition and smoldering fires occur.

Figure 4 is a schematic of a Purac reactor. The compost mix is spread uniformly across the top of the reactor by a series of conveyors. The composting material then moves by gravity to the bottom of the reactor, where it is removed by a horizontal auger that spans the reactor width and travels back-and-forth along the length of the reactor. Air is supplied to the bottom of the reactor through a series of perforated pipes and is exhausted through a manifold located about 2 meters below the top of the compost mass. The exhaust airflow rate is higher than the supply airflow rate, so some air is drawn from outside the reactor through the headspace and through the top 2 meters of the compost mass.

CLAYTON COUNTY, GEORGIA

Table 3 summarizes pertinent information for the Clayton County, Georgia facility, which uses two Taulman reactors. The facility has one first-step reactor, and one second-step reactor. Clayton County had the smallest design capacity of the facilities surveyed. Clayton County was processing all of the sludge being produced by the wastewater treatment plant. Both reactors are 8.8 meters tall and 7.6 meters in diameter. Material discharged from the reactors is stored in unaerated piles located in a covered shed.

Figure 5 is a schematic view of a Taulman reactor. The compost mix is uniformly spread across the top of the reactor by a "slinger" which throws the material out from the center. The composting material then moves by gravity to the bottom of the reactor, where it is removed by a horizontal auger that rotates around a center pivot. Air is supplied to the bottom of the reactor through a series of perforated pipes and is exhausted out the top of the reactor.

NEWBERG, OREGON

Table 4 summarizes pertinent information for the Newberg, Oregon facility, which uses two Ashbrook-Simon-Hartley (Ashbrook) reactors. Each reactor is 5.5 meters wide, 2.7 meters high, and 19 meters long. This facility has had major structural failure of the concrete walls of one of the reactors; the walls buckled due to excessive pressure inside the reactor. Neither reactor at Newberg is being operating while the failed reactor is being rebuilt. Operation is scheduled to begin in October 1989.

Figure 6 is a schematic of an Ashbrook reactor. Compost material is loaded into the reactor by a large hydraulically operated door or "ram". The ram pushes the material through length of the reactor, and a horizontal auger at the far end of the reactor removes the compost. The compost mass is moved forward about 0.25 meters with each push by the ram. Air is supplied to and exhausted from the reactor through a series of perforated pipes embedded in the floor of the reactor. Each set of three adjacent pipes can be either air supply or air exhaust as needed.

PLATTSBURGH, NEW YORK

Table 5 summarizes pertinent information for the Plattsburgh, New York facility, which uses two Fairfield reactors. The reactors are 35.4 meters in diameter and 3 meters deep. Plattsburgh was processing all of the sludge being produced by the wastewater treatment plant. Plattsburgh has had numerous odor complaints and is modifying the odor control systems.

Figure 7 is a schematic of a Fairfield reactor. The compost mix is placed around the perimeter of the reactor by a traveling bridge. The compost material is mixed and moved toward the center of the reactor by a series of 27 augers which hang from the traveling bridge. Material is removed from the center of the reactor. Air is supplied to the bottom of the reactor by a series of perforated pipes. Air is exhausted from the top of the compost into the building which encloses the reactors, and the building air is removed by exhaust blowers.

PORTLAND, OREGON

Table 6 summarizes pertinent information for the Portland, Oregon facility, which has two parallel process trains each using three Taulman reactors. Each process train has two first-step reactors, and one larger second-step reactor. The first-step reactors are 8.8 meters tall and 13 meters in diameter. The second-step reactors are 8.8 meters tall and 16 meters in diameter. Compost material is processed in either one of the first-step reactors, and then conveyed to the second-step reactor. Material discharged from the reactors is stored in unaerated piles located outside. Portland was the only facility composting digested sludge. It was also processing the largest amount of sludge. Portland is operating at reduced capacity due to a number of mechanical maintenance problems, mostly related to excessive wear. Portland has also had reactor fires caused by welding accidents.

SARASOTA, FLORIDA

Table 7 summarizes pertinent information for the Sarasota, Florida facility, which uses two Purac reactors. Sarasota was processing all of the sludge being produced by the wastewater treatment plant. Sarasota has also had odor complaints and has modified the odor control systems. This facility has one first-step and one second-step reactor. Both reactors are 8.2 meters tall, 7.6 meters wide and 15 meters long. Material discharged from the reactors is stored in unaerated piles located in a covered shed.

SCHENECTADY, NEW YORK

Table 8 summarizes pertinent information for the Schenectady, New York facility, which uses four American Bio Tech (ABT) reactors. There are two first-step reactors and two second-step reactors contained inside a building. All of the reactors are 7.5 meters tall, and have a volume of about 500 cubic meters. This facility has had several problems including compost fires caused by spontaneous combustion and minor structural failures of the reactor's exterior steel frame. However, odor complaints are the reason that the facility is not operating now. It will not be operated until a new odor treatment system, which is currently being constructed, is operational. Operation is scheduled to begin in October 1989.

Figure 8 is a schematic of an ABT reactor. The compost mix is spread uniformly across the top of the reactor by a series of conveyors. The composting material then moves by gravity to the bottom of the reactor, where it is removed by a horizontal auger that spans the reactor width and travels back-and-forth along the length of the reactor. The reactors are constructed of corrugated fiberglass walls that hang from an exterior steel frame. Air is supplied to and exhausted from the reactor through a series of "air lances" that hang from headers located at the top of the reactor. The operating mode of the headers alternates between air supply and exhaust, and the operating mode is reversed every 20 minutes. That is, one set of air lances supplies air for 20 minutes, and then exhausts air for 20 minutes, while the next set of air lances exhausts air for 20 minutes, and then supplies air for 20 minutes, and so on.

PROBLEM AREAS

MATERIALS HANDLING

Mixing appropriate proportions of sludge, amendment, and recycle is essential to create a desirable compost structure with respect to porosity, moisture content, and energy balance. Proper porosity promotes easier materials handling and thorough aeration. Proper moisture content and energy balance support the biological decomposition of the organic material. Table 9 shows the design and actual percent solids in the mix, and the design and actual volumetric mix ratios. All of the facilities have had to use more amendment or recycle than expected, and most have had to use a drier mix than expected, for several reasons. The reasons include: sludge with a lower percent solids than expected, difficulty conveying the mix at the design mix ratio, and difficulty with aeration of the mix in the reactor. These increases resulted in a greater volume of material to be handled than expected.

Mechanically and operationally, conveyance systems are the dominant physical features of in-vessel facilities. Conveyors were the second biggest cause of system problems at the sites (after odor production). Although a high degree of automation for conveyor systems is not required, most in-vessel composting facilities chose automation to minimize labor costs.

The type of conveyors used at these facilities have been used reliably in industry for years; none are experimental. Yet due to differences in

materials characteristics, conveyors at in-vessel plants have experienced problems with jamming, breakage, excessive wear, and spillage.

Screw conveyors and belt conveyors have been used with the most success. The few reported problems with screw conveyors were related to excessive wear and the accumulation of material on flights and housings. Regular cleaning and lining the housings with ultra-high molecular weight plastic were effective for preventing problems. Spillage has been the major problem with belt conveyors. Wet materials adhere to the belts and do not transfer completely, resulting in spills and labor intensive cleanup problems. These problems were minimized by using belt scrapers or by placing a layer of dry material, such as sawdust, onto the belt before adding a wet material.

Drag chain conveyors have not performed well, suffering from plugging and excessive wear. Modifications to these conveyors have included increasing the distance between flights, increasing the clearance between the flights and the conveyor housing, and decreasing the area of the flights themselves. These modifications allow the composting material to push backwards over or around the flight, so it won't be compacted. These modifications, however, also reduce the capacity of the conveyor.

Cleated belts have had a number of problems including excessive spillage of both wet and dry materials, and do not appear to be suitable for composting facilities. These belts, because of their cleated design, cannot be effectively scraped.

Reactor discharge devices are key conveyors in the materials-handling system; if they fail, the reactors cannot be used. If compost becomes too dry, it can form into hard chunks; if it becomes too wet, it can form a glue-like mass. Both conditions make it difficult to move the compost with a discharge device.

A few structural failures with discharge devices have been noted, but excessive wear of auger flights has been the major mechanical problem observed. Compost is a surprisingly abrasive material, especially when it is compressed at the bottoms of vertical reactors. Decreases of as great as 15 percent of flight thickness in 6 months because of wear on the flight face were reported. Discharge screws originally hard-surfaced with Triten steel on both the flight edge and face have not suffered excessive wear.

MOISTURE REMOVAL

There are three basic demands on the aeration system: (1) heat removal, (2) moisture removal, and (3) oxygen for biological metabolism. The aeration systems at the sites visited appeared to be adequately sized to meet the oxygen supply and heat removal requirements; only Schenectady had problems with elevated temperatures. Table 10 shows the total aeration capacity of the facilities, the unit capacity per cubic meter of empty reactor volume, and the length of the path that air must take to pass through the reactor.

The major problems have been related to moisture removal: condensation and fog formation above the reactors and formation of a "hardpan" layer. Inadequate aeration for curing/storage piles was a third major problem.

Condensation and Fog Formation

In-vessel composting systems must provide for adequate collection of air after it has passed through the compost. Air can hold four times as much water at 60°C than at 40°C. Therefore, if the temperature drops before the air leaves the reactor, it does not remove as much water, and condensation occurs within the reactor and composting mass. If the air collection piping extends into the composting mass, exhaust holes in the piping can become exposed due to natural settling and uneven subsidence. Such exposure can cause odors to escape, saturated air to cool, and condensate to form within the reactor.

Systems with open reactors discharge warm, moist air from the compost bed into a building or work space above the reactor. In these cases, the introduction of too much cold outside air into these same areas results in the condensation of water vapor and the formation of fog. The worst case was found at Plattsburgh, where the fog hampers the stabilization of the product, promotes corrosion of equipment, and is so thick that operators cannot see to work.

Hardpan Formation

At least three of the facilities have noted the formation of a "hardpan" layer at the bottom of the reactor that is nearly impermeable to air. Formation of a hardpan layer occurs when leachate forms because of inadequate moisture removal from the reactor. The leachate extracts solubles as it moves down through the compost and drains to the low points in the reactor. If these low areas are low in humidity, the leachate dries, cementing material in the bottom of the reactor into an impermeable layer. At Clayton County, leachate seeped into the one-centimeter unwashed river stone layer on the bottom of the reactor, cemented the particles together and plugged the holes in the aeration piping. The stone was replaced with two-centimeter washed round river rock, and the aeration piping was remounted in the middle rather than at the bottom of the gravel layer, separating it from the pooling leachate. At Cape May County a hardpan layer formed which was attributed to excessive leachate generation. The mix ratio and amendment specifications were changed to obtain a more porous initial compost mix, which significantly reduced leachate generation. Also, the aeration system was modified to give the operators better control of the airflow. The hardpan problem has not recurred since these changes were made.

CURING/STORAGE AERATION

The curing/storage process is an important step in the overall sludge composting process. The composting process does not stop when material is removed from the reactor(s); the compost continues to have an oxygen demand. Even where compost is processed in two reactors (at Clayton County, Portland, Sarasota, and Schenectady), the unaerated curing/storage piles have reheated, indicating the continuation of biological activity. Depending on the stability of the compost, oxygen demand may be elevated to the point where the available oxygen is consumed and anaerobic conditions develop. Moving the piles then causes odors to be emitted. At Akron and Cape May County the exterior curing/storage piles are aerated or will be aerated in the future because of odor problems. At Plattsburgh, the curing/storage piles are

odorous, and an evaluation is underway to determine whether they should be aerated.

ODOR CONTROL

In the U.S. effective odor control is a critical requirement for the success of a composting facility. Fear of potential odors can cause public opposition to proposed plant sites. Odor complaints can cause a regulatory agency to shut down or curtail the operations of an existing facility. Plant odors can also hurt product sales.

Odor control was the greatest problem found at the sites surveyed. Of the eight composting facilities surveyed in this project, six had received odor complaints from the local community. At five of these sites, the complaints were serious enough to cause the plant owners to retrofit the facilities with additional odor treatment equipment. Two of the five plants ceased operations while the new equipment was being installed. Two plants were operated at reduced capacity to limit odor production while the new treatment equipment was piloted, designed, and installed. The last plant continued to operate but at the same time was proceeding with the odor treatment system retrofits as quickly as possible.

Odors found at compost facilities are caused by a large number of compounds that originate in both the sludge and the amendment. In-vessel systems may produce more amendment-related odors than static pile or windrow systems because the sawdust used at most in-vessel facilities has a greater surface area than the wood chips usually used in the other systems.

Odor-causing compounds fall into a number of classes that differ in their physical and chemical properties. Some are acidic; some are basic. Some can be destroyed by oxidation; others cannot. These compounds are not equally soluble in water, nor equally amenable to adsorption. Moreover, the mix of compounds in the odorous gas stream can change over time, depending on the composting materials and the reactor operating conditions. For this reason, treatment systems must employ a broad spectrum of removal mechanisms to be effective. Treatment systems that rely on only one or two removal mechanisms such as single-stage wet scrubbers and ozone oxidation chambers have not been successful.

Table 11 summarizes the odor treatment methods that were found at the eight plants surveyed. Of all the methods shown in Table 11, bubbling air through wastewater appeared to be the most successful. Dilution alone was completely ineffective as a treatment method. Biofilters appeared to remove a large fraction of odors based on measurements taken at the site. However, the facility using the biofilters has still had numerous odor complaints and is planning to install wet scrubbers to replace the biofilters. Ozonation was ineffective and replaced by wet scrubbers. Activated carbon was tested at pilot scale at one location but also appeared to be ineffective.

At Portland, Newberg, and Clayton County, odorous air is bubbled through wastewater for treatment. Operators at Clayton County have received some odor complaints, but the odors are generally not a problem, partly because of the rural setting of the plant. The Portland and Newberg facilities have not

received any odor complaints, but both plants are located in outlying industrial areas and both compost relatively stable sludges. These mitigating factors make it difficult to determine the effectiveness of this technique. The design parameters associated with this technique are not well established, and a ground-level discharge is created.

At the time of the site visits, packed tower wet scrubbers were in use at three of the eight facilities visited. Table 12 lists the odor sources treated and the chemicals used in these scrubbers. Sarasota installed the system shown in Table 12 after the site visit. "DeAmine" is a proprietary chemical.

Single-stage wet scrubbers were ineffective for several reasons. In addition to not being able to treat a broad spectrum of compounds, reasons include overloading the scrubbers and using inappropriate scrubbing chemicals.

Two stages appear to be the minimum number needed for effective odor treatment. Table 13 lists the planned odor treatment facilities at several of the plants. Typically, one stage uses an acid, and the other stage uses an oxidizing agent. Cape May County is planning to try corn syrup as a surfactant for aldehyde compounds. Packed tower wet scrubbers were the only scrubbers observed at the eight plants studied. Mist-type scrubbers have been used successfully at the Montgomery County Composting Facility, a large enclosed static pile operation in suburban Washington, D.C.

Although this discussion of odors has concentrated on treatment systems, an effective odor control system must include more than just a treatment system. A complete odor control plan, including quantification and qualification of all odor sources, and proper control of the composting process, must be implemented. In addition, no treatment system is 100 percent effective; some concentration of odorous compounds will be discharged after treatment. To avoid undesirable effects on local residents, the odorous emissions must be diluted and/or dispersed so that ambient air concentrations of the odorous gases fall below detection thresholds.

COSTS

Capital costs for the eight facilities are shown in Table 14. The facilities are listed by design capacity, from smallest to largest. Costs shown are the original construction, including change orders, in millions of dollars; additional expenditures made since start-up, in thousands of dollars; and expeditures that the facilities anticipate in the near future, in thousands of dollars. Note that the majority of additional and future capital costs are for odor control equipment, and that these costs range up to 40% of the original cost.

Figure 9 shows annual operation and maintenance costs, expressed as dollars per dry metric ton, as a function of current operating capacity. This figure shows an economy of scale for these facilities. Clayton County, the smallest facility, appears to have low costs, but this facility obtains its amendment at no cost. Akron, the largest facility, appears to have high

costs, but the cost shown includes sludge dewatering costs which are not included for the other facilities.

CONCLUSION

In-vessel composting is a total system that comprises a number of integrally related components, including materials handling, reactors, an aeration system, exterior curing/storage facilities, an odor control system, and marketing of the compost product. Inadequate or improper design, management, operation, or maintenance of any of these components will seriously impact the total system, and most likely adversely affect the compost product.

Most of these facilites failed to realize that a sludge in-vessel composting system includes much more than just the reactor(s). The major problems encountered at these facilities were not related to the reactor itself, but to the other components, including materials handling and odor control. Indeed, many of the difficulties encountered resulted from a failure to recognize the totality of the system that must be managed.

Finally, it should be noted that because this study focused on problem areas, in-vessel composting appears to be plagued with problems. However, all but two of the facilites discussed were producing and marketing a compost product. All of the facilities were meeting EPA requirements for pathogen reduction, and all were meeting State limits on heavy metals and organics.

ACKNOWLEDGEMENT

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1988 In-Vessel Study

Figure 1. In-Vessel Locations Studied

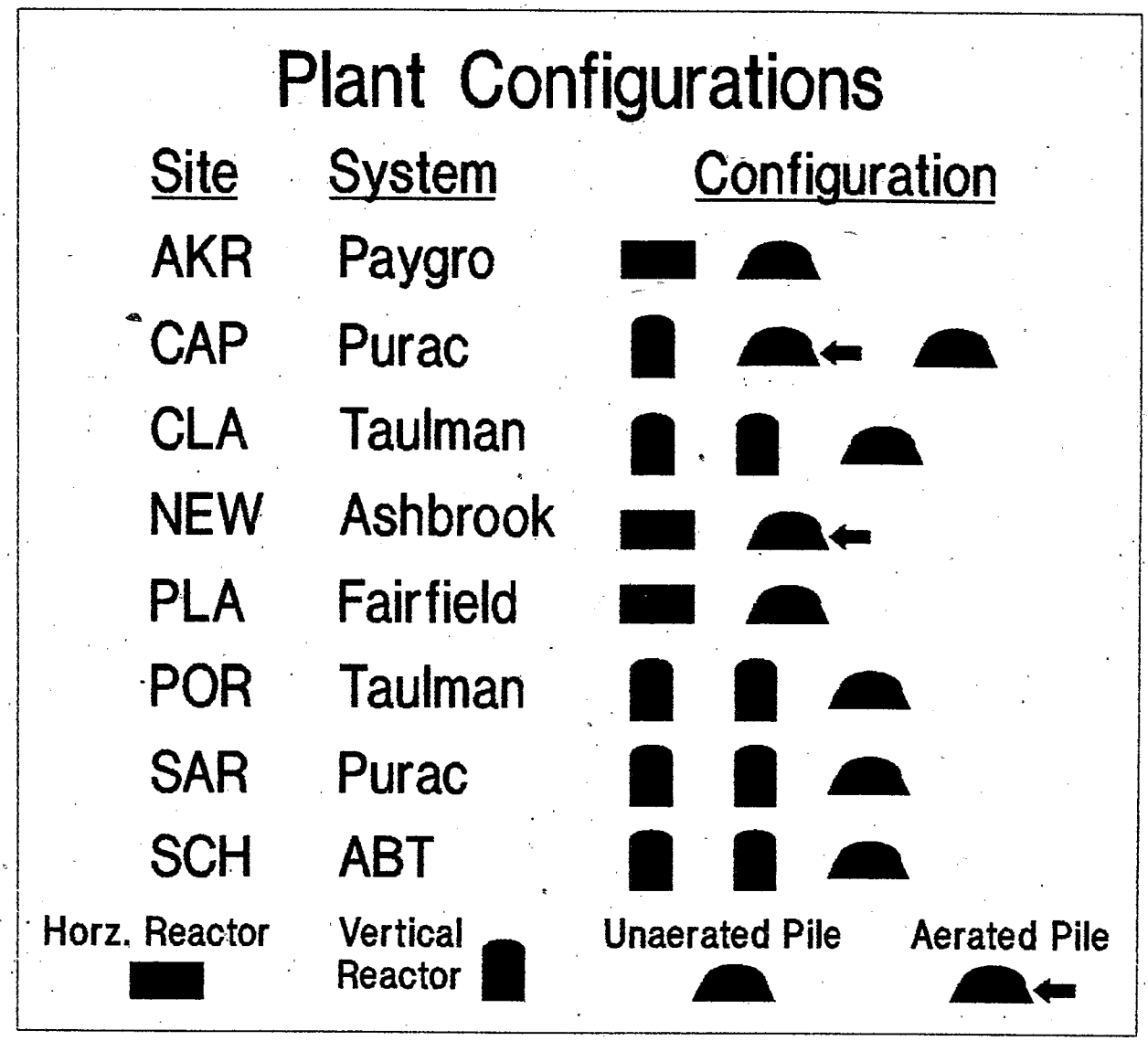
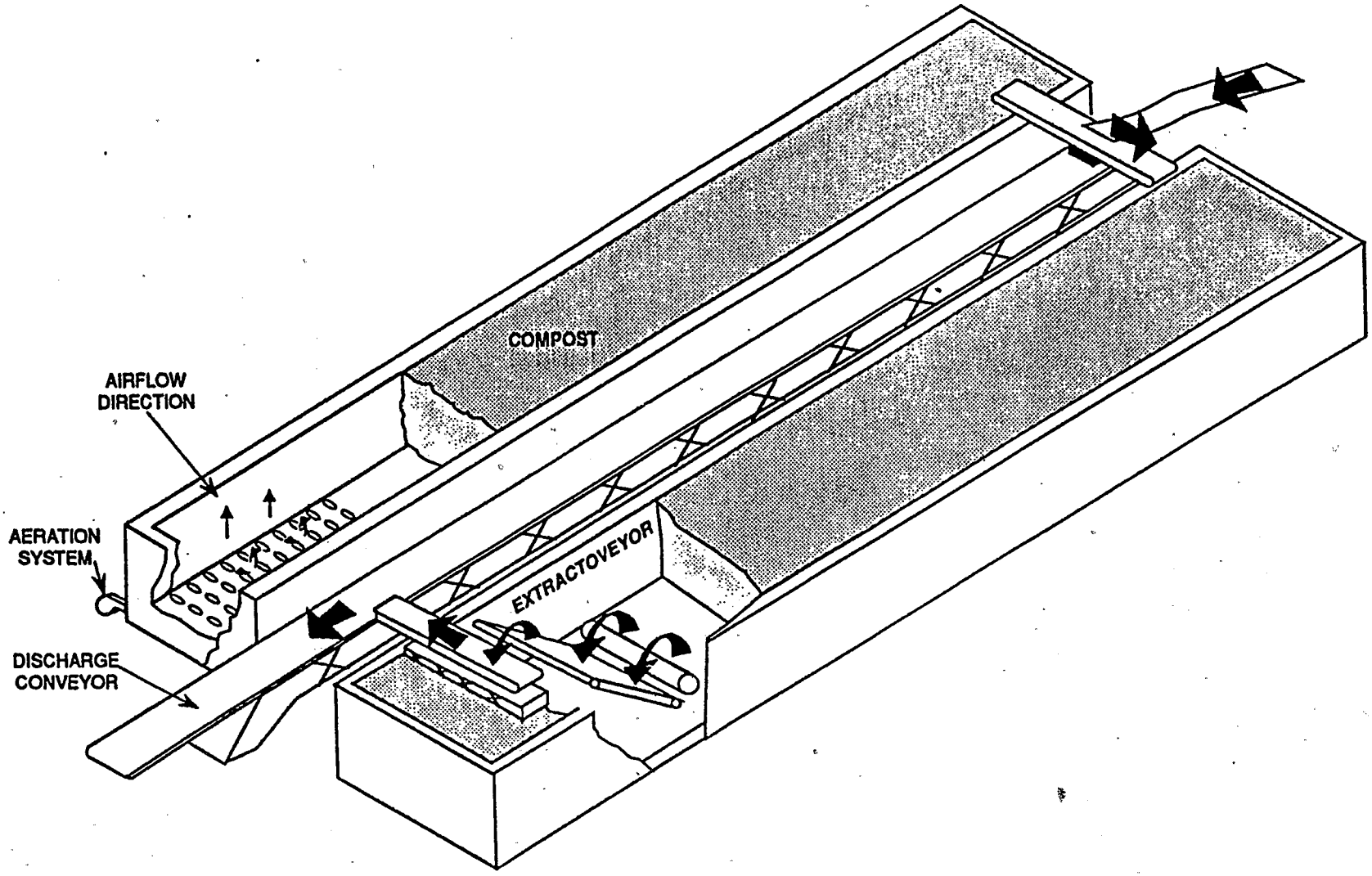


Figure 2. Plant Configurations

Figure 3. Paygro Reactor



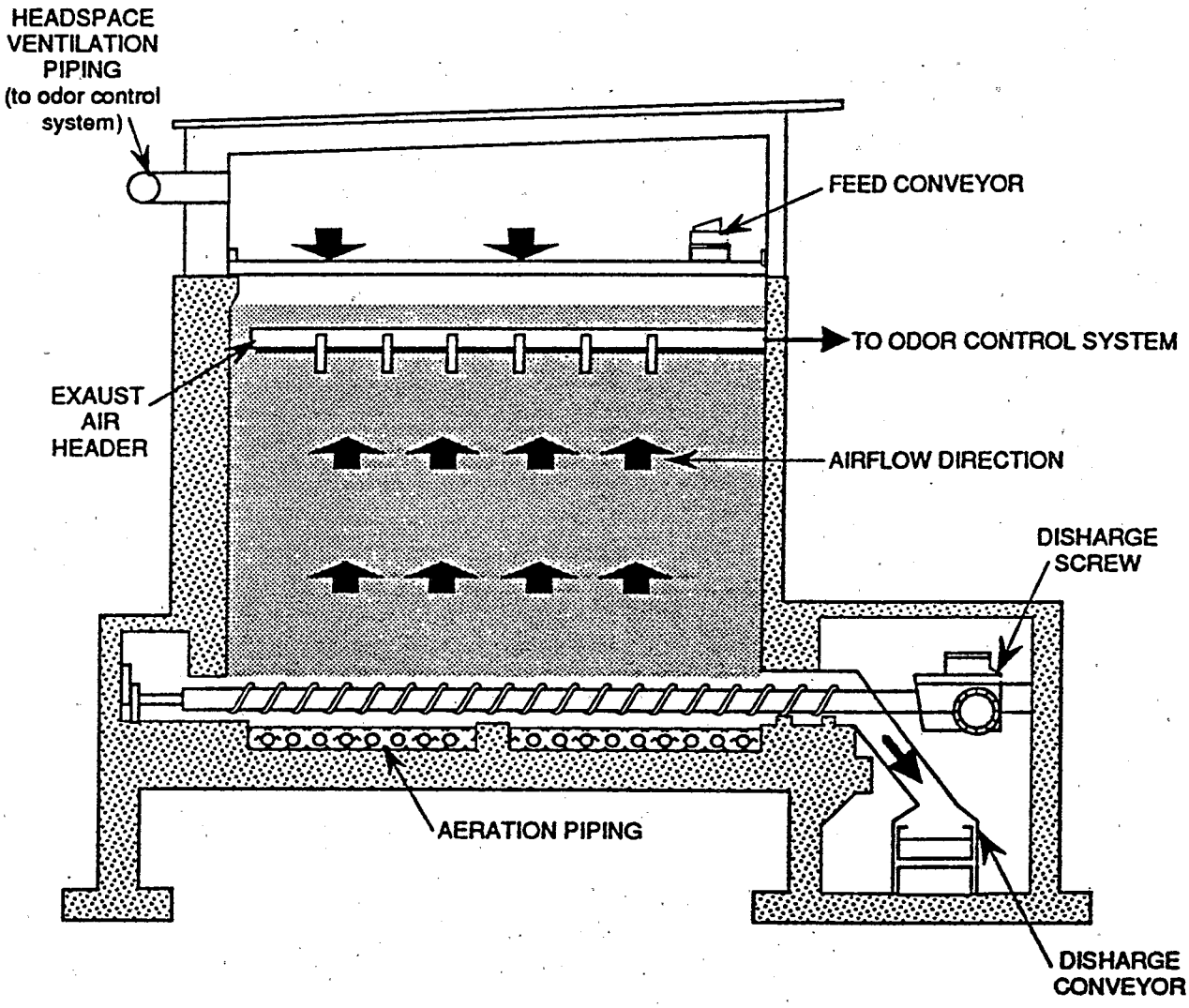


Figure 4. Purac Reactor

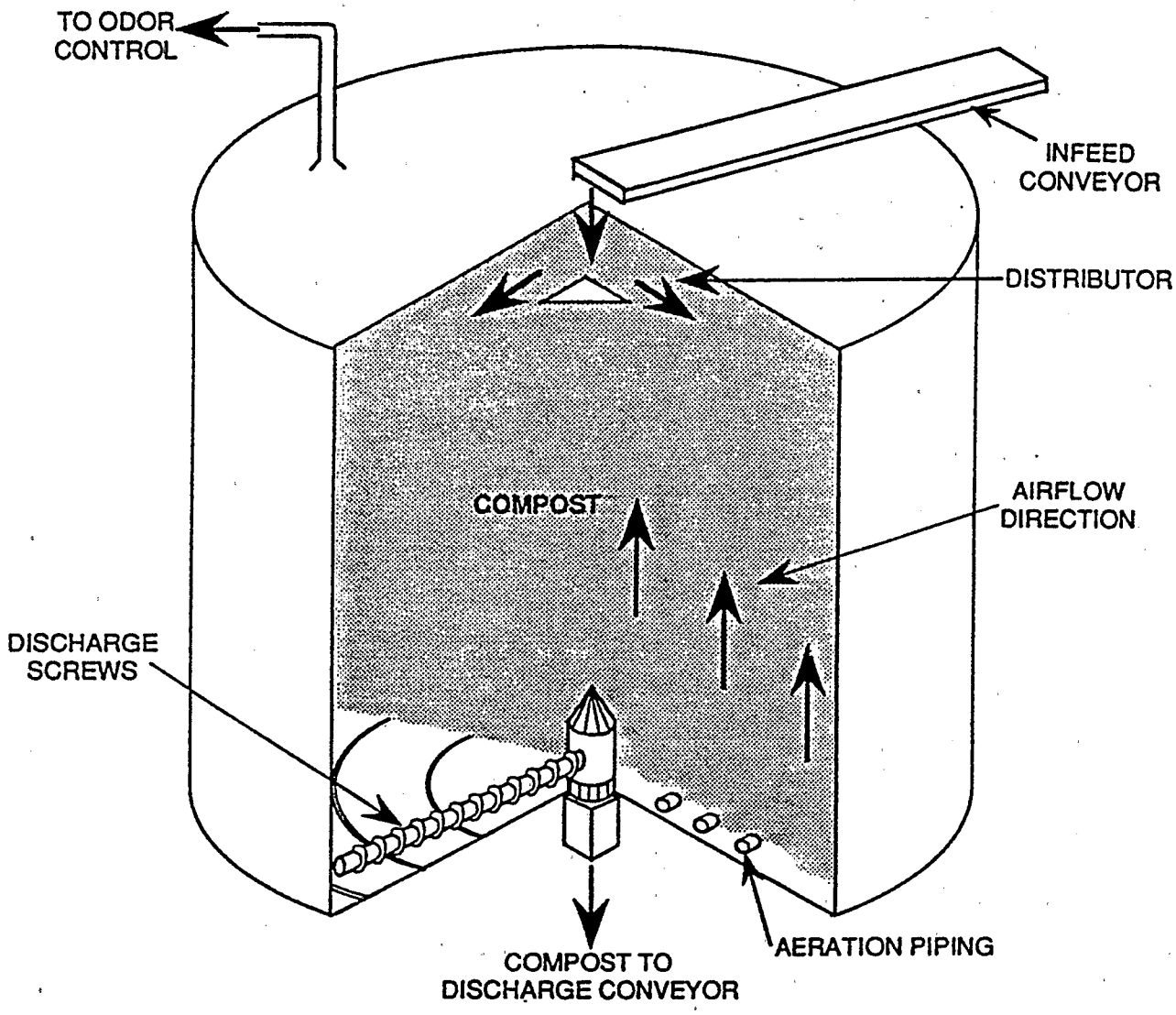
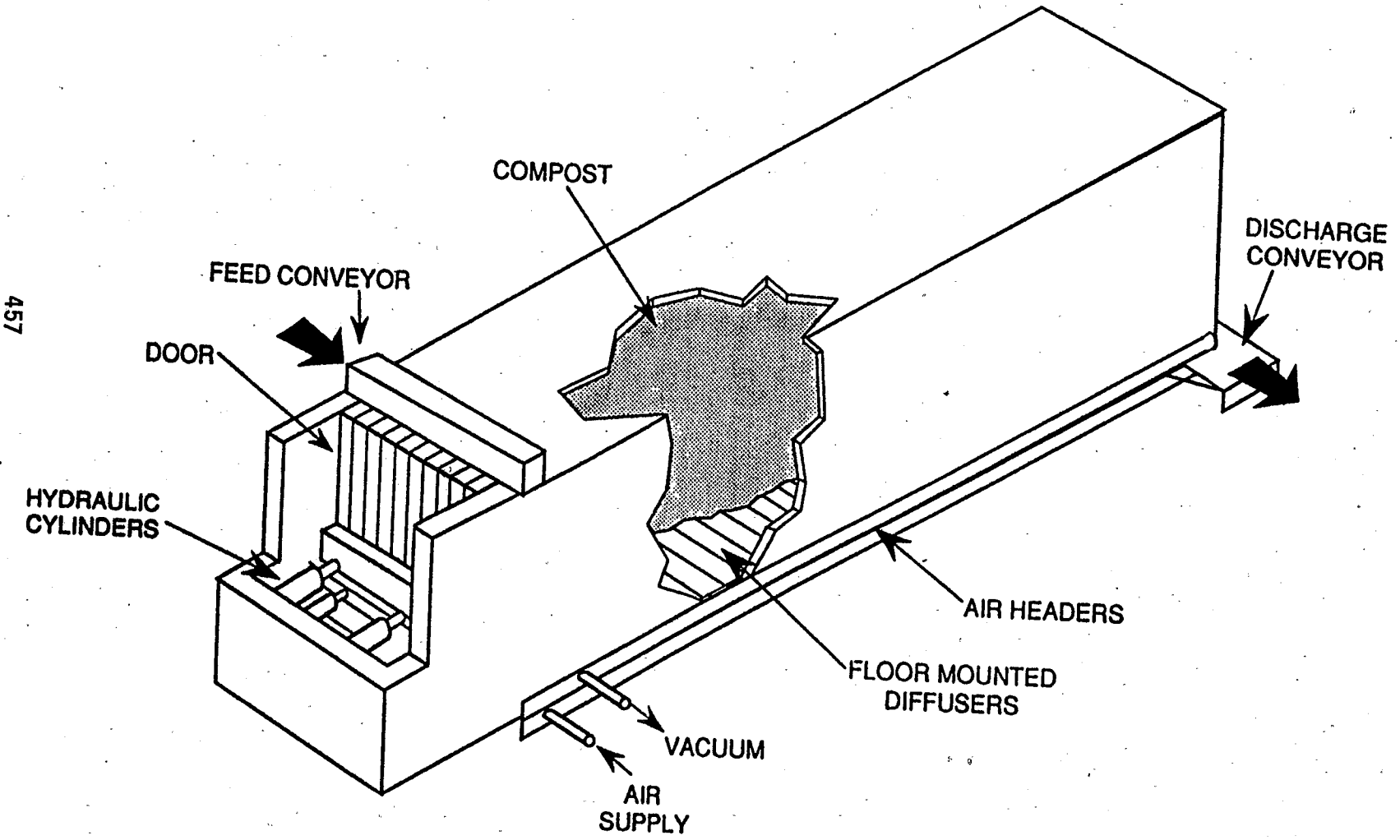


Figure 5. Taulman Reactor

Figure 6. Ashbrook - Simon-Hartley Reactor



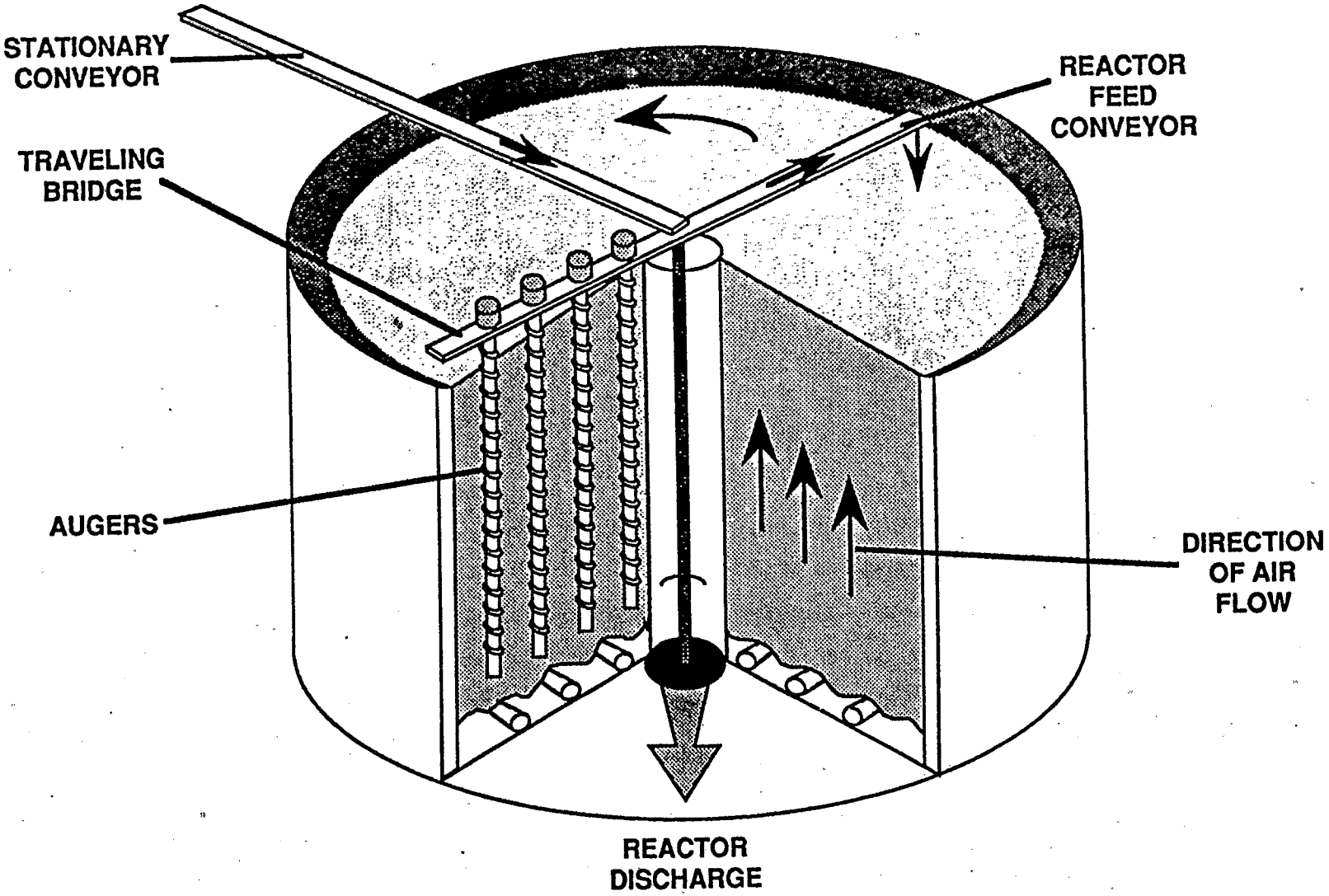


Figure 7. Fairfield Reactor

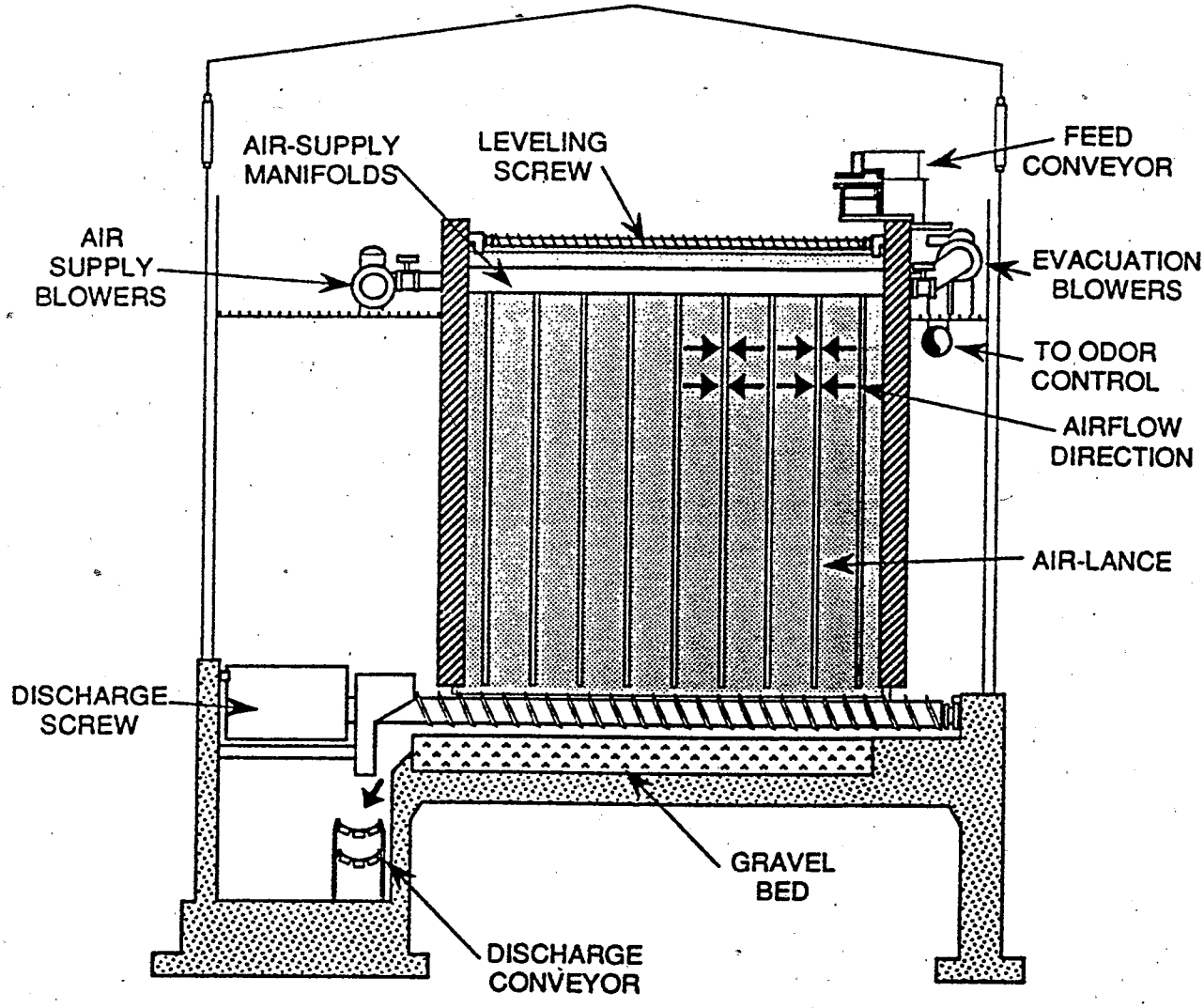


Figure 8. American Bib Tech Reactor

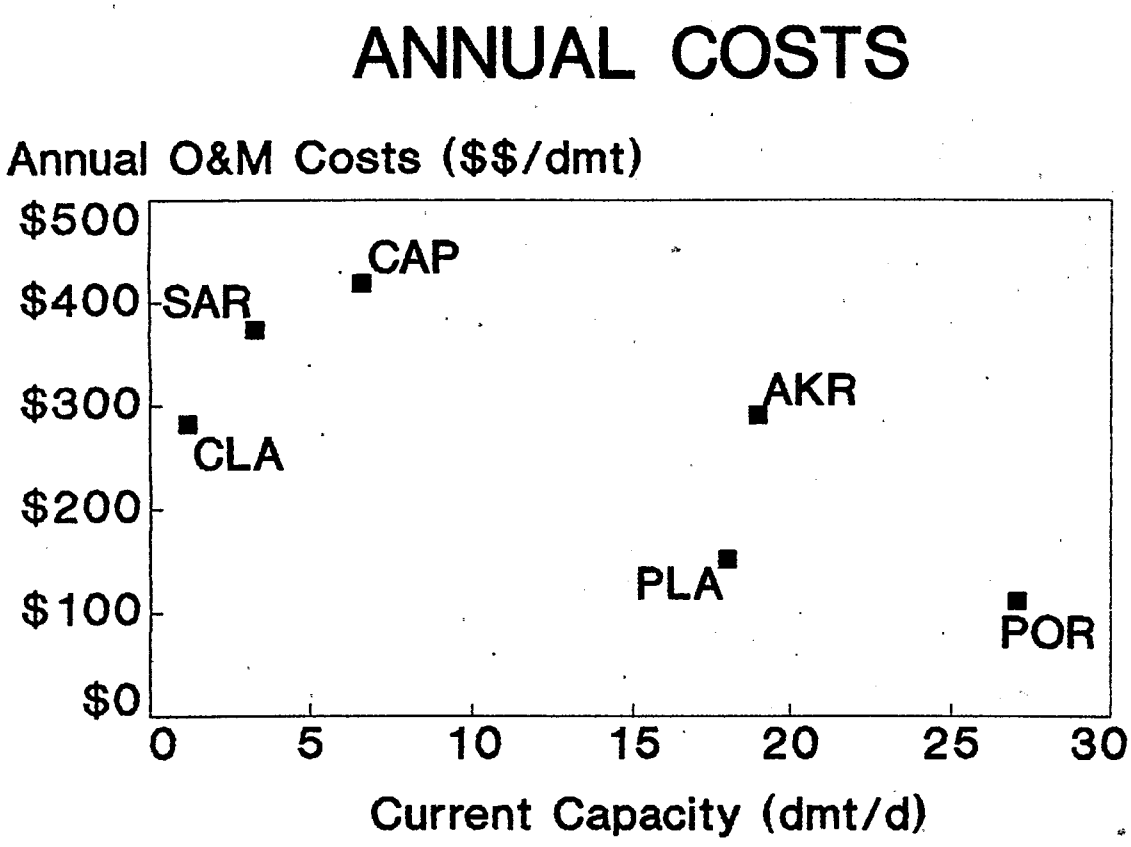


Figure 9. Annual Operation and Maintenance Costs

Akron, Ohio (AKR)

System:	Paygro
Design Capacity:	66 dmt/d*
Start-Up Date:	12/86
Type of Sludge:	Primary + Waste Activated 20 - 26% Total Solids > 70% Volatile Solids
Current Capacity:	19 dmt/d*
Current Status:	Operating at reduced capacity due to odors

* dry metric tons per day

Table 1. Summary of Akron, Ohio Facility

Cape May County, New Jersey (CAP)

System:	Purac
Design Capacity:	18 dmt/d*
Start-Up Date:	5/85
Type of Sludge:	Primary + Waste Activated
	20 - 30% Total Solids
	70 - 80% Volatile Solids
	$FeCl_3$ added to primary
Current Capacity:	6.6 dmt/d*
Current Status:	Operating at reduced capacity due to odors

* dry metric tons per day

Table 2. Cape May County, New Jersey Facility

Clayton County, Georgia (CLA)

System:	Taulman
Design Capacity:	2.6 dmt/d*
Start-Up Date:	8/86
Type of Sludge:	Waste Activated
	15 - 17% Total Solids
	68 - 79% Volatile Solids
	$Al_2(SO_4)_3$ added
Current Capacity:	1.2 dmt/d*
Current Status:	Processing all sludge produced by WWTP

* dry metric tons per day

Table 3. Summary of Clayton County, Georgia Facility

Newberg, Oregon (NEW)

System:	Ashbrook-Simon-Hartley (Ashbrook)
Design Capacity:	3.2 dmt/d*
Start-Up Date:	8/87
Type of Sludge:	Oxidation Ditch
	12 - 20% Total Solids
	--- Volatile Solids
Current Capacity:	--- dmt/d*
Current Status:	Not operating due to structural failure

* dry metric tons per day

Table 4. Summary of Newburg, Oregon Facility

Plattsburgh, New York (PLA)

System:	Fairfield
Design Capacity:	31 dmt/d*
Start-Up Date:	3/86
Type of Sludge:	Primary + Waste Activated
	20 - 25% Total Solids
	--- Volatile Solids
	$KMnO_4$ added to sludge
Current Capacity:	18 dmt/d*
Current Status:	Processing all sludge produced by WWTP

* dry metric tons per day

Table 5. Summary of Plattsburgh, New York Facility

Portland, Oregon (POR)

System:	Taulman
Design Capacity:	54 dmt/d*
Start-Up Date:	3/85
Type of Sludge:	Digested Primary + Digested Waste Activated 23 - 28% Total Solids 35 - 50% Volatile Solids
Current Capacity:	27 dmt/d*
Current Status:	Reduced capacity due to mechanical problems

* dry metric tons per day

Table 6. Summary of Portland, Oregon Facility

Sarasota, Florida (SAR)

System:	Purac
Design Capacity:	5.7 dmt/d*
Start-Up Date:	8/87
Type of Sludge:	Waste Activated 14% Total Solids 76% Volatile Solids
Current Capacity:	3.3 dmt/d*
Current Status:	Processing all sludge produced by WWTP

* dry metric tons per day

Table 7. Summary of Sarasota, Florida Facility

Schenectady, New York (SCH)

System:	American Bio Tech (ABT)
Design Capacity:	14 dmt/d*
Start-Up Date:	7/87
Type of Sludge:	Primary + Waste Activated 20 - 28% Total Solids 73% Volatile Solids
Current Capacity:	--- dmt/d*
Current Status:	Operating at reduced capacity due to odors

* dry metric tons per day

Table 8. Summary of Schenectady, New York Facility

Design vs Actual Mix

	Vol. Mix Ratio*		% Solids	
	Design	Actual	Design	Actual
AKR	1/0.66/1.33	1/1/2	40	44
CAP	---	1/0.6/1	40	37
CLA	1/1/1	1/1.4/1.4	--	34
NEW	1/0.35/5	1/0.5/10	38	38
PLA	1/0.67/1.67	1/1.4/3.3	40	40
POR	1/0.4/0.6	1/1.6/0.6	32	37
SAR	1/1/2	1/0.56/2.44	35	40
SCH	---	---	35	37

* Sludge/Amendment/Recycle

Table 9. Design and Actual Mix Ratios

Aeration Capacity

	Total Capacity (L/s)	Unit Airflow (L/m^3-s)	Path Length (m)
AKR	110,000	6.8	2.4-3.0
CAP	7800	4.1	6.4
CLA	710	0.88	7.9
NEW	2400	3.0	0.6-9.1
PLA	15,000	3.5	2.4-3.0
POR	3700	0.43	7.9
SAR	2500	1.2	7.0
SCH	4500	2.3	1.8-2.1

Table 10. Aeration Capacity of Reactors

Initial Odor Treatment Systems

AKR	Dilution (Process Air)
	Compost Biofilter (Non-process Air)
CAP	2-Stage Scrubber (Process Air)
	1-Stage Scrubber (Non-process Air)
CLA	Bubbling into Wastewater Pond
NEW	Bubbling into Oxidation Ditch
PLA	Single-Stage Wet Scrubber
POR	Bubbling into Channel at Headworks
SAR	Ozone Contact Chamber
SCH	Single-Stage Wet Scrubber

Table 11. Initial Odor Treatment Systems

Wet Scrubber Installations

	Stage	Source	Chemicals
CAP	1	Sludge storage	NaOCl+NaOH
	1	Mixing	NaOCl+NaOH
	1	Reactors	NaOCl+NaOH
	2		H_2SO_4
PLA	1	All	H_2SO_4+DeAmine
SAR	1	All	Cooling
	2		H_2SO_4
	3		NaOCl+NaOH
SCH	1	All	H_2SO_4+DeAmine

Table 12. Current Wet Scrubber Installations

Planned Wet Scrubbers

	Stage	Chemicals
AKR	1	H_2SO_4
	2	NaOCl+NaOH
CAP	1	Cooling
	2	H_2SO_4+Corn Syrup
	3	H_2O_2
PLA	1	H_2SO_4+DeAmine
	2	NaOCl+NaOH
SCH	1	Cooling
	2	H_2SO_4+DeAmine
	3	NaOCL+NaOH

Table 13. Planned Wet Scrubber Installations

Capital Costs

	Original ($$ Millions)	Additional ($$ Thous)	Future ($$ Thous)
CLA	2.0	80	--
NEW	2.8	--	??
SAR	4.2	--	200*
SCH	6.7	--	470*
CAP	8.4	3000*	--
PLA	13	--	5300*
POR	12	205	??
AKR	22	--	7500*

* Odor Control

Table 14. Capital Costs

THE DEVELOPMENT OF RISK ASSESSMENT METHODOLOGIES FOR USE IN REGULATION OF SEWAGE SLUDGE DISPOSAL

by

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INTRODUCTION

Risk assessment is a structured procedure for making quantitative estimates of the potential for harmful agents to cause adverse effects on human health or the environment. The use of risk assessment methods to determine appropriate limits for chemicals in environmental media such as air and water is well established. This paper will describe their use for estimating risk from chemicals in sewage sludge, and their potential use for estimating pathogenic risks as well.

The Environmental Criteria and Assessment Office of the U.S. EPA Office of Research and Development in Cincinnati, Ohio, has recently developed risk assessment methodologies which can be used to calculate criteria for protecting human health and the environment from toxic pollutants in municipal wastewater sludges (1-4). Since municipal sludges are managed utilizing a number of different reuse or disposal practices, separate assessment methods were developed to address land application (including distribution and marketing [D&M]), landfilling, land-based incineration, and ocean disposal of sludges. Assessment methods to address surface disposal (i.e., surface impoundment) of sludges are currently under development. Each methodology provides procedures for estimating the fate and transport of sludge-borne pollutants and for evaluating the potential for effects on humans or other organisms. The methods also describe data requirements for using these procedures to calculate criteria. The criteria are not necessarily numerical in all cases. In some instances, the risk assessment calculations can be used to determine appropriate controls on the design or operation of the management practice which achieve the same level of environmental protection. Such management controls are often preferred for their practicality. In many cases, however, the methodologies suggest that numerical limits on pollutant levels in the sludge should be calculated.

The risk assessment methodologies for each practice describe algorithms that can be used to calculate whether or not adverse effects would be expected from a given sludge constituent under a given set of conditions. Separate algorithms are presented for different pathways of potential exposure. For example, the methodology for sludge landfilling develops separate algorithms for contaminant movement to groundwater and for volatilization through the soil cap. The algorithms range from simple to complex. For example, in order to prevent phytotoxicity, the land application methodology states that soil levels toxic to plants should be determined from the literature for each chemical, and that sludge applications should be limited to prevent the occurrence of such levels. Minimal calculation is required. On the other hand, the algorithm for the

groundwater pathway for sludge landfilling requires the use of three computer models in succession to determine the effects of unsaturated zone transport, geochemical equilibria, and saturated zone transport, respectively, on potential for human exposure.

Using these procedures, conditions can be established which would prevent adverse effects. As mentioned previously, these conditions ("criteria") may consist of a set of recommendations for design or operation of the practice. For example, a maximum ground slope or minimum depth to groundwater could be recommended for land application practices in order to prevent surface or groundwater pollution, respectively. Often, however, numerical criteria which limit the chemicals themselves are derived. Depending on the algorithm, these numerical criteria may take the form of chemical concentrations in the sludge (on a dry or wet weight basis), rates of disposal (in terms of chemical mass disposed per unit area or per unit time or both), or some other form. The form of criteria dictated by each algorithm will be described for each reuse/disposal practice.

LAND APPLICATION (INCLUDING D&M)

Municipal sludges may be used to enhance productivity of agricultural or forest lands, to reclaim severely disturbed lands or may be applied to land which is dedicated to sludge disposal. Sludge may also be sold or given away in bag or bulk form to private or institutional users, a practice referred to as distribution and marketing (D&M). These practices combined were estimated to account for about 42 percent of municipal sludge disposal in the U.S. in 1980 (5), and were estimated at 24.7 percent in 1986 (6).

There are a wide variety of pathways by which chemicals in land-applied sludge can impact humans or other organisms. Human exposure potential is greatest when sludge is applied to agricultural lands or provided to the public for lawn or garden use. Uses on other lands pose less exposure potential initially, but in many cases the possibility exists of land-use conversion to agricultural or residential use.

Plant uptake of sludge-applied chemicals has been shown to increase as a function of amount of chemical applied. Criteria to protect humans or animals consuming these plants therefore will usually be expressed in terms of chemical mass applied per unit area, or kg/ha. For chemicals, such as many metals, which are not appreciably lost over time, the mass refers to the cumulative amount applied; no further additions would be permitted once the limit was reached. For chemicals that are lost through degradation or other processes, criteria are expressed as areal application per year, or kg/(ha•yr). The criterion is the amount which could be reapplied annually without ever exceeding a maximum cumulative amount.

In some situations, humans or animals may consume the sludge directly. Grazing animals may be directly exposed to surface-applied sludge which is not incorporated into the soil. Similarly, children are known to ingest

soil and could be exposed to pure sludge. For these types of exposures, the dry-weight pollutant concentration in the sludge is more pertinent than the pollutant application rate.

Offsite movement of sludge-borne chemicals can also occur through transport in surface runoff or by percolation to groundwater. These movements are treated as a function of annual areal pollutant application rate, rather than sludge concentration. Air concentrations resulting from volatilization may be a function of sludge concentration (wet weight) immediately upon application, but long-term average concentrations would again be a function of the annual areal application rate.

LANDFILLING

Landfilling was estimated to account for 15 percent of municipal sludge disposal in 1980 (5) and was estimated at 42 percent in 1986 (6). Most landfilled sludge is codisposed with municipal solid waste, while a lesser amount (1.3 percent of total sludge disposal in 1986) is disposed in sludge-only landfills, or monofills. The risk assessment methodology for landfilling (2) considers only the monofilling practice; codisposal will be dealt with in a separate assessment effort. It is assumed that site design features such as runon/runoff controls are adequate to protect local surface waters and that the daily application of cover material prevents any problems of windblown dust. Numerical criteria are suggested to prevent hazard from groundwater contamination: design criteria are not recommended since very few sludge landfills have liners and these liners may rupture. Numerical criteria are also suggested in order to limit the vapor concentrations downwind from the site.

Contaminant movement to groundwater and subsequent movement to a point of human use are predicted using a succession of transport and geochemical-equilibrium models. The required input to the first model (unsaturated zone transport) is a leachate concentration. In addition to leachate concentration, the total mass in the landfill of each pollutant must be estimated in order to determine the maximum length of time over which a leachate of a given concentration could be produced, since this duration can affect the predicted maximum concentration at the point of use.

Downwind movement of vaporized contaminant is estimated using a Gaussian air dispersion model. The vapor concentration immediately above the source during daily (uncovered) operation, or post-closure, is estimated from the leachate concentration using Henry's Law. Thus the leachate concentration is required for both of these transport pathways.

For organic compounds, the leachate concentration can be estimated from the whole-sludge concentration and the partition coefficient of the contaminant. Inorganic compounds are assumed to be present at the limits of their solubility, based on the maximum effluent or leachate levels observed in wastewater treatment plants.

Chemicals to be analyzed in the leachate or whole sludge, as determined by preliminary screening procedures, could include various metals, cyanide, nitrate, nitrosamines, chlorinated hydrocarbon pesticides, PCB and phthalate esters (7, 8).

SURFACE DISPOSAL

Placement of sludge in surface impoundments of some type is common; however, the extent of the practice is difficult to estimate because of the wide variety of impoundments and their reasons for use. Impoundments vary widely in size and construction. Use may be for treatment, such as aeration or dewatering, or for storage prior to ultimate disposal. Occasionally the impoundment itself becomes the long-term or ultimate repository, and therefore, is considered disposal. This form of disposal was estimated in 1986 to account for 2.6 percent of all U.S. sludge disposal (6).

Exposure pathways included in the methodology under development are the same as for landfilling, except that movement from ground water into surface water, and subsequent transport, are included because impoundments normally are located at treatment plants in the proximity of streams, rivers and estuaries.

INCINERATION

Incineration accounted for an estimated 27 percent of sludge disposal in 1980 (5) and estimated at 21.4 percent in 1986 (6). Incineration technologies in use include primarily the multiple hearth and fluidized bed furnace types. The risk assessment methodology for sludge incineration (3) examines potential risk to humans from incinerator stack emissions. Contaminant concentrations at ground level near the facility are estimated by the use of air dispersion models which account for aerodynamic downwash (resulting from low stack heights) as well as local terrain and meteorology. The required input to the model is the stack emission rate of the contaminant, in g/sec. Indirect pathways of exposure which could result from long-term deposition of pollutants in soils or waters near the incinerator, were not examined. New methods under development will eventually allow the inclusion of these indirect pathways.

Contaminants likely to be of concern vary according to incinerator type. Metals are more likely to pose a problem at the higher temperatures typical of fluidized bed furnaces. Loss of volatile organics from the upper hearths and incomplete destruction of other organics are likely to be more prevalent for multiple hearth furnaces. Classes of organics possibly requiring criteria include chlorinated hydrocarbon pesticides, polychlorinated biphenyls (PCB), polynuclear aromatic hydrocarbons (PAH) and phthalate esters (7, 8).

OCEAN DISPOSAL

Ocean disposal of municipal sludge by barging is currently permitted in only one location, the newly designated 106-Mile Ocean Waste Dump Site off the New York/New Jersey coast (9). Ocean disposal (at the previously designated "12-Mile Site") accounted for about 4 percent of U.S. sludge disposal in 1980 (5) and about 5.5 percent in 1986 (6).

The risk assessment methodology for ocean disposal of sludge is concerned with both protection of marine life near the site and protection of humans consuming seafood from that area. Water-column concentrations to which marine organisms are exposed are estimated from dispersion models. To calculate short-term concentrations within the dump site itself, where toxic effects on marine organisms are most likely, the sludge concentration (wet weight), as well as speed, dumping rate and positioning of barges, is the required input information. To estimate long-term, down-range concentrations, which are of more importance for assessing human exposure, only the total contaminant input to the dump site on a daily basis is required. Methods are not yet available to estimate sedimentation rates for sludge particles and potential for seafloor accumulation at the 106-Mile Site.

Chemicals potentially requiring analysis include metals, chlorinated hydrocarbon pesticides, PCB, PAH, phthalate esters and aromatic amines (7,8). In addition to analysis of individual constituents, whole-sludge bioassay methods may be used to determine the degree of dilution required to prevent aquatic toxicity. While not dealt with in these risk assessment methodologies, which focus on evaluating hazards of individual chemicals, these bioassay methods have been required for ocean-disposed wastes under other authorities (10), and are also useful for sewage sludge.

PATHOGENS

The U.S. EPA is currently reevaluating the existing, technology-based pathogen regulations, which define Processes to Significantly Reduce Pathogens (PSRP) and Processes to Further Reduce Pathogens (PFRP) (11). A feasibility study for performing risk assessments for pathogens was completed in 1985 (12, 13). Based on the feasibility study, the U.S. EPA is developing a quantitative risk assessment methodology for pathogens (i.e., bacteria, viruses, protozoa and helminths) in land-applied sludge. The methodology examines human exposures to pathogens that could result from the following pathways:

- aerosols generated by sludge application or incorporation
- wind-generated particulates from soil surface
- runoff of precipitation or irrigation
- direct contact with soil or crop

- movement through ground water to well
- use of contaminated ground water for irrigation
- consumption of crops or animal products

Because of substantial gaps and uncertainties in the data required to execute this methodology, it is anticipted that the initial uses will be to set priorities for needed research. Eventually, however, the use of such methodologies may enable the Agency to better determine the efficacy of the technology-based approach, or could indicate that a concentration-based approach, as used for chemicals, would be more appropriate.

CONCLUSION

The development of risk assessment methodologies, as this paper has described for sewage sludge, is an important first step in regulation development. When undertaken in a careful manner, divorced from policy and other accessory considerations, such methodologies can help to ensure that regulations will be consistent and scientifically appropriate. However, the steps involved in executing the methodology are equally important. Methodologies of the present type, which include multiple pathways of transport and exposure, require large amounts of data for execution. Thorough collection and accurate interpretation of data are essential. Sensitivity and uncertainty analyses are needed to interpret variability in calculated results.

In addition to the use of risk assessment to determine criteria for avoiding health and environmental risks, the consideration of economic impacts, technical feasibility and policy issues ordinarily is appropriate. It usually is best, however, to clearly delineate these latter issues from the former.

An especially important element when assessing the risks of disposal of a waste product such as sewage sludge is the comparability of assessment procedures for the various disposal alternatives. If a given disposal practice is examined in isolation from the others, and the existence of a risk is determined, that practice could be rejected without adequate determination of whether available alternative practices are more or less risky. Similarly, if the procedures for one practice are much more stringent than those for the others, then sludges may be forced away from that practice and into others that are more risky.

In the case of the methodologies for chemical pollutants in sludge, comparability was difficult to achieve because of the great differences in the exposure pathways involved from one disposal practice to another. An attempt was made to minimize this problem by using similar analytical frameworks for each practice, and by using the same methods and assumptions wherever possible. Additional work is needed, however, to provide better procedures for ensuring that all risks are comparably assessed.

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UV DISINFECTION

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INTRODUCTION

The use of ultraviolet (UV) radiation for the disinfection of treated wastewaters is an accepted practice in the United States. There is an increasing pace of installation at new plants and chlorine contact chambers at a number of existing plants are being retrofitted with UV equipment. This has been spurred by the acceptance of the technology as a viable and established process, and by the negative aspects of chlorination: very often dechlorination is required to meet increasingly stringent limits on residual chlorine levels; chlorination can result in the formation of toxic chlorinated organics; and there are significant concerns over the safety of transporting and storing chlorine and sulfur dioxide. Conversely, UV has no residual, does not affect chemical reactivity, and does not have safety concerns that would affect the surrounding community.

The USEPA disinfection program initiated research and demonstration efforts in the mid-seventies to develop the UV process as an effective alternative to chlorination. This support has continued, particularly in the construction grants funding under the EPA Innovative and Alternative Technology program. By 1987 forty-two plants had received I/A funding for UV facilities. Research efforts were directed to assessing the efficiency of the process, and to establishing and demonstrating protocols by which the process can be designed and evaluated. This design procedure will comprise the major portion of this presentation. Before this, however, some discussion will be given to how UV works and the status of UV installations in the United States today.

UV DISINFECTION

Ultraviolet radiation disinfection is a physical process, relying on photochemical changes brought about when far-UV radiation emitted from a source is absorbed by the genetic material of a cell (deoxyribonucleic acid, or DNA). The damage to the DNA results in the inability of the cell to replicate, a lethal effect. The most effective spectral region for germicidal activity corresponds to the maximum absorption spectrum of nucleic acids (Figure 1). This is between 250 and 265 nm.(1)

Very conveniently, low pressure mercury arc lamps are very efficient in generating UV light within this range: approximately 85 percent of their light output is monochromatic at 253.7 nm. They are typically made of Vycor glass, are 0.75 m or 1.5 m long, and 0.75 cm in diameter. Some disinfection applications exist that use alternative lamps. The Lewisburg plant in Ohio uses medium pressure mercury lamps, which have higher intensities and a broader spectrum of output. A modified low pressure lamp that is capable of higher intensities than the conventional low pressure lamp is being demonstrated at the Baldwin, Florida plant. Overall, however, the conventional low pressure lamps comprise the source of UV energy in effectively all disinfection systems today.

PHOTOREACTIVATION

Most organisms have an ability to repair and reverse the lethal effects of UV, a phenomenon broadly termed photoreactivation. It is typically a photoenzymatic mechanism, and requires concurrent or post-treatment with non-ionizing radiation in the near UV or visible range. The repair is of a constant fraction, independent of dose. This is clearly exhibited by data shown on Figure 2, developed from an EPA study conducted at a New York City treatment plant.(2) The figure shows the residual fecal coliform density as a function of UV loading (lpm/UV Watt). The data represent analyses of samples held in the dark after UV exposure, and the same sample exposed to sunlight for at least one hour. Photorepair accounted for approximately a 1 log increase in fecal coliform density.

Most organisms have the ability to repair, although there are some pathogens that will not, including the viruses. The total and fecal coliform groups will exhibit photorepair, while the enterococcus group does not. In the design of UV systems, it is important to understand that environmental conditions at a treatment plant will be conducive to photoreactivation. As such, the phenomenon must be considered in sizing the system; generally this suggests that the system must be sized to accomplish an additional log reduction beyond the design effluent conditions.

STATUS OF UV SYSTEM APPLICATIONS IN THE UNITED STATES

There are a number of ways in which the UV reactors have been designed to disinfect treated wastewaters. Basically the intent must be to minimize the loss of UV energy and to maintain a minimum exposure time for the wastewater. This requires close contact of the wastewater with the UV source, and plug flow conditions within the reactor itself.

Closed shell quartz reactors have the lamps (with quartz sleeves) fixed within stainless steel cylinders in full contact with the wastewater. The lamp spacing is typically 8 to 12 cm (centerline), with the wastewater flow directed parallel to the quartz. The systems are typically gravity flow, with piped inlet and outlet. These systems, because of the higher velocities at the entrance and exit points, tend to exhibit higher advective dispersion, affecting disinfection performance. The design also provides poor access to the lamps and quartz for maintenance and repair. These types of units tended to dominate the market through the mid-eighties, but are currently being used less frequently.

A "non-contact" configuration uses thin-walled Teflon pipes to carry the liquid. These are surrounded by unsheathed lamps, placed coaxially with the Teflon tube. The hydraulics of this arrangement are very good, with low dispersion plug flow through the long (typically 1.5 to 3 meters), small diameter (9 cm) pipes. Energy efficiency is generally lower, however, than the quartz systems, and the systems have been difficult to maintain. These are generally not considered for new applications.

Open-channel systems rely on the submergence of the quartz and lamp bundle in an open-channel. The lamps are either fixed in place, with flow directed perpendicular to the lamp, or inserted as modules, with the flow parallel to the lamp. The fixed open-channel units reflect the earlier designs and suffered problems with accessibility and maintenance; these are no longer installed.

The newer open-channel designs use a modular arrangement of quartz-sheathed lamps that can be easily inserted into and removed from the channel. The lamps are placed vertically into the channel, or horizontally, with the flow parallel to the lamps. Figure 3 shows an open-channel, horizontally placed, modular UV design. These systems represent the state-of-the-art in UV equipment, accounting for the major fraction of new installations and retrofitting of chlorine contact basins.

The open-channel modular design is best suited to UV process design. The lamps are placed in the liquid at relatively close spacing, affecting efficient use of the UV energy. Additionally, the open-channel design itself can be exploited to favor lower dispersion, plug flow conditions, by forcing a long liquid pathlength relative to its cross-sectional dimension. The modular design also allows easy removal of the units for cleaning and repair, the lack of which hampered the efficient operation of the earlier designs.

Figure 4 gives a perspective on how UV has progressed, particularly with the types of systems that are being installed. A survey(3) in 1984 identified 53 operating plants. Most were small; 80 percent had design flows less than 1 mgd (3,900 m^3/d). Nearly a third were the non-contact Teflon units, and half were closed shell reactors. The remainder were open-channel designs, but only one of these used the modular approach.

In 1988 there were close to 300 plants in operation. Of 177 for which information was available, 59 percent had design flows less than 1 mgd (3,900 m^3/d). Larger systems are being put on-line; however, 38 percent have design flows between 1 and 10 mgd (3,900 and 39,000 m^3/d) and several plants have capacities greater than 10 mgd (39,000 m^3/d). The largest operating plant is Madison, Wisconsin, with a design peak of approximately 100 mgd (390,000 m^3/d). Quebec City, Canada, has commissioned two plants with a combined design capacity of 230 mgd (900,000 m^3/d).

The distribution of the systems reflects the direction of the UV equipment field. Seventeen percent are Teflon, with only a few new installations since

1984. Only a quarter (26 percent) of the plants use the closed shell design; they are still being installed, but the trend is away from them for wastewater applications. There are no new fixed in-channel units. The trend is obviously to the open-channel modular systems, which after four-years now comprise greater than half the operating plants. Of these, 85 percent utilize the horizontal configuration and 15 percent the vertical placement.

DESIGN OF UV PROCESS

Through the mid-eighties, reflecting experience with "first-generation" systems, problems with UV installations related to severe component, fabrication and installation deficiencies (ballast failures, panel heating, electrical failures, and poor reliability). This has changed as the field is dominated more and more by "second generation" systems. There is a far greater degree of standardization, and the modular concept is allowing for a greater degree of flexibility and ability to affect quick repair and maintenance. More attention is being directed to design protocols that will assure proper sizing and can assist in the evaluation of existing systems.

A significant effort for the EPA Disinfection Program was the development and demonstration of a design protocol for the UV process. This was first described from a large scale pilot study conducted at the New York City Port Richmond plant[2], and subsequently detailed in the EPA Design Manual[1] and the Journal, WPCF.[4] Several in-field studies were also conducted to support specific aspects of the design protocol, and are described in the Manual. Since then, the design protocol has been used to develop the process sizing and layout of several new plants. Most recently, the USEPA has funded a pilot study to further evaluate UV process design. The field effort, conducted in Rehoboth Beach, Delaware, has been completed, and the project is now in the analysis phase.

DESIGN EXPRESSION

The inactivation of bacteria by UV radiation can be approximated by the first order expression (Figure 5):

$$N = N_o e^{-kIt}$$

where:

N = density after exposure (organisms/100 ml)
N_o = initial density (organisms/100 ml)
k = inactivation rate ($cm^2/\mu Watts\text{-}sec$)
I = intensity at 253.7nm ($\mu Watts/cm^2$)
t = exposure time (second)

This is ideal. However, under actual conditions there will be a distribution of exposure times about the mean, the intensity cannot be defined by direct measure, and particulates will set a base level for the effluent density. The modified model incorporates these design elements (Figure 6):

$$N = N_o e^{\left[\frac{ux}{2E}\left\{1 - (1 + \frac{4KE}{u^2})^{1/2}\right\}\right]} + N_p$$

where:

x = the characteristic length of the reactor, defined as the average distance traveled under direct UV exposure (cm)

u = the liquid velocity (cm/sec). This is calculated as

$$x/(V_v/Q)$$

where V_v is the void, or liquid volume in the reactor (L) and Q is the flow rate (Lps)

E = the dispersion coefficient (cm^2/sec), which quantifies the breadth of the residence time distribution about the mean time

K = the inactivation rate (sec^{-1})

N_p = the bacterial density associated with the particulates and unaffected by exposure to UV (org/100 mL

RESIDENCE TIME DISTRIBUTION

The dispersion coefficient is a key parameter defining the hydraulic behavior of the reactor, providing a measure of the spread of time about the mean residence time. It is estimated from a residence time distribution (RTD) analysis of a reactor, such as shown on part (a) of Figure 6. E is estimated by the expression:

$$E = \frac{ux\sigma^2}{2\theta^2}$$

where:

σ^2 = the variance of the RTD (sec^2)
u = reactor velocity (cm/sec)
x = characteristic length (cm)
θ = mean residence time (sec)

This assumes that the RTD determined for a reactor approximates a normal distribution. Under ideal plug flow conditions, E approaches zero; with increased mixing, E will approach infinity, which defines ideal complete mix. Generally, an E less than 100 cm^2/sec would be considered acceptable for effective design.

Figure 7 gives an example of a hydraulic analysis of a UV system in West Virginia. It is an open-channel, horizontal lamp unit, which originally had over and under baffles to break the velocity of the piped inlet to the channel. An RTD analysis of this channel showed very high dispersion, with E estimated at greater than 2,000 cm^2/sec. The ratio of t_{90} to t_{10}, which is the time for 90 and 10 percent of the tracer to pass, respectively, ranged between 2.2 and 6.4. Known as the Morrill Dispersion Index, this will also indicate significant dispersion at values greater than 2.0. Mixing was occurring in the reactor, preventing the unit from achieving expected disinfection efficiencies.

Removing the baffles provided some improvement, but not sufficient. The higher velocities from the inlet pipe caused a significant degree of dispersion to occur in the reactor. The E was estimated to be between 800 and 3,500 cm^2/sec, with a Morrill Index of 1.7 to 2.7.

A stilling plate was installed in place of the baffles. This is a uniformly perforated plate that is able to equalize the velocities entering the reactor lamp battery, encouraging plug flow. The E was reduced to less than 100 cm^2/sec in this case, indicating good plug flow conditions.

Thus, it is important to design a UV disinfection system for plug flow. This can be quantified by the dispersion coefficient, which has also been incorporated into the design expression. RTD analysis should be used to determine E (and other hydraulic indices), both from the design standpoint and to troubleshoot performance problems.

INACTIVATION RATE AND REACTOR INTENSITY

Returning to the design expression (Figure 6), the inactivation rate coefficient, K, is described as a function of the average intensity in the UV reactor:

$$K = a(I_{avg})^b$$

where I_{avg} is the average reactor intensity ($\mu Watts/cm^2$), and a and b are coefficients developed from a Log-Log regression of K and I_{avg}. These are specific to wastewater conditions, reflecting the sensitivity of the organisms to UV.

There is no method to directly measure the actual intensity within a complex, multi-lamp reactor. Rather, a computational scheme (point source summation method) is employed, based on the number of lamps in a reactor (UV density) and the absorptive properties of the liquid (UV absorbance coefficient). Solutions have been developed and are presented in the EPA Design Manual for a number of system configurations.

Figure 8 shows such an analysis for the symmetrical array of lamps employed by the modular, open-channel, horizontal lamp systems. Thus for units having a centerline spacing of 6.35 cm (typical), the UV density, D, is approximately 5.0 UV Watts/Liter; at an absorbance coefficient of 0.4 cm^{-1} (representative of a "secondary" effluent), the nominal average intensity is approximately 15,000 $\mu Watts/cm^2$.

For design purposes this nominal intensity should be adjusted to reflect deterioration of lamp output with time and fouling of the quartz surfaces. Both conditions will obviously attenuate the UV intensity. Typically, the lamp UV output will be at 70 percent of the nominal rating during the latter half of its operating life. The quartz surfaces will get dirty, generally from inorganic deposits such as iron, and calcium and magnesium carbonates. Routine maintenance of the quartz surfaces (which is a critical O&M task) can typically keep the transmissibility of the quartz at an average 80 percent of nominal. Thus, the design intensity can be 50 to 60 percent (70 percent x 80 percent) of the nominal reactor intensity.

Developing a rate coefficient for a specific application requires influent/effluent bacterial sampling over a range of intensity levels. This is typically done by operating two UV pilot units, each with differing average intensities. Figure 9 shows an analysis conducted at the Port Richmond WPCP in New York City.(2) In this case,

$$K = 1.45 \times 10^{-5} (I_{avg})^{1.3}$$

This procedure was demonstrated at four additional plants as a special EPA sponsored project in 1985. The results are summarized in reduced form on Figure 10. There was sufficient variability to suggest that this type of information may best be developed for a specific site (particularly larger plants), although using the combined expression may be adequate for estimating purposes. The combined data, shown by the heavy dark line, yielded an expression

$$K = 1.38 \times 10^{-5} (I_{avg})^{1.28}$$

EFFECT OF WASTEWATER PARTICULATES

Bacteria that are occluded in particulates will not be effected by UV radiation, and, as such, will comprise the base level of effluent density that can be achieved for a specific application. The term N_p in the design expression quantifies this effect (Figure 6) by relating it to the effluent suspended solids:

$$N_p = c\ SS^m$$

where SS is the suspended solids (mg/L) and c and m are coefficients derived from a Log-Log regression of effluent bacterial density and suspended solids.

These data must be generated under very high dose levels, such that the residual density can be attributed to those bacteria which were occluded from the radiation. An example is shown on Figure 11 for data developed during the Port Richmond project.(2) The log effluent fecal coliform density is plotted against the log effluent suspended solids. A linear regression analysis yields the expression (when transformed):

$$N_p = 0.26\ SS^{1.96}$$

Studies conducted at several plants show considerable variability, although it appears that these coefficients provide a reasonable estimate of N_p.

Calibrated Disinfection Model

The design expression takes on the following format, incorporating the coefficients required for model calibration (Figure 12):

$$N = N_o \exp\left[\frac{ux}{2E}\left\{1 - \left(1 + \frac{4E(aI_{avg}^{\ b})}{u^2}\right)^{1/2}\right\}\right] + c\ SS^m$$

The coefficients a, b, c and m are specific to a given wastewater application. These can be determined by pilot testing; alternatively, one may

use approximations determined from earlier studies, recognizing that they may not be fully representative of actual conditions. In this case, the suggested coefficients are:

a = 1.38×10^{-5}
b = 1.28
c = 0.26 (fecal coliform); 0.9 (total coliform)
m = 1.96

To confirm the model calibration, measured effluent densities can be compared to the residual predicted by the model. An example is given on Figure 13, developed from an evaluation of the Northfield, Minnesota plant[5], showing the observed versus predicted fecal coliform densities. The calibrated model was subsequently used to determine equipment requirements to upgrade the system.

SYSTEM SIZING USING THE DESIGN MODEL

The design model is a very effective tool in the sizing of a full-scale facility and assessing alternative configurations. This is especially the case when the model has been calibrated by direct testing.

Figure 14 is an example of design curves developed from the model for a wastewater treatment plant. The figure gives the log N'/N_o as a function of the reactor loading, which is defined as the flow (Lpm) per unit UV watt (W). The sizing assumes using a horizontal lamp, open-channel system, with a 6.35 cm centerline spacing, and 1.5 m lamps. The solutions are for two banks of lamps in series and a dispersion coefficient (E) of 100 cm^2/sec. The curves represent different inactivation rates that are defined by varying levels of the UV absorbance coefficient in the wastewater (recall that these affect the UV intensity in the reactor).

Consider sizing for the maximum six-hour design condition:

Q	= 42 mgd (160,000 m^3/d)
TSS	= 30 mg/L
N_o	= 600,000 FC/100 mL
UV Absorbance Coefficient	= 0.55 cm^{-1}
Design Effluent N	= 800 FC/100 mL

The inactivation rate, K, at this condition was estimated to be 1.0 sec^{-1}. The performance requirement (Log N'/N_o) is determined by first accounting for the particulate coliform density, N_p:

$$N_p = 0.26 \cdot (30 \text{ mg/L})^{1.96} = 204 \text{ FC/100 mL}$$
$$N' = N - N_p = 800 - 204 = 596 \text{ FC/100 mL}$$

$$\text{Log } \frac{N}{N_o} = -3.03$$

From Figure 14, the loading is estimated to be 2.0 Lpm/UV Watt. The number of lamps is computed on the basis of the design flow and a lamp specification of 26.7 UV Watts/Lamp. This yields 2,070 lamps for the maximum six-hour design condition.

With two units in series, the lamp requirement is 1,035/unit. At 8 lamps per module, approximately 140 modules are required per unit. This would be divided to parallel channels to provide flexibility and control of operations under varying conditions. A simplified layout is given on Figure 15. This shows 5 channels, each with 2 banks in series and 26 eight-lamp modules per bank. A sixth channel is provided as standby. Each channel is hydraulically independent, with liquid level control gates at the effluent end of the channel. The control panels and wireways are generally located near each bank.

SUMMARY

In summary, UV disinfection is an established technology that is rapidly gaining acceptance and is being put into operation at a growing number of

plants. The technology configuration that represents current state-of-the-art is the open-channel system with modular, horizontally placed lamp banks. Design protocols have been developed that address key process elements, including the dimensional configuration of the unit, UV intensity, reactor dispersion and RTD characteristics, and key water quality parameters. The modeling approach is generic and can be used to characterize most reactor configurations and wastewater conditions.

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5. HydroQual, Inc. Report to the City of Northfield, Northfield, Minnesota, "Evaluation of the UV Disinfection Process at the Northfield Water Pollution Control Plant," Mahwah, New Jersey, December 1984.

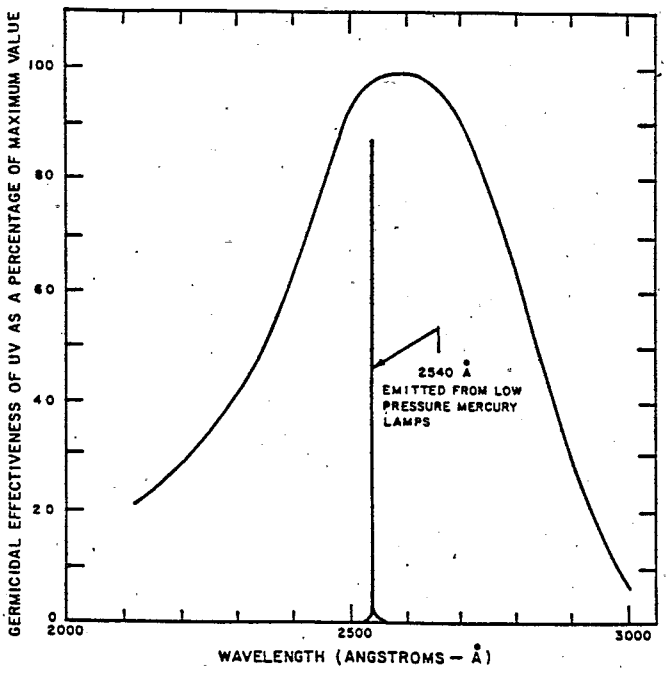


FIGURE 1. RELATIVE GERMICIDAL EFFECTIVENESS AS A FUNCTION OF WAVELENGTH (1).

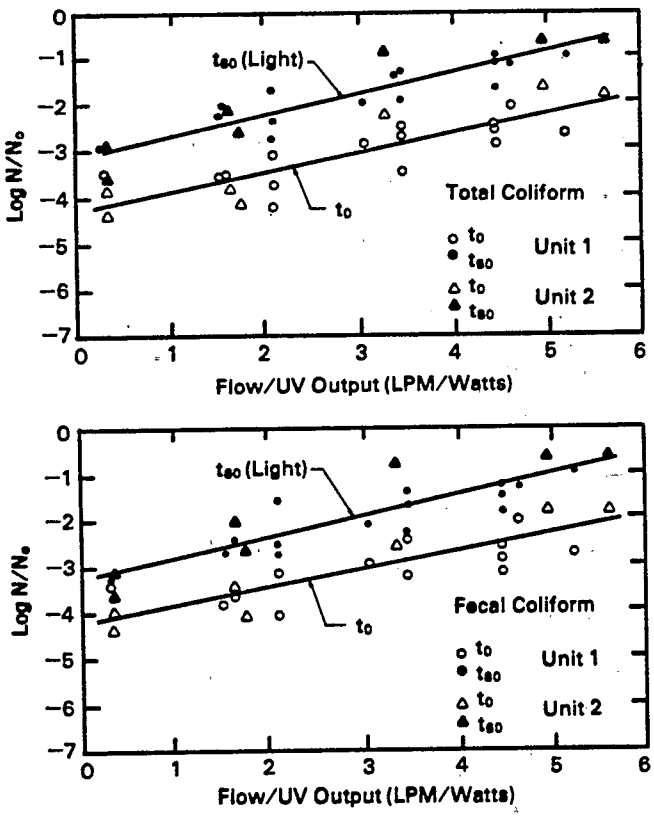
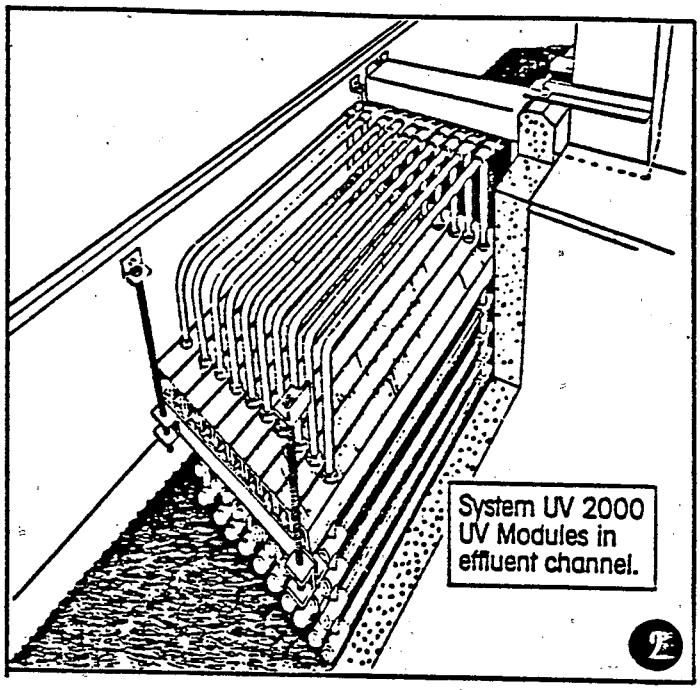


FIGURE 2. PHOTOREACTIVATION EFFECTS FOR TOTAL AND FECAL COLIFORM AT PORT RICHMOND (2).



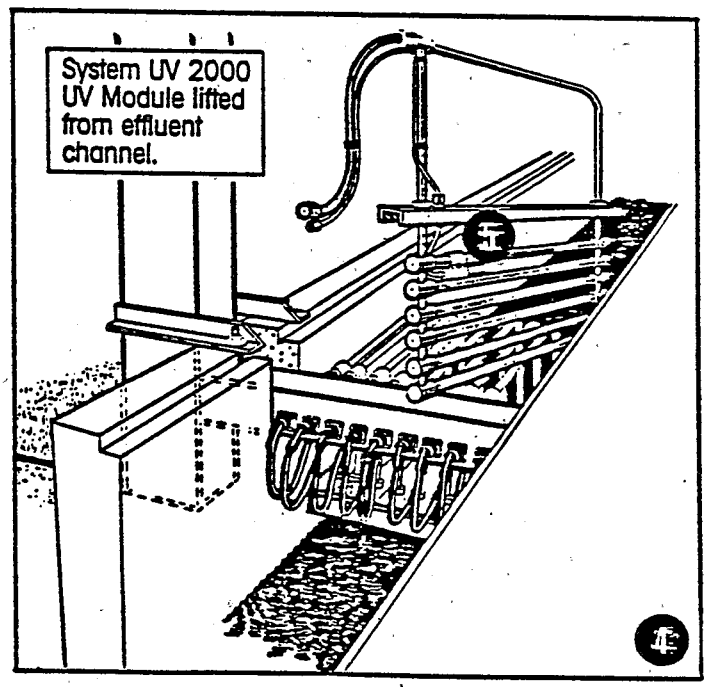


FIGURE 3. SCHEMATIC OF OPEN-CHANNEL, MODULAR UV SYSTEM (COURTESY OF TROJAN ENVIRONMENT, INC., LONDON, ONTARIO, CANADA).

1984

- OPERATING PLANTS
  - 80% WITH Q < 1.0 MGD
  - 35% TEFLON UNITS
  - 49% CLOSED SHELL REACTORS

1988

- ESTIMATE OF 250 TO 300 OPERATING PLANTS
- OF 177 PLANTS SURVEYED
  - 59% < 1 MGD
  - 38% 1 TO 10 MGD
  - 3% > 10 MGD
- TYPE OF SYSTEMS
  - 17% TEFLON
  - 26% CLOSED VESSEL
  - 4.5% OPEN CHANNEL (FIXED)
  - 52.5% OPEN CHANNEL (MODULAR)

FIGURE 4. STATUS OF UV INSTALLATIONS

$$N = N_o e^{-KIt}$$

N = effluent density (organisms/100 mL)

N_o = initial density (organisms/100 mL)

k = inactivation rate ($cm^2/\mu Watts\text{-}sec$)

I = intensity at 253.7 nm ($\mu Watts/cm^2$)

t = exposure time (sec)

It = Dose = $\mu Watt\text{-}sec/cm^2$

FIGURE 5. UV DISINFECTION EQUATION

$$N = N_o e^{\left[\frac{ux}{2E}\left\{1-\left(1+\frac{4KE}{u^2}\right)^{1/2}\right\}\right]} + N_p$$

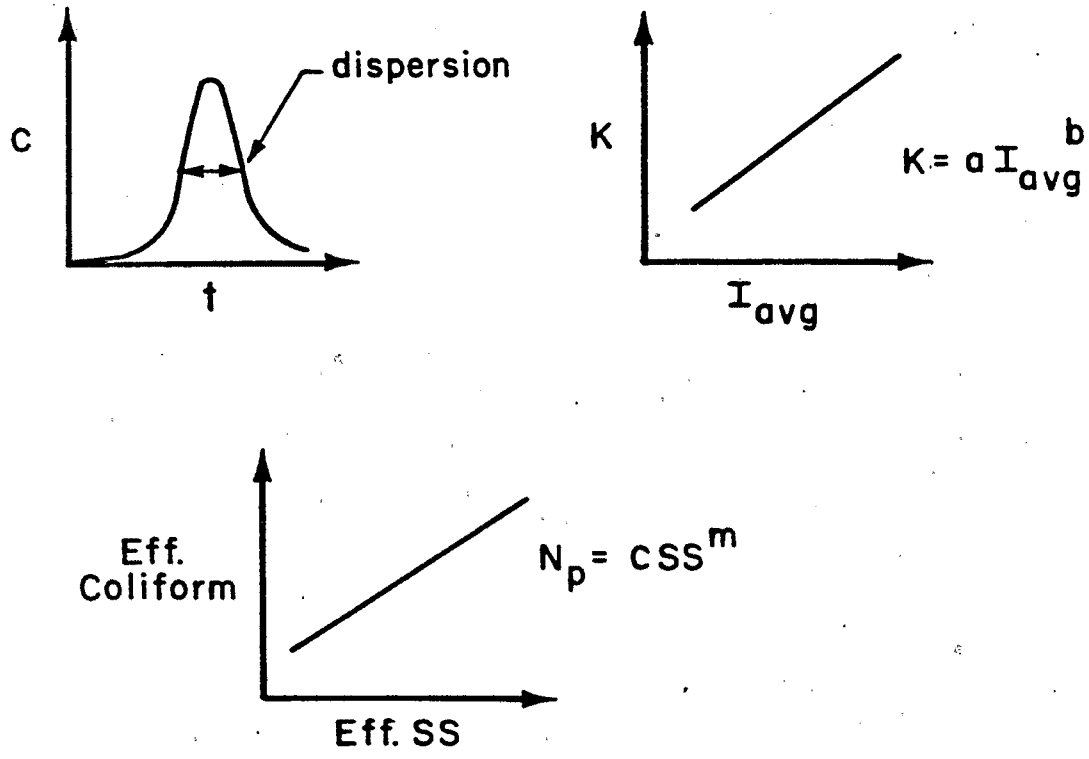
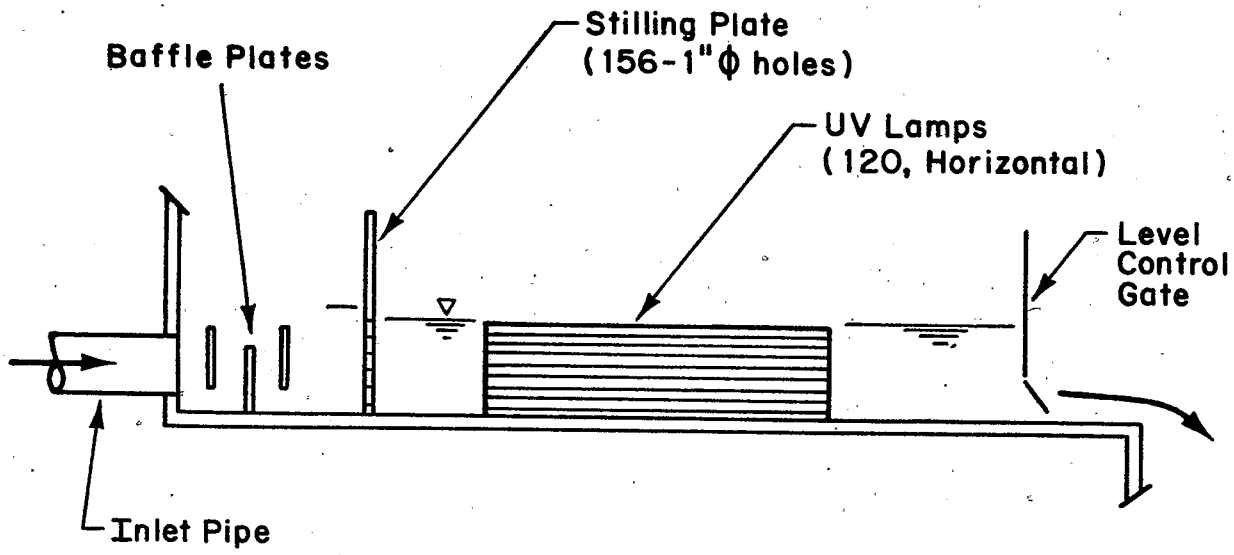


FIGURE 6. CONCEPTUALIZATION OF UV DESIGN MODEL



Conditions

	Baffles Only	No Baffles No Stilling Plate	Stilling Plate
E (cm^2/sec)	> 2000	800 - 3500	20 - 90
t_{90}/t_{10}	2.2 - 6.4	1.7 - 2.7	1.5 - 2.2

FIGURE 7. WILLIAMSON, WEST VIRGINIA HYDRAULIC ANALYSIS

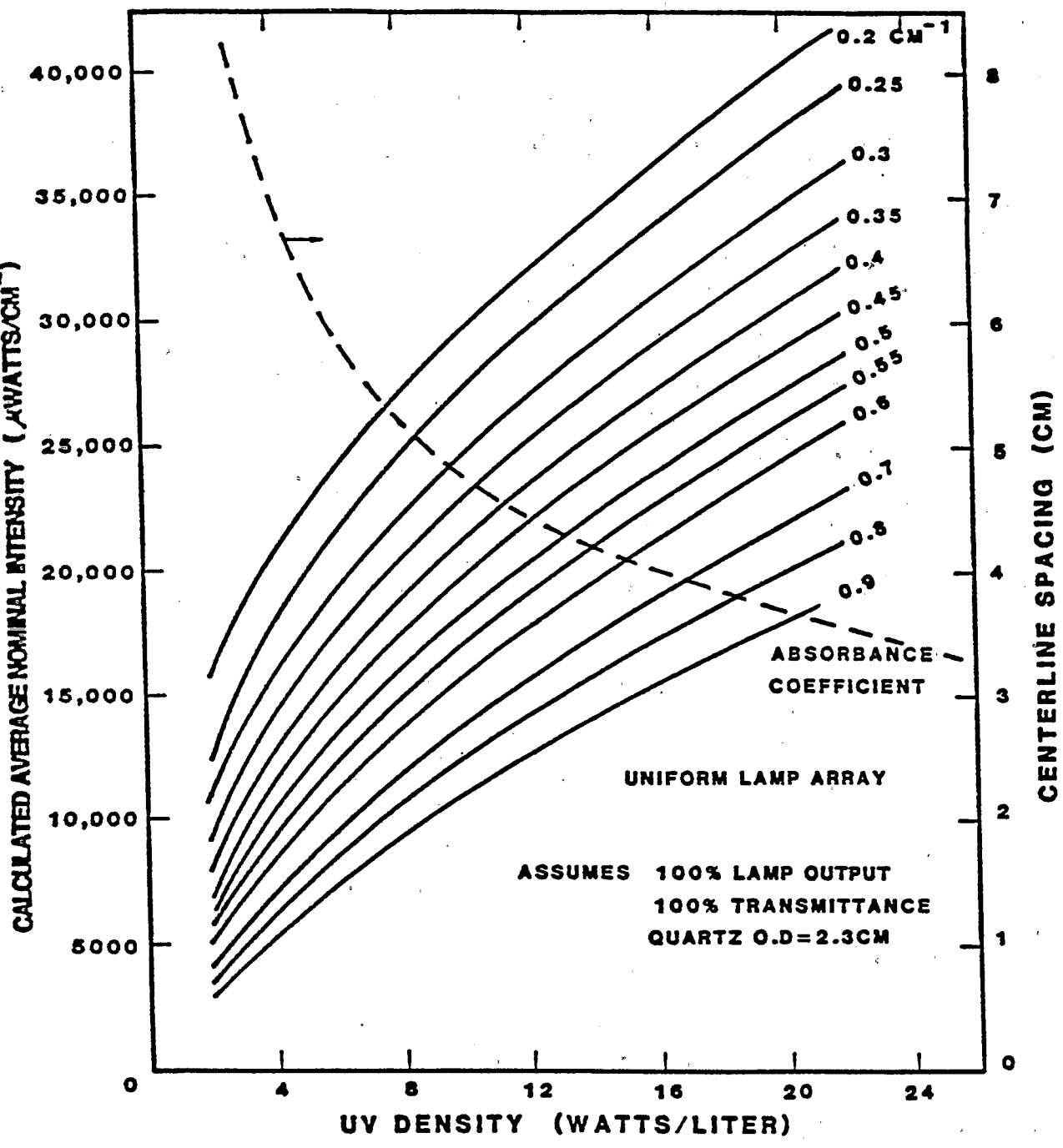


FIGURE 8. NOMINAL INTENSITY SOLUTIONS FOR SYMMETRICAL, UNIFORM LAMP ARRAY.

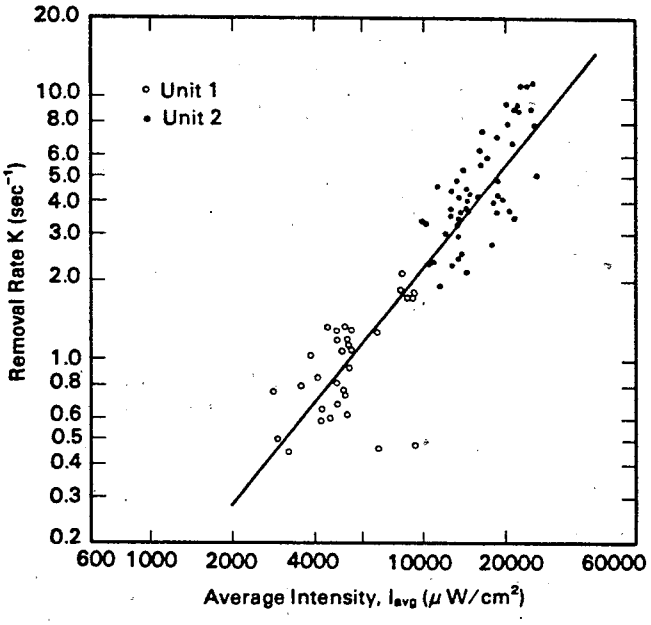


FIGURE 9. AN EXAMPLE FOR DERIVING AN ESTIMATE OF THE INACTIVATION RATE FOR FECAL COLIFORMS AS A FUNCTION OF THE CALCULATED AVERAGE INTENSITY (2).

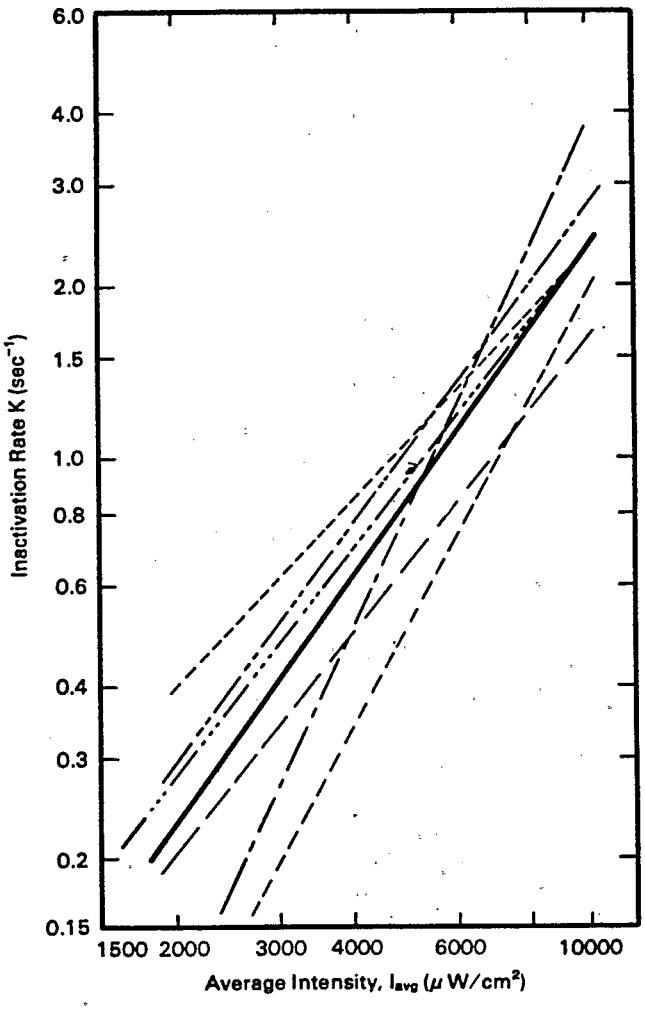


FIGURE 10. COMPARISON OF INACTIVATION RATE ESTIMATES FROM SEVERAL WASTEWATER TREATMENT PLANTS.

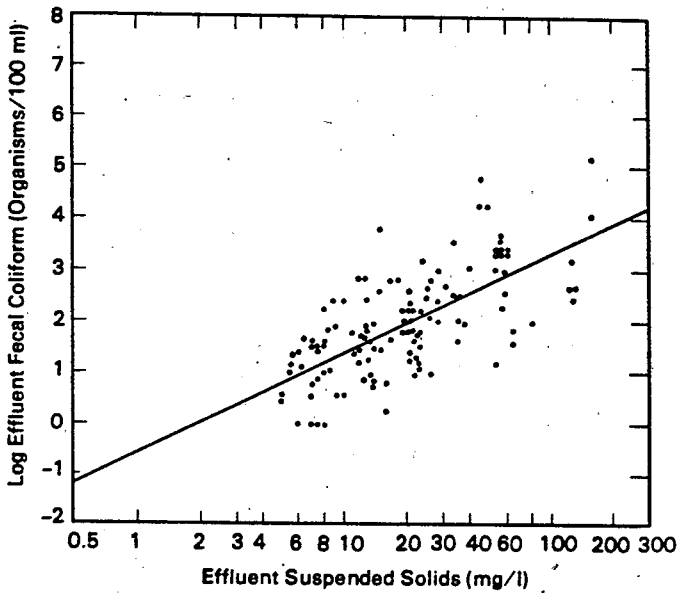


FIGURE 11. EXAMPLE OF DERIVING AN ESTIMATE OF THE RESIDUAL FECAL COLIFORM DENSITY ASSOCIATED WITH PARTICULATES AS A FUNCTION OF SUSPENDED SOLIDS (2).

$$N = N_o \exp\left[\frac{ux}{2E}\left\{1 - \left(1 + \frac{4E(aI_{avg}^{b})}{u^2}\right)^{1/2}\right\}\right] + c\ SS^{m}$$

FIRST APPROXIMATION:

$a = 1.38 \times 10^{-5}$

$b = 1.28$

$c = 0.26$ (fecal coliform)
$= 0.9$ (total coliform)

$m = 1.96$

FIGURE 12. DESIGN EXPRESSION INCORPORATING THE COEFFICIENTS REQUIRED FOR CALIBRATION

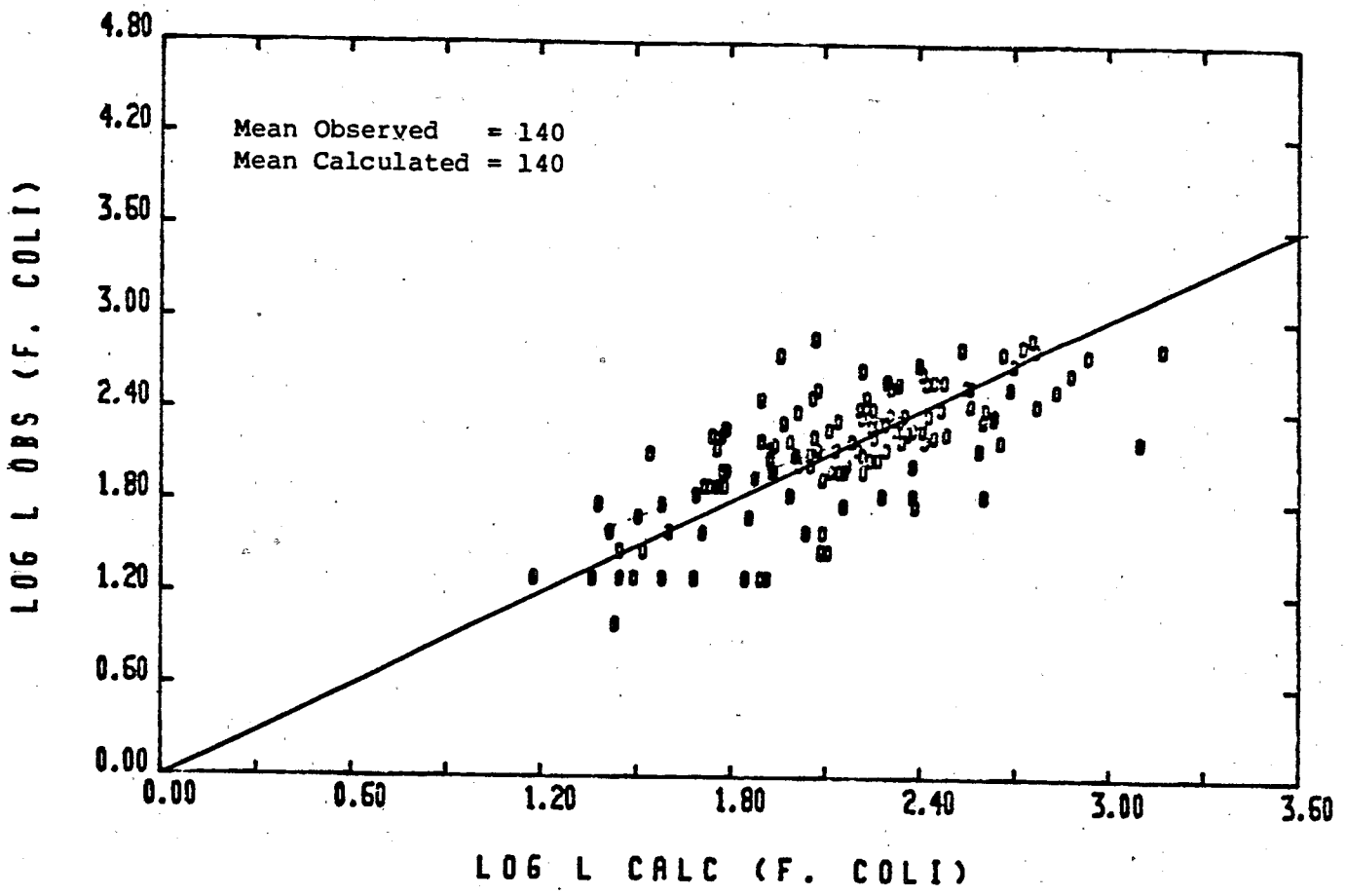


FIGURE 13. EXAMPLE OF COMPARISON OF OBSERVED DENSITIES WITH DENSITIES PREDICTED BY MODEL (NORTHFIELD, MN).

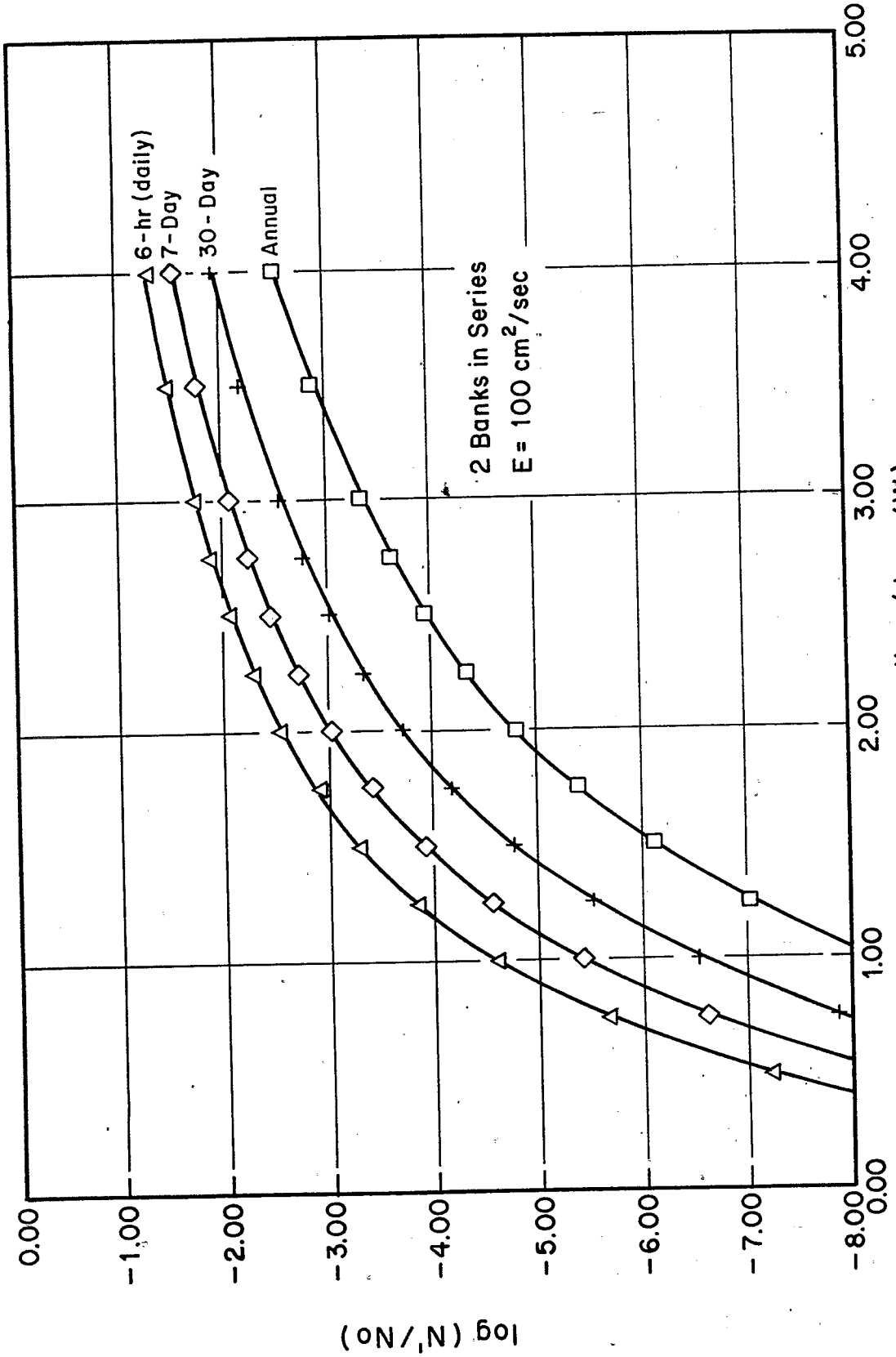


FIGURE 14. DESIGN EXAMPLE: PERFORMANCE VS. LOADING RATE

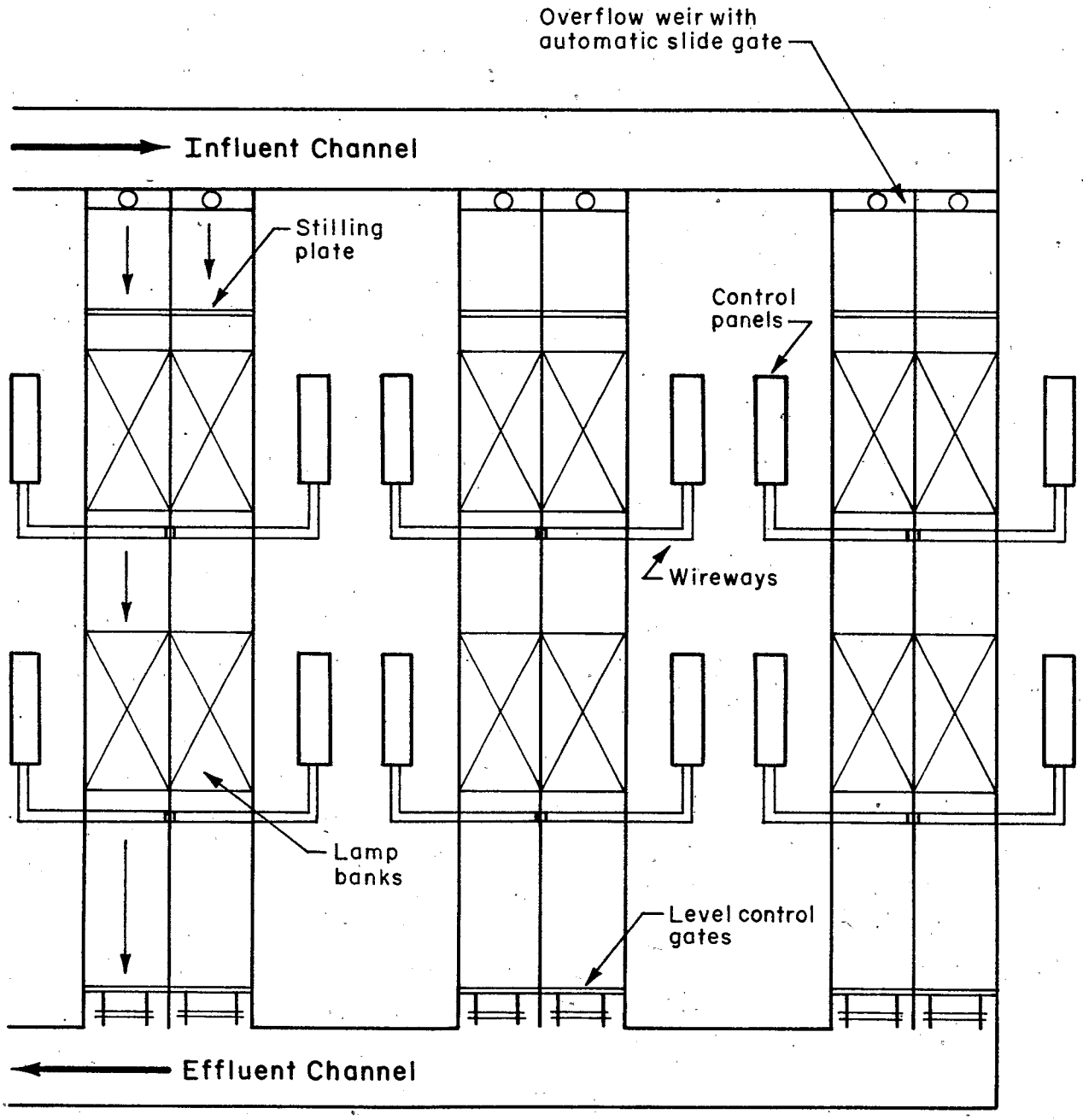


FIGURE 15. SIMPLIFIED LAYOUT FOR DESIGN EXAMPLE

A PRELIMINARY ASSESSMENT OF HIGH BIOMASS SYSTEMS

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This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

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INTRODUCTION

The Federal grants program for the construction of municipal wastewater treatment plants will end on 30 September 1989. With its termination, municipalities will be seeking plant modifications which will allow for increased capacity of their existing facilities with minimal capital outlay. To assist municipalities in their search for cost-effective solutions to their plant capacity dilemma, the U.S. Environmental Protection Agency (U.S. EPA) has undertaken a search for new technologies which may allow for the realization of this goal. One such technology thus far identified is what has been termed high biomass systems.

High biomass systems allow for the growth of both fixed film and suspended biomass in the aeration tanks of activated sludge systems. The suspended growth concentration is controlled by adjusting the amount of MLSS recycled in the underflow from the secondary clarifier. Inert support media is placed in the aeration tanks and serves as the locus for fixed biomass growth. This fixed film growth increases the concentration of biomass in an activated sludge plant's aeration tanks which, in turn, lowers the food to mass ratio (F/M) and increases the solids residence time (SRT) of the system. It also lowers the concentration of biomass being conveyed to secondary clarifiers since the fixed film biomass is retained in the system's aeration tanks. Such retention allows for enhanced levels of solids-liquid separation, especially in hydraulically overloaded clarifiers. In addition to the above, the presence of the fixed film biomass has been shown to allow for enhanced wastewater nitrification as well as to yield a better settling secondary sludge.

CURRENTLY AVAILABLE HIGH BIOMASS SYSTEMS

At the present time, there appear to be at least six (6) high biomass systems commercially available which can be incorporated into conventional aeration tanks. Linde AG of the Federal Republic of Germany (FRG) and Simon-Hartley of Great Britain (Ashbrook-Simon-Hartley in the USA) both use small, highly reticulated sponge pads as their inert support media. Bio-2-Sludge (FRG) and Smith & Loveless, Inc. (USA) use racks of synthetic trickling filter media to effect fixed film growth. Ring Lace (Japan) employs a looped string material as the inert support media and a Chinese firm uses tassels of a synthetic material attached to a string for fixed biomass attachment and growth. The latter two inert support media are tied to racks which are placed in the plant's aeration tanks.

SYSTEM DESCRIPTIONS

Linde AG/Ashbrook-Simon-Hartley

The basic concept of systems offered by these two manufacturers is to use highly reticulated sponges as the inert support media for the fixed film biomass growth. The systems are marketed as Linpor (Linde AG) and as Captor (Simon-Hartley). Linpor sponges are approximately cubical with sides of about 10 to 12 mm whereas Captor sponges are about 12x25x25 mm. The Linpor system includes solids recycle from the secondary clarifier whereas a Captor system normally does not include secondary solids return to the aeration basin. Volume of sponge in Linpor secondary systems approximates 10-30 percent of the aeration tank volume. Biomass accumulation in the sponges may range from 9-37 g/L (of sponge), thus accounting for 25-60 percent of the total biomass in Linpor secondary system aeration tanks. Measurements of the fixed film biomass in Linpor systems have indicated equivalent mixed liquor suspended solids (MLSS) concentrations of 1,200-3,800 mg/L. Linde AG's cost estimate for the conversion of an existing conventional activated sludge plant to a high biomass system is in the range of $\$250\text{-}350/m^3$ of aeration tank volume.

The inert support media in the Linpor system are kept in the aeration tanks by means of retention screens located approximately 0.3 m upstream of the effluent launder as shown in Figure 1. Openings in the retention screen are 8-mm holes placed on 12-mm centers. Forward flow of wastewater through the aeration tank can cause a maldistribution of sponges in the tank, with higher concentrations occurring at the retention screen than at the influent end of the tank. To normalize the concentration, Linde AG has installed air lift pumps to convey sponges from the discharge to the influent end of aeration tanks at installations where concentration maldistributions have occurred.

Ashbrook-Simon-Hartley has developed a process to squeeze fixed biomass from sponges in their systems. An air-lift pump lifts the pads to a conveyor system where they are transported through pressure rollers. This has been viewed as a means of removing excess biomass from the pads and producing concentrated waste secondary sludge solids. The Linde AG system relies on naturally occurring sloughing of fixed biomass from the sponges and wasting excess biomass from the return activated sludge (RAS) system.

Ring Lace

The basic concept of a Ring Lace system is the installation of a polyvinyl chloridene (PVCE) string on racks in an aeration tank as shown in Figure 2. Clear spaces above and below the racks approximate 0.5 m each to allow for normal side roll mixing in the aeration tanks. Each string has numerous attached loops of the

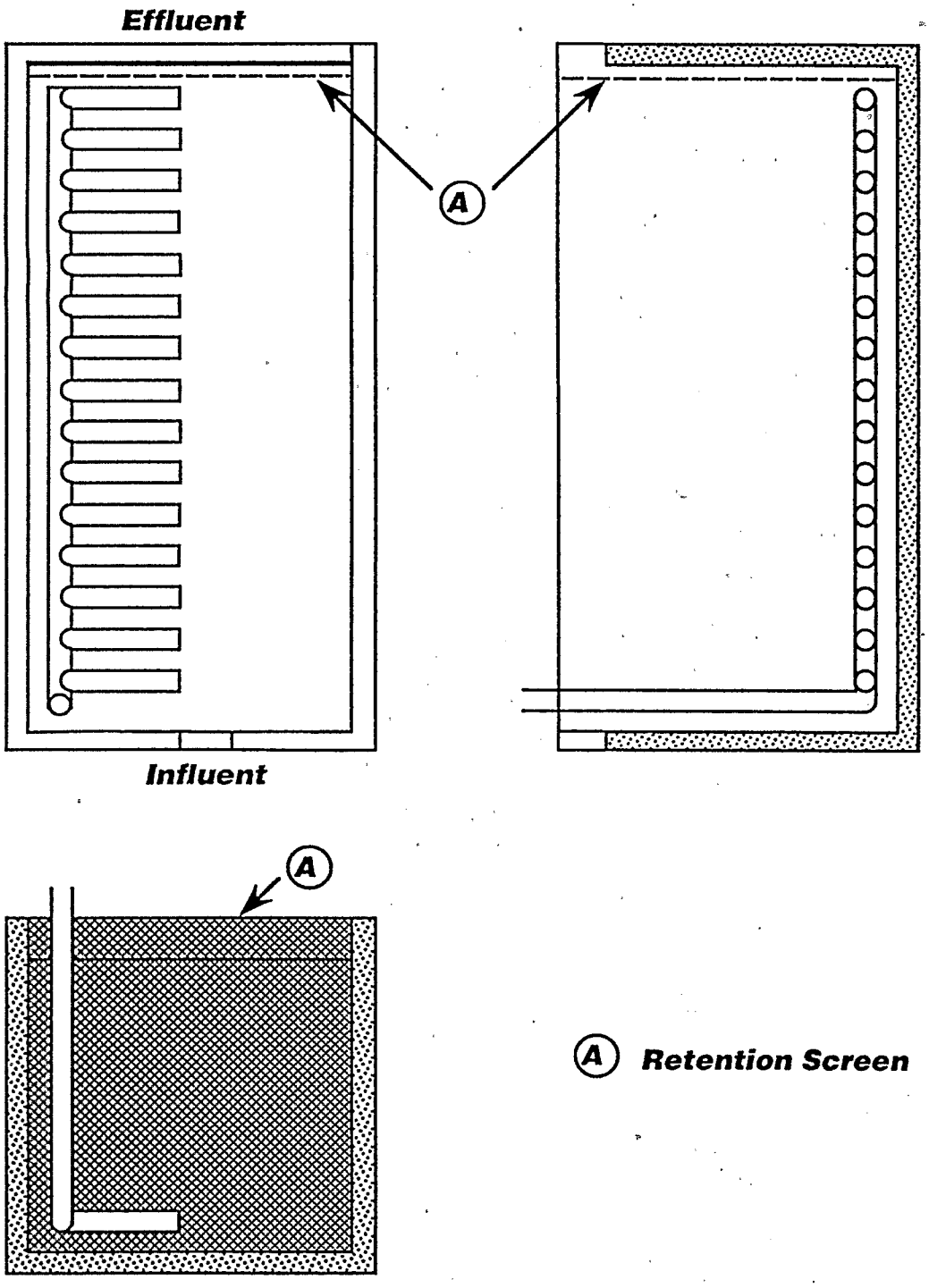

Figure 1. Basic Layout of a Linde AG System.

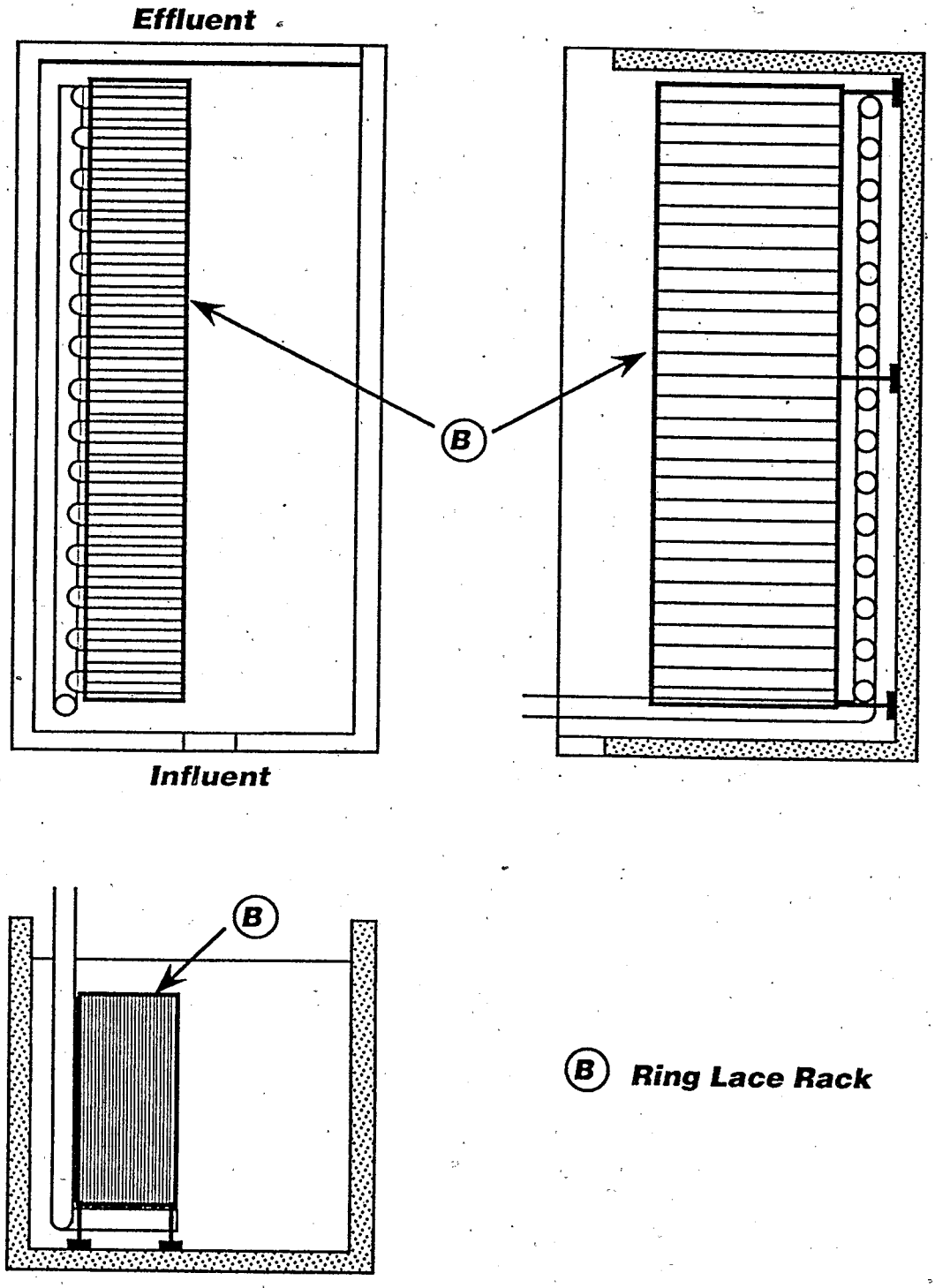

Figure 2. Basic Layout of a Ring Lace System.

same material, thus greatly increasing the surface area available for fixed film growth. The European distributor indicated that typically 25-50 percent of an aeration tank's total volume is occupied by the Ring Lace racks, and the racks are spaced to yield a density in this volume of 120-300 m of Ring Lace/m^3. One distributor in the US indicated that the racks would typically occupy from 50 to 80 percent of the basin volume with a density of about 120 linear meters per m^3 of the entire basin volume. Retrofitting a conventional activated sludge plant in Olching Germany to a Ring Lace system was at a cost of about $2.65/m of Ring Lace material installed.

Fixed film growth on the inert media has been reported to be 6-7 g/m and thus the equivalent MLSS of fixed biomass in a Ring Lace system could vary from roughly 200-1,100 mg/l of total tank volume. As the fixed film biomass continues to grow, natural sloughing occurs. The sloughed material then becomes a part of the MLSS in the system and is removed by conventional wasting practices. The level of suspended MLSS is controlled by the amount of solids recycled from the secondary clarifiers.

Bio-2-Sludge/Smith & Loveless, Inc.

The basic concept of these systems is the inclusion of synthetic trickling filter media to provide the surface on which fixed film biomass growth can occur as shown in Figure 3. Volume of media used in Bio-2-Sludge systems approximates 25 percent of the aeration tank volume while in Smith & Loveless systems, it is close to 75 percent. In contrast to the Bio-2-Sludge systems, the Smith and Loveless system does not recycle settled sludge from the secondary clarifier back to the aeration tank.

Bio-2-Sludge systems use just about any synthetic media as long as the openings in the media are at least 2x2 cm. Typical surface area of the media is in the range of 90-120 m^2/m^3. Racks to hold the media are constructed to provide approximately 0.5 m of clear space above the air diffusion system as well as about 0.5 m of clear space between the top of the racks and the liquid surface in the aeration tanks. This arrangement allows for normal side roll mixing in the tank. If the media begins to plug from excess fixed film growth, a common practice is to turn on an additional blower to induce surplus biomass sloughing. Retrofit costs for Bio-2-Sludge systems have been reported to be in the vicinity of $300-350/$m^3$ of media installed.

Chinese system

As mentioned above the Chinese system uses tassels of a synthetic material which are tied to strings and then mounted in the aeration tanks. At the present time, there is no additional information available on this system.

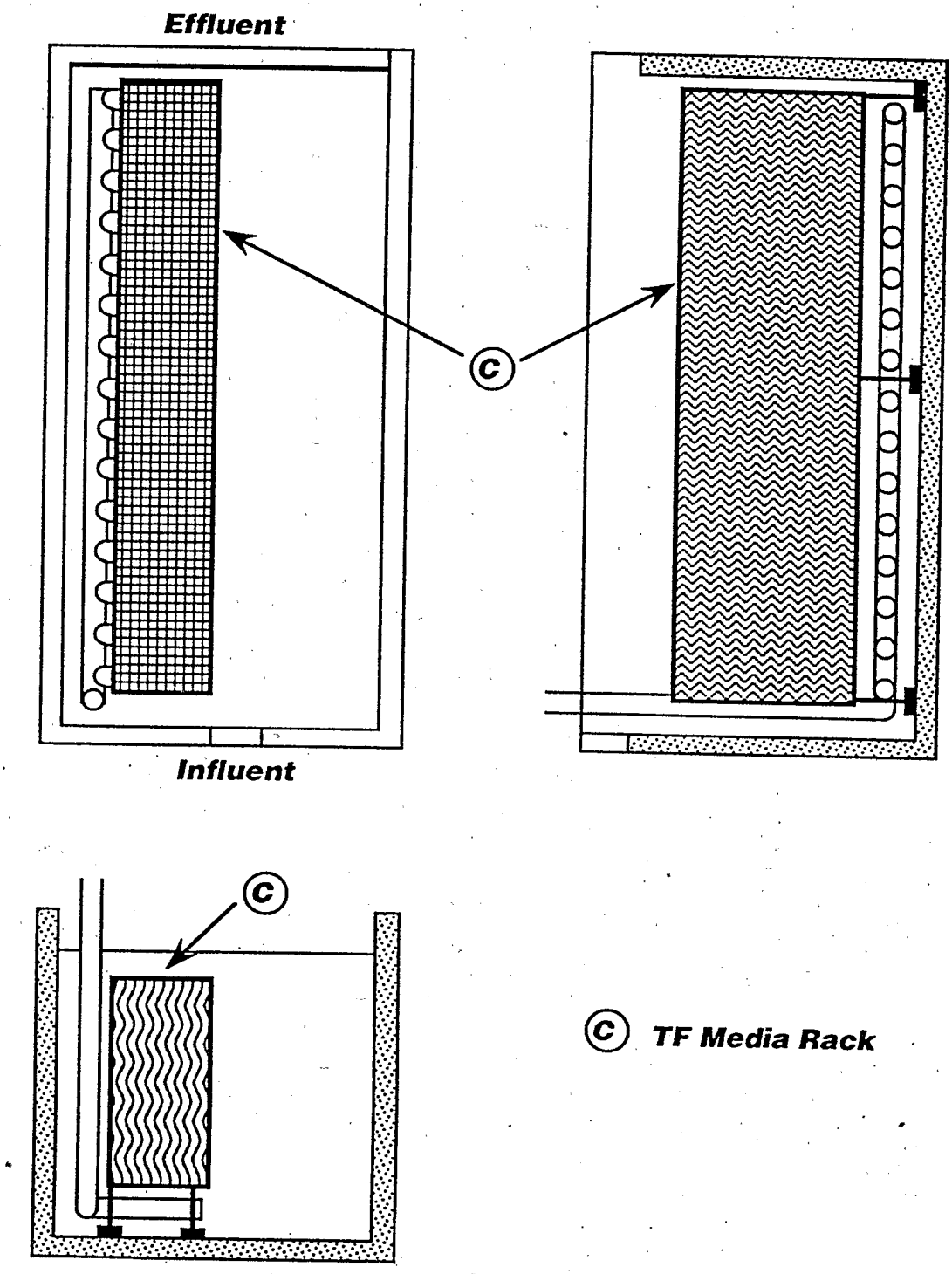

Figure 3. Basic Layout of a Bio-2-Sludge System.

SITE VISITS

During the late spring of 1988, a U.S. EPA evaluation team visited a number of full scale high biomass systems in the Federal Republic of Germany. The purpose of the site visits was to view the systems in operation, collect operational and performance data, learn of system design details from the system manufacturers, and discuss process operational and maintenance concerns with the treatment plant staffs.

Facility locations and types visited included, among others, those presented in Table 1.

Table 1. Site Visit Basic Information.

Type of System	Location
Linde AG	Freising Munich
Ring Lace	Olching
Bio-2-Sludge	Schomberg Calw/Hirsau

Summarized below are data and information on the various types of high biomass systems from certain of the above facilities.

Freising (Linde AG System)

The Freising activated sludge wastewater treatment plant was converted to a Linde AG high biomass system in 1984 after a series of pilot plant studies. It appears that there were three primary reasons for the conversion: frequently occurring poor sludge settleability, limited space at the plant site and higher costs of alternative technologies. Operational and performance data were collected for a short period of time before and after conversion and these are summarized in Table 2.

Before conversion to the high biomass system, which employed a sponge volume equal to 20 percent of the aeration tank volume, the plant could only maintain a MLSS concentration of about 2,600 mg/L. Following conversion, a much higher MLSS concentration could be maintained which dramatically lowered the F/M of the system and greatly improved the secondary sludge settleability.

As a follow-up to this historic information, data for calendar year 1987 were collected and analyzed. A summary of this information is presented in Table 3.

Table 2. Historic Freising Data.

Parameter	Conversion Before	Conversion After
Solids, g/L		
Suspended	2.6	5.1
Fixed	---	1.7
Total	2.6	6.8
F/M, kg/kg-d	0.47	0.17
SVI*, ml/g	485	85
Effluent BOD5, mg/L	48	4

*Sludge Volume Index

Table 3. Freising Data Averages for 1987.

Parameter		Value
Secondary Influent		
Flow, m^3/d		10,100
BOD5, mg/L		273
Secondary Effluent		
BOD5, mg/L		5
Biomass, g/L		
Suspended		2.59
Fixed		3.75
Total		6.34
Aeration Tank		
Det'n Time, hr		5.5
F/M, kg/kg-d		0.19
SVI,* ml/g		114
D.O.*, mg/L	1.7	
Aeration Rate		
m^3/kg BOD5 Removed		64
Secondary Clarification		
SOR**, m/hr		0.34
R/Q***, %		41

*Dissolved Oxygen
**Surface Overflow Rate
***Ratio of return/influent flow

In 1987, the Freising plant was operated at 78 percent of its hydraulic capacity and 80 percent of its BOD5 capacity. Instantaneous influent pH, because of industrial discharges to the

system, ranged from 6.2- 12.0; effluent pH ranged from 6.8-7.5. An average of 65 percent nitrification was also achieved over a wastewater temperature range of 10-17 C, even though the D.O. concentration averaged only 1.7 mg/L.

Munich (Linde AG System)

The Munich Grosslappen plant was retrofitted with the Linde AG high biomass system to allow for additional treatment while a parallel activated sludge plant of equal capacity is being constructed. Sponge media were also installed to allow for full scale evaluation of the process for possible inclusion in the new treatment plant. The existing Munich plant has three banks of aeration tanks and during the site visit, two of the banks had varying quantities of sponges in place. Data from the plant are summarized in Table 4.

Table 4. Munich Data for 1987.

Parameter	Sponge Volume in Tanks, %		
	0	10	30
Secondary Influent			
Flow, m^3/d	165,100	165,060	145,800
BOD5, mg/L	186	183	183
Secondary Effluent			
BOD5, mg/L	21	18	14
TSS, mg/L	20	16	13
Biomass, g/L			
Suspended	2.81	2.90	2.77
Fixed	--	1.20	1.84
Total	2.81	4.10	4.60
Aeration Tank			
Det'n Time, hr	1.91	1.90	2.15
F/M, kg/kg-d	0.84	0.57	0.46
kg BOD5/m^3-d	2.39	2.34	2.11
Sludge Age, days	1.3	2.0	2.3
SVI, ml/g	145	95	90
D.O., mg/L	2.1	2.2	2.2
Aeration Rate			
m^3 Air/m^3 WW	5.4	6.1	8.5
m^3/kg BOD5 Removed	33	37	50
Secondary Clarification			
SOR*, m/hr	1.08	1.08	0.95
SLR*, kg/m^2-hr	3.05	3.15	2.82
R/Q, %	76	73	49
RAS, g/L	6.24	6.86	6.94
Waste Activated Sludge			
kg/kg BOD5 Removed	1.08	1.02	1.03

*Surface Loading Rate

Of interest to note with the Munich data are: 1) the lowering of the system's F/M by the presence of the fixed film biomass, and 2) the required increase in aeration rate to address the demand of additional biomass in the system.

Olching (Ring Lace System)

During the spring of 1988, the Olching plant was operating at 85 percent of its design hydraulic capacity of 35,000 m^3/day and 83 percent of its design organic loading which was based on a population equivalent of 240,000. In late 1987, conversion of the plant to a high biomass system began by adding a new 4,000 m^3 denitrification basin and also by adding 252,000 m of Ring Lace material to each of four 2,070 m^3 aeration tanks. The lead denitrification basin, which contains no Ring Lace and had not yet been placed in operation at the time of the site visit, is equipped with paddle mixers and also contains aeration equipment for nitrification if necessary. A portion of the mixed liquor from the Ring Lace basins will eventually be recycled to the lead denitrification basin. At the time of the site visit, influent flow was sent directly to the aeration basins containing the Ring Lace.

The Ring Lace material was strung on racks, with the individual strands being separated by approximately 50 mm. Spacing between the racks, which occupied 31 percent of the aeration tank volume, was about 60 mm. A tensioning system was built into the racks in case the Ring Lace material elongates with time. In addition to the above, the equipment supplier had to provide a 10-year process performance and equipment guarantee. Required process performance was based on the plant's discharge requirements of 15 mg/L BOD5, 20 mg/L TSS and 10 mg/L NH4-N, the latter equating to an approximate 75 percent level of nitrification.

Because the Olching Ring Lace plant had been in operation for less than one year and the required performance test had not been completed, no operation or performance data were released to the U.S. EPA evaluation team. However, the following general observations on the Ring Lace system were communicated from the treatment plant staff.

- Before conversion to the high biomass system, the maximum operational MLSS concentration that could be achieved was about 1,500 mg/L, which resulted in the plant operating at a F/M of 0.6-0.70 kg BOD5/kg MLSS-day. At this loading rate the MLSS were gray in color and had very poor settling characteristics. In addition, no substantial nitrification could be achieved.

- After conversion, the suspended biomass varied from 3,500 to 4,500 mg/l. The fixed biomass on the Ring Lace material was estimated at 6.5 g/m which is equivalent to an additional 790 mg/l of equivalent basin MLSS. This

increase resulted in a MLSS with a rich brown color that had greatly improved settling characteristics. Operating F/M of the system decreased to about 0.2 kg/kg-day and desired levels of nitrification were being achieved.

- Other high biomass systems were evaluated at the pilot plant level by the Olching treatment plant staff and they all yielded effluent qualities equivalent to that realized by the Ring Lace system. However, the Linde AG system was not selected because the staff felt that this system might require additional electrical power for proper operation, the sponge cubes would be subject to wear by abrasion and, upon extended levels of mineralization, the sponge cubes might have a tendency to settle in the aeration tanks. The Bio-2-Sludge process was not selected because the plant staff felt that the synthetic trickling filter media might be subject to plugging by biomass and/or large particulates, either of which might lead to anaerobic zones in the media. Such zones, if they developed, were felt to potentially limit the nitrification process.

Schomberg (Bio-2-Sludge System)

The Schomberg wastewater treatment plant was recently expanded in throughput capacity and simultaneously converted to a Bio-2-Sludge high biomass system. Both aeration tank volume and final clarifier surface area were more than doubled and a denitrification section was added to the aeration tankage. In the spring of 1988, the plant was operating at 20 percent of its design hydraulic capacity and 57 percent of its design BOD5 capacity.

Prior to conversion, the plant was neither hydraulically nor organically overloaded, but could not meet discharge permit requirements primarily because of very poor settling MLSS. The poor settling characteristics would allow the maintenance of only 1,000-1,500 mg/L in the system's aeration tanks in spite of the very high RAS pumping rate employed. After conversion, overall plant performance greatly increased as indicated in Table 5.

Installation of the Bio-2-Sludge system at Schomberg along with the other plant modifications greatly enhanced MLSS settleability as evidenced by the SVI of 82 ml/g after conversion. Although the RAS rate at this plant could have been adjusted to about 50 percent, a return rate of 100 percent was used. It appears that this is a common design/operational practice at smaller treatment plants in certain areas of Germany. Of interest to note is that this plant, in addition to achieving an effluent containing 5 mg/L BOD5, realized nearly complete nitrification.

Table 5. Schomberg Data Averages for 1987.

Parameter	Value
Secondary Influent	
Flow, m^3/day	3,161
BOD5, mg/L	120
NH4-N, mg/L (Approximate)	40
Secondary Effluent	
BOD5, mg/L	5
NH4-N, mg/L	0.4
Biomass, g/L	
Suspended	3.96
Fixed (Estimated)	1.74
Total	5.70
Aeration Tank	
Detention Time, hr	6.7
F/M, kg/kg-d	0.08
Kg BOD5/m^3-d	0.45
SVI, ml/g	82
D.O., mg/L	3.2
Secondary Clarification	
SOR, m/hr	0.18
SLR, kg/m^2-hr	1.42
R/Q, %	100

SUMMARY AND CONCLUSIONS

A number of high biomass systems have been installed at wastewater treatment plants throughout the Federal Republic of Germany. These installations were designed to effect improved effluent quality and it appears that this goal is being realized.

Reasons for selecting high biomass systems over construction of additional aeration tanks and clarifiers (or other secondary treatment processes) include reduced space requirements, increased process stability and capital/operating cost savings.

High biomass systems call for installation of supplemental equipment over that contained in a conventional activated sludge plant. More installed equipment generally implies more maintenance and, to a certain extent, this occurs with some of the high biomass systems. In addition, the presence of both suspended and fixed biomass forms and higher biomass concentrations may require a certain level of additional operator time to achieve optimum system performance.

The presence of inert support media and higher biomass concentrations in these systems can increase overall power consumption. To achieve desired mixing patterns and velocities in retrofitted aeration tanks, power input may have to be increased to overcome additional headloss brought about by the installation of the inert support media. Due to endogenous respiration rate alone, the presence of additional biomass will increase system oxygen requirements which, in turn, will require additional power input. In addition, high biomass systems generally yield lower effluent BOD5 concentrations and/or higher levels of nitrification which also can effect overall power consumption. Such factors should be addressed when analyzing high biomass system operating costs.

Certain system design limitations have been identified in the past, but many of these have been corrected by the system manufacturers and/or operators. For example, influent hydraulic surges at Linde AG plants have caused blinding of the retention screens by the sponge media. This has been partially corrected by increasing the pumping rate of the sponge return system. Also, abrasion of the sponge media was stated to result in losses of less than 5 percent per year, but there does not appear to be a corrective action for this system limitation beyond adding new sponges as required. In Ring Lace systems, a question on the extent of media stretching over time persists. Construction of self-tensioning media racks may address this potential limitation. Plugging of the synthetic trickling filter media in Bio-2-Sludge systems has been a concern of certain individuals. It appears that turning on an additional air blower periodically for a short period of time will induce additional sloughing of the fixed film biomass, thus negating this potential problem. There were other potential problems areas communicated to the U.S. EPA evaluation team, but their relative importance was deemed minor because solutions had, for the most part, already been developed and field tested.

Cost of retrofitting a 3,785 m^3/day activated sludge plant with high biomass system equipment appears to be $77,000-260,000, depending on the process selected. These estimated costs are based on an activated sludge plant with six hours of contact time in the system's aeration tanks.

Additional throughput capacity realized by conversion to a high biomass system cannot be ascertained from the data and information thus far amassed. However, it appears conversion can allow for at least the doubling of the biomass concentration in a system and for reducing by at least one-third the required RAS pumping rate compared to a conventional activated sludge plant. These two factors alone indicate the potential for increased throughput capacity.

Actual capacity increase which can be realized by conversion to a high biomass system should be established from the conduct of pilot plant studies. Representatives for both Bio-2-Sludge and Linde AG stated that to obtain accurate design and operational information, a pilot plant should have an aeration tank with a volume of at least 100 m^3.

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EXPERIENCE WITH INTRACHANNEL CLARIFIERS IN THE U.S.A.

by

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INTRODUCTION

The modern oxidation ditch (OD) which is depicted in Figure 1, is a high-sludge-age variation of activated sludge or suspended growth biological treatment system which emanates from the work of Dr. Pasveer of The Netherlands in 1954. Since the time of its introduction in 1973 more than 9200 such plants have been constructed in the U.S., and are widely perceived as the state-of-the-art treatment in small-to medium- sized communities (populations/capacities of 5,000 to 50,000 (0.5 to 5.0 million gallons per day (MGD)). This status has been attained owing to the OD's simplicity of operation, reliability of performance and overall cost-effectiveness.

Intrachannel clarifiers (ICCs) as depicted in Figure 2, were introduced during the 1980's to replace the conventional external clarifiers which have been commonly employed for OD facilities in the U.S. These systems are marketed by a variety of commercial entities and are of a variety of configurations (1,2), but in general are physically placed in the ditch and do not employ return sludge thickening and pumping. Consequently, ICC purveyors have claimed:

1. Lower capital costs owing to elimination of return activated sludge (RAS) pumping and piping facilities

2. Lower operation and maintenance (O&M) costs owing to the elimination of RAS facilities and external clarifier control problems

3. Reduced land area requirements by elimination of separate aeration and sedimentation structures.

These claimed advantages served the purveyors well during the early 1980's as an EPA-sponsored study (2) in 1987 revealed approximately 54 systems in design, construction or operation, of which 26 were in the last category. The majority of these systems were approved under the innovative technology provisions of the USEPA Construction Grants Program whereby they received an increased Federal grant with potential guarantees of performance. In fact, the ICC systems were the second most popular technology approved under the innovative technology provisions, second only to ultraviolet light disinfection.

The most widely employed version of an ICC is known as a BMTS™ system. Descriptions of this ICC can be found in references 1 and 2. For the purpose of this discussion, it is sufficient to describe the BMTS system as one which has generally been field constructed integral with the ditch structure, as shown in Figure 3. The second and almost equally popular ICC system is the BOAT™ clarifier. It is described in more detail in references 1 and 2. The BOAT, as opposed to the BMTS, is factory fabricated and delivered to the site for installation in a more conventional OD geometry, and is schematically displayed in Figure 4. Since these two types of ICC clarifiers constitute 46 of the 54 systems noted in 1987, much of the remainder of this discussion deals with these two systems as subcategories of ICC technology.

With any degree of popularity a proportionate degree of attention is naturally drawn, in this case in the form of investigative studies and reports. Early literature citations were generally from the commercial purveyors and already committed clients and were quite positive in their nature regarding the costs and perceived benefits of the technology. Several articles appeared (3,4,5,6) which provided some performance and comparative cost data, but the circumstances and assumptions for the data presented were generally undefined. For example, Parsons (5) described a Florida BOAT system which was estimated to provide 20 percent capital cost savings over an external clarifier. Some articles (3,4) published by BMTS advocates addressed the results of short-term studies which were subsequently adopted as full-scale design criteria and the basis for cost savings and other claims which provided the stimulus for system adoption.

USEPA STUDIES

The USEPA had participated in some of the conceptual and development studies which led to the widespread adoption of the ICC concept in its role of assisting new technological ideas. However, the nature of these developmental projects merely identified a few crucial areas of concern, but failed to quantify the extent of potential problems in full-scale application. Likewise an early attempt to generically assess the ICC technology was abandoned owing to difficulties encountered in providing an unbiased assessment without full-scale information on systems which approached the ICC concept from significantly different directions. In a 1984 USEPA Seminar Series an attempt was made to assess the status of development of the commercially available ICC options and to restate generic advantages and disadvantages of the generic technology (7). The advantages were cited to be:

1. Elimination of separate clarifiers
2. Elimination of sludge return pumps
3. Reduced O&M associated with #1 and #2
4. Elimination of sludge septicity problems
5. Reduced capital and O/M costs.

Disadvantages were listed as:

1. Minimal control over sludge blanket depths and return sludge flows
2. Less operational flexibility
3. Difficulty in adding flocculant aids
4. Generally unthickened MLSS wastage
5. Unknown response to high peak flows
6. Unknown control of bulking sludges
7. Unknown scum control effectiveness.

In a 1986 USEPA Seminar Series on ICC technology, an assessment based several short-term site visits was presented (8). Since most installations had been operational for only a short period, the findings were presented in a manner designed to identify potential problems which designers and operators might minimize through awareness. The most serious problem identified was the inadequate mixing and aeration resulting in below critical (1.0 foot/sec.) horizontal velocities and inadequate dissolved oxygen in the mixed liquor. The latter problem was correctable, since the total aeration capacity should be the same for an ICC systems as it is for a conventional OD. The other common problem identified was the general undersizing of sludge handling facilities, e.g., thickeners and drying beds. Other manufacturer-specific problems noted were a need for better scum removal for ICCs which completely baffle the ditch water surface, e.g., BMTS, and the need to improve the design understanding of propeller mixers which impart horizontal velocity in systems which separate this function from aeration.

With the advent of significant population of operating ICC systems a more thorough technology evaluation was initiated in 1987. This evaluation involved the development of a master list of available information from the operating facilities. From this list eleven systems were selected for one-day site visits to collect more extensive information and to assess O&M and laboratory procedures, competence, etc. The purpose of this evaluation was to evaluate the true comparative costs of ICC systems, the design tradeoffs identified by Bender (1), and the relationships between these factors.

RESULTS

Costs

The initial effort on cost comparison dealt with the cost-effective analyses supplied by the designers/municipalities to justify their choice of an ICC system over an OD design. In the Construction Grants regulations a 15 percent cost savings based on life cycle costs or a 20 percent net energy savings is generally required to justify designation as a innovative technology. For a list of eleven projects, comprised of facilities employing only ICC equipment from the two major ICC purveyors, capital cost savings of from 4 to 45 percent were claimed by the engineers. Interpretation of these data was impossible because of the unequal components. For example, in some cases a 10- to 20-hour ICC aeration time was compared to a 24-hour aeration time in the OD, in others no additional cost for the handling of MLSS vs thickened sludge from a conventional clarifier was provided, several others used lower aeration/mixing costs for the ICC than the OD despite the fact that the ICC creates additional headloss in the channel, and one even omitted preliminary treatment in the ICC.

Even with bid tabulations from 26 projects indexed to a common date the widespread practice of package bidding and lumping together of individual contractual subdivisions made quantitative comparisons impossible. Generally, bid prices (indexed to the second quarter of 1987) indicated that small (<0.5 MGD) ICC-based total treatment systems varied between $2.00 and $5.00/gallon of daily capacity, while larger (>2.0 MGD) systems varied from $1.00 to $1.50/gallon. Between 0.5 and 2.0 MGD the scatter reflects the entire spectrum from $1.00 to $5.00/gallon.

In order to attempt to compare the real capital costs of the two technologies it was decided to perform quantity takeoffs and detailed cost estimates from some ICC system designs and replace those components with conventional OD components using accepted construction guides supplemented by vendor-installed prices. This technique was applied to three BMTS plants with capacities from 0.8 to 1.9 MGD and resulted in the construction costs of conventional OD being consistently less expensive by 5 to 17 percent. A similar analysis for two BOAT systems (0.25 and 0.88 MGD) reveals an even greater ICC plant cost. However, comparison of bid tabulations on two recent projects with nominal plant capacities from 2.0 and 2.4 MGD showed the capital costs of BOAT systems to be less than a conventional OD by only 3 to 5 percent. Although, there is inconsistency in these data, they do indicate that the capital cost of properly designed ICC systems can be expected to be similar to a conventional OD system.

Performance

Performance of operating ICC facilities was compared to previously analyzed OD facilities to roughly assess whether significant differences in BOD_5 and TSS were apparent. The ICC data were based on the 17 plants which were apparently operating at a steady state. The OD data were based on the performance of 29 facilities (8). The data shown on Figures 5 and 6 indicate that performance is generally comparable, with slightly higher effluent concentrations likely attributable to the newness of the ICC technology in terms of developing operational staff understanding and problems owing to first-generation design shortcomings in sludge thickening and aeration/mixing systems. Subsequent analyses of components provide insight into the latter problems. Although clarifier loadings for both sets of data are comparable, e.g., actual/design clarifier overflow rates (gal/sq ft-day) are 357/485 for the OD plants and 357/500 for the ICC plants, influent wastewater characteristics are not documented. Therefore, the above comparisons are only qualitative, but lead one to the conclusion that design overflow rates should be made at least equal to those of external clarifiers.

Design Tradeoffs

Bender (1) has previously identified the key design tradeoffs which should be quantified to fully evaluate the ICC technology against the conventional OD alternative. These key tradeoffs were identified as:

- Waste sludge concentration - Must the ICC system provide external thickening to compensate for the internal thickening provided in an external clarifier to be fully comparable?
- Restriction in aeration channel flow - How much additional mixing power must be employed to compensate for the increased headloss of the ICC placed in the channel?
- Aeration channel and clarifier maintenance - Owing to the interdependence of the ICC and the aeration channel, will increased redundancy in design (and cost) be required to satisfy effluent permit requirements while performing major O/M tasks?
- Operational flexibility - Without sludge return facilities, how can the O&M staff monitor control and chemically treat (as necessary) return sludges to control the activated sludge system?

The present study was initiated to provide some field experience for the purpose of quantifying these issues in addition to analyzing costs and performance information.

Waste Sludge - -

Early promotional literature (3,4) dismissed this issue by asserting that the required surface area of a thickener is determined only by the mass of solids entering the unit, and was therefore identical for the MLSS from an ICC system and the partially thickened (1 to 1.5 percent solids) underflow from an external clarifier. Although other references (9,10,11) generally confirm this analysis for determining surface area of the thickener, they also note limitations on hydraulic loading and on thickener volume which would indeed equate to additional construction costs for treatment of significantly more dilute suspensions. Also, in cases where sludge is removed from the plant directly without thickening, additional costs will be associated with ICC systems owing to increased sludge volume.

This problem, i.e., inadequate sludge thickening capacity, was noted as a significant performance limitation at 3 of 11 facilities visited and was reported as a significant problem at several of the other plants contacted by telephone with subsequent information submission. At one ICC facility visited, the engineer increased thickener capacity such that with the added capital cost of thickening the ICC was 67 percent more costly than for the external clarifier alternative. At the time of the visit the enlarged thickener was receiving 2,000 to 4,000 mg/L MLSS and yielding an underflow of 2 percent solids.

The above information, when combined with the fact that the average ICC plant was being loaded at 53 percent of design organic loading, provides qualitative support for Bender's concern over the relative thickening requirements for thin MLSS wasted directly from the ditch or from the ICC with minimal thickening. Additional analysis of this design issue is underway.

Channel Headloss -

Of all the issues investigated in the present study, this one appears to be the most serious due both to the fact that manufacturers tend to dispute that there is a problem and the general uncertainties on how to properly address the issue. Actual channel headlosses have not been adequately characterized. In systems which employ propeller mixers and diffused air aeration the wall of rising bubbles also represents an additional

headloss on which there is little quantitative data. Aeration energy required for an ICC or conventional OD should be the same for the same detention time and operating conditions, but the horizontal-shaft mechanical aeration systems commonly employed, i.e., surface rotors and disks, provide both mixing and aeration in a single mechanical function, making total power requirements to reach the required operating conditions the only means of comparison between the two applications. The newer propeller mixer/diffused aeration designs theoretically permit direct analysis of each of the two functions.

Six of the ICC plants investigated in this study which employed horizontal-shaft mechanical aeration had aerator energy data. The total power required in these six plants varied from 6.2 to 91.6 W/m^3 of channel volume. Only one plant was in steady-state operation with reasonable (75% of design loading) organic loading, and it required 12.2 W/m^3. Ettlich (8) suggests that 10 to 25 W/m^3 is required for ODs to maintain a proper D.O. (1.0-2.0 mg/L) and V min (1.0 ft/sec), while Hartley (12) estimates 5 to 10 W/m^3 with full baffling. Measurement of total energy use by 30 O.D. systems in the field yielded a range of 4 to 78 W/m^3, which reduced to a range of 6 to 30 W/m^3 in the 11 plants which had reasonable organic and hydraulic loading (13). Koot and Zeper (14) performed a theoretical hydraulic analysis of OD systems using the Chezy formula and estimated that 5.0 to 7.5 W/m^3 are required to provide 1.0 ft/sec velocity and adequate aeration. In comparing the above operating ICC facility with these ranges, no quantitative determination is possible, especially since the average channel velocity in this plant has been measured to be as low as 0.7 to 0.8 ft/sec and the larger power-use system operated at 82 hours of detention time.

However, the placement of an ICC in the ditch/channel will undoubtedly increase headloss and, therefore, the power required to maintain any specified velocity. One promotional report for the BOAT unit describes studies showing a 15-30% velocity reduction (15) by their own measurement based on average velocities in the channel with and without the ICC. Using the manufacturer's figures and the horizontal-shaft aerator relationship (12) between power input and average velocity, i.e., average velocity varies as the 0.55 power of the power per unit volume, the increase in power required to overcome a 15 to 30% average velocity loss is about 25 to 50% of the effective power. Therefore, to restore average channel velocity to the original level, approximately 25 to 50% more power must be applied to the horizontal shaft aeration/mixing system. Since the effects of these constrictions on oxygen transfer are unknown, this increase is ascribed only to maintaining required channel velocity.

The theoretical analysis of Koot and Zeper indicated that mixing, i.e., maintaining a velocity of 1.0 ft/sec, was approximately 10 percent (0.3-0.4 W/m^3) of the total energy required for shallow OD systems and about 1 percent (0.1 W/m^3) for deep, short submerged turbine systems, e.g., the Carrousel™ system. Analysis of the operating ICC facilities employing propeller/mixer these systems yielded only five facilities operating at reasonable organic loadings (0.3-1.3 of design loading) with channel detention times between 17 and 29 hours, i.e, reasonably conventional OD systems. The mixing power demand for these five facilities varied from 4.6 to 29.6 W/m^3, or about 10 to almost 100 times the theoretical mixing requirement computed by Koot and Zeper. The ratio of mixing power to total power demand varied from 23 to 71 percent, and total power/unit volume varied from 18 to 54 W/m^3.

Even in the early promotional literature on ICC systems inadequate concern over minimum velocities in the presence of large channel headlosses was apparent. Average channel velocities were shown to meet the 1.0 ft/sec standard and the presence of 0.5 ft/sec velocities at a certain point in the channel was acknowledged and dismissed (4). Attainment of 1.0 ft/sec velocity in the ditch prior to insertion of an ICC has also been promoted as a proper test of adequate mixing (6,15). Field experience at some ICC installations has shown the value of adjustable mixing systems not only in adapting to different operating conditions, but also in providing periodic, e.g., once per week, scouring velocities of at least 1.3 ft/sec to clear settled solids from the bottom of the ditches in ICC systems which have locations where local velocities are significantly less than 1.0 ft/sec. Numerous problems from poor distribution of velocities within the channel have been reported by the operating ICC facilities.

Clearly, the increased headloss in the OD channel due to the presence of the ICC has substantially increased the power requirement for maintaining adequate channel velocities to prevent biological solids deposition. However, more data will be needed to fully quantify this effect. Since the original claim of energy savings was based on RAS pumping elimination, the above power increase must be compared against the power saved by elimination of RAS pumping. Middlebrooks, et al, (16) indicate that the power required for return and waste activated sludge pumping and conventional clarifier sludge collection are about 3 percent of aeration requirements for 1-MGD extended aeration systems. Also, Wesner, et al (17) indicate that RAS pumping and extended clarifier mechanisms require approximately 6 percent of aeration energy requirements. Therefore, energy savings claimed for ICC systems are apparently non-existent or extremely minor in nature.

Additional O&M Problems

Although several unanticipated O&M problems were discovered with ICC systems, many were due to design deficiencies. In addition to those items described above, preliminary treatment deficiencies were abundantly apparent in most facilities employing propeller mixers. Early systems employed a conventional propeller design which became fouled with rags and debris and resulted in power demand increases, vibrational imbalances and frequent removal and cleaning efforts by the O&M staff. Although replacement with "weedless" blade designs has reduced this problem, the use of conventional OD preliminary treatment, e.g., coarse screens and comminutors, has resulted in scum removal and orifice/baffle opening plugging problems with certain ICC designs. Improved preliminary treatment with fine screens with approximately 0.5-inch openings would be beneficial in reducing high O&M staff requirements for existing systems and preventing these problems in new facilities. Also, the need to provide good flow distribution across the channel crossection has become increasingly apparent. Although problems due to this inadequacy have been proven soluble, systematic prevention in the form of design criteria are still forthcoming.

Other design deficiencies in ICC systems which have resulted in excessive O&M demands include:

1. ICC bottom baffle spacing in BMTS systems - The value of wider spacing is under investigation at the largest BMTS system. A frequent problem found at these facilities has been the buildup of solids on top of the bottom baffles. This buildup apparently results in an inordinate percentage of the MLSS being accumulated in the ICC instead of freely returning to the channel. These conditions cause an apparent reduction in effective solids retention time (SRT) in the channel and result in deteriorating system performance until corrected.

2. Effluent orifice pipe design in BMTS - The outlet orifices of the BMTS system are submerged and can result in significant water level changes, e.g., a 9-inch difference was recorded at one small BMTS plant. This water level swing can make slotted scum trough removal systems ineffective and increase O&M labor requirements. This problem can be solved by designing auxiliary weirs in the effluent collection box or by modifying scum collection troughs for a varying water level.

3. Placement and design of propeller mixers and diffusers - The performance of propeller mixing systems in ICC systems has been a learning experience for both mixer suppliers and design engineers. The headlosses represented by the ICC systems and the banks of diffusers must be accounted for in determining power requirements to attain a 1.0 ft/sec minimum ditch velocity. For systems which vary ditch dimensions such that velocities vary over a wide range, the deposition of solids has been controlled through periodic higher ditch scour velocities.

SUMMARY AND CONCLUSIONS

ICC systems appear to be a viable alternative to a conventional oxidation ditch. Technologically, the ICC concept can provide secondary treatment if the same overflow rates are employed as are used in the design of external clarifiers. Early problems with ICC systems appear to be amenable to improved design procedures which provide better preliminary treatment to minimize rags and debris, proper mixing to insure a maintenance of at least a 1.0 ft/sec minimum velocity, and adequate sludge handling and thickening capacities. Cost savings are not inherent with ICC designs, so the relative advantages and disadvantages of this technology must be thoroughly analyzed for each potential application.

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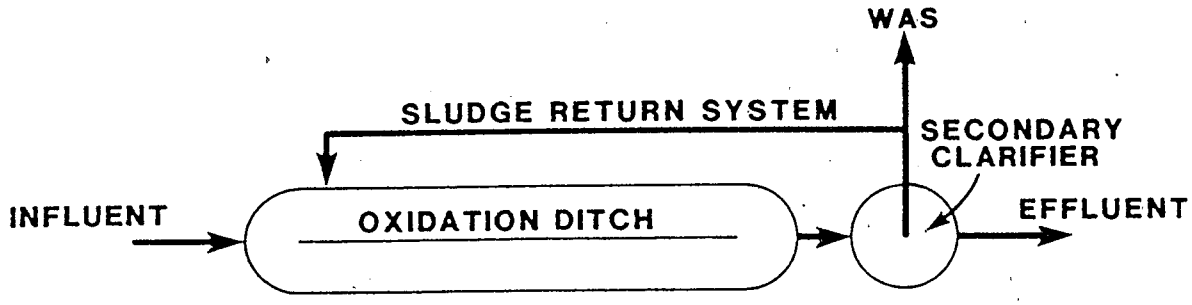


Figure 1. CONVENTIONAL OXIDATION DITCH SYSTEM

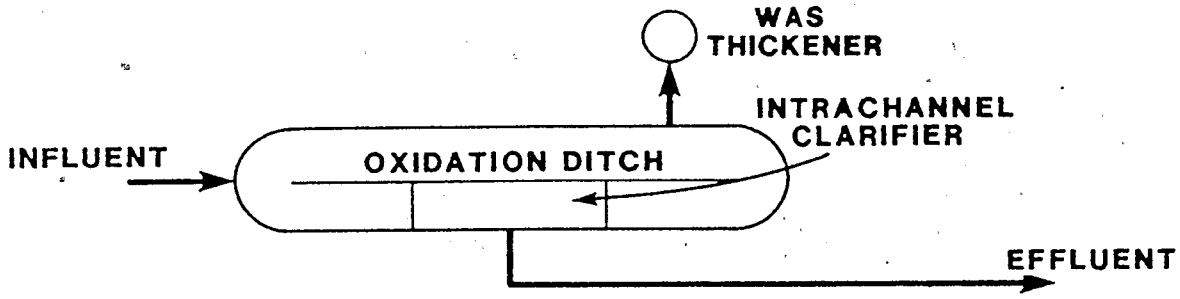
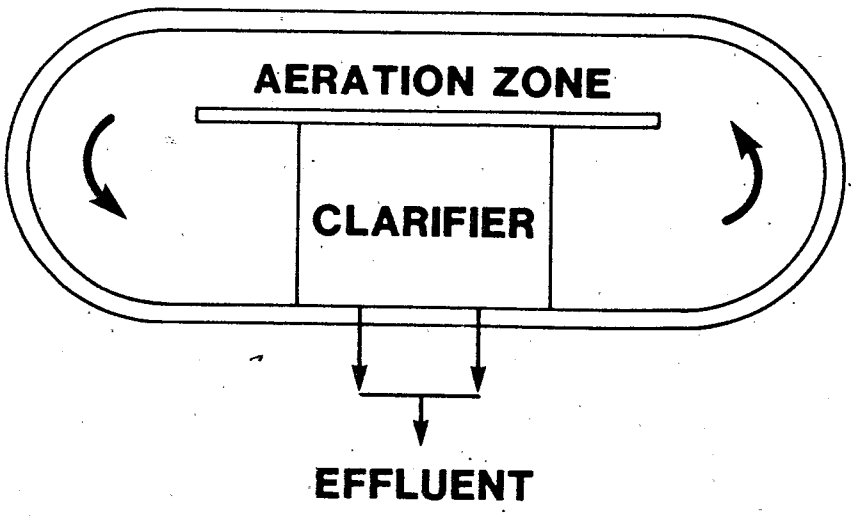


Figure 2. OXIDATION DITCH WITH INTRACHANNEL CLARIFIER



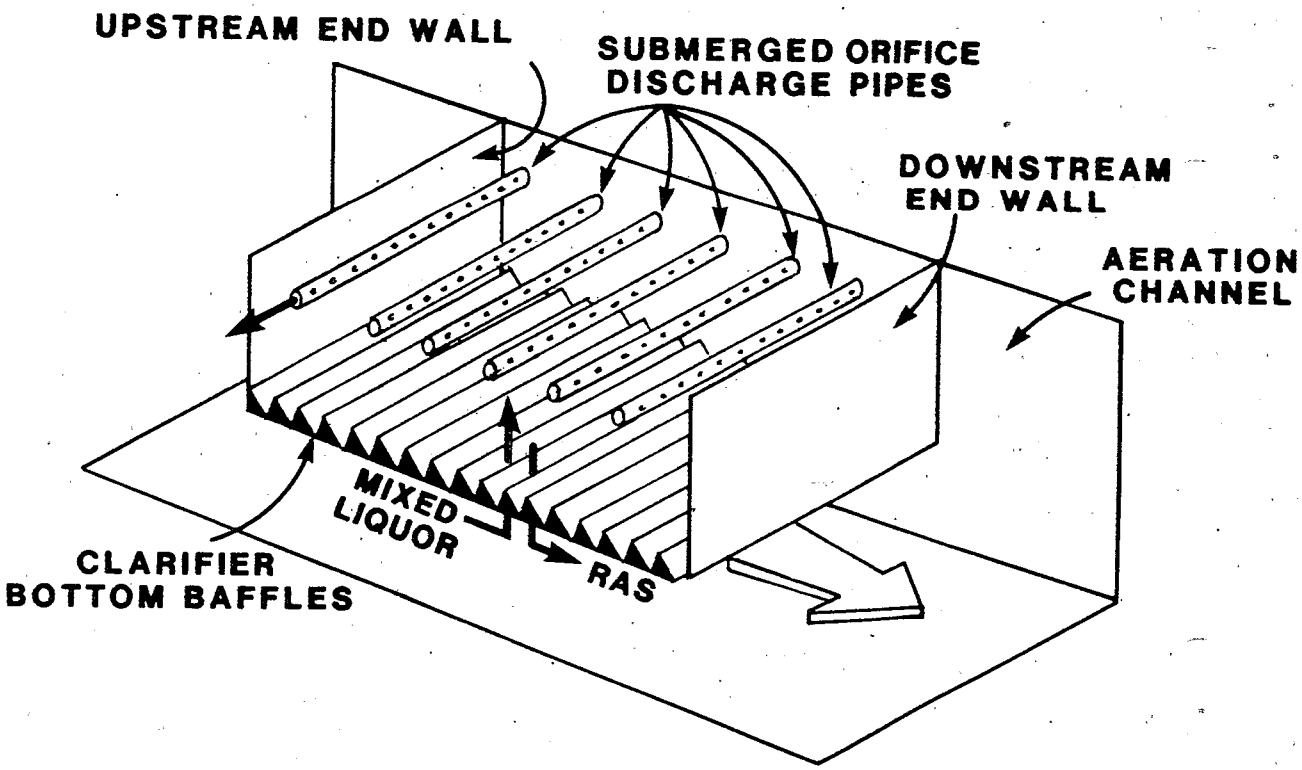


Figure 3. **BMTS SYSTEM** (2)

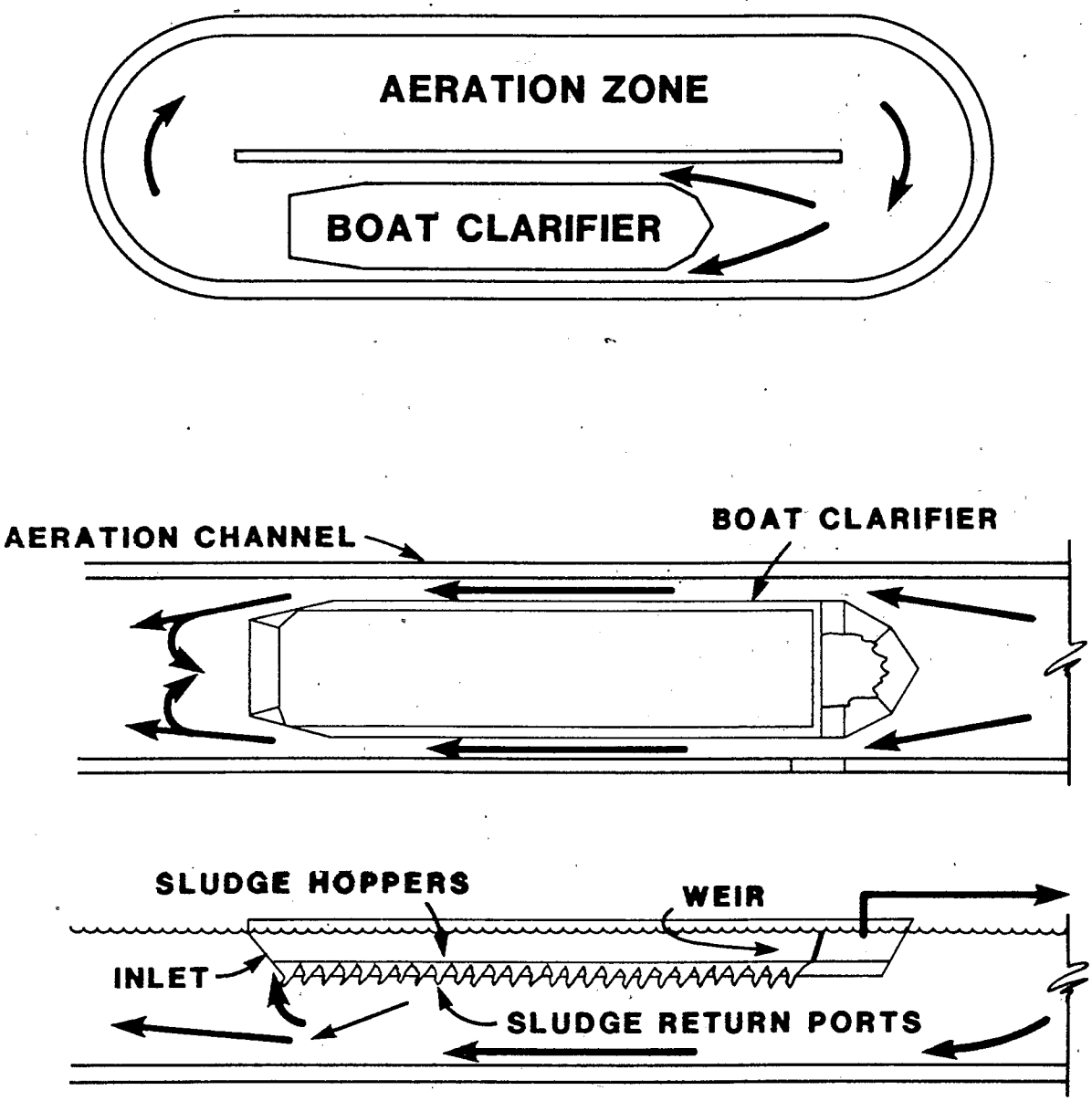
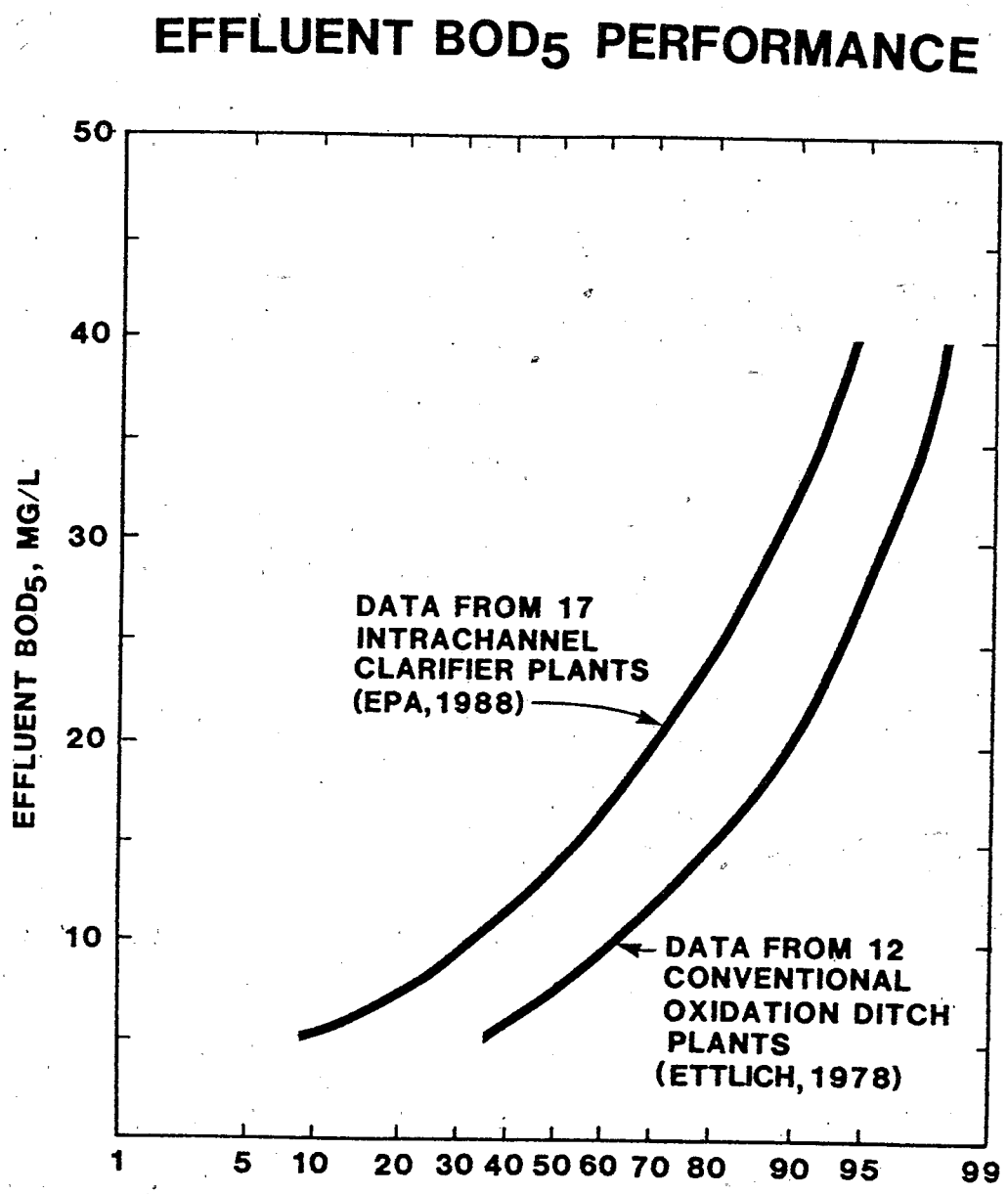


Figure 4. **UNITED INDUSTRIES BOAT CLARIFIER** (2)


EFFLUENT BOD5 PERFORMANCE
EFFLUENT BOD5, MG/L
50
40
30
20
10
DATA FROM 17 INTRACHANNEL CLARIFIER PLANTS (EPA, 1988)
DATA FROM 12 CONVENTIONAL OXIDATION DITCH PLANTS (ETTLICH, 1978)
1
5
10
20
30
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70
80
90
95
99
PERCENT OF TIME VALUE WAS LESS THAN OR EQUAL TO

Figure 5.

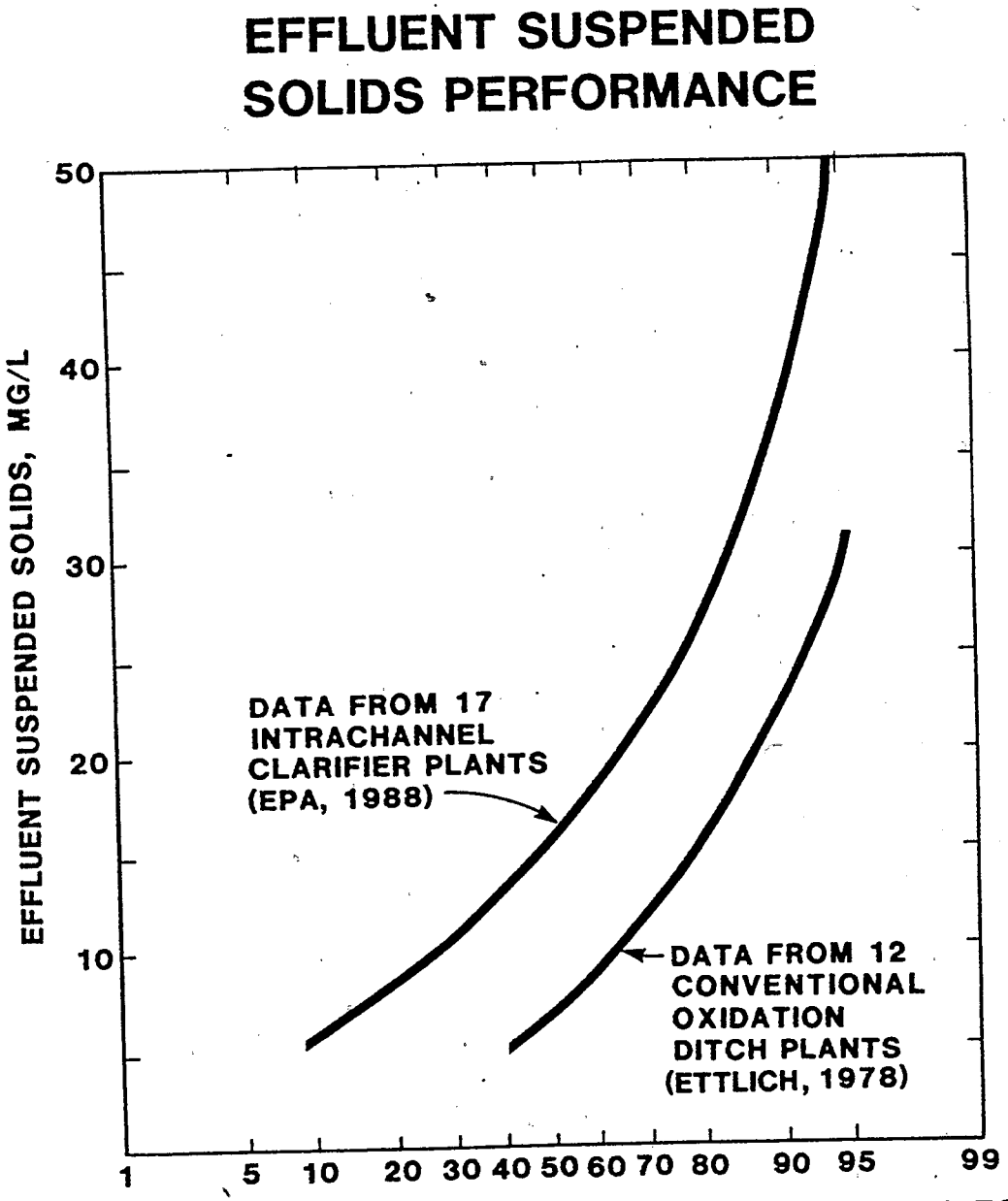

Figure 6.

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