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THE DESIGN AND TESTING OF A WASTEWATER TREATMENT PROCESS SUITABLE FOR ARCTIC APPLICATION

THE DESIGN AND TESTING OF A WASTEWATER TREATMENT PROCESS SUITABLE FOR ARCTIC APPLICATION

BY

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ABSTRACT

Design and Testing of a Wastewater Treatment Process Suitable for Arctic Application

The purpose of the project was to design and test the components of a potential package sewage treatment plant suitable for the Arctic environment. The designed treatment plant was simplistic in concept while utilizing the advantages of both biological and physical-chemical treatment.

The system was tested over a ten-week period. The emphasis of the testing was on treatment of high-strength wastes. In addition the effects of temperatures and system start-up-shutdown conditions were assessed. The results indicated that the system operated well under most of the conditions tested. Both the physical-chemical and biological processes complemented each other with the former providing the necessary treatment at start-up and the latter providing low cost treatment once the system had been in operation for a month.

With the high degree of treatment performance achieved, it was felt the process was feasible for Arctic application. However, further testing was recommended to develop the complete package unit.

PREFACE

The following report documents an experimental project performed in partial fulfillment of the requirements for the degree of Masters of Engineering at McMaster University, under the direction of Dr. Andrew Benedek.

The writer was interested in the design and operation of package sewage treatment plants. Realizing the inevitable development of the Canadian Arctic and the need for municipal services, a project was initiated which developed and tested a sewage treatment process suitable for Arctic application.

The project has not only been an education, but a benefit to this writer's career. The completion of the project would not have been possible without the guidance of Dr. A. Benedek and the facilities provided by McMaster University. A sincere thank you is certainly in order.

Finally, the writer would like to dedicate this report to his wife, Irene, for her continued support, patience and reassurance.

David E. G. Bromley

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1.0 INTRODUCTION

With the inevitable depletion of natural resources and, in particular, fossil fuels, the untapped Arctic cannot help but develop at a far greater pace than ever before.

To ensure that this development does not result in irreversible environmental damage, it is essential that future problems and methodology to avoid these problems be identified now. One of the many problems man is faced with is the disposal of his sanitary wastes. Hence, it is the purpose of this thesis to address this particular challenge.

Not until recent decades has Arctic water supply, solid waste disposal, air pollution control or any other environmental concern, been given any more than sporadic attention.

In the past, sewage disposal was either:

1) ignored;
2) accomplished with modified tools and methodology borrowed from temperate climate experience;
3) accomplished with tools and methodology fully favourable to the site.

Economists generally did not consider the value of preserving the environment and thus economics rarely encouraged anything more than the first approach mentioned above.

At present, most Arctic work camps use conventional methods for wastewater treatment. Since these camps are well financed and maintenance personnel are readily available, operation of these systems has not been a severe problem. However, this is not the case with the small northern community, and as a result, they rely on very crude methods of waste treatment.

Severe climate, weak local economy and the absence of a management system resulting from non-existent local governments have all had a deterring influence on the availability of a proper waste treatment system.

The operation of temperate climate systems in the Arctic has been difficult because of inadequate transportation facilities, community remoteness, and the lack of available operator personnel. In addition, the shortage of energy, high construction costs, high material costs and poor soil conditions for drainage stability have all had an inhibiting effect on the use of conventional temperate climate methods.

Arctic inhabitants, whether they live in a work camp or a small community, want modern sanitary waste disposal services, not only for convenience but also to avoid degradation of their surrounding environment. Providing this disposal source is the problem this project discusses and attempts to solve.

2.0 PRESENT KNOWLEDGE

2.1 The Arctic

The Arctic is defined in many ways, depending on the viewpoint of those defining it. The most useful definitions, in terms of their meaningfulness to man and his environment, are those based on climatic factors. A typical definition of the Arctic as suggested by J. E. Slater (1969) is a climatic area in which the temperature for the warmest month is less than 10°C and for the coldest month is below -3°C.

However, characteristics other than temperature are also very prominent in defining the Arctic. This project is concerned specifically with those areas in Canada's north which are habitable but are exposed to such restrictions as a sensitive environment, poor drainage, remoteness as well as cold temperatures.

2.1.1 The Arctic Topography

The Arctic's most prominent topographical characteristic is its extensive areas of permafrost. Permafrost is due to the soil's excessive exposure to cold temperature which freezes an area below the surface of the ground. During summer periods when temperatures are warmer, an active zone occurs at the ground surface. Since permafrost underlies this zone, moisture cannot escape, resulting in poor soil stability. Occasionally during the summer small pockets in the permafrost thaw and create "thaw bulbs" (Thorton, 1973). These thaw bulbs create areas of weakness within the soil structure and hence cause a stability and drainage problem.

Northern soils characteristically contain high amounts of water. Because of the permafrost structure and a certain amount of capillary action, the water content increases as the depth decreases (Thorton, 1973).

2.1.2 Present Arctic Water Quality

Very little is known about Arctic water quality except that extreme variations in the water characteristics are a common occurrence. Such extremes would not be tolerated by aquatic life found in a temperate climate. The dissolved oxygen (DO) level in rivers and lakes can fall as low as 1 mg/l in the winter. This is far below the accepted limit of 4 mg/l for fish life in temperate climates (Greenwood and Murphy, 1972). The great sensitivity of the Arctic environment lies mainly in this characteristic alone. Although the Arctic's aquatic organisms may be able to survive these low DO levels, no oxygen-dependent organism can survive on zero DO levels and therefore any pollutant loading on the water courses would be extremely toxic. In the winter it is not unusual for large rivers to have no flow as they are covered with six feet of ice, and therefore, the rivers have very little dilution capabilities. It is known that during spring flooding, water courses carry an enormous amount of suspended sediment. Thus, aquatic life must not only survive through low DO levels, but also high suspended solids. Finally the Arctic aquatic organism must be capable of metabolizing at temperatures less than 0°C. Only very hardy species would be capable of withstanding these harsh conditions.

2.1.3 Available Water Resources

Although the Arctic rivers produce massive runoffs, there is a water supply problem. (Heinke, 1973) The coastal plain area is a virtual desert as it receives only 2 inches of precipitation annually. Water does not percolate through the ground anywhere in the Arctic because of the permafrost hence, groundwater is minimal. Also there are major variations in both quality and quantity of waters found in this area. It is not abnormal for water conditions to change drastically over one year and therefore there is both a supply and a dependency problem.

Most of the coastal plain has a high artificial water table because of the soil's impermeability. This creates abundant swamps with muskeg vegetation. These swamps are a poor water supply source as they have very high levels of organic and iron content.

Some communities can obtain their water from ground storage below the permafrost regions. However, there are several problems with this source: (i) there is difficulty in drilling through the permafrost; (ii) the wells usually freeze; (iii) the water usually has a high salt concentration (i.e. greater than 2,000 mg/l).

Most lakes and rivers are impossible to use for a potable water source as only large water bodies do not freeze up completely. The best source of water is the scarce alluvial aquifers which are found in alluviums surrounding large water bodies.

Due to this deficiency in potable water sources the average cost in 1973 was $.35/gallon and rose as high as $1.00/gallon in some northern locations (Heinke, 1973).

2.2 Wastewater Sources and Collection

2.2.1 Black Water

Black water is the term given to toilet wastewater. Quantities of this waste stream can vary significantly depending on the system employed. The various systems available are discussed in the following section. (Grainge *et al*, 1973)

2.2.1.1 The Conventional Toilet

The conventional toilet is the same toilet that is installed in most homes throughout North America. The system contains a bowl and a flush water tank. When the toilet is flushed a plug is triggered in the tank which allows the flush water to enter the bowl. When the tank is emptied the plug falls back into place and the tank fills up. This system uses 3 to 5 Imperial gallons of water per flush.

2.2.1.2 The Bucket Toilet

This type of toilet is of the portable type. It consists of a sheet metal bucket inside a sheet metal container. The vented outside container has a toilet seat and is lined with a plastic bag to collect the waste. After the bucket is emptied a quart of water plus a biological inhibiting chemical (i.e. creosote) is added. This prevents septicity.

The method produces a very concentrated waste which is contained in plastic bags commonly called "honey bags". These bags are collected on a regular basis for disposal in a sludge pit (see Section 2.4.3.1). This method is very simple and inexpensive but is aesthetically objectionable as the honey bags are unsightly and a potential health hazard.

2.2.1.3 Vacuum Toilet

Vacuum toilets are part of a total vacuum system which is explained in Section 2.2.3.3. Basically the vacuum toilet is comprised of a toilet seat and bowl. At the base of the bowl is a plug which leads to a 3 inch pipe. When the toilet is flushed the plug is removed and a vacuum pressure of 7.5 psi draws the contents through the pipe.

Approximately 10 to 15 percent of the water used for a conventional toilet is used for the vacuum toilet (i.e. 0.5 gallons or 1.5-2.0 litres per flush). Most of the water is used to clean the bowl as very little is needed for transport. (Averill, 1974). Hence this system produces a low flow of highly concentrated "black water".

A study was made (Edwards, 1973) on the use of vacuum toilets and recirculating chemical toilets. It was found that the vacuum toilet was greatly preferred over the chemical toilet. The report also found that the only significant complaint users had when comparing the vacuum toilet with the conventional toilet was that the vacuum toilet was noisier. The noise of a vacuum toilet is similar to the noise of chemical toilets used on airplanes.

2.2.1.4 The Recirculating Chemical Toilet

The recirculating chemical toilet employs the contents of a waste holding tank for flushing purposes. The wastewater from the bowl is flushed into a tank. The tank's contents are made up of a lye which aids in liquefying the solid fraction of the waste, a chemical inhibitor, a blue dye, and a perfume. These contents are then used for flushing.

However, the system is useless if the pumps break down. In addition, the holding tank must be pumped out every day as it will emit very strong odours after a few days. Although the system is a recycle system and very little water is employed, it would be impractical in an Arctic community, for the reasons stated above.

2.2.1.5 Incinerator Toilet

Essentially this self-contained unit is a toilet bowl and incinerator. The bowl has a plastic liner which drops into an incinerator below. The source of heat in the incinerator is either propane or electricity. The waste is totally oxidized and no further treatment is necessary except in disposing of the residual ash.

This system is excellent aesthetically but has extremely high capital and operating costs. Since the system is basically a treatment method it is discussed in more detail in Section 2.4.3.4.

2.2.2 Grey Water

Grey water is the term given to wastewater generated from household sources and is usually 50 to 59 percent of the total waste stream. Some grey water sources are washing and bathing (65%), kitchen use (10%), washing clothes (9%), household cleaning (7%) and miscellaneous (9%). (Heinke, 1973)

Household liquid wastes are more dependent on the user than "black water". In other words, if the person is a slow washer, he will use more water.

There are methods that are available for reducing amounts of "grey water". Low pressure systems can be installed in wash facilities so that great amounts of water are not available. Showers can be installed instead of baths and timers can be put on showers so that the shower only runs for five-minute periods. This makes a break in the user's shower which tends to make him stop. Wash basin taps can also be spring loaded for automatic shutoff. These types of taps are commonly seen in gas station restrooms.

2.2.3 Wastewater Collection and Removal Methods (Heinke, 1973)

Presently in many small northern communities, wastes are trucked to dumping locations. However, in some larger communities, sewer systems have been introduced. The following is a brief summary of the various transporting systems available.

2.2.3.1 Trucking of Holding Tank Contents and/or Honey Bags (Heinke, 1973)

Removal of waste by trucking occurs in communities where honey bags or holding tanks are employed to collect the waste. Honey bags are plastic bags placed in bucket toilets to collect wastes. These bags are supposed to be placed outside each day and picked up by a truck which takes them to a dumping site. Holding tanks such as those placed below septic toilets are also used to collect wastes. Their contents are pumped into a truck and also disposed of at a dumping site.

Holding tanks are far more hygienic than honey bags as the latter tend to break occasionally. However, holding tanks have a high capital cost (i.e. between $600.00 and $700.00 per household). Both systems are extremely crude and certainly present a health hazard.

2.2.3.2 Conventional Gravity Sewer System

The conventional sewer system transports wastewater by gravity induced flow. This method has been used for centuries with very few changes. It is simple, dependable, and usually economical. However, gravity sewers have problems under certain conditions. First they require large quantities of water which in turn require large pipes. In the Arctic, pipes must be placed above ground, because of the high water table and permafrost. Above-ground piping must be protected from freezing, hence insulated pipes are heated by steam or electricity. These heated pipe systems are called utilidors. Utilidors are extremely large in diameter (i.e. 4.0 to 4.5 ft diameter in Inuvik, NWT) and are expensive to install. The cost estimate in 1972 was $180.00/ft and $250.00/ft. (Heinke, 1973)

In addition, since piping systems must be placed above ground, one would like them to follow the contours of the topography.

However, with gravity systems this is not always possible and the result is large obstructive support systems.

2.2.3.3 The Vacuum Sewer

The vacuum sewer is an alternative to the conventional gravity system but is used only for short distances. The fundamental principal of the vacuum sewer (Averill, 1974) is to transport wastewater by air pressure in a plug form rather than by gravity flow.

The vacuum sewer system is usually connected to a vacuum toilet as explained in Section 2.2.1.3, a plastic pipe, a collection tank and a vacuum pump.

The vacuum pump provides a vacuum of 7.5 psi in the system. When the discharge valve in the vacuum toilet is activated, the wastewater is pushed into the pipe by the atmospheric pressure. A liquid plug is forced by the air along the pipe to the collection tank. However, the liquid plug occasionally breaks down due to wall friction and the air behind breaks through. Therefore, along the piping system, transport pockets are placed as shown in Figure 2-1. The liquid flows by gravity into these pockets and reforms itself as a plug and then air again pushes the plug on along the pipe.

Under Arctic conditions the vacuum sewer system has some very important advantages. Its low water usage is not only ideal for an area of restricted water supply, but it also reduces the chance of freezing within the pipes as compared to the gravity system. In addition, because it uses a pressure differential, the system can be placed above ground to follow the contours of the topography. Finally, because of its low water requirements, smaller pipe sizes can be employed and less insulation is needed which results in reduced costs. Averill (1974) reports that in Europe and the Bahamas, where locations favoured the use of the vacuum system, the conventional gravity system cost 50 percent more.

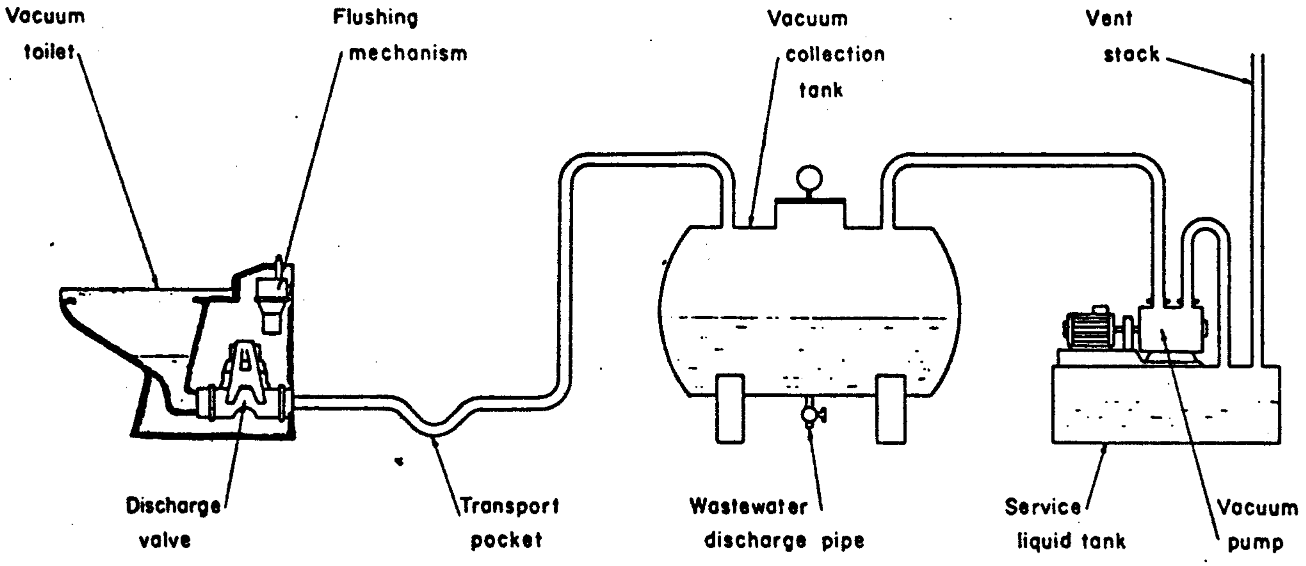


Fig. 2-1 BLACK WATER VACUUM SYSTEM - SCHEMATIC
(FROM AVERILL , 1974)

The only criticisms found in the literature were that the vacuum system was noisy, as explained in Section 2.2.2.4, and that it was good only for short transport distances (Averill, 1974).

2.2.3.4 Pressure Sewers

The pressure system is similar in theory to the vacuum system but has some important differences. Waste is pumped by a pressure pump to a pressure main which has a continuously circulating flow of concentrated waste. This continuous flow prevents freezing in the system.

In addition, before any waste passes to the pressure pump, a macerator or settling tank removes all large solids so that blockage does not occur. The pressure system can be used for longer distances than the vacuum system and because it uses a continuous flow, insulation costs are less than the vacuum system.

2.3 Arctic Sanitary Waste Characteristics

Both Heinke and Grange et al (1973) reported that in the Arctic, one can expect a sanitary waste flow from a conventional system of approximately 200 l (45 Imperial gallons) per day per capita. Heinke (1973) also reported that if a low water usage system was used the expected flow would be reduced to 130 l (29 Igal) per day per capita (see Appendix 2). Grainge et al (1973) indicated that at a prison in the Arctic, the flow varied from 200 l (45 Igal) to 330 l (75 Igal) per day per capita.

Bayley et al (1972) claimed that in Britain, where waste flows were similar to those estimated for the Arctic, chemical oxygen demand (COD) concentrations in the feed varied from 450 mg/l to 750 mg/l and suspended solids (SS) varied from 350 mg/l to 550 mg/l. As shown in Appendix 2, Heinke indicated that for conventional systems and low water usage systems, the expected COD concentrations in Arctic sanitary wastes would be 700 mg/l and 1,100 mg/l respectively.

2.4 Northern Wastewater Treatment

There are many methods available for wastewater treatment in the north. However, they are generally either very crude or very sophisticated.

In 1973 (Greenwood et al, 1972) all but four of 17 Arctic oil related work camps had extended aeration plants. Of the other four, two were physico-chemical treatment plants, one was an aerated lagoon and the fourth was an incineration process. In the larger Arctic communities (e.g. Yellowknife), wastewater treatment is generally carried out in sewage lagoons or extended aeration plants. However, in the smaller communities, treatment is either non-existent or at a crude and primitive stage.

In this section the treatment processes already in use or available for use are briefly summarized.

2.4.1 Biological Treatment

The basic principle of biological wastewater treatment is to encourage micro-organisms to feed on the nutrients (i.e. carbon, nitrogen and phosphorous) in a controlled environment. As these micro-organisms feed on the nutrients more micro-organisms are produced in addition to the by-products H_2O and CO_2. Prior to releasing the treated wastewater to a receiving stream, the micro-organisms are separated from the liquid waste and are either allowed to die or are reused. The result is a final effluent which will demand far less oxygen from the receiving stream.

Micro-organisms generally go through several growth phases within the controlled environment. The log growth phase and the endogenous growth phase are two particularly important phases as shown in Figure 2.2.

In the log growth phase the substrate is in excess and the micro-organism reproduces at an exponential rate. As a result, the micro-organism decay rate is insignificant.

In the endogenous growth phase the substrate is limited and therefore the micro-organisms feed on themselves. As a result, reproduction is minimal and therefore the decay rate becomes important.

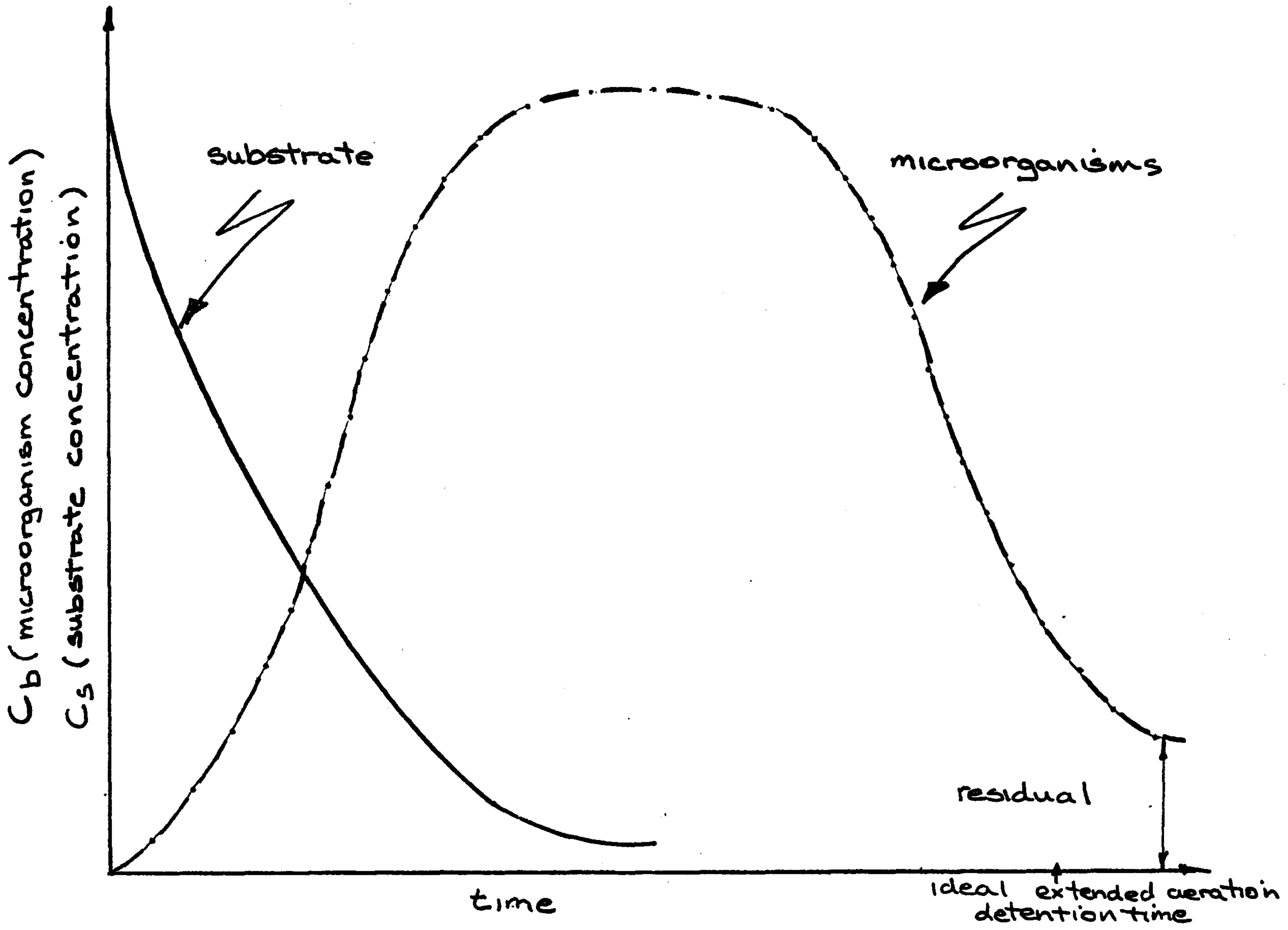


FIGURE 2-2 SUBSTRATE AND MICROORGANISM CONCT^n^ VS TIME

2.4.1.1 Temperature Characterization of Bacteria

Microbiologists generally agree that three thermal groups of bacteria exist. These are psychrophiles, mesophiles and thermophiles. The optimum growing temperature of each is shown in Table 2-1. In cold climates psychrophiles would obviously be the predominant group of bacteria in any biological system.

It is believed (Henry et al, 1972) that the major reason these groups grow at various temperatures is due to the degree in which their fatty acids are saturated. Bacterial membranes are constructed from fatty acids and protein (i.e. three layers; fatty acid, protein, fatty acid). Fatty acids can either be of the saturated type (i.e. single carbon to carbon bonds) or of the unsaturated type (double carbon to carbon bonds). Unsaturated fatty acids have a lower melting point than saturated fatty acids. Psychrophiles likely have more unsaturated fatty acids in their membranes than mesophiles or thermophiles. As a result the fatty acids in the psychrophile do not crystallize at the lower temperatures and thus permit food to be transported through the membrane. Mesophiles have a high amount of saturated fatty acids and therefore at lower temperatures the fatty acids crystallize. This restricts food transport and the micro-organism dies. This same phenomenon would also occur with thermophiles, which have

TABLE 2.1

TEMPERATURE CLASSIFICATION OF MICRO-ORGANISMS

Group \ Temperature (°C)	Minimum	Optimum	Maximum
Thermophiles	40-45	55-75	60-86
Mesophiles	10-15	23-35	35-50
Psychrophiles			
Obligates	(-5)-(+5)	5-10	15-20
Facultative	(-5)-(+5)	12-16	35-40

an even higher percentage of saturated fatty acids.

Microbiologists also agree that thermophiles grow at a faster rate than mesophiles which grow at a faster rate than psychrophiles.

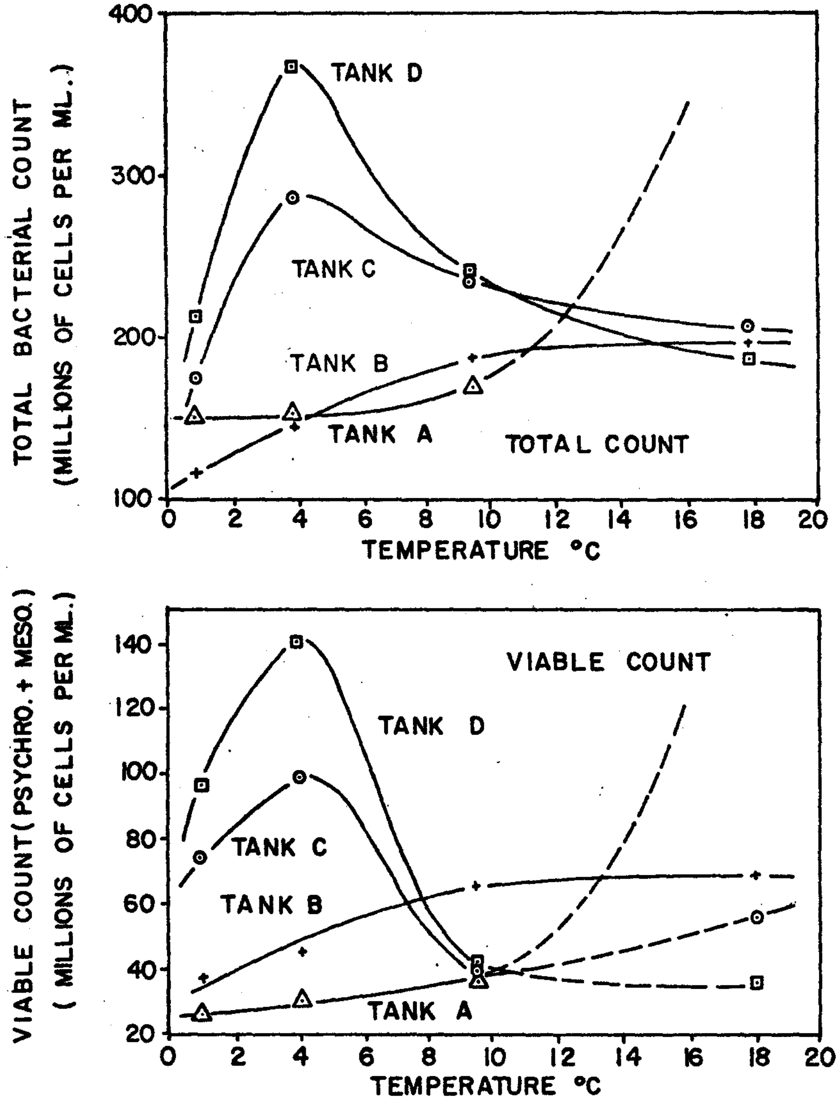
2.4.1.2 Cold Temperature Effects

Henry (1972) performed some experiments to assess temperature effects on the biological environment. He looked at a conventional biological waste treatment system and varied the detention times and temperatures. Henry's observations showed that there was psychrophilic growth but it was only detectable after long reactor detention times, i.e. 24 hours and 96 hours. This can be seen in Figure 2-3.

In addition, Henry felt that psychrophiles were the least sensitive of any group, since his observations indicated that psychrophiles would grow over a broader temperature range than either mesophiles or thermophiles. It is believed that since psychrophiles' fatty acids are not as saturated as the mesophiles or thermophiles, not only can psychrophiles live at low temperatures for reasons previously stated, but they can also live in the mesophilic temperature range.

Benedek (1970) claimed that temperature did not affect gm of O_2/g of biological oxygen demand (BOD) removed. This would mean that psychrophiles would not

FIGURE 2-3



DETENTION	TIME
TANK A	2·4 HRS.
TANK B	9·6 HRS.
TANK C	24 HRS.
TANK D	96 HRS.

EFFECT OF TEMPERATURE ON NUMBERS OF BACTERIA
(FROM HENRY, 1972)

require more oxygen than the other thermal groups. Since dissolved oxygen saturation increases as temperature increases, one would not expect a cold climate treatment system to require a greater oxygen supply than temperate climate systems.

Friedman (1972) showed that temperature had little effect on nitrogen or phosphorous removal in a system. However, he did state that the best nitrogen removals occurred at approximately 28°C. This would seem acceptable since it is believed anaerobic denitrifying bacteria are normally mesophiles.

Finally, Friedman also found that temperature had a minimal effect on the sludge volume (SVI). As temperature decreased the SVI increased slightly which reduced the sludge settling capabilities.

2.4.1.3 Sewage Lagoons

Sewage lagoons or oxidation ponds provide waste treatment by the biological process. The lagoon is a large pond, into which sewage flows. Sunlight and CO_2 released from decomposing organic waste helps the growth of algae which, through the process of photosynthesis, provides sufficient amounts of oxygen.

According to Grainge *et al* (1973), lagoons placed in the north can expect 90 percent BOD removal in the summer and 45 percent removal in the winter. Solids removal is approximately 90 percent all year round. The lower BOD removal in the winter is due to the lack of sunlight during this part of the year. As a result, algal oxygen production is decreased, and BOD removals are reduced significantly. For this reason it is essential that Arctic lagoons have a 1-year detention time so that all sewage will have summer treatment.

Many problems with construction and operation of such lagoons have been reported. Some of these problems include:

(i) Difficulty in placing and constructing the lagoon. In 1973 earth-moving costs were approximately \$2.20/yd^3 (Grainge et al, 1973) and proper berm (embankment) material was difficult to obtain. Also, as mentioned in Section 2.1.1, the existence of thaw bulbs makes the soil unstable and therefore creates a problem for embankment stability and permeability. In addition, effluent from the lagoon must be discharged properly. If a warm effluent leaves the lagoon and crosses a permafrost area, thawing takes place in the soil causing a gully to form. If effluent is discharged into a water course, the receiving stream must be capable of good mixing and dilution.

(ii) Poor process control. A lagoon might be operating adequately one week and extremely poor the next. Also, in the winter, the system goes anaerobic and unpleasant odours can result.

(iii) Difficulty in conveying the wastes to the lagoon. Lagoon treatment requires a piping system to feed the lagoon. This is not only costly but operationally it is a problem because of freezing.

2.4.1.4 The Extended Aeration Plant

As of 1972 oil related Arctic work camps employed the extended aeration system far more often than any other system. As with the sewage lagoon, the extended aeration plant utilizes the biological process but in a different way. The system is essentially comprised of two components, an aerated reactor and a clarifier. Oxygen is supplied to the reactor by mechanical means to feed a micro-organism population. Within the reactor the hydraulic detention time is 18-30 hours. The contents of the reactor continuously flow into a clarifier where the micro-organisms are settled out as a sludge. The settled sludge is then returned to the reactor where the micro-organism concentration is maintained at a high level, i.e. between 2000-10,000 mg/litre.

These types of biological systems, where the sludge is returned to the reactor, are called activated sludge systems.

The extended aeration system is different from other activated sludge systems as it operates on the principle of total oxidation of organics.

As seen in Figure 2-3 the plant maintains long reactor detention times. In the early stages of the reaction the substrate is oxidized and micro-organisms, CO_2 and H_2O are produced. As time progresses the substrate is depleted and the micro-organisms begin to feed on themselves until only an inorganic residual remains. The result is a minimal amount of solids. However, it is impossible to oxidize all the organics within a reasonable length of time, therefore someorganic substance is left.

In all biological systems, energy is produced when oxidation takes place. Some of the energy is stored and some is wasted. The energy which is wasted is generally given off as heat. The extended aeration plant, by attempting to oxidize all organics, produces as much heat as possible, however, with the long mixing times, extensive heat losses also occur.

Also, using data from Henry et al (1974), the extended aeration plant allows for growth of psychrophiles in cold climates because of its long detention times. This permits adequate micro-organism concentrations to occur at the low temperatures. Short detention time units do not provide adequate time for the psychrophiles to grow and hence poorer treatment results. In addition, the system does not use any costly chemicals and when running at optimum conditions, provides excellent removals, i.e. <90 percent BOD removal, <90 percent SS removal as reported by Grainge _et al_ (1973).

However, there are problems with the extended aeration system as related to the Arctic. Firstly, as reported by Greenwood _et al_ (1972) one cannot expect the plant to reach optimum operation conditions for the first two months. This is probably due to the slow rate of growth of phychrophiles. Due to this slow start-up, the plant is incapable of providing wastewater treatment for the temporary camp that may only operate for a two-month period. A system for these types of camps or communities would have to operate at a high efficiency immediately upon start-up.

There have been, and still are, many package systems which use the extended aeration process. Many of these systems have been sold to northern oil related camps. Heachert (1972) made a study of two of these package systems which were installed by Imperial Oil on two artificial island camps in the Mackenzie Delta. The plants were standard extended aeration plants containing comminutors, aerators, clarifier with sludge return and hypochlorinator. Both plants were housed and were built for a 22 hour detention time. Flush water for both plants was sea water.

The package system ran very poorly. They only achieved 80 percent BOD_5 removal and 63 percent removal of COD. In addition, there were problems with oxygen deficiency and the plant was not capable of handling shock loads. Finally, the plants were highly dependent on an operator, and since both plant operators were not familiar with the plant, the treatment systems never ran efficiently. Also one should note that these plants did face rather high and abnormal COD concentrations in the feed. It was not unusual for the plant to have a feed with COD greater than 2,500 mg/l, which apparently exceeded the design capacity of this package system.

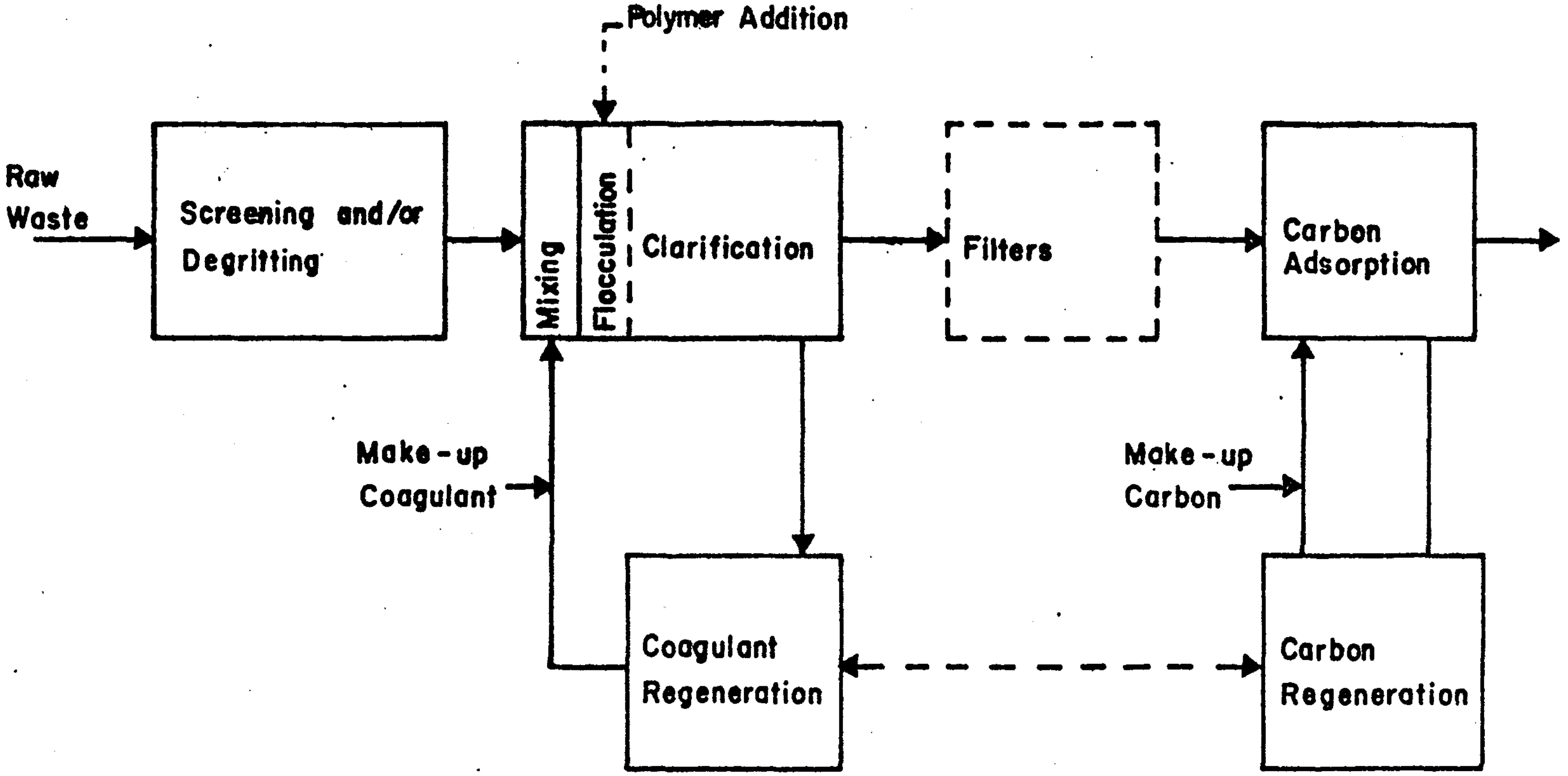
2.4.2 Physical-Chemical Treatment

This method of treatment depends strictly upon the use of chemical addition and various physical processes for the treatment of the wastewater. A typical physical-chemical treatment flowsheet is shown in Figure 2-4.

Chemicals are added to treat the waste in different ways (i.e. flocculation, disinfection, oxidation). Chemical

FIGURE 2-4 : GENERALIZED FLOW SCHEME FOR PHYSICAL - CHEMICAL TREATMENT

(FROM MAQSOOD, 1973)



agents such as alum, ferric chloride, polymers or lime are commonly used for coagulation and flocculation. Chlorine is commonly added for disinfection and sometimes oxidation. Ozone and permanganate are also added occasionally for oxidation of waste organics.

In addition physical processes play a large role in waste treatment. Such processes as ion exchange or ultrafiltration are well utilized in the industrial field and sometimes for municipal use. However, the most prominent among these processes for municipal wastes it the use of activated carbon for its organic adsorptive capabilities. Activated carbon can be utilized in two forms, the powdered form (PAC) or the granular form (GAC). PAC is usually added to a contact tank in a similar fashion to alum. The PAC slurry is then directed to a clarifier for settling. GAC, on the other hand, is usually placed in a column. Normally, clarifier effluent passes through the column where dissolved organics are either adsorbed on the immersed pore surface area of the activated carbon particles or removed by the biological activity occurring on the surface of these particles. In addition, the column's filtration effect encourages a further removal of suspended solids.

2.4.2.1 Cold Temperature Effects

Temperature effects on chemical addition or physical processes should be minimal in comparison to its effect on biological activities. It is believed (Maqsood, 1973) that temperature will adversely affect the solubility of coagulants in fluids. Mohtadi (1973) confirmed this theory when he reported that temperature affects the optimum pH for coagulant dosage when alum is used. He found that as the temperature decreased the optimum pH increased, but no extra dosage of alum would be needed if the dosage was applied at the optimum pH.

Temperature effects on the physical processes were also found to be negligible. Maqsood and Benedek (1973) reported that low temperatures had little effect on the adsorption capacity of powdered activated carbon. Granular activated carbon was more temperature-sensitive with regard to the rate of adsorption. However, Maqsood noted that althoughtemperature had an effect on adsorption, it was not as great as temperature's effect on the biological activity within the column.

Maqsood was in general agreement with Bishop _et al_ (1972) who also felt that temperature had little effect on the adsorptive capacity of activated carbon.

Fraser (1972) reported on the use of cold climates to separate solids from the effluent by freezing the wastewater. The idea was that as the upper level of the waste froze, solids would be forced to the bottom of the liquid where they would be concentrated. However, the system failed at high solids levels and therefore was deemed unfeasible for most wastewater treatment systems.

2.4.2.2 Package Physical-Chemical Treatment Systems

There have been a number of package plant systems which have incorporated physical-chemical processes. These package systems are ideal for northern camps because of their ability to operate efficiently at start-up. Also, they have the ability to handle shock loads easily and hence, are suited to camps which frequently change in size. However, they are in some regards not suitable for the small community as they depend upon costly chemicals and the transport of these chemicals to the site. In addition, they require constant operator attention.

Marland Environmental Systems developed a marine sanitation system which was tested and reported upon by Kaminsky (1973). The system attempted to employ a total recycle system (i.e. treated effluent was used as toilet water, etc.) where only solids were wasted. For this reason the system was of interest to this report. The process first removed suspended solids in a two-step physical operation, i.e. large solids removed by a vibrating screen and smaller solids by a centrifuge. The liquid waste was then chlorinated prior to entering the carbon columns. In the carbon column most of the dissolved organics were removed and the column's effluent was post-chlorinated. All solids were stored in a solids holding tank. The system was tested first as a flow-through system and secondly as a total recycle system.

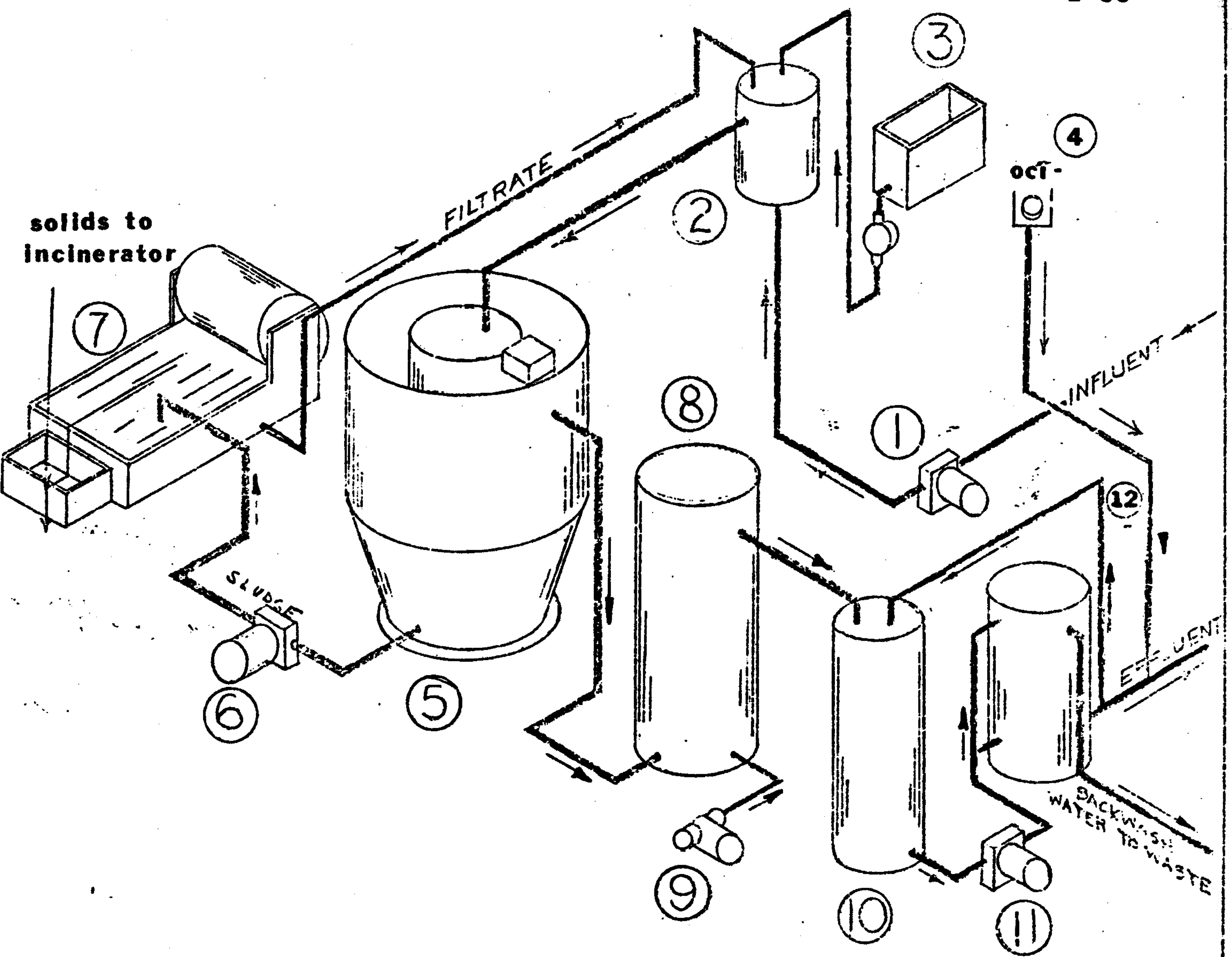
The system consistently produced an effluent with suspended solids and BOD_5 less than 50 mg/litre when tested as a flow-through system. When the system was converted to a recycle system, SS were less than 50 mg/litre. BOD_5 on the first day was 50 mg/litre but increased to 140-400 mg/litre by the sixth day. The recycled flush water was slightly milky in colour on the first day and then changed to a grey colour by the fifth day.

The system did not provide for an adequate method of solids removal. Solids reduction was poor since there was only a small amount of digestion in the anaerobic tank. As a result the solids holding tank had to be pumped out frequently. Also, the system contained a great deal of mechanical equipment (e.g. centrifugre, vibrator screens) which required maintenance. Finally, the system went anaerobic in sections. One particular spot was in the carbon columm which led to the formation of H_2S and some odour problems.

A second physical-chemical package system which recycled effluent for toilets was developed in Canada by Elsam Yarrow Limited (Port Colborne, Ontario) and was reported upon by Heinke (1973). The system incorporated a pressure breakdown tank where chlorine was added, a comminutor, coagulant addition tank, clarifier and pressurized effluent storage tank. Sodium hydroxide was used as the coagulant and also created a high pH which acted as a disinfectant to kill bacteria.

Finally, a third physical-chemical package plant of interest to this report was developed by Met-Pro Systems Inc. (Lansdale, Pennsylvania). Testing of the system was carried out by the United States Environmental Protection Agency and a report was prepared by H. Coutts (1972).

A prototype installation was operated consecutively at two different oil drilling camps (i.e. the Nabor camp and the Arco camp) near Prudhoe Bay in April and in May of 1972. A block flow diagram of a 7,000 gpd unit is shown in Figure 2-5 (Met-Pro Dwg. Number 0003902). Raw sewage is collected and pumped from a sump below the housing trailer floor



1 RAW WASTE PUMP
2 FLASH MIX TANK
3 COAGULANT FEEDER
4 DISINFECTANT FEEDER
5 FLOCCULATOR CLARIFIER
6 SLUDGE PUMP
7 DISPOSABLE MEDIA FILTER (OPT)
8 ADSORBER
9 ADSORBER AERATOR
10 SURGE TANK
11 FILTER/BACKWASH PUMP
12 PRESSURE FILTER

Figure 2-5

Process Flow Isometric

MET-PRO WATER TREATMENT CORP.
LANSDALE, PA. 19446

TITLE SERIES 14000
IPC WASTEWATER TREATMENT

into a 3,500 gallon aerated equalization tank.

From the equalization tank the sewage and coagulant are separately pumped into a flash mixer. From the flash mixer, flow is by gravity into the flocculation (centre) section of an upflow clarifier. Effluent (supernatant) from the clarifier then flows along with a small airstream (i.e. for fluidization and to prevent anaerobiosis) up through a carbon adsorption column (upflow) and into an 80 gallon surge tank. Liquid from the surge tank is pumped through a pressure sand filter. Effluent from this filter is chlorinated by use of hypochlorite tablets and intermittently siphoned onto the tundra.

The chlorinator consisted of a small dissolution box into which four 3-inch slotted pipes (hypochlorite tablet containers) are standing. Effluent flowing around and through the slots leaches hypochlorite from the tablets. The baffled chlorine chamber has a theoretical design detention of 0.5 hour.

Sludge from the bottom of the clarifier is pumped onto a moving paper filter. Filtrate is pumped to the flash mixer. Sludge solids collected on the filter paper are incinerated with the paper.

When activated by the high liquid level switch in the feed equalizer tank, the unit operates continuously at about 5 gpm design capacity until shut-off by a low level switch. This physical-chemical process operated only at

100 percent of design capacity since the equalization tank absorbed sewage flow fluctuations.

The unit was installed along with the chlorine contact tank, the feed equalizer tank, sewage sump and incinerator in a skid-mounted trailer, approximately 10 x 40 feet, so that the whole waste handling system could be loaded into a C-130 aircraft (Hercules) and flown to the North Slope. The disadvantage to such a compact enclosure was the limited access room for easy operation and maintenance. The only access to the unit with more than one foot clearance from any wall was along the control panel side. Clearance on that side was limited to less than two feet.

Access to the surface of the clarifier was severely limited by the domed trailer roof; so limited, in fact, that there was not enough room for the mixer to be installed on the flash-mix tank.

At the Nabor camp the plant treated 2,000 USgpd of bathroom and personal laundry wastewater. Kitchen wastes were separately sewered and discharged into the lagoon. At the Arco site the plant treated all wastewater. This resulted in an average flow of 4,100 USgpd or a 120 gal per day per capita. The most important process parameters are summarized in Table 2-2.

In April the plant operated well with good COD and SS removals, i.e. 95 percent and 96 percent respectively.

TABLE 2-2

Data Summary

Physical-Chemical Unit on the North Slope, 1972

	April Average Nabors Site	May Average Arco Site
FEED: Waste Water Source Rate	Bathroom and Personal Laundry Est. 1000-2000 gpd	All Domestic Camp Sewage 4100 gpd
COD mg/l	1257	846
Total Nitrogen mg/l	127	48
Total Phosphorus mg/l	9.3	15
(Number of samples)	(3)	(7)
Clarifier Overflow:		
COD mg/l	178	439
Suspended Solids mg/l	40	119
(Number of samples)	(8)	(4)
Effluent:		
COD mg/l	61	312
Suspended Solids mg/l	26	155
Total Nitrogen mg/l	127	32
Total Phosphorus mg/l	0.7	3.4
(Number of samples)	(3)	(7)

In May the treatment was poor with COD and SS removals averaging 63 percent and 72 percent respectively. The poor results in May could have been due to either the inclusion of kitchen wastewater or because the sensor light in the sludge level indicator became inoperable.

Problems which occurred with the operation of the unit as reported by the Environmental Protection Agency study were as follows:

1) Excess amounts of coagulant (i.e. ferric sulphate) were used because the operator did not observe the colour of the clarifier effluent nor was he familiar with the jar test.

2) The carbon columns air-fluidizing line became clogged with carbon particles. A diffuser screen should have been placed over the air outlet.

3) The supply of filter paper for the vacuum filter was exhausted and caused the operator to use paper towels. This resulted in a poorly dewatered sludge.

4) Air within the trailer became contaminated as the disinfection process encouraged the effervescing of chlorine gas. In addition,

aerosols were created from aerating the equalization tank. This problem could have been avoided if the tank had been covered or aeration had been reduced.

5) Excessive noise occurred when the compressor was in operation. An inlet filter silencer on the air compressor would have alleviated this problem.

6) The sludge level sensor was activated by surface scum rather than by the sludge blanket. This problem could have been solved if a dimmer switch on the sensor had been employed.

7) The 3/4" feed line became clogged and caused the pump to overheat, which resulted in the relay switches tripping. To avoid this difficulty a larger feed line could have been used or the sewage could have been macerated, prior to entering the feed line.

8) The chemical feeder pump's head needed rebuilding. This was either due to the grit in the ferric sulphate or ferric sulphate's high corrosive nature. It is felt that if

alum had been used this problem would not have occurred. However, it should be noted that operators of P-C units in the Arctic seem to prefer the use of ferric salts as a coagulant. Some advantages of ferric salts as compared to lime or alum are that they operate over a wide concentration range, they dissolve easier and a rust clouded effluent may indicate excessive use.

9) During shutdown a freeze-up occurred in the sand filter. This created a problem at start-up which resulted in the excessive loss of media in the sand filter. The problem could have been avoided if the system had been completely drained at shutdown.

The EPA study concluded that the system was capable of providing good treatment when operating as intended. However, considerable effort must be devoted to the design, set-up, hardware selection and operation if P-C units are to be utilized in the future for isolated Arctic installations.

2.4.3 Miscellaneous Treatment Methods

There are many other treatment methods now used or proposed for use in the Arctic. The following is a summary of these methods.

2.4.3.1 Dumping

One of the oldest methods of waste treatment all over the world, including the Arctic, is dumping the waste in a remote location with the intent of hoping it shall disappear. In the Arctic the usual location was on the ice covered sea and when the spring break-up arrived, the waste would settle in the ocean.

Fortunately, this practice is not allowed any more. However, dumping is still not uncommon today in small communities where no sludge pits or lagoons are present and dumping is the only method of waste removal. As reported by Heinke (1973) the dumping is usually done in an area away from the garbage dump and least detrimental to the water supply. There is, however some beneficial uses of dumping in the Arctic as stated by Fahlman _et_ _al_ (1973). He claims that sewage can be used to aid restoration of the land where Arctic soils of low nutrient content exist (i.e. an area where there has been construction). He cautions though that dumping or land disposal should not be used as a regular method of waste removal because of the permafrost. In permafrost soils the water table is very high and in many places is surface exposed. Land disposal will contaminate this groundwater which will affect wildlife and eventually water courses.

2.4.3.2 Sludge-pits

Sludge-pits are another common method of waste removal in small Arctic communities. "Honey bags" as defined in section 2.2.3.1, and holding tank contents are disposed of in sludge-pits (lagoons). When the pit is filled it is covered with earth and a new one is dug. Sometimes a lagoon as explained in section 2.4.1.3 is located beside the pit so that the liquid from the pit can run into the lagoon.

The problems with the sludge-pit are the same as those associated with a lagoon with regard to construction and placement. These problems are noted in section 2.4.1.3. In addition, the sludge-pit goes anaerobic, resulting in an odour problem. It also can severely contaminate groundwater.

2.4.3.3 Sludge Disposal Stations

Sludge disposal stations or "Honey bag" disposal stations are presently being used in Greenland (Heinke, 1973). Essentially these stations are two-storey buildings. The contents of the bags are unloaded into a tank and then the bags are burnt. The tank's contents are then piped to the sea. Although the practice of dumping to the sea is not allowed in Canada, this disposal procedure of honey bags and holding tank contents is superior to any method available in Canada. A method such as this could be used if the stations' contents were treated. However, the feasibility of such an approach is limited since honey bags and holding tanks are not suitable methods for sanitary waste collection and will likely become extinct.

2.4.3.4 Incineration

This method has become very popular for self-contained units and large municipalities. The process either incinerates liquid and solid wastes coming directly from the household, or first concentrates the waste to a sludge form followed by incineration. In either case, the waste is reduced to a small quantity of ash.

The idea behind the process (Grainge et al, 1973) is that little or no fuel is required as the organic material in the waste, once oxidized, will produce an adequate amount of heat to continue the oxidation process.

Unfortunately this is not the case and the process depends on a constant fuel source. Fuels used for the system include propane, oil mixed with the waste, and electricity.

The advantages to incineration are as follows:

(i) low installation cost
(ii) the small quantity of sludge to handle

(iii) almost total destruction of the waste

However, there are also a number of disadvantages. The main disadvantage is the high operating cost resulting from the high amounts of fuel required, and the high capital cost. Transportation of fuel to remote Arctic communities is difficult and therefore the system is likely to be inoperable much of the time. Odours emanate from the incinerating units if they are operated improperly and air pollution may be a problem if stacks and scrubbers are malfunctioning.

There have been a number of package systems which incorporate the incinerator process. The popular incinerator toilet was described in Section 2.2.1.4. Most users of this unit are satisfied with it. However, the costs are extremely high. According to Grainge _et al_ (1973) the installation cost of the incinerator toilet is 3.5 times greater than the low water usage flush toilet, and it has only half the life expectancy. Also, he found that the operating costs of the incinerating toilet are 3.5 times greater than the low water usage toilet. One should note that this cost comparison was made in 1972 and therefore the difference would be even greater today.

An incinerator package system was tested in 1973 at Fort Simpson, Northwest Territories. This system consisted of two trailers; one trailer as a wash car (i.e. basins, showers, toilets) and the other as an incineration complex. The toilets in the wash car were of the vacuum and chemical type as explained in section 2.2.1. The wash car waste was transported by vacuum sewer to a holding tank. When the holding tank was full, it was emptied by an utilidor gravity system which ran to the incinerator trailer. Here, the wastewater was macerated, and finally incinerated.

However, as with other systems, there were problems. Again the cost was extremely high, as electricity was used for the incinerating process. In addition, extensive maintenance was required and during these breakdowns, serious back-ups occurred in the sewage system.

CHAPTER 3

3.0 SYSTEM RATIONALE

3.1 Design Objectives

The state of the art review in Chapter 2 indicated some of the concerns and problems associated with the disposal of sanitary wastes in the Arctic. Small Arctic communities utilize very crude disposal practices, which are both aesthetically and hygienically poor.

Arctic work camps utilize more sophisticated systems but as a result the are very costly, require extensive operator time, and often still provide poor treatment efficiencies (see section 2.4).

The treatment systems discussed in Chapter 2 were either dependent on biological activity or physical-chemical processes. The main disadvantage to biological systems is the poor treatment of the wastewater when the system initially starts up. Indications are that the extended aeration process is the most suitable biological process for cold temperature application as it allows sufficient time for the micro-organisms to metabolize (see section 2.4.1.2). Physical-chemical systems provide excellent treatment at start-up but are costly to operate and maintain (see section 2.4.2).

The literature review also noted the lack of available potable water sources (see section 2.1.4). As a result there was concern over minimizing water usage and its effect on a treatment system (i.e. concentrated wastes, see section 2.2).

Generally soil conditions were assessed as being very poor because of the permafrost and high water table (see section 2.1.1). In addition, these poor soil conditions, along with cold temperatures, remoteness and lack of trained personnel make construction of treatment facilities and their collection systems both expensive and difficult (see section 2.2.3.2).

The objective of this thesis was to develop a basic process capable of functioning under constraints such as those mentioned above. To accomplish this, the following design objectives were formulated.

1) The treatment system should incorporate both biological and physical-chemical processes. This combination will ensure adequate and economical treatment at all times. The physical-chemical processes will provide the necessary treatment when the system first starts up. However, once biological activity has reached its optimum level, the physical-chemical processes can be phased out, resulting in reduced operational costs.

2) The system should not require any subterranean construction (e.g. septic tank tile field, etc.).

3) The system's concept and operation must be as simple as possible.

4) The system must be housed within a heated containment area. In addition to minimizing the need for costly outdoor piping, the system should be utilized by only a small number of dwellings which are in close proximity to each other.

5) The system should be capable of providing a high quality effluent suitable for recycle (see Appendix 1). The effluent should have "desired" COD and SS values less than 20 mg/l and 5 mg/l respectively, and "not to exceed" values of 50 mg/l and 15 mg/l respectively (see Appendix 1 for determination of these values).

3.2 Treatment System

To satisfy these objectives, it was felt that the actual treatment system should consist of the following five components:

(i) an equalization tank

(ii) a combined biological reactor and clarifier (hereinafter called "the reactor-clarifier")

(iii) an adsorption-filtration column

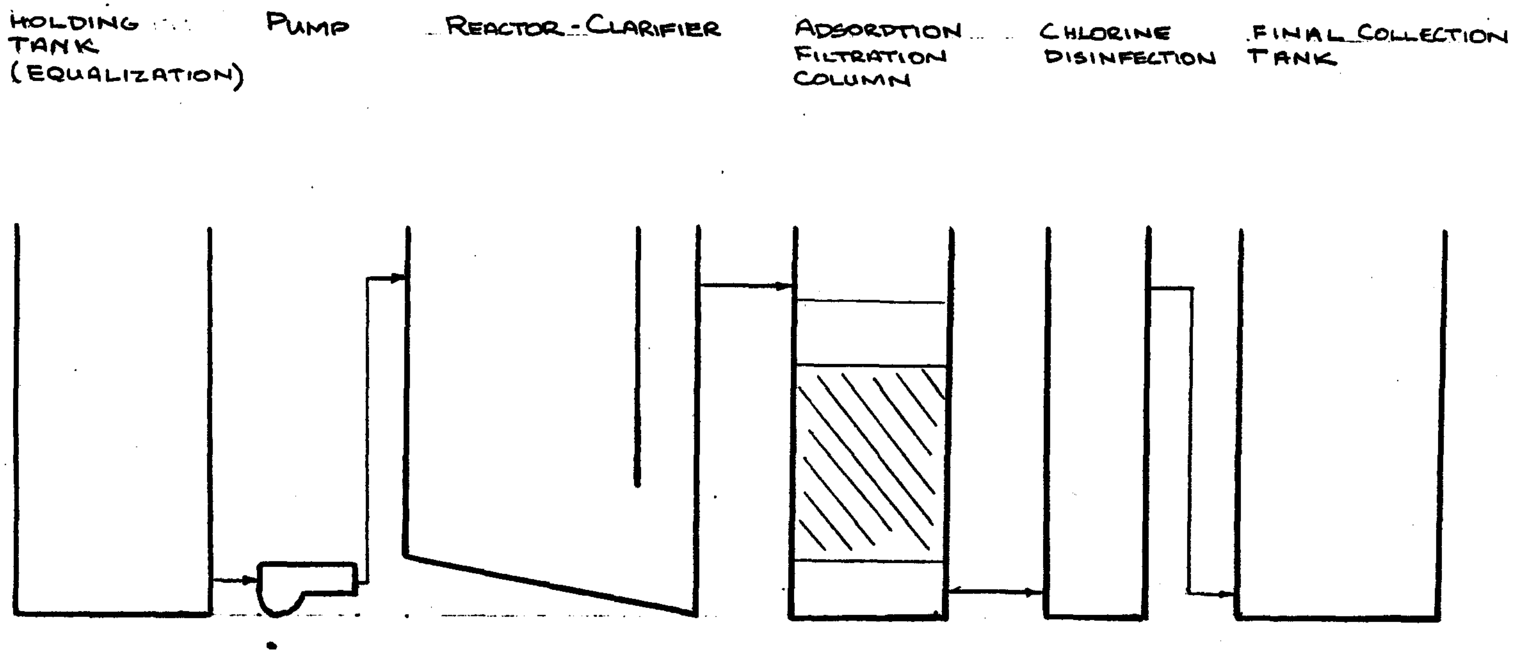


FIGURE 3-1 SCHEMATIC OF TREATMENT SYSTEM

(iv) a disinfection contact chamber
(v) a final storage tank

In an actual installation the equalization tank would be used to prevent surge flows from entering the treatment system. All household wastewater would be directed to the equalization tank. The contents of the tank would be pumped at a constant rate to the reactor-clarifier.

Within the reactor-clarifier the waste stream would be first exposed to biological treatment in the form of an extended aeration process. If necessary (i.e. at system start-up) coagulant could be added at the upstream end of the reactor. Mixing in the reactor would be provided by coarse air diffusers which would also supply the necessary oxygen for the biological activity. After passing through the reactor the waste stream would enter an integral clarifier which would contain inclined settling plates to improve the solids removal. The integral clarifier would be separated from the reactor by an underflow baffle. This arrangement would allow solids to return to the reactor and thereby provide the necessary sludge return. In addition, unlike the conventional separate clarifier unit, the rapid solids return from the integral clarifier should prevent the sludge from going anaerobic (Brucker, 1975).

After the reactor-clarifier the waste stream would enter an adsorption-filtration column. Since the column utilizes downflow hydraulics and activated carbon as its media, it would be capable of providing both filtration and organic adsorption. This obviously improves the water quality and again ensures proper treatment at start-up (i.e. at start-up the dissolved organics would pass through the relatively inactive biological reactor, but would be adsorbed by the activated carbon in the adsorption-filtration column).

Once the waste stream has passed through the adsorption-filtration column, disinfection would be necessary. This would be accomplished by a chlorine contact chamber which essentially is an upflow vertical tube in which chlorine is added at the bottom.

Finally, the treated water would be directed to a final collection tank so that it could be used within the household as toilet water.

It is felt the treatment system should be a package unit contained within a heated structure adjacent to one of the dwellings it serves. The minimum operating temperature within this structure would likely be 3-5°C, as this would prevent freezing, while minimizing energy consumption.

Finally, the system should be designed and manufactured in such a way as to facilitate installation. Design features such as common-wall construction, plastic piping, light-weight material, gravity flow, etc. should be used to their maximum benefit.

3.3 Pilot Plant

The pilot plant consisted of the same five components indicated in section 3.2, and was placed in a temperature-controlled room for testing. The design of each component is reviewed below. It should be noted that emphasis on design and testing was placed on the two main components of the treatment system; the reactor-clarifier and the adsorption-filtration column.

3.3.1 Design Parameters

The pilot plant was designed to treat sanitary waste typical of that produced by one man in an Arctic environment. Using the information obtained from the literature, the following design parameters were used:

(i) flow = 45 Igal (200 l)/day
(ii) COD concentration = 700 mg/l
(iii) SS concentration = 500 mg/l

3.3.2 The Equalization Tank

The equalization tank had a 75 Igal (340 l) capacity. This capacity was based on the highest flow which would pass through the pilot plant during the experimental period.

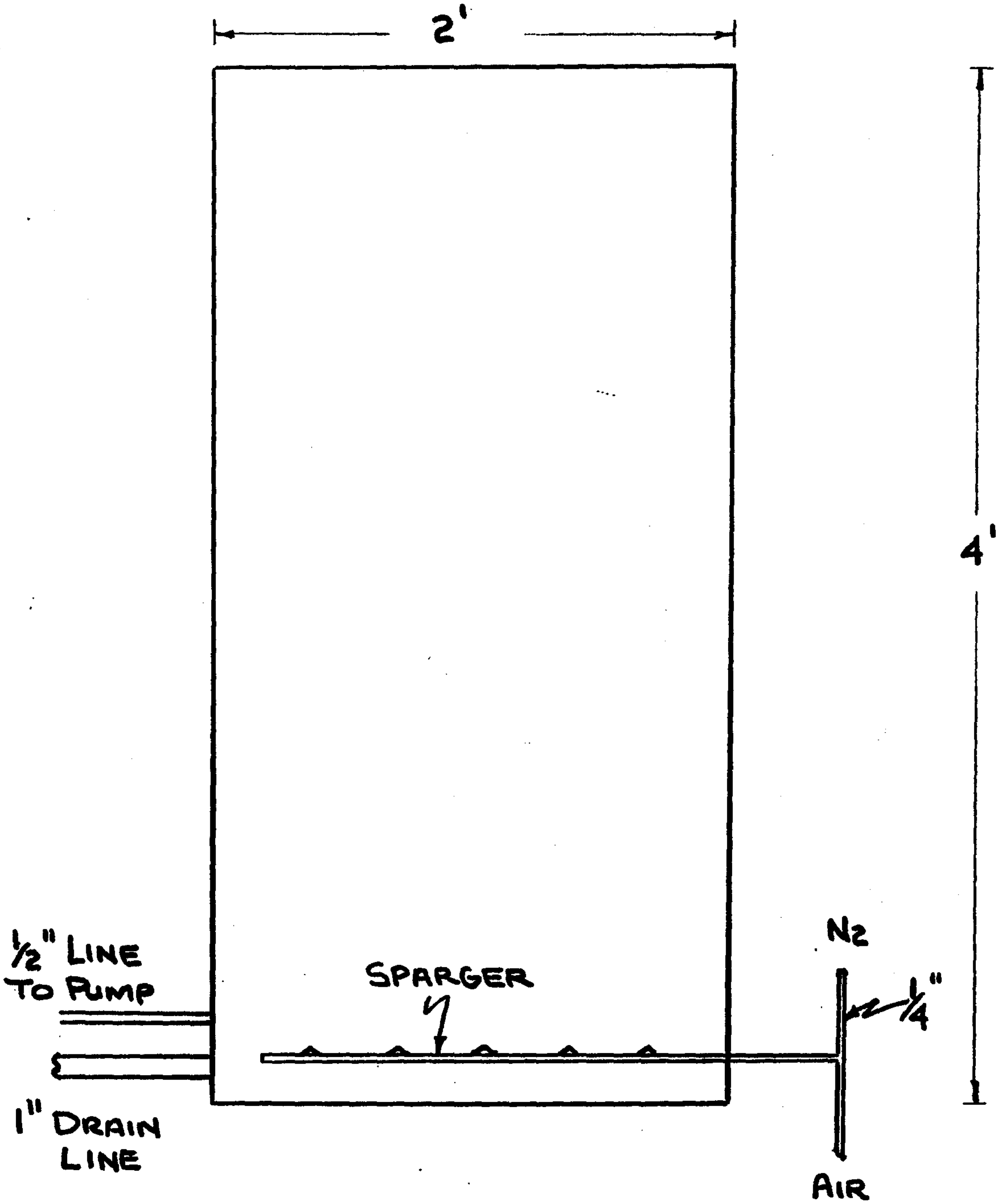


FIGURE 3-2 EQUALIZATION TANK

A peristallic pump with a Master Flex head (No. 7018) directed the contents of the equalization tank to the reactor-clarifier via 1/2 inch diameter tygon tubing.

The sewage within the equalization basin was mixed continuously to ensure a homogenous feed was directed to the treatment system. Mixing was provided by air and nitrogen gas at a nitrogen to air ratio of 1:8.

The nitrogen gas was fed from a nitrogen cylinder and used to minimize dissolved oxygen in the equalization tank's contents. This reduced the oxidation of organics prior to the influent entering the reactor-clarifier. In an actual installation air alone would be used to mix the contents of the equalization tank. In the present case nitrogen was used for mixing and only the amount of oxygen necessary to avoid septicity was supplied.

The equalization tank, as shown in Figure 3.2, had two 1/4 inch diameter gas inlets where nitrogen and air were mixed. In addition, a 1 inch diameter drain line was provided. All fittings were brass,and tygon tubing was used for all lines.

3.3.3 Reactor-Clarifier

When designing the reactor-clarifier the following five constraints were considered:

(i) Flow from the reactor-clarifier to the remaining components in the pilot plant system must be induced by gravity.

(ii) The air spargers in the reactor should be placed as far away from the integral clarifier as possible so as to minimize any disturbance to the clarification process. This meant the unit should be as long as possible.

(iii) The clarifier must be wide enough to allow for the placement of inclined settling plates.

(iv) To provide extended aeration treatment, the reactor should have a detention time of approximately 24 hours.

(v) The bottom of the tank should be inclined so that the sludge wastage line is at the deep end. This will encourage the sludge to concentrate before being wasted. Since the solids are being removed from the wastewater in the clarifier, the added depth and the sludge wastage were placed at the clarifier end of the tank.

As a result of these design constraints, the reactor-clarifier was sized as shown in Figure 3-3. The inside

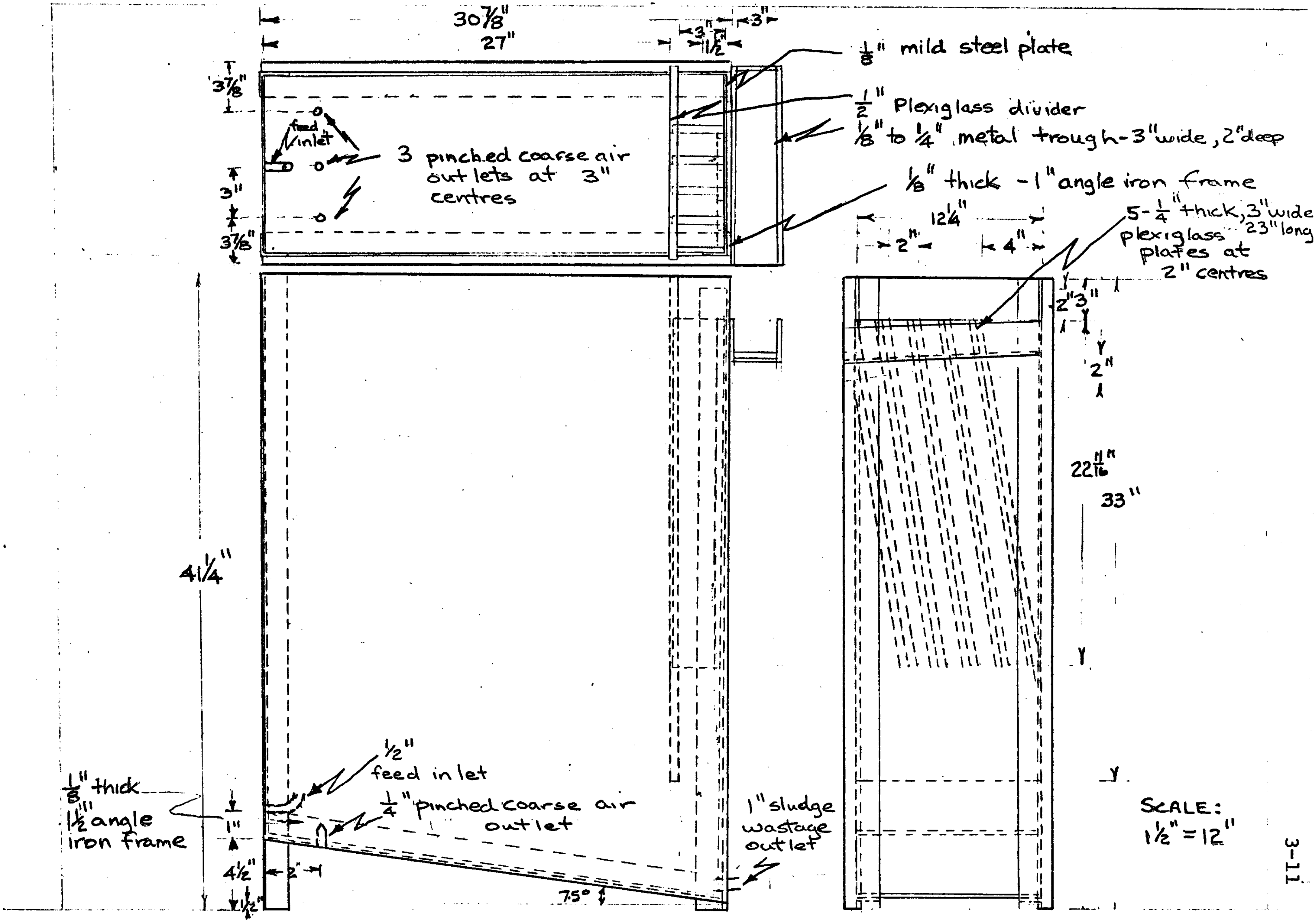


FIGURE 3-3 REACTOR - CLARIFIER TANK

dimensions were 27 inches long, 12 inches wide and 36.5 inches deep at the inlet end, and 40 inches deep at the clarifier end. The tank had a sloped bottom and a total volume of 46 Igal (207 l). The feed inlet was placed at the bottom of the upstream side and was turned up to reduce short circuiting in the reactor. A sludge wastage valve was placed immediately below the clarifier on the downstream side of the tank.

Mixing was provided by coarse air diffusers which were located on the bottom of the upstream half of the reactor as shown in Figure 3-3. The placement of the coarse air diffusers force the contents of the reactor to circulate in a clockwise direction and thereby encourage solids to be removed from the clarifier and returned to the reactor. A constant flow of 6000 cm^3/min was injected into the reactor. This figure was believed to be excessive to satisfy the reactor's oxygen requirements (i.e. estimates indicated that 4500 cm^3/min would be required to satisfy oxygen needs). However, this air flow provided sufficient circulation to ensure adequate sludge return from the clarifier.

The air source originated from the university's pressurized system and flowed through a coarse pressure regulator, a glass fibre filter, a fine pressure regulator and finally a rotameter before entering the reactor through three valved coarse bubble diffusers.

Coagulant (i.e. alum) was added to the reactor by a 1/4 inch copper line at the upstream end just above the wastewater feed inlet. A constant head bottle (Marriett bottle) was used to control the coagulant flow.

Three valves were installed on the feed line from the constant head bottle. One valve was used for closing the feed line when the constant head bottle was being filled. The remaining valves could then be left untouched while the bottle was being filled. The second valve was a needle point valve and was used to control flow. A 3/8 inch ASCO solenoid valve controlled by an "Eagle Signal" timer followed the needle point valve. The timer was set to open the solenoid valve for 2 sec every 10 minutes.

When designing the clarifier two parameters; overflow rate and detention time, were considered.

The clarifier detention times reported in the literature vary from 1 hour (Bayley, 1972) to 4 hours (Betze, 1973). A detention time of 2 hours was used in this design as it was expected to encourage adequate flocculant settling and yet not allow the clarifier to become anaerobic.

Design parameters concerning the installation of inclined settling plates also had to be considered. Generally settling plates are used to increase the allowable overflow rate and thus reduce the surface area required. In this particular case, settling plates were used to reduce the actual overflow rate to a very low value, i.e. 100 Igpd/ft^2. The reason for this was to improve the solids removal in the clarifier and thus lessen the burden of filtration on the adsorption-filtration column.

Culp (1974) stated that the actual surface area for a clarifier containing tilted plates was the total of the projected plate area in the horizontal plane.

After considering the various parameters, the clarifier was designed as shown in Figure 3-3. A baffle was placed 3 inches upstream from the clarifier's effluent outlet. The baffle extended 33 inches into the tank. With the position of the baffle the detention time within the clarifier was approximately 2 hours at the design flow.

The plates were 23 inches long and were placed at an inclination of 80° from the horizontal. This permitted the placement of 5 plates in the clarifier. As a result the settling area was doubled when compared to the same size clarifier without plates. At the design flow the overflow rate was 100 Igpd/ft^2 of available settling area (i.e. after adjusting for plate area).

A galvanized trough was used to collect the effluent from the clarifier and convey it to the adsorption-filtration column. The trough was designed for a maximum flow of 73 Igal (330 l)/day. The method of design (Weber, 1972) is shown in Appendix 3.

The bottom and ends of the reactor-clarifier were constructed from 1/8 inch steel plate and were treated with epoxy resin. The sides were constructed from 3/4 inch plexiglass. All fittings were brass, tygon tubing was used for the air feed and the 1 inch diameter sludge drain line, and pinched copper piping for the coarse air diffusers.

3.3.4 The Adsorption-Filtration Column

The adsorption-filtration column was a downflow column (i.e. the waste stream enters at the top and exits at the bottom). The column's filtering media was activated carbon followed by a graded support filter.

The activated carbon was a product of Calgon Corporation (Pittsburgh, Pennsylvania) called Filtrasorb 400. This type of carbon has been used in numerous applications for domestic wastewater treatment (e.g. Lake Tahoe, Pomona, California). The properties of Filtrasorb 400 are shown in Table 3.1.

As with the reactor-clarifier the adsorption-filtration column was designed for a flow of 45 Igal per day and a COD feed concentration of 700 mg/l. When designing the adsorption-filtration column the following constraints had to be considered:

(i) there was a height restriction on the column to ensure that gravity flow directed the clarifier's effluent to the adsorption-

Table 3-1

Specification of Filtrasorb 400

	Units	Properties of Filtrasorb 400	Reference
1. Raw Material		Bituminous Coal	(Benedek, 1974)
2. Surface Area	m^2/g	950-1050	(Benedek, 1974)
3. Backwashed and Drained Density	lb/ft^3	26	(Benedek, 1974)
4. Average Particle Size	mm	.8-.9	(Benedek, 1974)
5. Uniformity Coefficient		1.9	(Benedek, 1974)
6. D_{10}	mm	.7	(Culp, 1974)
7. Approximate loading for municipal waste	$\frac{\text{lb of COD}}{\text{100 lb of carbon}}$	80	(Benedek, 1975)

filtration column.

(ii) the hydraulic loading on the carbon column should not be more than 4 Igpm/ft^2 (Benedek, 1974).

(iii) there should be sufficient carbon to remove approximately 75% of the dissolved COD in the column's feed. In addition, the carbon should be capable of providing adequate dissolved COD removal over an extended period of time (i.e. 1 year). Benedek (1975) reports that approximately 80 lb of COD/100 lb of carbon can be used as a design parameter when using activated carbon for municipal sewage treatment.

Upon consideration of these constraints a 10 inch diameter column with a 12 inch deep activated carbon bed was employed as the adsorption-filtration column (see Figure 3-4).

The activated carbon layer was supported by a gravel filter. This filter provided a constant permeability for the flow of water through the column, but at the same time prevented activated carbon particles from being washed out of the column.

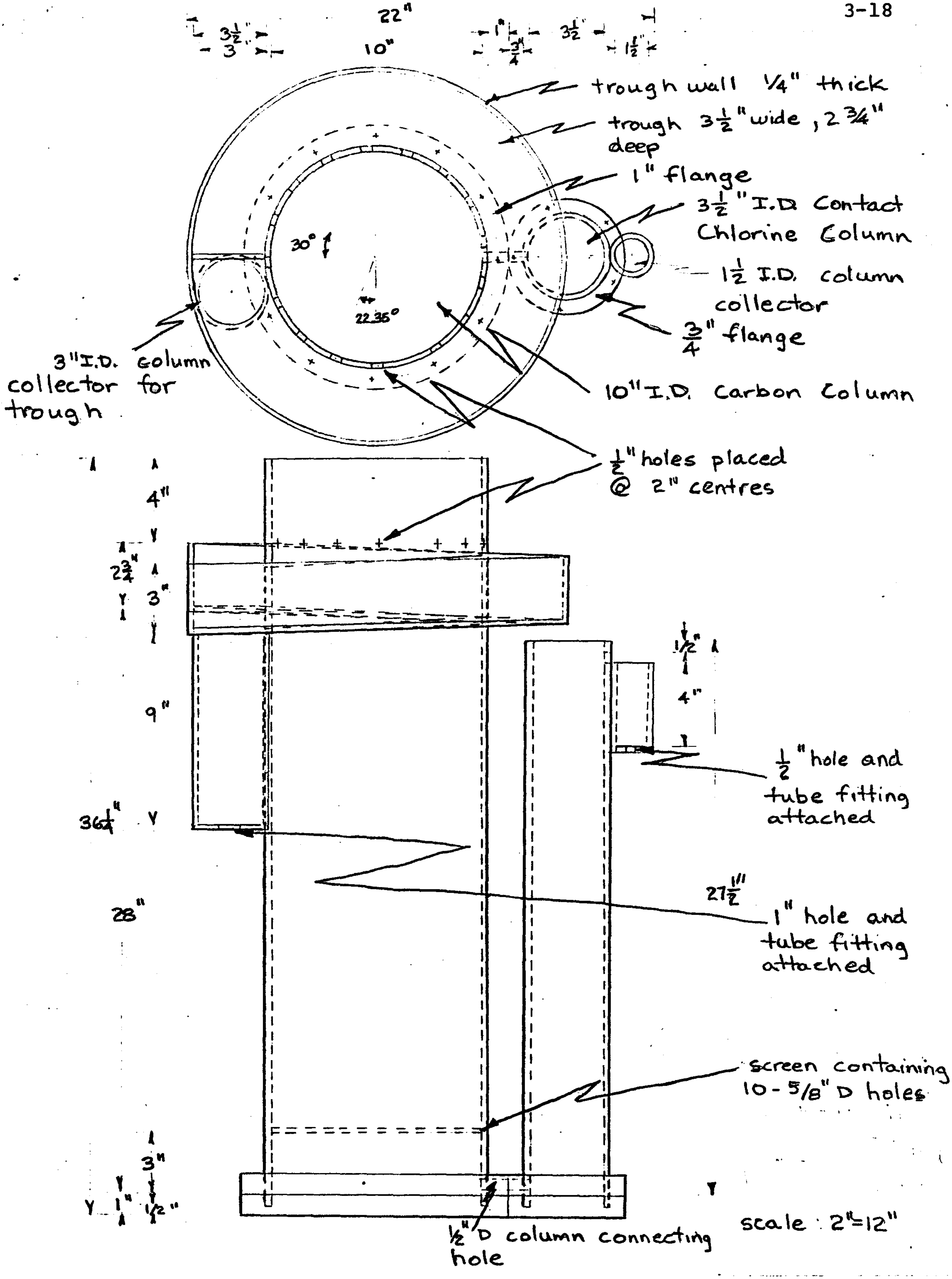


FIGURE 3-4 CARBON COLUMN & CHLORINE CONTACT

DB

The support filter was composed of two layers, a 3" thick fine filter and a 4" thick coarse filter. The design constraints for each layer were as follows:

a) Piping prevention: the grain size of the filter being designed should be small enough to prevent particles from the upstream layer washing through the system

$\therefore$ D_{15} of filter ≥ 5 D_{85} of upstream media.

b) Permeability guarantee: particles of the filter being designed should be large enough to encourage adequate percolation of water

$\therefore$ D_{15} filter ≥ 5 D_{15} carbon

Knowing the characteristics of Filtrasorb 400, a design for a supporting filter, based on the above criteria, was completed as shown in Appendix 4. The results are shown in Table 3.2 and are plotted on the grain size chart in Figure 3-5.

The coarse filter is supported by a filter plate with 10-1/2 inch diameter orifice openings in the plate. The design of the plate ensured that the orifice openings were half the D_{85} of the supported media.

The 10 inch diameter column provided a surface area of .55 ft^2. This resulted in a hydraulic loading of

Table 3-2

Support Filter Requirements

Fine Filter $D_{15} = 4.0$ mm = 5/32"
$D_{85} = 12.0$ mm = 15/32"

Uniformity coefficient = 2.3

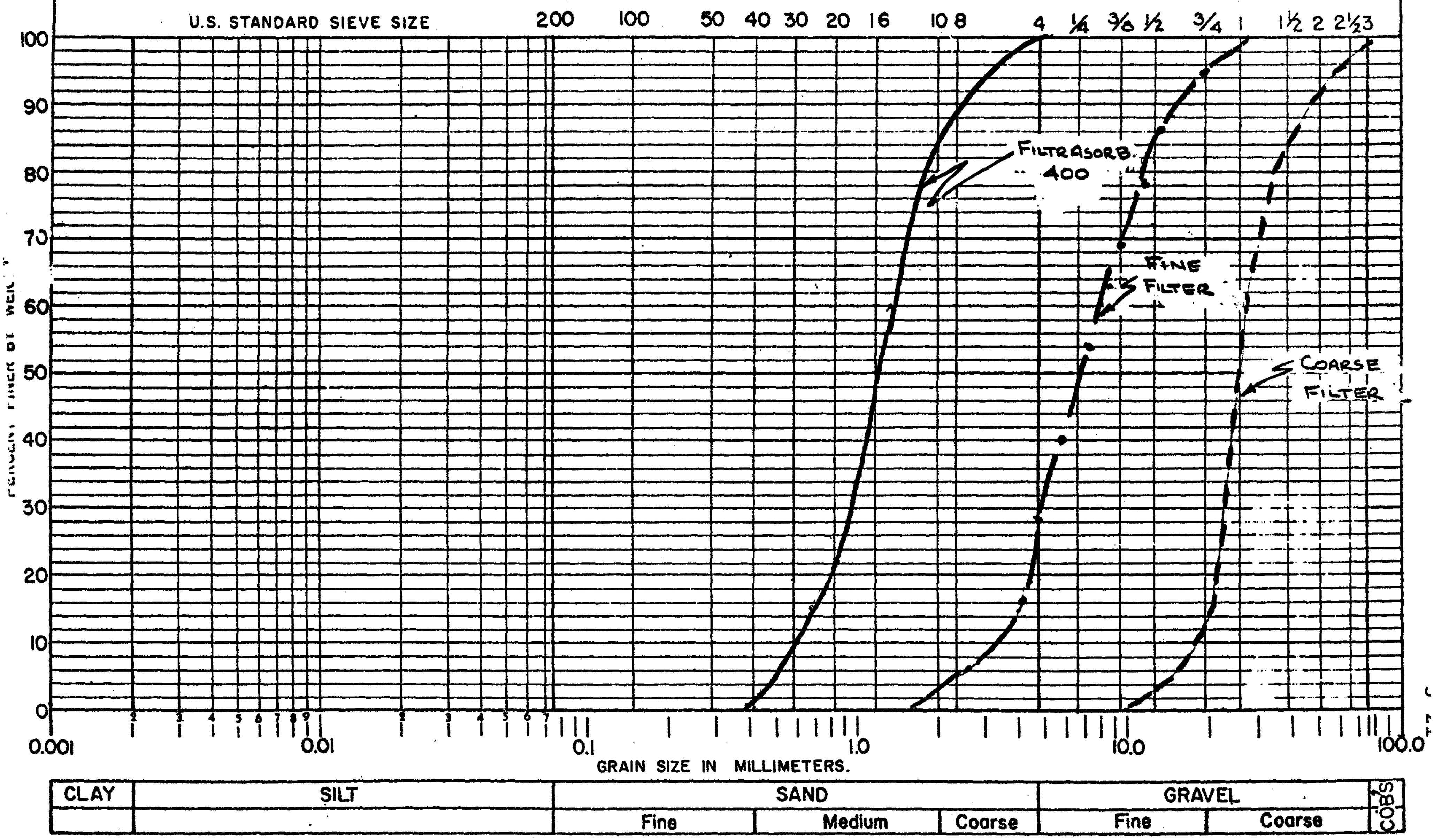
Coarse Filter $D_{15} = 20$ mm = 25/32"
$D_{85} = 30$ mm = 15/32"

Uniformity coefficient = 1.23

GRADING CURVES

PROJECT

Figure 3-5 Grain Size Chart For Adsorption - Filtration Column



.06 Igpm/ft^2. It should be noted that this loading was much less than the suggested maximum value of 4.0 Igpm/ft^2.

The backwash flowrate required to obtain 50 percent expansion of the column's media was calculated to be 10.8 Igpm/ft^2 (see Appendix 5). With a column surface area of .55 ft^2, a backwash flowrate of 6.0 Igpm was required.

The backwash water was collected in a trough which completely surrounded the adsorption-filtration column. This trough was designed on a similar basis as the trough designed for the clarifier effluent.

Attached to the downstream end of the trough was a collector for the backwash water. The purpose of the collector was to provide sufficient head so that 6.0 Igpm would flow through the backwash drain line. The result was a 2.5 inch diameter cylindrical collector, 9 inches deep (see Appendix 6).

3.3.5 Chlorine Contact Chamber

The design of the chlorine contact chamber was based on the following three criteria:

(i) provide a 30 minute detention time for the design flow of 45 Igal/day

(ii) encourage plug flow

(iii) place the outlet from the chlorine column at a height which ensures a 4" layer of water above the activated carbon media in the adsorption-filtration column (see Figure 3-4). The purpose of the layer of water was to enhance equalized hydraulic loading on the column's surface and thereby prevent channelling.

As a result of these three criteria, a 3.5 inch diameter cylinder with the outlet 26 inches above the base of the adsorption-filtration column was constructed. Sodium hypochlorite was added to provide a chlorine (Cl) concentration of 15 mg/l. The disinfectant was fed to the bottom of the cylinder by 1/4" copper tubing and the flow was regulated by a constant head bottle arrangement similar to that used for coagulant addition (see section 3.3.3).

3.3.6 Final Collection Tank

The final collection tank was used to collect a composite sample of the run's final effluent.

This tank had a storage capacity of 27 gallons and was equipped with a feed line, a valved drain line and a valved effluent line.

All drain lines emanating from any of the system's components were connected to a 1" diameter manifold. From the manifold the plant's discharges were directed to a building drain.

CHAPTER 4

4.0 EXPERIMENTAL

4.1 Apparatus

The apparatus used during the experimentation was as described in section 3.3. Any analytical apparatus will be identified in the forthcoming sections.

4.2 Procedure

The apparatus described in Chapter 3 was operated on a batch daily basis. The daily experimental procedure is shown in Table 4-1. In general, it consisted of obtaining primary sludge from the Dundas, Ontario sewage treatment plant, preparing the system's daily feed solution, taking samples from the various components, and analysing the samples for the various parameters.

Each experimental run lasted four days. On the fifth day the system was prepared for the next run and analyses of that week's run were completed. The pilot plant was then shut down for the remaining part of the week (i.e. 3 days).

Table 4-1

Daily Testing Procedure for Each Run

<u>DAY 1</u>	8:00 a.m.	Obtain 10 gal of primary settled sewage from Dundas Sewage Treatment Plant.
	8:30 a.m.	After macerating primary sludge in blender, prepare three dilutions of waste 30:1, 60:1 and 90:1.
	9:00 a.m.	Prepare COD solutions according to Standard Methods, i.e. 3 dilutions and a control.
	9:30 a.m.	Place COD solutions on digestion equipment.
	10:00 a.m.	Weigh and add appropriate amount of yeast extract.
	10:30 a.m.	Check treatment system.
	11:00 a.m.	Take sample of reactor's contents and measure MLSS and MLVSS.
	11:30 a.m.	Turn off COD Digestion Equipment and allow COD solutions to cool.
	1:00 p.m.	Measure COD's. Add appropriate amount of yeast and primary sludge to system's feed tank and dilute to proper volume of feed (see Appendix 8 for yeast ccalibration test).

Table 4-1
(cont'd.)

	1:15 p.m.	Start system's feed.
	1:30 p.m.	Take sample of system's feed.
	2:00-4:30 p.m.	Measure COD and SS according to Standard Methods.
<u>DAY 2</u>	8:00 a.m.	Same as Day 1.
	8:30 a.m.	Same as Day 1.
	8:45 a.m.	Obtain sample from system's clarifier to measure previous day's COD and SS removal in reactor-clarifier (approximately 19 hr after start of previous day's feed)
	9:00 a.m.	Same as Day 1. In addition to the three dilutions, measure COD of sample taken at 8:45.
	9:30-1:30 p.m.	Same as Day 1 with the exception that at 1.00 p.m. take a sample of the end of the previous day's feed from the equali-ation tank.
	1:30 p.m.	Take sample of this day's feed from feed tank and previous day's final effluent from final collection tank.

Table 4-1
(cont'd.)

	2:00-4:00 p.m.	Measure COD of samples taken at 1:00 p.m. and 1:30 p.m., and measure SS of all samples taken. Clarifier effluent and final effluent are stored in a refrigerator for DCOD measurements, which are done on Day 5.
<u>DAY 3</u>	8:00 a.m. - 4:30 p.m.	Same as Day 2.
<u>DAY 4</u>	8:00 a.m. - 4:30 p.m.	Same as Day 2.
<u>DAY 5</u>	8:45 a.m.	Same as Day 2.
	10:00 a.m.	Prepare for operation of technicon autoanalyser which is used to analyse DCOD.
	1:00 p.m.	Take samples of previous day's feed, final effluent and reactor contents.
	1:30 p.m.	Shut down system completely except for diffused air.
	1:45 p.m.	Measure COD of feed, clarifier effluent and final effluent and SS of all samples taken during the day. In addition, measure MLVSS of reactor's contents.

Table 4-1
(cont'd.)

2:00 p.m.	Use technicon autoanalyser to measure DCOD from day 1-4 clarifier effluent and final effluent. In addition, one day's feed in each run was measured for DCOD.
5:00 p.m.	Waste sludge from reactor if necessary.

4.3 Analyses

4.3.1 Parameters Monitored

Daily plant performance was measured by three parameters: chemical oxygen demand (COD); dissolved chemical oxygen demand (dissolved COD), and suspended solids (SS). Mixed liquor suspended solids (MLSS) and mixed liquor volatile suspended solids (MLVSS) were measured at the beginning and end of each run. In addition, dissolved oxygen (DO) was measured at least once each run.

COD was measured as an indicator of the relative oxygen demand of the pollutants in the wastewater. By measuring the reduction of COD between system components, the treatment capabilities of the component could be evaluated. Hence COD's were measured in the feed, clarifier effluent and composite collection tank.

Dissolved COD was measured to rate the performance of the activated carbon and the biological component within the system.

Daily measurements of dissolved COD were made in the adsorption-filtration column influent and effluent so that an accurate performance study could be made on the activated carbon.

Suspended solids (SS) were measured to give an indication of solids removed by each of the system's components and evaluate the system's total production of sludge residual.

For this reason solids were measured daily at the same location as COD's and also in the reactor before and after each run. Solid measurements in the reactor provided values for the MLSS and the MLVSS. This in turn gave measurements of solid production and micro-organism growth. Suspended solids were measured in the backwash from the adsorption-filtration column for the last four runs. This allowed a mass balance to be carried out on the treatment system.

Finally, DO levels were measured at least once each run in the reactor and the influent to the adsorption-filtration column. These measurements were performed to:

(i) check on the DO level in the reactor

(ii) check whether septicity was occurring in the adsorption-filtration column

(iii) evaluate the effect of the clarifier's collection trough in reaerating the clarifier's effluent prior to it entering the adsorption-filtration column

4.3.2 Method of Analyses

COD's were measured according to Standard Methods and dissolved COD's were measured on the Technicon autoanalyser according to the procedure suggested in the Technicon autoanalyser methodology report number 27-69W.

SS and MLVSS were also measured according to Standard Methods. A combination of a Fiberglas prefilter and a .45μ Membran filter were employed with a vacuum bottle to provide the necessary filtration of the various samples.

Once the filtration was completed, the media were placed in crucibles and then put in an oven for drying at 103°C. If MLVSS were measured the media were volatized in an oven at 600°C.

A filter ash loss test was carried out prior to any experimentation. The purpose of this test was to check on the weight loss in the filters and the crucibles during the drying and volatizing procedures. The test and its results are discussed in Appendix 13.

DO was measured with a portable Yellow Springs Instruments, Model No. 54, dissolved oxygen meter. The procedure followed was according to the operator's manual provided with the instrument.

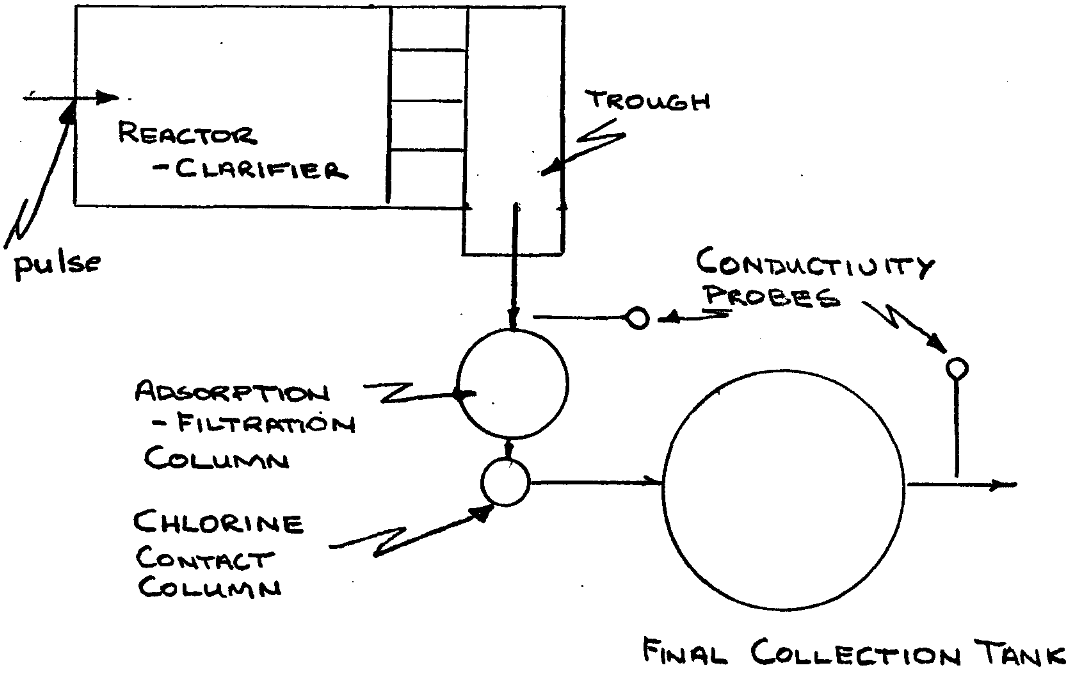
4.3.3 Residence Time Distribution

A residence time distribution (RTD) allows one to assess the general flow characteristics within a reactor. It is generally performed by adding a distinct pulse at the influent to the reactor and then monitoring that pulse as it exits from the reactor.

In this particular project RTD's were performed to study the flow characteristics of the overall system and especially the reactor-clarifier and the adsorption-filtration column.

Two different set-ups were required to produce the RTD's mentioned above. The first set-up utilized two charted conductivity meters. The first meter was placed immediately after the trough attached to the clarifier and produced an RTD for the reactor-clarifier. The second was placed at the effluent outlet of the collection tank and produced an RTD for the overall system.

Figure 4-1 First Residence Time Distribution Set-up



A second set-up was required to establish an RTD for the carbon column. A charted meter was placed at the base of the chlorine contact column so that conductivity of the carbon column's effluent could be measured.

Various impulse concentrations, volumes and places of injections were experimented with in the first set-up. Concentrations of salt such as 1,000 mg/litre and 10,000 mg/litre with volumes of 25 ml and 50 ml were injected into the reactor's tygon feed line by syringe. However, the mixing and dilution effect of the reactor made it impossible for the meters to produce a reading of any significance. It was felt that with the 46 Igal (207 litres) in the tank, and assuming total dilution, an impulse volume of 1 litre and a salt concentration of 200,000 mg/litre would produce a reasonable signal. However, because of the length of time to inject one litre into the feed line by syringe, the litre of salt solution was placed in a covered container and lowered to the bottom of the reactor at the upstream end (i.e. at the influent inlet). The cover was removed and the litre of salt solution was released immediately into the reactor.

In the second set-up 500 ml of 20,000 mg/litre salt solution was dumped into the outlet trough off the clarifier. The impulse immediately entered the adsorption-filtration column and the conductivity of the column's effluent was measured producing an RTD for the column. (see section 5.1 for results).

4.3.4 Activated Carbon Isotherms

Following the actual testing of the system, isotherms were prepared on the activated carbon to establish the exhaustion of the carbon's adsorptive capacity.

An isotherm was prepared on the clean unused carbon identical to that used in the system and on the used carbon obtained from the system's adsorption-filtration column. The procedure followed was the same as that recommended by Benedek (1974).

The difference in the two isotherms indicated the reduction in the used carbon's adsorptive capacity. (See section 5.2.3.2 for results.)

CHAPTER 5

5.0 RESULTS AND DISCUSSIONS

5.1 Residence Time Distribution (RTD) Studies

As indicated in Section 4.3.3, a residence time distribution was performed on the overall system, the reactor-clarifier and the adsorption-filtration column.

The results of the RTD are tabulated and plotted in Appendix 9. The RTD vessel dispersion numbers are shown in Table 5-1.

The results indicate that:

(i) the reactor-clarifier was providing immediate mixing of the influent and was closer to operating as a completely stirred tank reactor (CSTR) than a plug flow reactor. This is evident from the high dispersion number (see Levenspiel, 1972 - Figure 9-13).

(ii) the components after the reactor-clarifier encouraged plug flow. This is indicated by the similarity in the RTD curve emanating from the reactor-clarifier and the curve obtained from the system's final effluent. In addition, the dispersion numbers for

Table 5-1

Residence Time Distribution - Summary

	*Dispersion Number
Overall	.20
Reactor-clarifier	.18
Adsorption-filtration column	.05

*Dispersion numbers are a dimensionless group which measure the extent of axial dispersion. The dispersion number equals D/(UL) where "D" is the axial dispersion coefficient, "U" the velocity of the fluid, and "L" the length of the reactor. An increasing dispersion number quantitatively represents an increasing deviation from plug flow behaviour.

these two curves are similar, which indicates there was little dispersion of the stream after the reactor-clarifier.

(iii) the adsorption-filtration column exhibited plug flow behavior. This is evident from the small dispersion number.

5.2 Experimental Plan

The system's operational capabilities were evaluated for the following five conditions:

(i) high-low pollutant concentration
(ii) high-low flow
(iii) coagulant and no coagulant addition
(iv) high-low temperature
(v) effectiveness at start-up after a system shut-down

As shown in Table 5-2 the first four runs were designed to study the first two conditions given above. To ensure that the mass loading parameter, which is dependent on both flow and concentration, did not interfere with the evaluation of these conditions, all four variations of the two parameters were tested.

The high-low values used for flow were 45 Igal (200 l)/day and 73 Igal (330 l)/day, and for COD concentrations, 550 mg of COD/l and 1,100 mg of COD/l. These values were based on information obtained from literature, as shown in Section 2.4.

Alum was added to the reactor in the first four runs at a rate of 150 mg/l. This alum dosage was developed through jar tests which are described in Appendix 14.

After the first four runs, flow and COD concentrations were held at the low values for the remaining runs. In Run 5 coagulant addition was suspended. This allowed one to assess the effect of not using a coagulant after the system was in operation for a reasonable length of time, and the biological activity was at a peak level. If the system operated well, Runs 6 and 7 would also be run without coagulant.

Run 6 was used to study the effect of temperature. As a result the temperature of the system was raised to 15-20°C.

Finally, Runs 7 and 8 were utilized to check the effect of a seven-day system shutdown on treatment efficiency (i.e. the system was left untouched for seven days between Runs 7 and 8). In Run 8 coagulant was added, but at a reduced rate as compared to the previous runs (i.e. 75 mg/l instead of 150 mg/l). This permitted an evaluation of the biological activity at start-up which was available to compensate for the reduced coagulant dosage.

Table 5-2

General Schedule

July 24/75 - run system with COD around 700 mg/l for four days

July 28/75	run 1	COD	1200 mg/l
		Flow	73 Igal (330 l)/day
		Temperature	4°C
Aug. 5/75	run 2	COD	400-500 mg/l
		Flow	45 Igal (200 l)/day
		Temperature	4°C
Aug. 11/75	run 3	COD	1200 mg/l
		Flow	45 Igal (200 l)/day
		Temperature	4°C
Aug. 18/75	run 4	COD	400-500 mg/l
		Flow	73 Igal (330 l)/day
		Temperature	4°C
Aug. 25/75	run 5	COD	400-500 mg/l
		Flow	45 Igal (200 l)/day
		Temperature	4°C
		*no coagulant added	
Sept. 2/75	run 6	COD	400-500 mg/l
		Flow	45 Igal (200 l)/day
		Temperature	15°C
Sept. 8/75	run 7	COD	400-500 mg/l
		Flow	45 Igal (200 l)/day
		Temperature	4°C

7 DAY SYSTEM SHUTDOWN

Sept. 19/75	run 8	COD	400-500 mg/l
		Flow	45 Igal (200 l)/day
		Temperature	4°C

It should be noted that prior to the actual testing of the data, a preparatory run was carried out on the pilot plant. This preparatory run fed varying COD concentrations and flows to the system for four days. The purpose of the run was to assess and correct any operational problems within the system. In addition, it ensured that all test runs would not encounter virgin activated carbon.

5.3 System Performance

All data recorded during the testing of the system are tabulated in Appendix 10 (see Table A10-1).

5.3.1 Overall System

5.3.1.1 COD and SS Removals

The overall system's performance (i.e. the system's influent compared with the final effluent) is displayed in Figures 5-1 through 5-4 and Tables 5-3 to 5-7.

The results indicated that:

(i) generally, although there were different influent COD and SS concentrations and flow rates, there were no differences in the final effluent with regard to the parameters mentioned. This observation is displayed in Table 5-7 where the statistical difference in the measured parameters of the final effluent were compared for various runs. As one can see, in most of the run comparisons there was no difference in the quality of the final effluent.

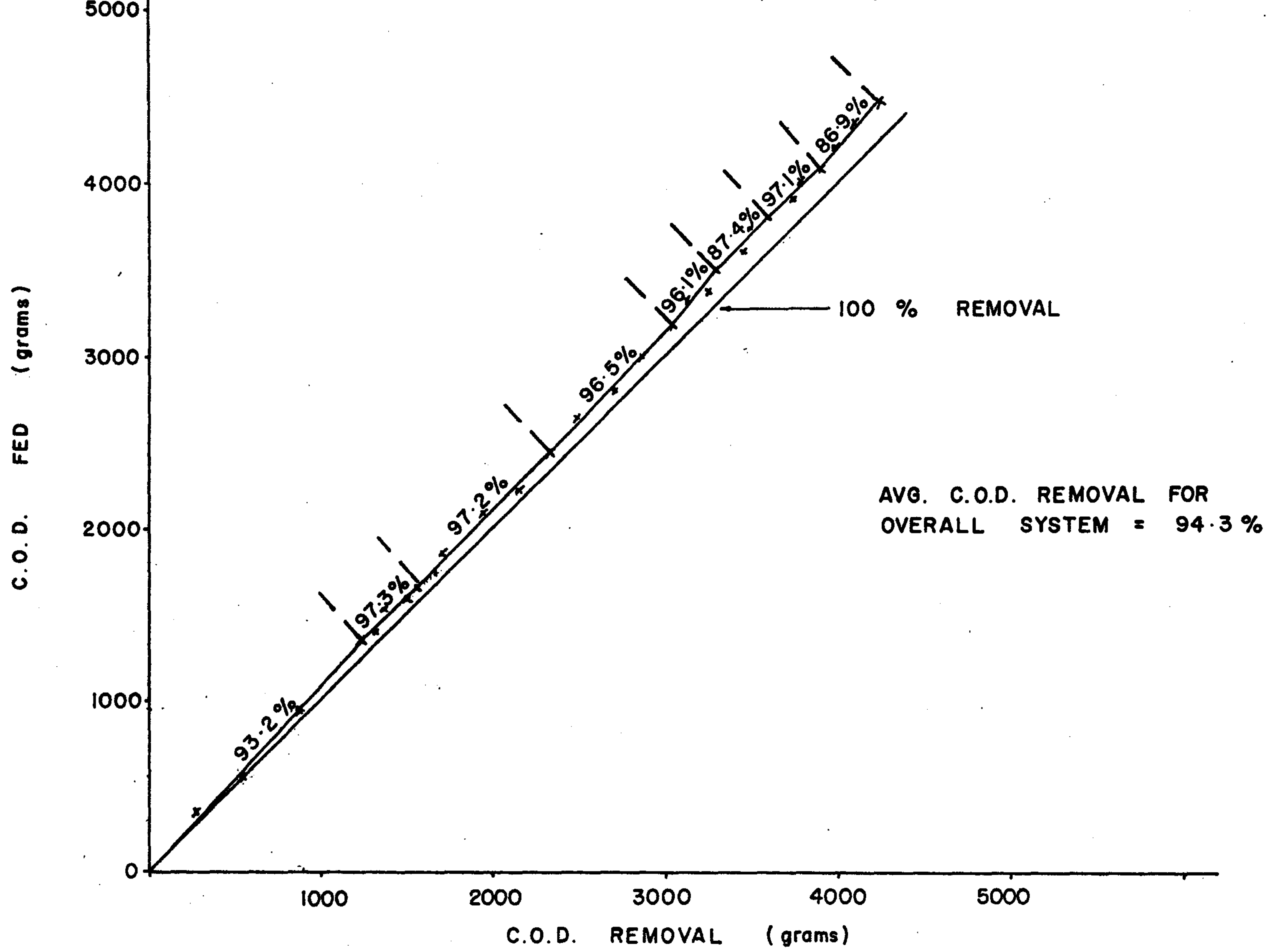
(ii) the average COD removal for the overall treatment system was 94.3 percent, with an average COD concentration in the final effluent of 37.2 mg/l.

(iii) the average SS removal for the overall system was 96.4 percent, with an average SS concentration in the final effluent of 13.8 mg/l.

(iv) 81 percent of all COD concentrations in the final effluent were below the "not to exceed" objective of 50 mg/l and 41 percent of the COD concentrations were below the "desired" objective of 20 mg/l.

(v) 78 percent of all SS concentrations in the final effluent were below the "not to exceed" objective of 15 mg/l and 34 percent of all final effluent SS concentrations were below the "desired" objective of 5 mg/l.

FIGURE 5-1
C.O.D.
AMOUNT FED vs. AMOUNT REMOVED
(OVERALL)



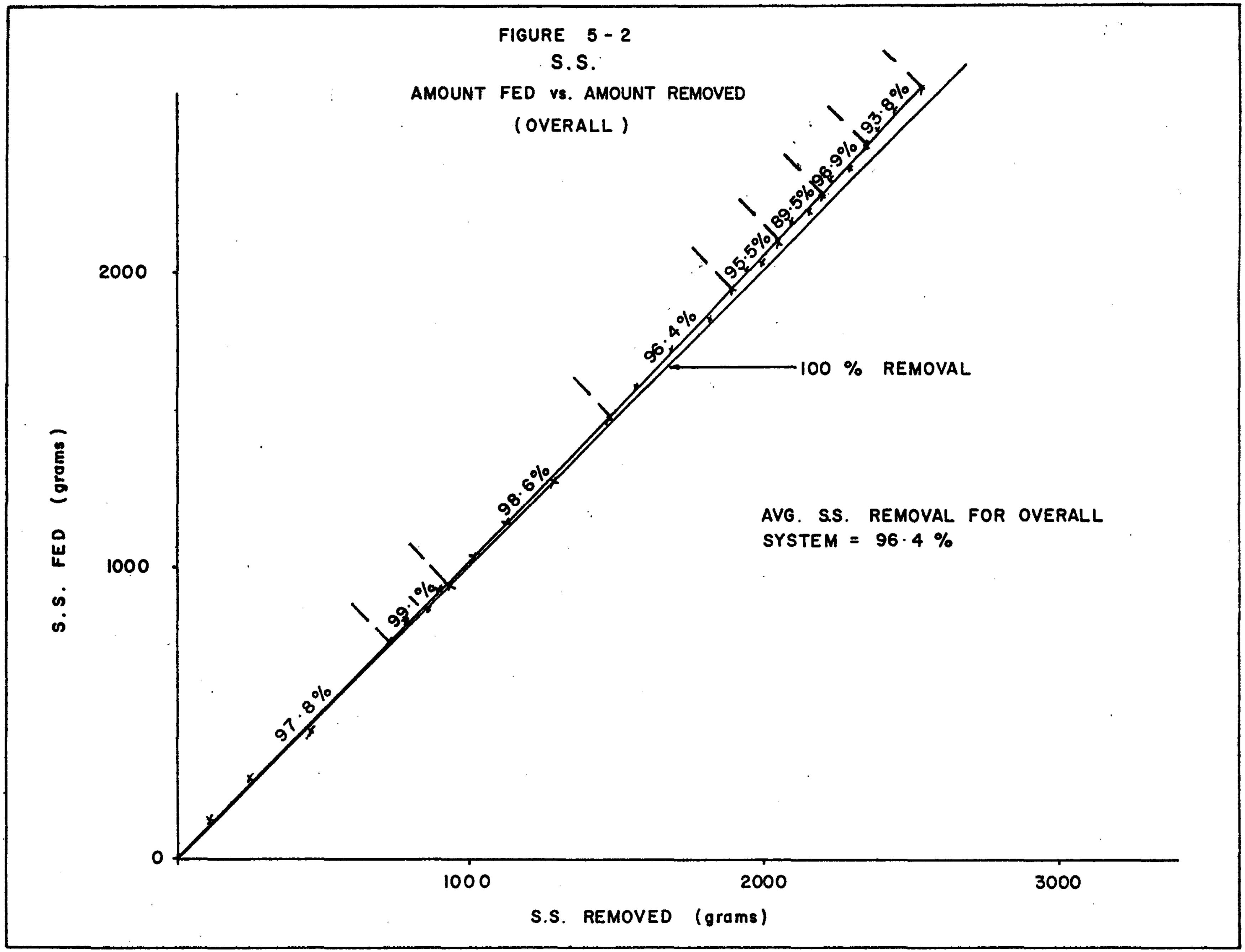
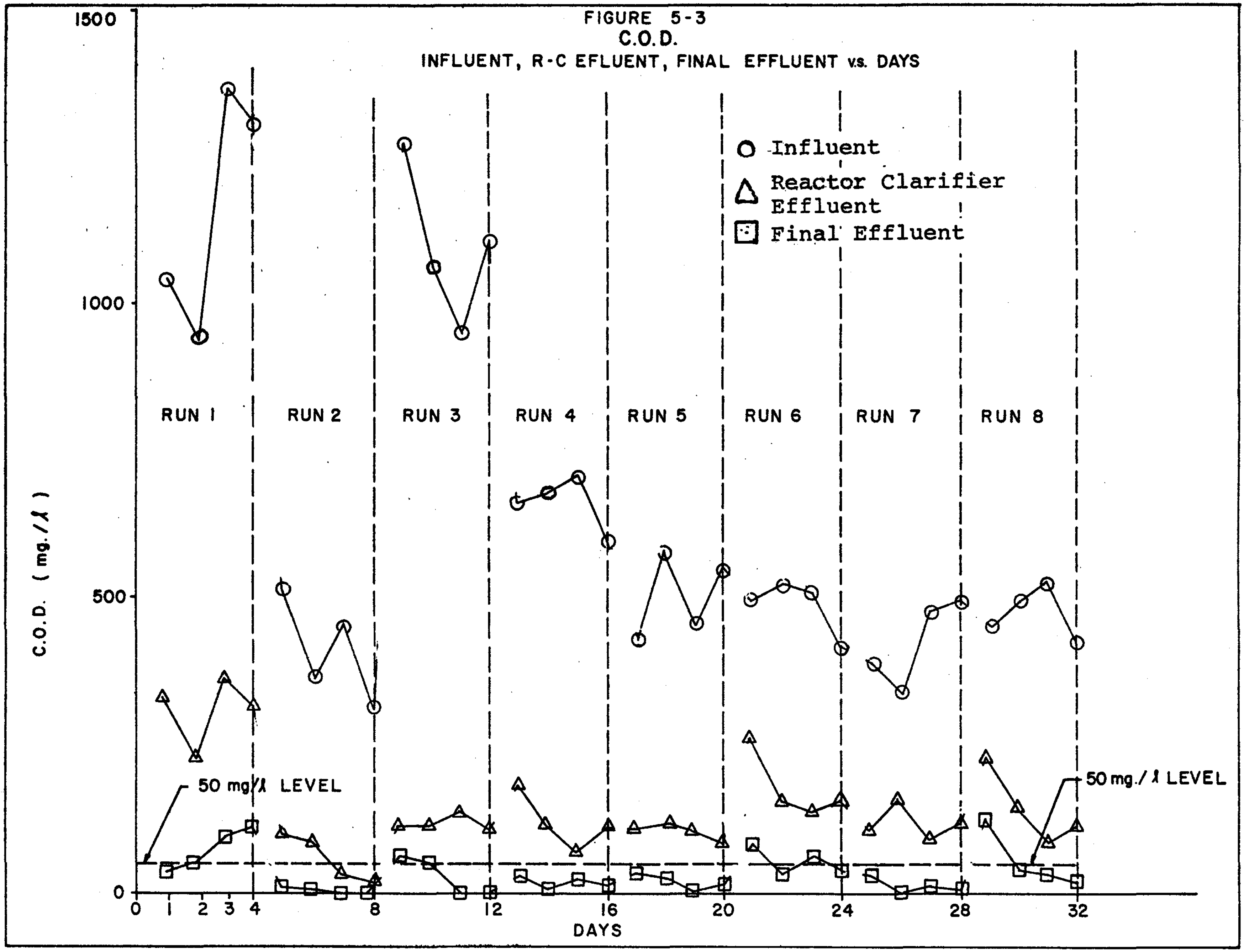


FIGURE 5-2
S.S.
AMOUNT FED vs. AMOUNT REMOVED
(OVERALL)

AVG. S.S. REMOVAL FOR OVERALL SYSTEM = 96·4 %

FIGURE 5-3
C.O.D.
INFLUENT, R-C EFLUENT, FINAL EFFLUENT vs. DAYS



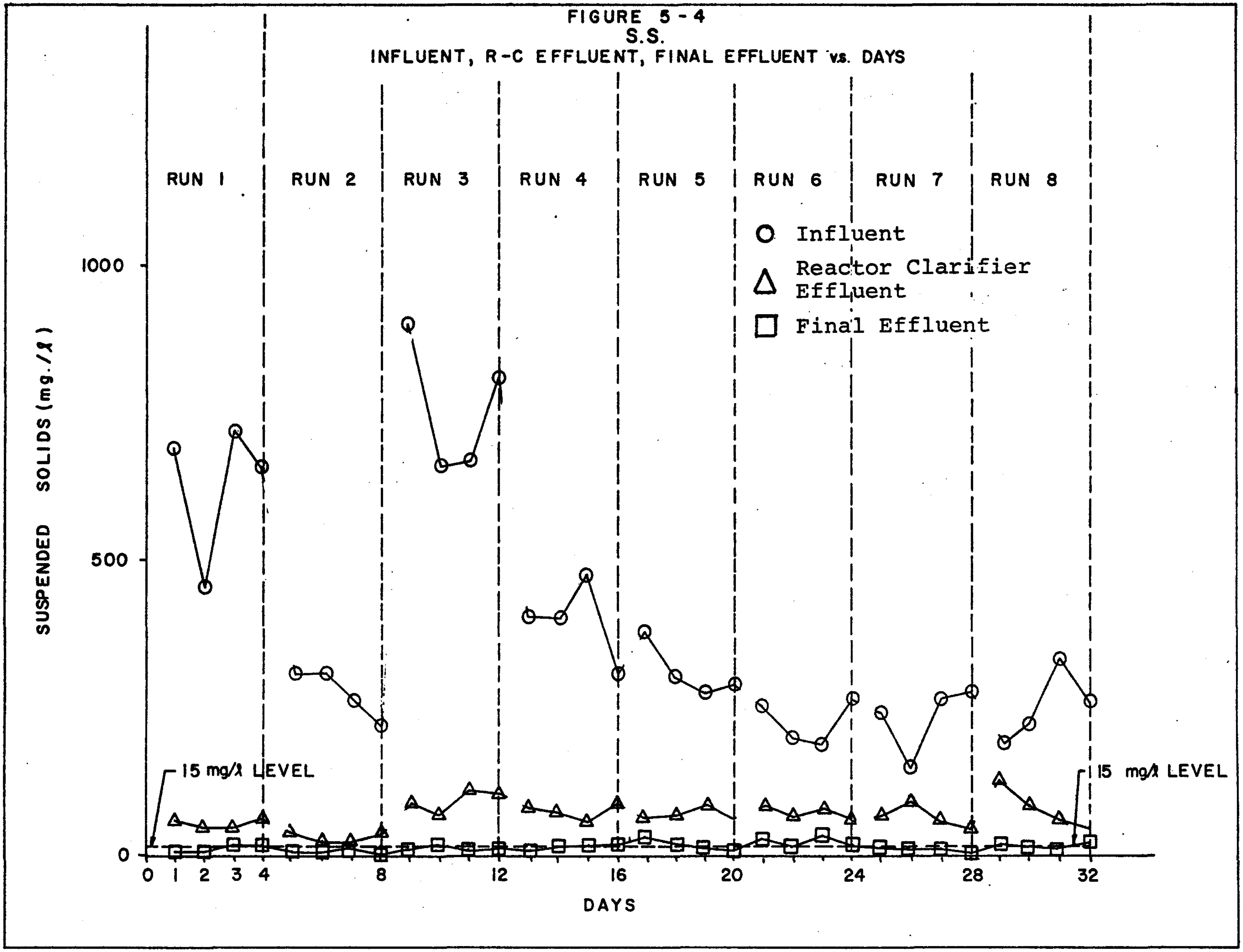


FIGURE 5-4
S.S.
INFLUENT, R-C EFFLUENT, FINAL EFFLUENT vs. DAYS

Table 5-3

Arithmetic Means and Standard Deviations of Data

	SUSPENDED SOLIDS			COD			DISSOLVED COD		
	mean ($\bar{x}$)	standard deviation (σ)	$\frac{\sigma}{\bar{x}}$	mean ($\bar{x}$)	standard deviation (σ)	$\frac{\sigma}{\bar{x}}$	mean ($\bar{x}$)	standard deviation (σ)	$\frac{\sigma}{\bar{x}}$
Influent	386 mg/l	195 mg/l	.505	656 mg/l	302 mg/l	.460	-	-	-
Reactor-clarifier Effluent and Adsorption-filtration Column Influent	68 mg/l	24 mg/l	.353	143 mg/l	82 mg/l	.580	32 mg/l	17 mg/l	.53
Final Effluent	13.8 mg/l	8.8 mg/l	.638	27.3 mg/l	34 mg/l	.914	7.6 mg/l	3.1 mg/l	.4

Table 5-4

Average COD Levels and Standard
Deviations for Each Run (mg/l)

	INFLUENT		REACTOR-CLARIFIER EFFLUENT		FINAL EFFLUENT	
	mean ($\bar{x}$)	standard deviation (s)	mean ($\bar{x}$)	standard deviation (s)	mean ($\bar{x}$)	standard deviation (s)
Run 1	1162	202	313	56	79	37
Run 2	410	92	62	47	6	2
Run 3	1095	132	118	17	33	32
Run 4	661	49	116	46	23	14
Run 5	505	76	103	18	24	14
Run 6	490	50	178	52	58	21
Run 7	431	74	107	37	15	15
Run 8	475	54	144	62	61	48

Table 5-5

Average SS Levels and Standard Deviations for Each Run (mg/l)

	INFLUENT		REACTOR-CLARIFIER EFFLUENT		FINAL EFFLUENT	
	mean ($\bar{x}$)	standard deviation (s)	mean ($\bar{x}$)	standard deviation (s)	mean ($\bar{x}$)	standard deviation (s)
Run 1	636	124	55	3	13	6
Run 2	274	40	30	12	3	3
Run 3	756	115	94	16	11	6
Run 4	399	74	74	13	16	5
Run 5	311	47	70	12	19	11
Run 6	224	34	75	24	25	20
Run 7	233	64	69	23	8	18
Run 8	250	59	80	44	15	5

Table 5-6

Average Dissolved COD Levels and Standard Deviations for Each Run (mg/l)

	COLUMN INFLUENT		COLUMN EFFLUENT	
	mean ($\bar{x}$)	standard deviation (s)	mean ($\bar{x}$)	standard deviation (s)
Run 1	-	-	-	-
Run 2	46	26	5	1
Run 3	31	30	6	9
Run 4	34	19	9	1
Run 5	28	8	5	4
Run 6	27	5	8	3
Run 7	21	3	6	4
Run 8	36	9	16	10

Table 5-7

Check for Statistical Differences Between Data of Various Runs Using Student T-Test, Significance Level = 5%

	System Influent	Reactor-Clarifier Effluent	Final Effluent		System Influent	Reactor-Clarifier Effluent	Final Effluent
Run 1 vs Run 2				Run 3 vs Run 4			
COD	D	D	D	COD	D	ND	ND
SS	D	D	D	SS	D	ND	ND
Dissolved COD		-	-	Dissolved COD		ND	ND
Run 1 vs Run 3				Run 5 vs Run 2			
COD	ND	D	ND	COD	ND	ND	ND
SS	ND	D	ND	SS	ND	D	D
Dissolved COD		-	-	Dissolved COD		ND	ND
Run 1 vs Run 4				Run 6 vs Run 5			
COD	D	D	ND	COD	ND	D	D
SS	D	D	ND	SS	*D	ND	D
Dissolved COD		-	-	Dissolved COD		ND	ND
Run 2 vs Run 3				Run 7 vs Run 8			
COD	D	D	ND	COD	ND	ND	ND
SS	D	D	ND	SS	ND	ND	ND
Dissolved COD		ND	ND	Dissolved COD		D	D
Run 2 vs Run 4							
COD	D	D	ND				
SS	D	D	D				
Dissolved COD		ND	D				

Legend: ND = no significant difference in data between runs
D = significant differences in data

*Difference in this run was in favour of Run 6 where influent SS levels were less than Run 5. After reactor-clarifier there was no difference and in final effluent Run 6 had higher SS levels

The arithmetic means and standard deviations for the data were calculated and are tabulated in Table 5-3 to 5-6. In Table 5-3 the ratio of standard deviation to arithmetic mean was also calculated. The purpose of the latter calculation was to study whether fluctuations in the various parameters, with respect to the mean, were being reduced as the waste stream passed through the system (i.e. the lower the value, the less the fluctuation).

One should note that the standard deviation to arithmetic mean ratios for all three influent parameters increased in the final effluent. In addition, as shown in Figures 5-3 and 5-4 the fluctuations in SS and COD levels in the influent did not correspond with the fluctuations of these parameters in the effluent.

These observations will be discussed in further dedail in Sections 5.3.2 and 5.3.3.

5.3.1.2 Solids Reduction

The solids within a sanitary waste can be considered as either non-biodegradable (inert) or biodegradable. The non-biodegradable component generally passes through a biological treatment system unchanged. Part or all of the biodegradable component is oxidized to form micro-organisms, CO_2 and H_2O. In addition, the micro-organisms themselves become oxidized to form more CO_2 and H_2O. The CO_2 leaves the system in a gaseous form and the H_2O leaves the system in a liquid form. The result is a reduction from the original biodegradable solids component which entered the system and, of course, an overall solids reduction.

In this experiment, the overall solids reductions were measured for the last four runs only. As shown in Table A10-2, the average solids reduction for the overall system was 51 percent of the system's incoming solids. These reductions are discussed in further detail in section 5.3.2.3.

5.3.1.3 Feasibility of Recycle

In Appendix 11 and Appendix 12 calculations were made on the resulting concentration increases of COD and SS in the final effluent if both backwash and final effluent are recycled back into the system.

On the basis of Appendix 11 it is evident that if treatment is based on the average values and the average values (i.e. removal percentages) do not change when backwash and final effluent are recycled to the head of the system, then the SS concentration in the effluent should become constant after day 4 at a level of 12 mg/l.

In Appendix 12 similar results are seen with COD concentrations. Here, however, COD values did not level out until day 6, when the concentration reached 39 mg/l. These levels would satisfy our design objectives.

Although the calculations were based on averages and the assumption that recycle would not affect these averages, they are a good approximation of what may happen if recycle of backwash and effluent is utilized.

From the results of these calculations it would seem that recycle is feasible. One should not expect in an actual system such a rapid approach to equilibrium as indicated, but eventual equilibrium is probable. Nonetheless, if this level is very high for either SS or COD, then recycle of backwash and/or final effluent would be unacceptable for aesthetic and/or hygienic reasons.

5.3.2 Reactor-clarifier

5.3.2.1 COD and SS Removals

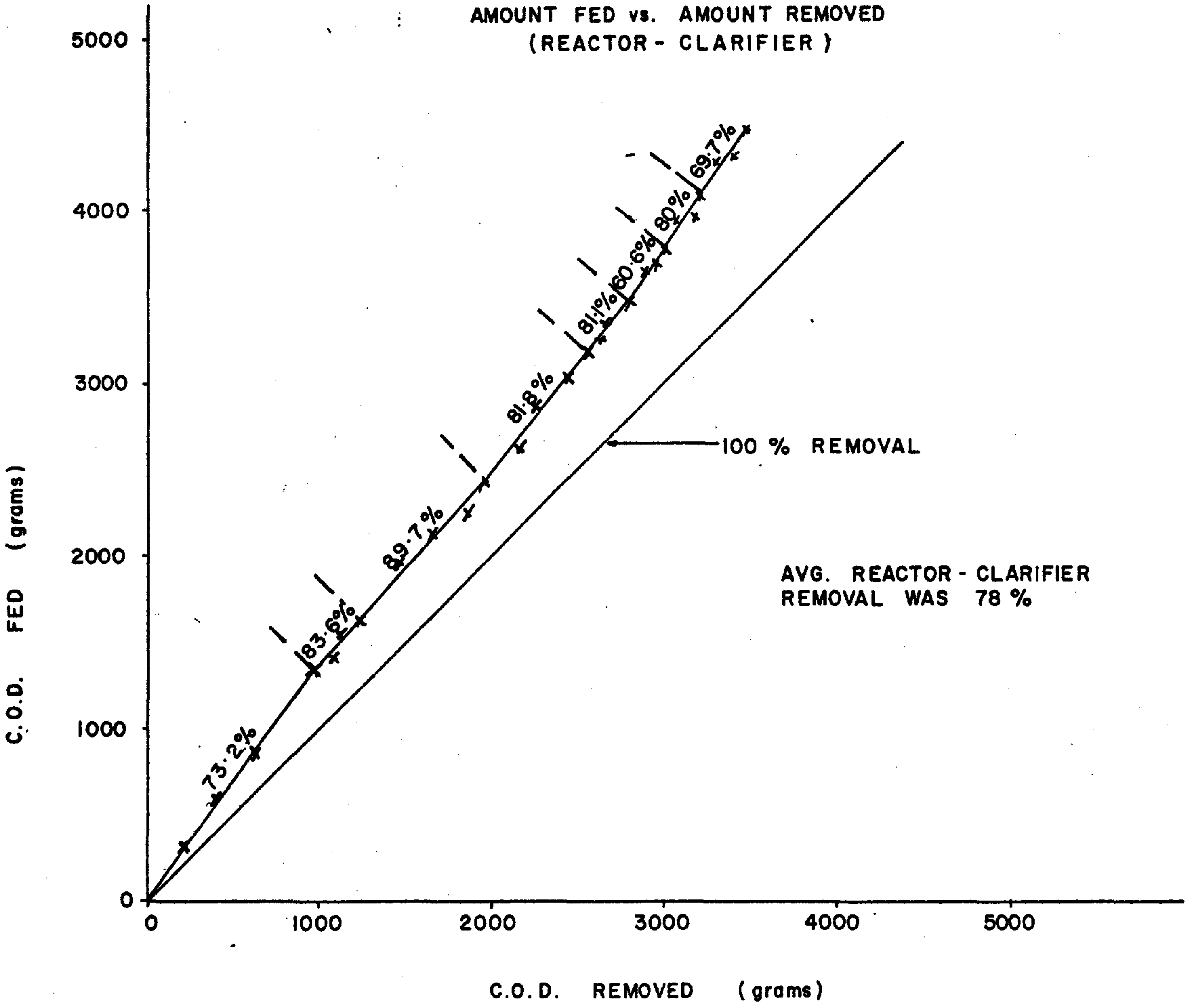
The reactor-clarifier's performance is displayed in Figures 5-5 and 5-6 and in Table 5-3 to 5-6.

The results indicate that the average COD removal in the reactor-clarifier was 78 percent, with an average COD value in the clarifier effluent of 143 mg/l. The average SS removal in the reactor-clarifier was 83.4 percent, resulting in an average SS of 68 mg/l in the clarifier effluent.

The following observations are relevant:

(i) Run 1 displayed good SS removals, but poor COD removals (see Figures 5-5 and 5-6).

FIGURE 5-5
C.O.D.
AMOUNT FED vs. AMOUNT REMOVED
(REACTOR - CLARIFIER)



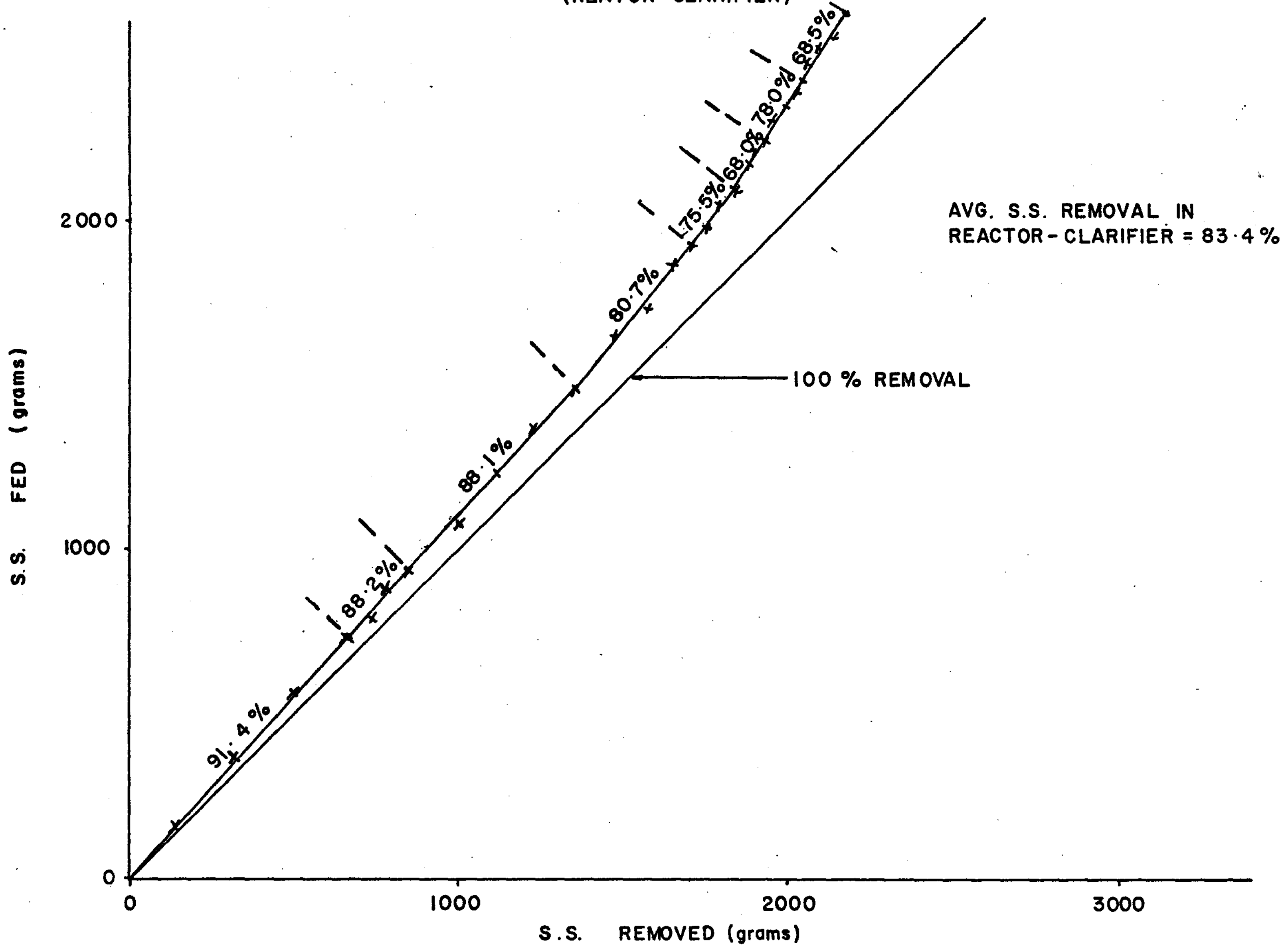
AVG. REACTOR - CLARIFIER
REMOVAL WAS 78 %

FIGURE 5-6

S.S.

AMOUNT FED v.s. AMOUNT REMOVED

(REATOR-CLARIFIER)



AVG. S.S. REMOVAL IN REACTOR-CLARIFIER = 83·4 %

It is believed that the coagulant added during this run provided the necessary SS removals, but the biological component of the system had not developed sufficiently for the removal of dissolved COD.

(ii) Run 6 had both poor COD and SS removals when comparing to Run 5 (see Table 5-7 and Figures 5-5 and 5-6). Run 6's poor performance was likely due to three prominent operating characteristics: the temperature increase, the low DO level in the reactor (less than 2 mg/l - see Table A10-7), and the MLSS level of 4,000 mg/l. The high temperatures caused a thermal shock on the micro-organisms and coupled with the low dissolved oxygen levels and the high MLSS levels, a decrease in bioactivity resulted. This caused floc break-up and a higher colloidal or pin-floc content. Because of the low bioactivity, dissolved COD removals were poor. Thus, due to the high colloidal content of the solids, the SS removals in the clarifier were poor and because of the poor dissolved COD removals, the COD levels in the effluent were high.

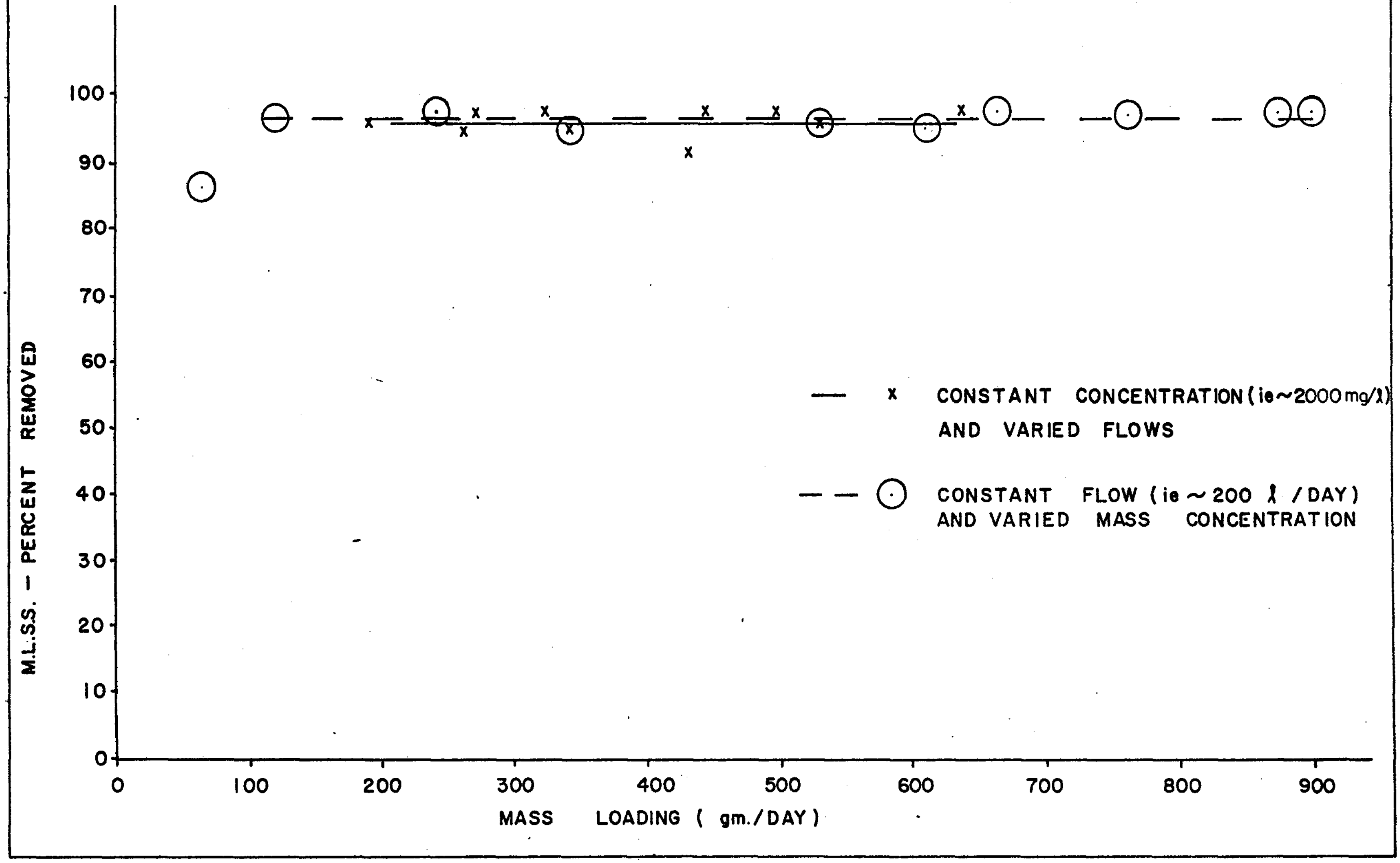
(iii) Run 8 showed poor COD and SS removals at the beginning of the run (see Figures 5-5 and 5-6). Once again this can be attributed to poor biological activity. The poor biological activity resulted from the seven-day shutdown which occurred prior to this run. However, as shown in Figure 5-14, dissolved COD levels emanating from the reactor-clarifier reduced rapidly. This caused overall COD levels to reduce to adequate levels after the second day (see Figure 5-3). All of these observations would tend to indicate the biological activity was poor at the beginning of the run, but recovered rapidly to improve the COD and SS removals.

(iv) Coagulant was not necessary once bioactivity had reached an optimum level. No coagulant was added in Runs 5 and 7. Table 5-7 shows that there was no difference in the COD levels of the reactor clarifier's effluent in Run 2 (alum added) and Run 5 (no allum added). However, there was a difference in the SS levels between these runs. Although SS removals were not quite as high in Run 5

FIGURE 5-7

M.L.S.S.

CLARIFIER MASS LOADING EFFECT



as in Run 2, Figure 5-4 shows that the SS removals for those runs (i.e. 5 and 7) in which alum was not added were sufficient to satisfy the objectives. We believe these observations indicate that alum is only required when the system is incapable of maintaining a suitable level of bioactivity.

(v) Fluctuations in the influent COD and SS do not correspond to the fluctuation of these parameters in the reactor-clarifier's effluent (see Figures 5-3 and 5-4).

5.3.2.2 The Effects of Mixed Liquor Suspended Solids (MLSS) Concentrations, Flows and Mass Loading on the Clarifier

The effects of mass loading, flow and influent MLSS concentration on the clarifier were studied. In Figure 5-7 mass loading was plotted against MLSS percent removed. The results indicated that:

(i) there was no decrease in the percentage of SS removals as the mass loading, with respect to MLSS, was increased.

(ii) neither flow nor MLSS concentrations affected suspended solids removal.

Generally the clarifier will be affected by flow since detention time and the overflow rate are affected. These two parameters are important in the operation of a clarifier. Since, in this particular case, the overflow rate was very low (i.e. 100 Igpd/ft^2 - see Section 3.3.3), flow through the clarifier had an undetectable effect on the clarifier's performance.

5.3.2.3 Yield and Solid Reductions

The yield of micro-organisms produced per pound of COD removed in the reactor-clarifier are shown in Table A10-2. These yields are considered as "observed yields" and not "actual yields". The difference between these two terms can be explained by the following micro-organism population equation:

$$\frac{dx}{dt} = y\,\frac{ds}{dt} - bx$$

where x = micro-organism concentration

s = substrate concentration

t = time

b = micro-organism decay coefficient (mainly due to endogenous respiration)

y = factor at which substrate is converted to micro-organisms

Observed yields (i.e. $\frac{dx}{ds}$) is the yield factor after accounting for micro-organism decay. Actual yield (i.e. y) is the term given to the fraction of the substrate actually converted to micro-organisms.

Over the length of the experiment the average yield after accounting for alum addition was .3. Ideally, yields should be .1 to .2 for extended aeration plants (Metcalf and Eddy, 1972). However, these yields are based on BOD yields. Eckenfelder (1970) claims the BOD/COD ratio for raw domestic waste is approximately .4. As a result yield values of .1 to .2 based on BOD removals are comparable to yield values of .04 to .08 based on COD removals. Thus the yield values found in this system were higher than the ideal extended aeration values and were closer to those found in conventional activated sludge systems.

Solid reductions, as defined in Section 5.3.1.2, for the reactor-clarifier, are tabulated in Table A10-2.

One should note that during Run 1 a baffle was placed upstream from the clarifier to create a sludge blanket. This is explained in greater detail in Section 5.3.2.6. When calculating solid reductions it is necessary to measure the daily influent SS concentrations, the MLSS concentrations, and the clarifier's SS concentrations. The blanket contained a significant proportion of the system's solids. Therefore when measuring solids in the reactor the solids in the blanket were not considered in the balance. As a result the reduction reported was much too high.

The average reduction over the length of the experiment was 30 percent after adjusting for Run 1. As expected solid reductions were high when yields were low (i.e. micro-organisms are in the endogenous phase) and low when yields were high (i.e. micro-organisms are in the log growth phase).

5.3.2.4 Food to Micro-organism Ratio

The food to micro-organism ratio (F/M) is defined as the mass of COD degraded divided by the mass of micro-organisms degrading the COD over a finite period of time. The F/M ratios for each day are summarized in Table A10-5 and displayed in Figure 5- 8. The average F/M ratio was .31 after adjusting for the problems encountered with the temporary baffle in Run 1 as discussed earlier. In addition, this F/M ratio provided the best COD removal as shown in Figure 5-8. F/M ratios for extended aeration plants are generally between .05-0.15 (Metcalf and Eddy, 1972) based on BOD removals. For COD removals these F/M ratios would have corresponding values of .12-0.38. Thus one can conclude that the average F/M ratio found in this system was similar to most extended aeration plants.

One should note, as indicated in Figure 5-8, that when F/M ratios were as high as 3.0, the COD removal was still approximately 70 percent. It is believed these COD removals were due to the alum addition and would likely not be this high in totally biological systems.

5.3.2.5 Mean Cell Residence Time (Sludge Age)

The mean cell residence time (θ_c) is defined as the total active microbial mass in a treatment system divided by the total quantity of active microbial mass withdrawn over a finite time period. Sludge ages for extended aeration plants are generally 20 to 30 days. The sludge ages for each run of this study are summarized in Table A10-6 and displayed in Figure 5-9. The average sludge age was 10 days and a sludge age of approximately 5 days provided the best COD removal. These values concur with those reported in the literature for conventional activated sludge plants which usually have sludge ages varying from 5 to 16 days (Water Pollution Control Federation, 1974).

Mean cell residence time can be equated to F/M and actual yield in the following way (Metcalf and Eddy, 1972):

$$\frac{1}{\theta_c} = Y_a U - b$$

where θ_c = mean cell residence time
y_a = actual yield
U = F/M ratio
b = microorganism decay coefficient, time^{-1}

As explained in Section 5.3.2.3 "observed yield" accounts for micro-organism decay, thus

$$\frac{1}{\theta_c} = Y_o U$$

where Y_o = observed yield

In the present study the measure average values of θ_c, Y_o and U were 10 days, .3 and .31 respectively.

$$\text{Left side} = \frac{1}{\theta_c} = \frac{1}{10} = .1$$

$$\text{Right side} = y_o \times u = .3 \times .31 = .093$$

$$\therefore \text{Left Side} \simeq \text{Right Side}$$

The fact that the above equations are satisfied by the measured parameter levels indicates that the values obtained are reasonable. One should note, however, that the measured values indicate that the system operated similar to a conventional activated sludge system rather than the intended extended aeration system.

FIGURE 5-8

F/M vs. PERCENT C.O.D. REMOVED IN REACTOR-CLARIFIER

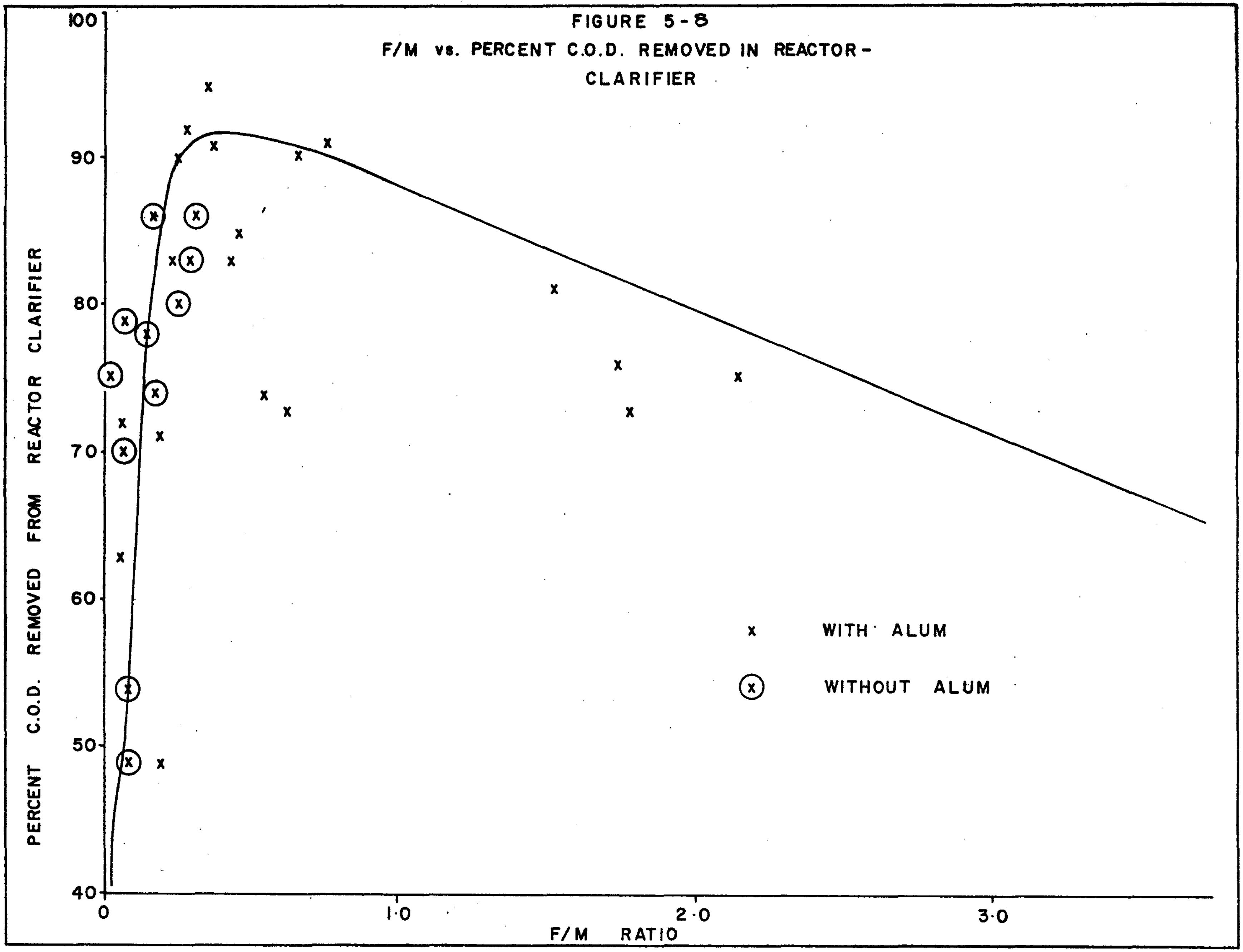
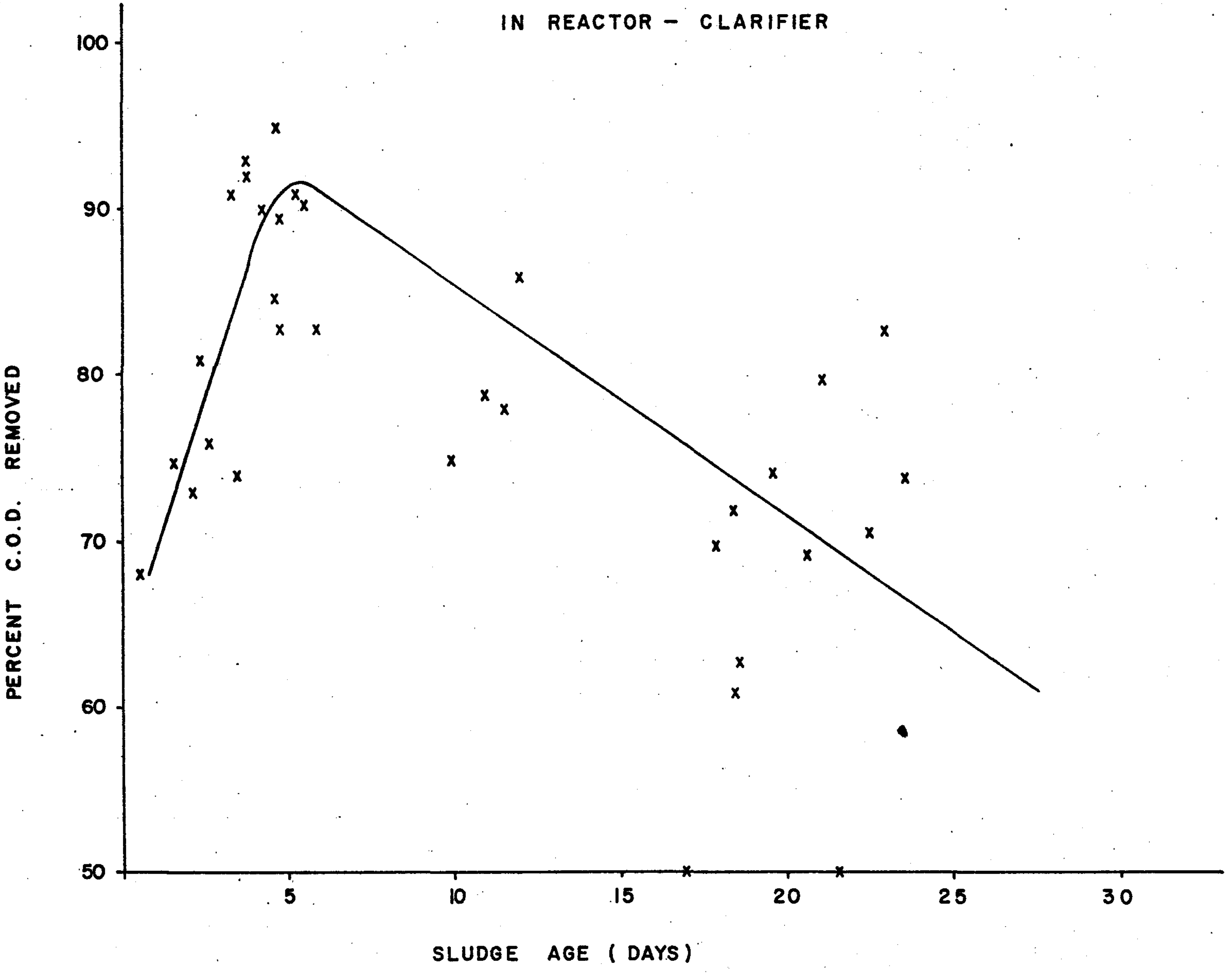


FIGURE 5-9

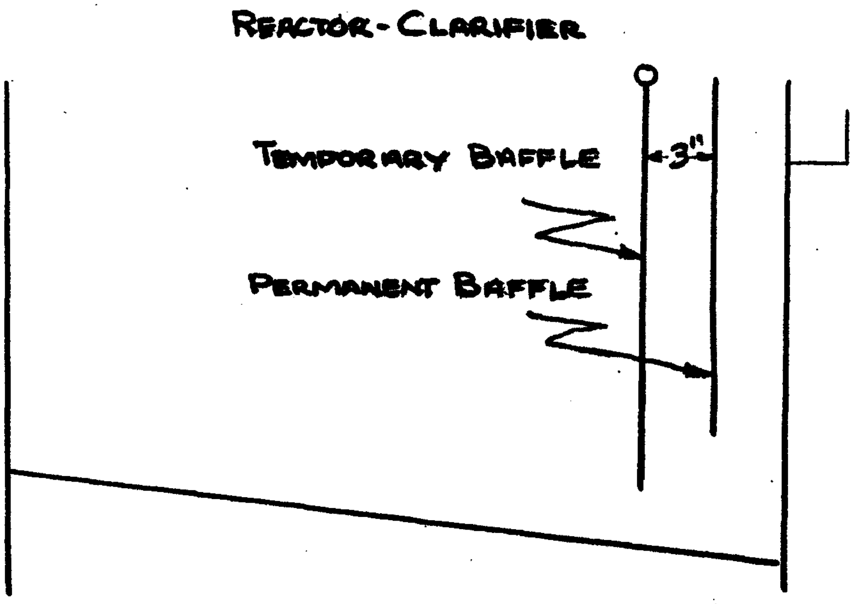
SLUDGE AGE vs. PERCENT C.O.D. REMOVED IN REACTOR - CLARIFIER



5.3.2.6 Mechanical Operation

The reactor-clarifier did not encounter any serious mechanical difficulties, however, the following should be noted:

(i) In run 1 a baffle was placed immediately upstream from the clarifier as shown below. The purpose of the baffle was to form a sludge blanket in the clarifier and thereby aid in solids removal.



Placement of Temporary Baffle

As expected, the sludge blanket formed and aided in the removal of solids. However, it hindered organic removal. Since there was little sludge return, there was an increase in the F/M ratio, which resulted in poor COD removals. For this reason the temporary baffle was removed for the remaining runs.

(ii) Occasionally the feed line to the reactor clarifier became clogged. However, this problem was alleviated by macerating the sewage sample each day.

(iii) The coarse air diffusers operated well. All measured dissolved oxygen levels, with one exception, were above 2 mg/l (see Table A10-7). The diffusers became plugged during the 7-day shutdown prior to run 8. This problem would be alleviated in the actual system by using the non-clog type coarse air diffusers.

(iv) The circulating current within the reactor resulted in an efficient sludge return system from the integral clarifier. This was evident by the small amount of sludge build-up within the clarifier for all runs except for run 1 as discussed earlier.

(v) After the system's 3-day shutdown prior to run 5, a black-grey scum on the surface of the clarifier was observed. However, as the run continued, the scum disappeared. This same observation occurred in the following runs.

It is expected that there was an abundant amount of DO in the reactor which caused nitrification (i.e. the formation of nitrates and nitrites). Since there was no flow through the reactor-clarifier during shutdown periods the clarifier became anaerobic. As a result, the nitrates and nitrites were employed as oxidizing agents in the clarifier and nitrogen gas was formed. With the nitrogen gas bubbling to the top of the clarifier a flotation effect occurred causing the scum to form.

(vi) The clarifier collection trough provided reaeration of the clarifier effluent prior to entering the adsorption-filtration column (see Table A10-7).

(vii) The only manual operation in connection with the reactor-clarifier was sludge wastage. In an actual system this operation could be automated by the use of a photocell which, at certain concentrations, would trigger a relay to open a sludge wastage drain line valve.

5.3.3 Adsorption-Filtration Column

5.3.3.1 Dissolved COD, COD and SS Removals

The adsorption-filtration column's performance is displayed in Figures 5-10 to 5-17 and Table 5-3 to 5-7.

The results indicated that the average suspended solids removal in the column was 80 percent of the solids entering the column. This contributed, on the average, a further 14 percent to the overall suspended solids removals. As a result, the average SS concentration in the final effluent was 13.8 mg/l.

The average COD removal was 74 percent of COD entering the column which contributed 16 percent to the overall COD removals. As a result the average COD level in the effluent was 37.2 mg/l.

The average dissolved COD removal in the adsorption-filtration column was 76.3 percent which resulted in a final effluent average dissolved COD level of 7.6 mg/l.

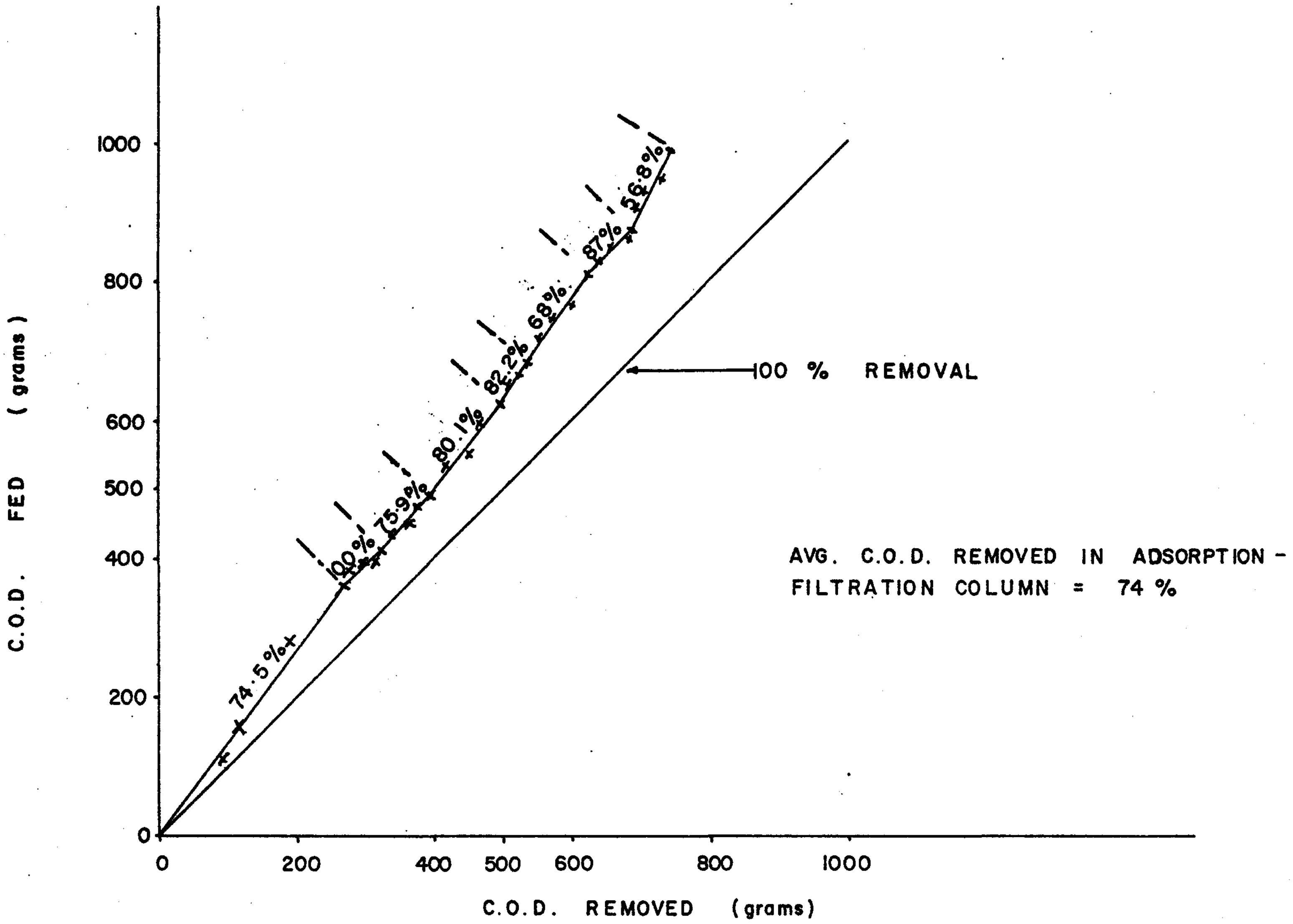
In addition, the following observations are worth noting:

FIGURE 5-10

C.O.D.

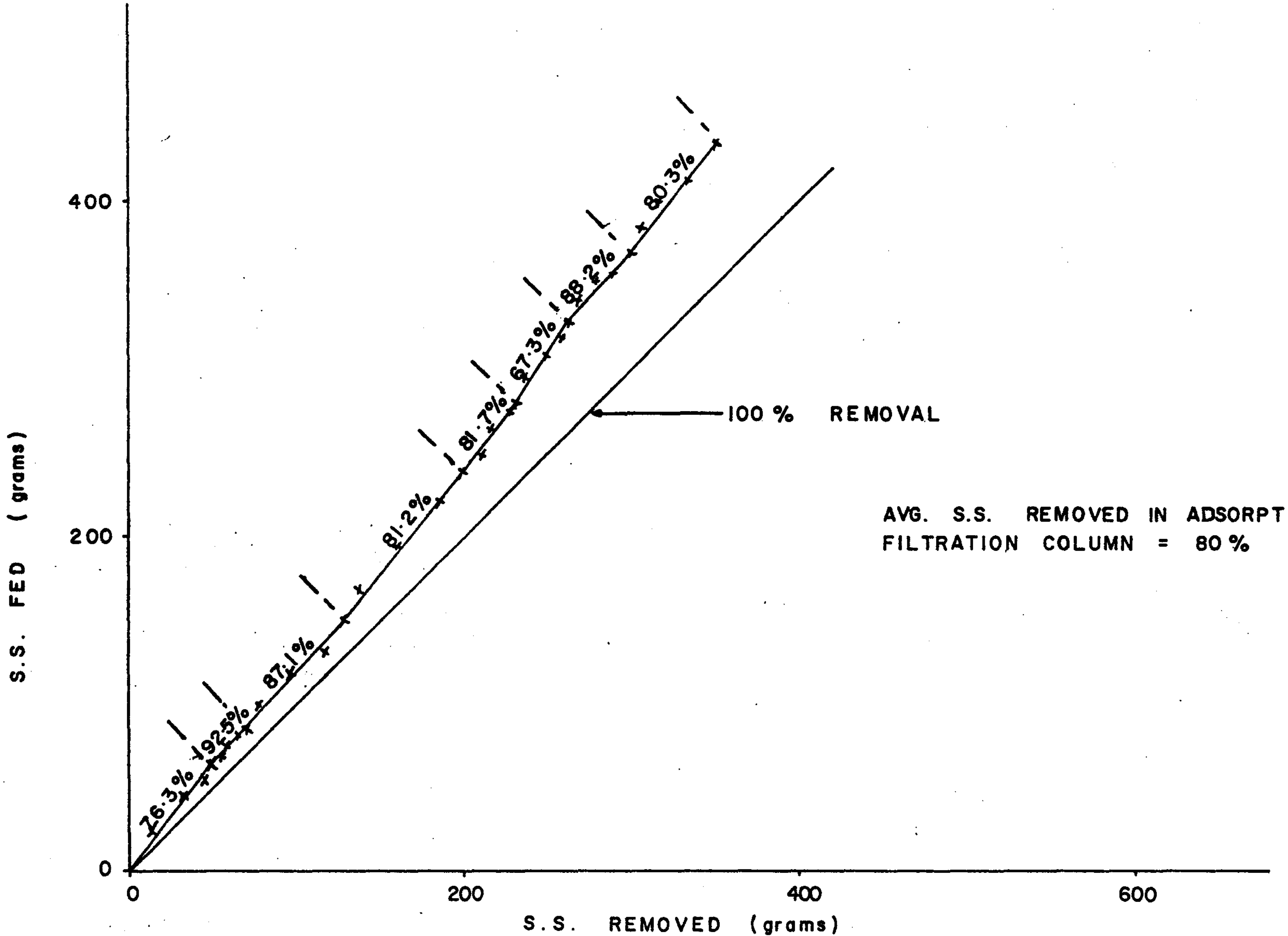
AMOUNT FED vs. AMOUNT REMOVED

(ADSORPTION - FILTRATION COLUMN)



AVG. C.O.D. REMOVED IN ADSORPTION - FILTRATION COLUMN = 74 %

FIGURE 5-11
S.S.
AMOUNT FED vs. AMOUNT REMOVED
(ADSORPTION-FILTRATION COLUMN)



AVG. S.S. REMOVED IN ADSORPTION-FILTRATION COLUMN = 80 %

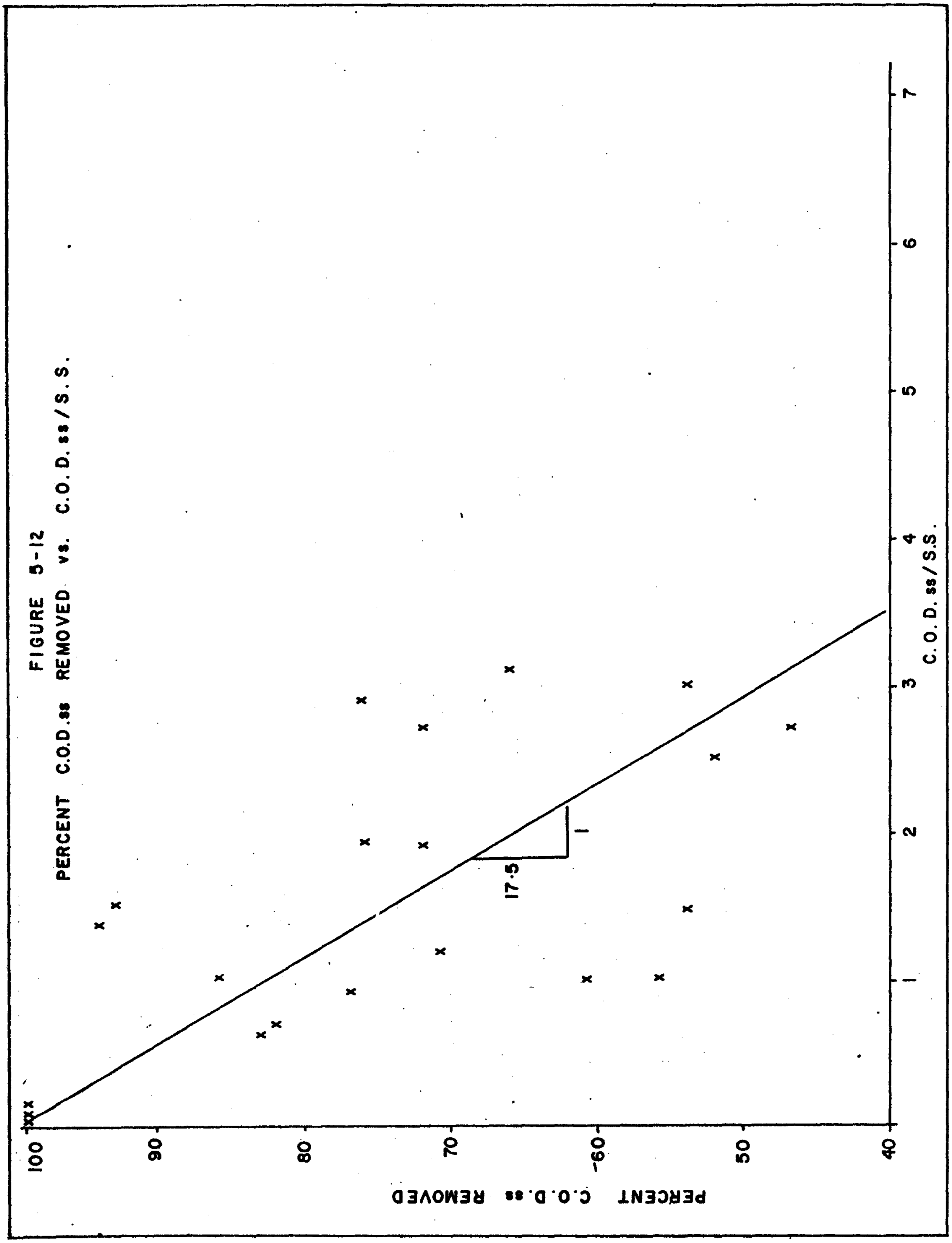
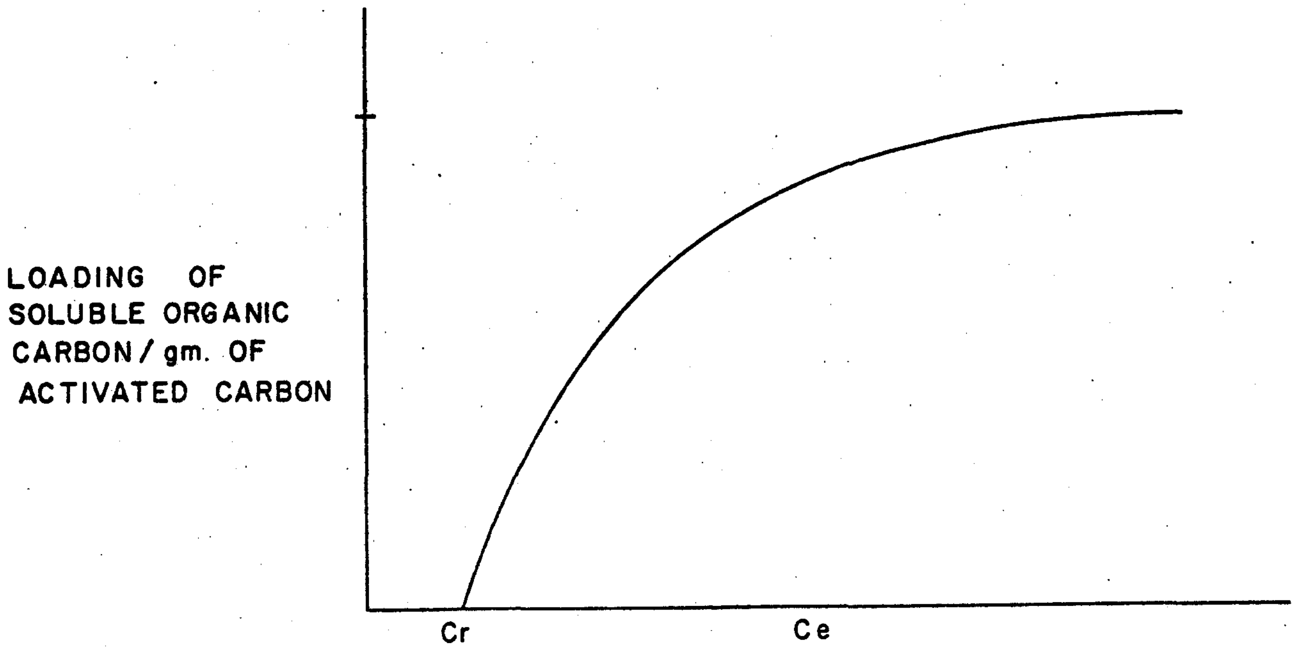


FIGURE 5-12

PERCENT C.O.D.ss REMOVED vs. C.O.D.ss/S.S.

FIGURE 5-13

TYPICAL ACTIVATED CARBON ISOTHERM



Cr – RESIDUAL CONCENTRATION
Ce – EQUILIBRIUM CONCENTRATION

FIGURE 5-14
PERCENT DISSOLVED C.O.D. REMOVED IN THE ADSORPTION-FILTRATION COLUMN vs. DAY

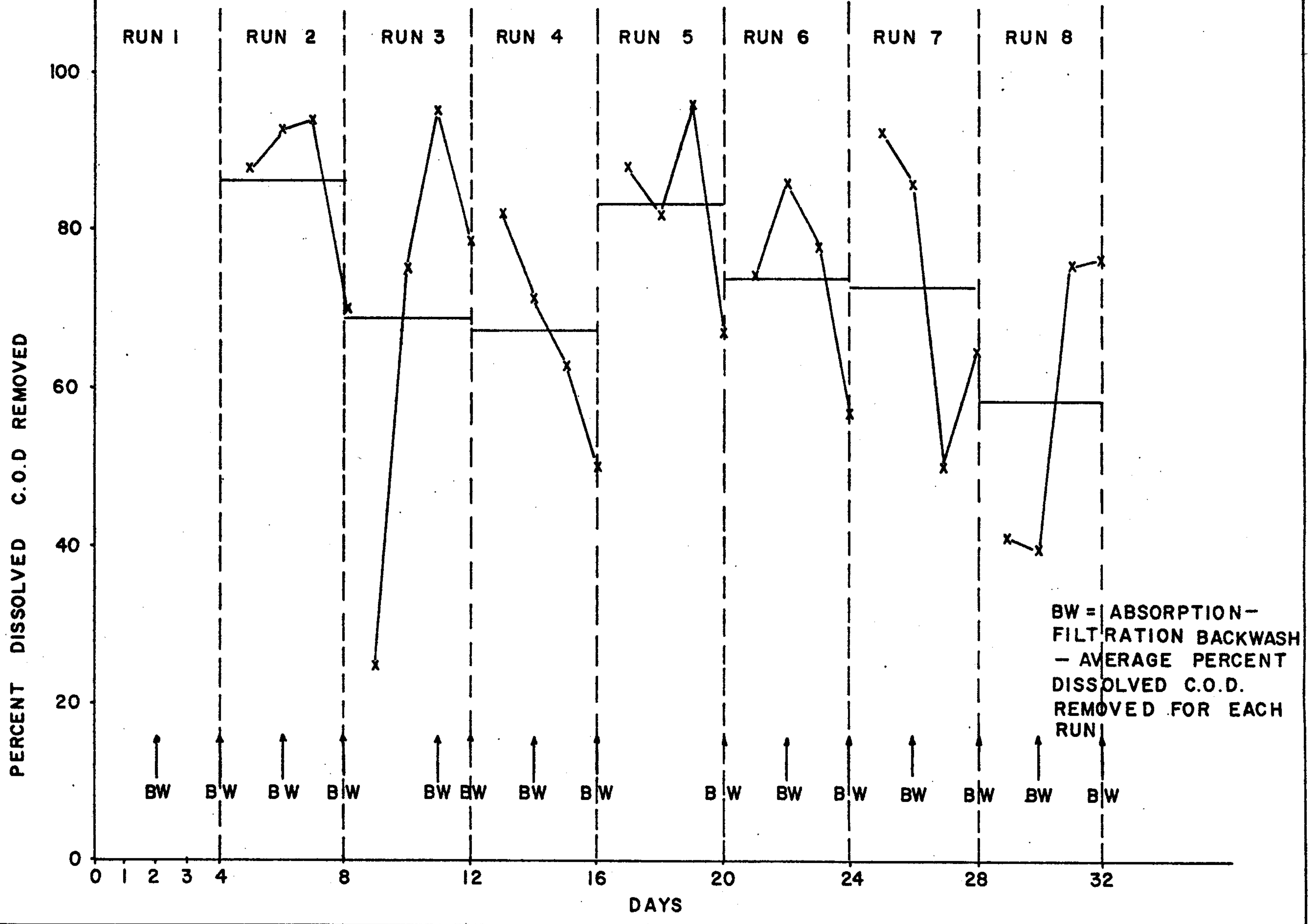


FIGURE 5-15

DISSOLVED C.O.D. IN THE ADSORPTION-FILTRATION COLUMN INFLUENT AND EFFLUENT vs. DAYS

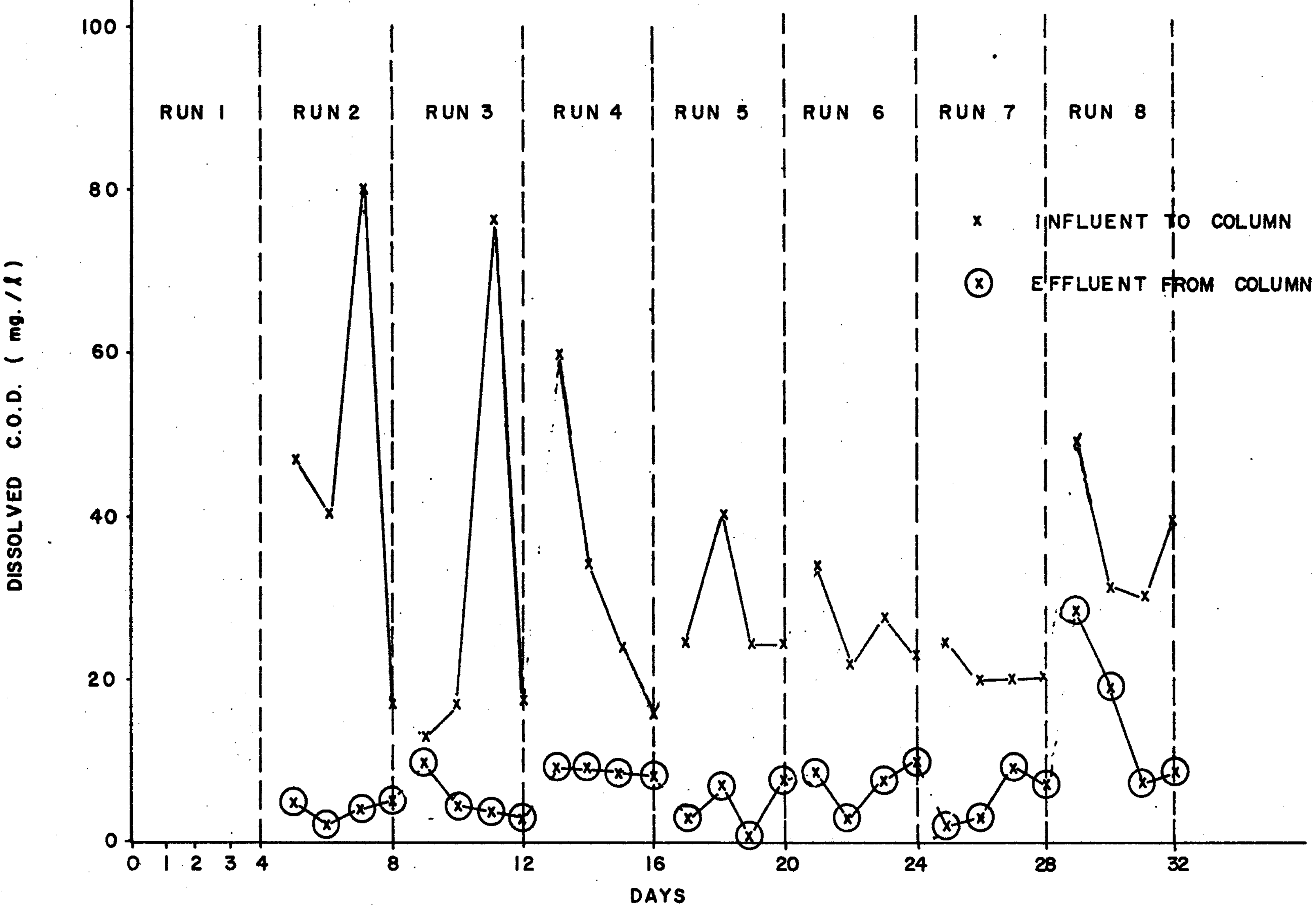


FIGURE 5-16
DISSOLVED C.O.D. – ADSORPTION FILTRATION COLUMN
AMOUNT FED / gm. OF ACTIVATED CARBON vs. AMOUNT REMOVED / gm OF ACTIVATED CARBON

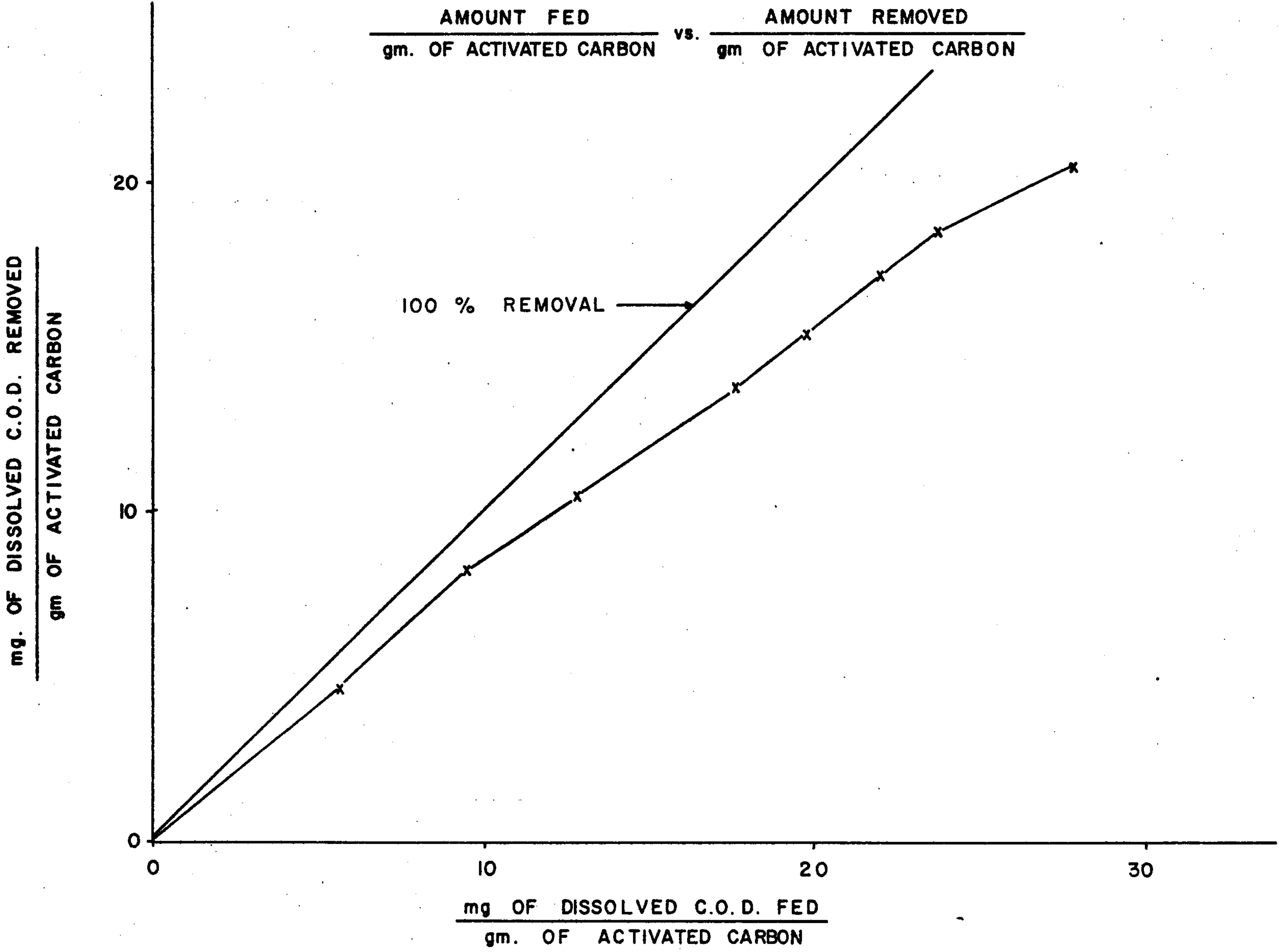
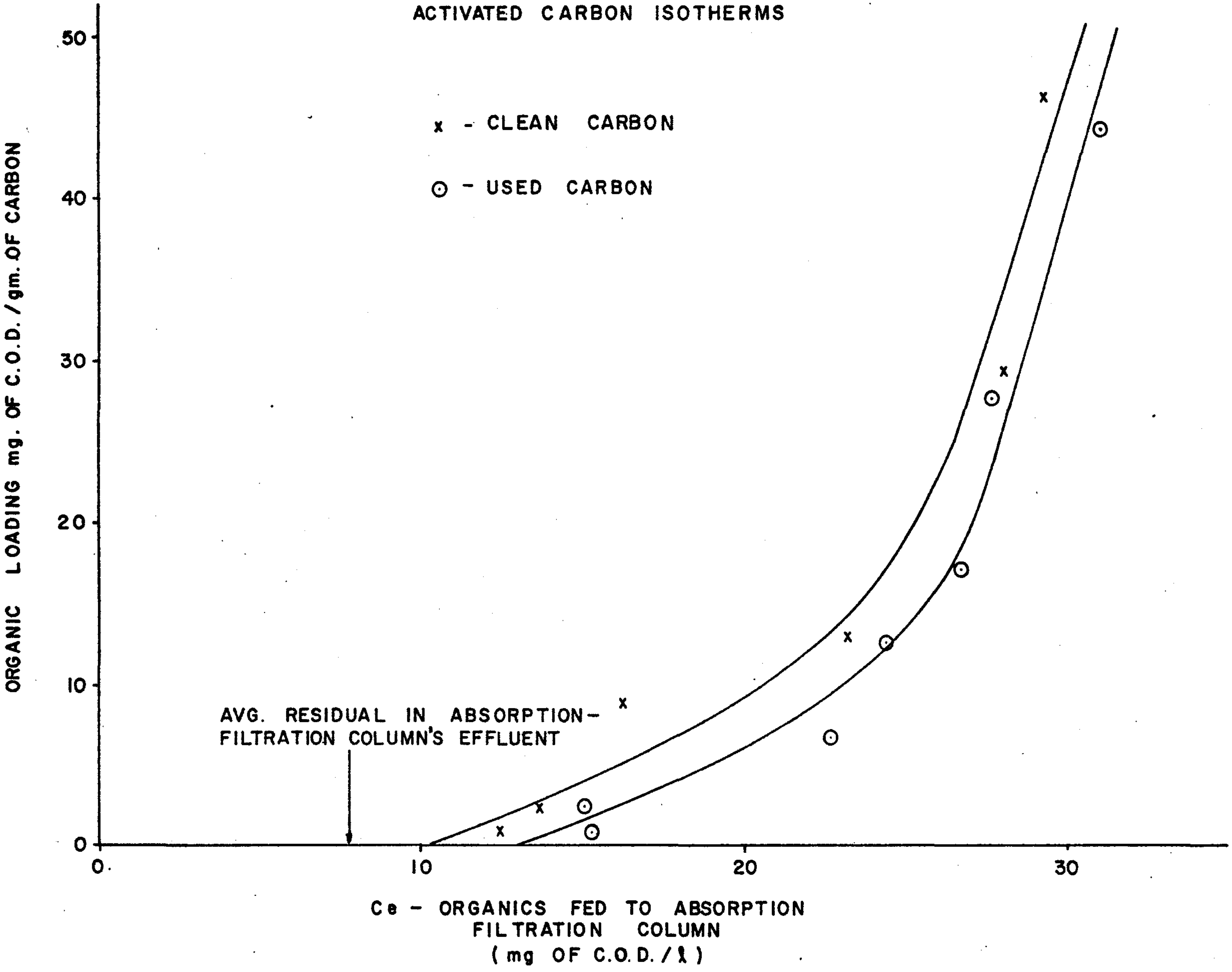


FIGURE 5- 17
ACTIVATED CARBON ISOTHERMS



(i) The adsorption-filtration column was capable of compensating for the treatment deficiencies of the reactor-clarifier and ensured a consistent final effluent. As indicated in Table 5-7 in situations where there were differences in the quality of the reactor-clarifier effluent between various runs, the adsorption filtration column generally provided the extra treatment necessary to produce a final effluent of similar quality.

(ii) As shown in Figure 5-10 and 5-11 Run 1 displayed SS and COD removals which were slightly below average. This was probably due to the poor biological activity in Run 1. Many of the solids which escaped the clarifier were likely colloidal in size and had a high COD_{SS}/SS ratio (see Figure 5-12). As a result, some of the colloidal particles passed through the column causing COD and SS removals to be below average.

(iii) In Run 6 dissolved COD removals were relatively high but SS and COD removals were low. These observations can also be attributed to the poor biological activity which occurred in the reactor during Run 6.

Dissolved COD removals were good because of the adsorptive capabilities of the carbon. A typical isotherm for activated carbon is shown in Figure 5-13. As one can see, when the equilibrium concentration of dissolved COD increases, the loading does also, until a certain point. In Run 6 the higher dissolved COD levels (see Table A10-1) emanating from the reactor-clarifier forced the activated carbon loadings to increase, resulting in high dissolved COD removals.

(iv) After the seven-day shutdown dissolved COD and COD removals were poor at the beginning of the run. It is believed that by the end of Run 7, removal of dissolved COD in the carbon column was mainly due to the biological activity within the column. The seven-day shutdown hindered the bioactivity, resulting in poor dissolved COD and COD removals on Day 1 of Run 8 (see Figure 5-16). However, on Day 2 of Run 8 these removals improved significantly, indicating a rapid recovery of micro-organism metabolism within either the reactor-clarifier or the adsorption-filtration column, or both.

(v) As shown in Figure 5-3 COD fluctuations in the column effluent followed a similar pattern to the fluctuations in the column influent.

However, this did not occur with the fluctuation of SS and dissolved COD concentrations (see Figure 5-4 and 5-17).

The magnitude of COD and SS fluctuations in the effluent (see Table 5-3) were higher than in the column's influent. However, the magnitude of dissolved COD fluctuations was reduced. These observations are possibly due to poor bioactivity. As explained in Item (ii) above, the solids emanating from the clarifier during periods of bioactivity likely had a high composition of colloidal solids with high COD_{SS}/SS ratios. The result was high concentrations of SS and COD in the column's effluent which caused high fluctuations with respect to the arithmetic mean. Since high COD levels occurred in the clarifier's effluent during poor bioactivity periods, the COD fluctuation pattern in the column's influent was similar to the column's effluent. This pattern did not develop with suspended solids because the reactor's poor biological activity caused poor dissolved COD removals which affected COD levels. Poor biological activity affected suspended solids removals in the reactor-clarifier also, but not to the same extent because alum was added during these periods.

The magnitude of fluctuations for dissolved COD in the column's influent was lower in the effluent because of the capability of activated carbon to increase organic adsorption when the influent organic concentration increases (see Item (iii) above). This particular characteristic caused a dampening effect on the dissolved COD fluctuations.

Thus, the reasons for the observed data fluctuations are likely due to the poor biological activity which occurred in some runs and the activated carbon adsorptive characteristics which dampened dissolved COD fluctuations.

5.3.3.2 Activated Carbon Adsorbed Quantities

The data as plotted in Figure 5-16 indiates the amount of dissolved COD removed per gram of carbon gradually decreased over the duration of the testing period.

The reason for this occurrence can possibly be explained in the following way:

(i) At the beginning of the testing period the activated carbon was fresh. Thus, the bacterial slime layer which usually encompasses a carbon particle was insignificant and most of the carbon's surface area was available for adsorption. As a result the transfer resistance through a bacterial film was negligible. This permitted high rates of adsorption. As the experiment continued the bacterial film on the carbon became thicker. The transfer resistance increased causing the adsorption rate to decrease. On the other hand, the bio-oxidation in the developing biofilm may compensate for this reduction in the adsorption rate.

(ii) As the adsorptive capacity of the carbon was gradually exhausted, the driving force, resulting from differences between concentrations on the carbon surface and liquid,

decreased causing a reduction in the
adsorption rate of dissolved COD.

After the 10-week testing period isotherms were performed on the activated carbon in the adsorption-filtration column and compared with clean unused carbon. The results are plotted in Figure 5-17.

The average C_e (dissolved COD in the column's feed) was 30 mg/l. At this concentration the clean carbon could take a loading of 49 mg of dissolved COD/gm of activated carbon and the used carbon could only take a loading of 40 mg/gm. Therefore 9 mg of dissolved COD/gm of activated carbon was adsorbed.

As indicated in Figure 5-16 the accumulated dissolved COD removed per gram of activated carbon was 21 mg/gm.

This may indicate that 12 mg of dissolved COD/gm of activated carbon (57 percent) was removed biologically and 9 mg of dissolved COD/gm of activated carbon was adsorbed (43 percent).

When designing the adsorption-filtration column it was intended that the activated carbon provide adequate dissolved COD removals for one year. The total amount of dissolved COD removed over a 32 day period was 158.3 g. This represents 5 g of dissolved COD removed per day. Thus, over a one year period 1,825 gm would have to be removed.

The isotherm showed that 49 mg of dissolved COD could be removed per gram of carbon. Since there was approximately 6,500 g of activated carbon in the column, this would indicate that the carbon in the column could adsorb 319 g.

In addition to this, there would be biological degradation in the column. Again extrapolation from these runs indicate the column would be capable of removing 741 g of dissolved COD. This means the quantity of carbon present would have only been sufficient for approximately five months. Improvement in bioactivity, however, may extend acceptable removal beyond 5 months.

The design of the column was based on 80 lb of COD removed per 100 lb of activated carbon (Benedek, 1975). In this particular system the exhaustion loading on the carbon is estimated to be 11.5 lb/100 lb. However, it should be noted that the above design parameter was recommended for the treatment of municipal sewage. One should expect lower loading rates when polishing a secondary effluent.

5.3.3.3 Mechanical Operation

As with the reactor-clarifier, the adsorption-filtration column operated well without any major mechanical problems. However, the following should be noted:

(i) Column backwashes were generally needed every two days during a run. The length of time between backwashes was expected to be longer when the system was designed. It is probable that the lack of solids penetration in the column media, due to the low hydraulic loading, caused extensive head losses and thus frequent backwashes.

The necessary backwash rates were monitored in runs 5 to 8. The results are tabulated in Table A10-8. When designing the system, a backwash rate of 10.0 Igpm/ft^2 was believed necessary to obtain 50 percent expansion (see Appendix 6). Since the column's surface area was .55 ft^2 it was expected the actual flow required would be 6.0 Igpm. As indicated in the results, the average flowrate for backwashes where the proper expansion occurred, was 6.33 Igpm and adequate backwashing could be accomplished in approximately 4 minutes and 30 seconds.

Thus, if backwash was recycled back to the head end of the system, the expected weekly volume increase would be 57 Igallons, or approximately another day's flow.

(ii) Reaeration of the waste stream in the clarifier collection trough prevented the column from going anaerobic. This prevented the formation of odourous gases (e.g. H_2S) in the column.

CHAPTER 6

6.0 CONCLUSIONS

From the testing that was performed one can conclude the following:

1) The system operated well as a treatment process. Generally the system provided consistent final effluents and was capable of treating wastes with high organic and solid concentration such as those concentrations employed during the testing procedure.

2) The overall system obtained an average SS and COD removal of 96.4% and 94.3% respectively, which resulted in an average SS and COD level in the final effluent of 13.8 mg/l and 36.5 mg/l respectively. As a result the system satisfied the "desired" (i.e. SS = 5 mg/l, COD = 20 mg/l) and "not to exceed" (i.e. SS = 15 mg/l, COD = 50 mg/l) objectives 78% and 34% of the time for SS and 81% and 41% of the time for COD.

3) The reactor was closer to operating as a completely stirred tank rather than a plug flow reactor. However, the downstream components encouraged plug flow behaviour.

4) Overall solid reductions were approximately 51%. The average observed yield factor in the reactor was .3 based on COD removals. The average sludge age was found to be 10 days, and the average F/M ratio was found to be 0.31. These values indicated that the system operated closer to a conventional activated sludge system rather than the intended extended aeration system.

5) The system operated well at start-up. The use of coagulant ensured good treatment in the reactor at the beginning of the testing period as did the activated carbon in the adsorption-filtration column. After one month, when the biological activity was at an optimum level, the overall system did not require coagulant addition. Also, it was shown that the biological activity recovered rapidly after the system was shut down for 7 days.

6) Based on projected values, the activated carbon in the adsorption-filtration column would have been exhausted in 6 months. The maximum loading for the carbon, including both adsorption and biological degradation, was 11 lb of COD per 100 lb of activated carbon.

7) The system ran without difficulty through the testing period. Operator time was not excessive, however, there are many areas within the system where automatic control (i.e. sludge wastage, backwash, etc.) could reduce operator time significantly.

7.0 RECOMMENDATIONS

Since the test results indicated the treatment system could provide more than adequate treatment, further experimentation seems warranted. To improve the operation and the treatment efficiency of the system, the following recommendations should be noted.

1) The system should encourage the residence time distribution in the reactor to be more plug flow. Since the present reactor was closer to a CSTR the feed concentration was lowered immediately upon entering the reactor. Since first order kinetics are believed to occur at the lower feed concentrations it is felt that if the incoming feed concentration was allowed to stay at a higher level the removal rate would be greater and the detention time could possibly be reduced (i.e. thereby reducing the size of the reactor). Therefore, it is recommended that the reactor be divided into two aerated compartments such that in the first compartment the feed concentration would be higher and the removal rate would be greater. The effluent from this cell would then overflow to the

second compartment where the solids concentration would be higher due to solids returning from the clarifier. With the lower substrate concentration and the higher micro-organism population, endogenous respiration would be more prominent and solids reduction should be greater. One caution, however, is that there may be a problem with floc break-up in the second chamber because of the low F/M ratio. This in turn may cause poor removal of solids in the clarifier.

2) The mixing in the reactor should be changed. It is recommended that varied air flows be tested to see if floc formation can be increased and a check should be made to observe the minimum flow required to maintain a DO in the reactor of 2 mg/litre.

3) There should be better control of F/M ratios and MLSS levels in future runs such that the system can be tested as an extended aeration process.

4) Run 6 in this project should be run again at the low temperature. This would indicate whether the system could operate at a low temperature with a MLSS level of 4,000 mg/litre and whether the applied air was sufficient for biological activity at this MLSS level.

5) Some experimentation should be carried out on the sludge handling (i.e. freezing wasted sludge to dewater it).

6) The system should be revised so that its capabilities to provide total recycle (i.e. final effluent and backwash) can be observed.

7) It is recommended that the head difference between the adsorption-filtration column and the chlorine column be increased so that there would be a longer time interval between backwashes.

8) The backwash collection trough should be designed to collect the carbon washout. This can be done in the following way:

   (i) provide enough surface area to accommodate the settling velocity of the carbon particles

   (ii) provide a horizontal flow of approximately 1.0 ft/sec so that all organics are carried out of the trough

   (iii) provide an overflow weir at the downstream end

In addition, the backwash trough should be raised to 10-11" above the carbon surface instead of the 8.5-9.0" in the present system. This will allow less carbon to enter the backwash collection trough as well as encouraging a longer period between backwashes.

8) Sludge wastage and backwash should be automated. Sludge build-up in the clarifier could be monitored using a photo cell. Once a certain solids concentration is reached, a relay would open a valve on the sludge wastage line. Backwash could be automated by monitoring the head above the carbon in the adsorption-filtration column. At a preset level, a valve on the backwash line would open for a certain length of time, allowing the column to be properly backwashed.

APPENDICES

APPENDIX 1

Effluent Objectives

It was decided that the effluent should attempt to have a value less than 20 mg/litre (mg/l) COD and 5 mg/l SS (i.e. "desired" levels), and have an absolute allowable maximum of 50 mg/l COD and 15 mg/l SS ("not to exceed" levels). The lower values of each parameter were the Government of Ontario guidelines for surface water quality (i.e. less than 5 mg/l SS, 5 mg/l BOD and ≈20 mg/l COD) (Eckenfelder, 1970).

Therefore the literature was searched for alternate criteria which led to the choosing of the upper values. In a marine recycle system tested by the United States Environmental Protection Agency (EPA) the effluent was required to have a BOD and SS less than 50 mg/l. However, with these criteria they had problems with the recycle (see section 2.3.2.2. At these levels, the wastewater is far from biologically stable. Therefore the upper allowable values were chosen to be 15 mg/l BOD_5, ≈50 mg/l COD and 15 mg/l SS.

The upper values correspond to the government of Ontario requirement for discharges into natural water courses.

APPENDIX 2

Calculation of Typical Arctic Flows and COD Concentrations Based on G.W. Heinke's Data (Heinke, 1972)

Heinke claimed that conventional toilets use 3-5 gallons of water per flush and that the total water consumption with people using conventional systems is 180-270 litres per day per capita, or 40-60 Imperial gallons per day per capita. Fifty percent of this water, he claimed, originated from conventional toilet systems, commonly called black water, and the other 50 percent came from household wastes, commonly called grey water.

In an actual study carried out by Heinke, he found that 1.3 litres of human waste was produced per day and if this wastewater was diluted by 7.5 litres of water, it had a COD of 12,000 mg/l. He also reported that man flushes a toilet 7-8 times per day. Assuming in the conventional toilet 3 gal/flush and 8 flushes/day/capita, 109 litres of water would be used for dilution. This results in a dilution factor of 12.5 (110.3/8.8) times Heinke's value of 12,000 mg/l of COD. This means that the COD of "black water" can be found in the 1,000 mg/l area.

Heinke reports that grey water contains 300 mg/l of COD. Thus a conventional water system in the Arctic will produce flows of 200 l/day/capita with a COD concentration of approximately 700 mg/l. This seems to be in general agreement with other values obtained from the literature (Edwards et al, 1972).

However, it seems probable that a low water usage sewer system, i.e. vacuum toilets and sewers, would be a better suited system for an area of limited water resources. Hence any system tested would have to take into account a more concentrated sewage than might be expected in the conventional water system. If a low water usage system was installed, vacuum toilets would likely be utilized. These toilets are known to employ 10-15 percent of the water required by the conventional toilet. Hence, one might expect 0.4 gal or 2 l/flush from a vacuum toilet. Therefore, 8 flushes/day/capita results in the use of 16 litres of water/day/capita for dilution of the black water. As stated earlier, Heinke claimed that in 8.8 litres of black water (i.e. 1.3 litres of waste plus 7.5 litres of water) there was 12,000 mg/l COD. However, if 16 litres are used for dilution the dilution factor is twice Heinke's dilution which means that the 17.3 litres of black water in the low water system would have a COD of 6,000 mg/l. The amount of household water would not change significantly from the conventional system and again would be approximately 110 litres and have a COD of 400 mg/l. This means that in the low water usage system one might expect a flow of 130 l/day/capita at a COD of 1,100 mg/l.

APPENDIX 3

Collection Trough Design (Weber, 1972)

Find critical depth, D_c, when trough is 3" wide

$$D_c = \left[\frac{Q^2}{g\ Wl^2}\right]^{1/3}$$

where Q = flow

g = 32.2 ft/sec^2

Wl = 3" = .25 ft

therefore

$$D_c = \left[\frac{1.81 \times 10^{-8}\ \text{cf}}{32.2\ (.0625)}\right]^{1/3} = 2.2 \times 10^{-3}\ \text{ft}$$

With a negligible slope the upper depth in the trough will be the greatest, therefore

Upper depth of trough = D_u

= .0038 ft

= .046"

rather insignificant for a 3" wide trough

APPENDIX 4

Design of Support Filter

Design of fine filter

D_{15} filter ≤ 5 D_{85} of carbon

D_{85} carbon $=$ 1.5 mm

therefore

D_{15} filter ≤ 6 mm

Permeability Guarantee

D_{15} filter ≥ 5 D_{15} carbon

D_{15} of carbon $=$ .8

D_{15} filter ≥ 4 mm

Design of Coarse Filter

Piping Prevention

*D_{85} of fine filter $=$ 12.0 mm

*D_{15} of coarse filter ≤ 60 mm

Permeability guarantee

*D_{15} fine filter $=$ 4 mm

*D_{15} coarse filter ≥ 20 mm

APPENDIX 4 (cont'd.)

The grain size distribution has been shown in Figure 3-2.

$^*D_x$ = diameter of particle for which x percentage of the remaining particles in media are less than

APPENDIX 5

Systems Hydraulic Head Loss During Normal Operation

1) <u>Column's Hydraulic Head Loss</u>

Head Loss Through Activated Carbon Layer

Interval(%)	Mean dia $(\frac{mm}{ft})$(d)	Reynold's #	C_D	$C_D \frac{x}{d} 10^4$
10	.65 mm $2.1\text{x}10^{-3}$ ft	.012	2000	9.52
10	.9 mm $3.0\text{x}10^{-3}$ ft	.016	1500	5.00
10	1.15 mm $3.8\text{x}10^{-3}$ ft	.02	1200	3.15
30	1.4 mm $4.6\text{x}10^{-3}$ ft	.025	960	6.26
20	1.8 mm $5.9\text{x}10^{-3}$ ft	.033	727	2.46
20	2.3 mm $7.5\text{x}10^{-3}$ ft	.042	571	1.52
				$27.91\text{x}10^4$

Shape factor (Ø) = .73 for pulverized coal

Void ratio (e) = .5

Drag Coefficient (C_D) = $\frac{24}{R_e}$ for $R_3 < 1.9$

APPENDIX 5 (cont'd.)

Velocity through column (V_S) = 2.45×10^{-4} ft/sec for 330 l/day

Viscosity (μ) at 2°C = 11.2×10^{-4} lb_m/ft sec

Length of filter media (L) = 1 ft

Acceleration due to gravity = 32.2 ft/sec^2

$$\text{Head Loss } (h_L) = \frac{1.067}{\emptyset} \frac{L}{g} \frac{V_s}{e^4} \Sigma \frac{C_D x}{d}$$

$$= \frac{1.067}{.73} \frac{1 \text{ ft}}{32.2 \text{ ft/sec}^2} \times \frac{(2.45 \times 10^{-4} \text{ ft/sec})^2}{(.5)^4}$$

$$\times \; 27.94 \times 10^4$$

$$= 0.13 \text{ ft} = 1/8"$$

Although a similar check for the gravel support filter layer and the support plate could have been made, it was not warranted as the head loss would have been even less than the carbon layer, since they are not as thick nor as fine.

2) Head Loss Through the Orifice Which Enters the Chlorine Contact

For a 1/2" diameter opening

Column Surface Area (A) = 1.36×10^{-3} ft^2

Flow through column (Q) = 1.35×10^{-4} ft^3/sec (330 l/day)

Orifice Factor (C°) = .6

Head Loss Through Orifice (h_f)

$$Q = C^\circ A \sqrt{2 g h_f}$$

$$h_f = 4.2 \times 10^{-4} \text{ ft}$$

APPENDIX 5 (cont'd.)

It is expected that head loss through the tubing connecting the carbon column to the chlorine column will be very small.

APPENDIX 6

Backwash Requirements in Carbon Column

The 10% size of the carbon = .8 mm

$$\frac{D_{60}}{D_{10}} = \text{uniformity coef} = 1.9 = >D_{60}$$

$$= 1.52 \text{ mm}$$

$$\upsilon_f(\text{gpm/sf}) = \frac{.00381(d_{60})^{1.82}\ [W_s(W_m-W_s)]^{.94}}{\mu^{.88}\ \text{centipoises}}$$

W_m = specific wt of carbon lb/ft^3

= 1.3* x 62.4 = 81.12 lb/ft^3

W_s = specific wt of water lb/ft^3 = 62.4 lb/ft^3

*carbon's specific gravity

μ = for 2°C = 1.67

υ_f = (minimum fluidization velocity)

$$= \frac{.00381\ (1.52)^{1.82}\ [62.4(18.72)]^{.94}}{(1.67)^{.88}}$$

$$= \frac{(.00381)(2.14)(764.6)}{1.57} = 3.96 \text{ gpm/ft}^2$$

$$= 0.11 \text{ fps}$$

APPENDIX 6 (cont'd.)

$$Re = \frac{le \ \upsilon \ d_{60}}{\mu}$$

where le = density = slugs/ft^3

$= 1.94$ lb sec^2/ft^4

$d_{60} = 1.52$ mm x $3.28 \times 10^{-3} = 4.98 \times 10^{-3}$ ft

$\mu = 3.5$ lb sec/ft^2

$$Re_f = \frac{1.94 \times 4.98 \times 10^{-3} \times 1.1 \times 10^{-2} \text{ ft/sec}}{3.5 \times 10^{-5}} = 3$$

According to Weber (Weber, 1972) no correction has to be applied to this Re

Unhindered settling velocity =

$\upsilon_s = 8.45 \ \upsilon_f = .93$ fps

Re_s for unhindered settling velocity $= 8.45 \ Re_f$

$= 25.4$

Expansion coefficient = ne = 3.22

$$\text{Constant } Ke = \frac{\upsilon}{(\Sigma)^{ne}}$$

$$= \frac{3.96 \text{ gpm/ft}^2}{(.5)^{3.22}}$$

$$= 36.9 \text{ gpm/ft}^2$$

where Σ is unexpanded porosity = .5

$\upsilon = \upsilon_f$

APPENDIX 6 (cont'd.)

Calculate desired porosity at 50% expansion

$$\frac{De}{D} = \frac{1-\Sigma}{1-\Sigma} = >1.5 = \frac{1-.5}{1-\Sigma} = >1.5-1.5\bar{\Sigma} = 4$$

where $\bar{\Sigma} = .73$

Now to find backwash rate reapply this ¯ to

$$\upsilon = Ke\ (\bar{\Sigma})^{ne}$$
$$= 36.9\ (.67)^{3.22}$$
$$= 13.5\ \text{gpm/ft}^2$$

Total Head Loss During Backwash

1) Head loss in line feeding under drain system

Assume 1" diameter line and since the surface area of column = .55 ft^2, total flow = 7.5 gpm

$$\text{Find velocity} = \frac{1.19\ \text{ft}^3/\text{min}}{\text{area}} = \frac{1.19\ \text{ft}^3/\text{min}}{5.44 \times 10^{-3}\ \text{ft}^2}$$
$$= 218.2\ \text{ft/min} = 3.64\ \text{fps}$$

From Water Supply page 12-8

$$H_f = 4.67\ (Q_{cfs}/C)^{1.85}\quad 1/d^{4.87}$$

Assume C = coefficient of hydraulic capacity - 130 approxi-

APPENDIX 6 (cont'd.)

mately for plastic

h_f = .45 ft for 6 ft of line
or .075 ft/ft of line

2) Head loss through carbon

Δh_f = 1 ft (1-.06) (1.3-1)
= (1) (.4) (.3)
= .12 ft

3) Calculate loss through supporting gravel

Assume gravel remains unfluidized as desired and it has a porosity of .6 and a shape factor of 7.5 (Ref. Weber)

υ = 1.8×10^{-5} ft^2/sec
J = 6
g = 32.2 ft/sec^2
γ = .036 fps
$_s$ = shape factor = 7.5 = .6

Using equation 4.2 in Weber = $>h_f/l = J\,\frac{\upsilon}{g}\,\frac{1-\Sigma^2}{\Sigma^3}\,\left(\frac{s}{d_p}\right)^2$

APPENDIX 6 (cont'd.)

Layer Position	Thickness	Mean Size	dp^2	$\frac{h_f}{L}$	h_f
Top	4"	13 mm = .507"	$1.78x10^{-3}\ ft^2$	$1.71x10^{-3}$	$= 5.7x10^4$
Bottom	4"	1.35"	$1.24x10^{-2}\ ft^2$	$2.45x10^{-4}$	$= .81x10^4$
				TOTAL	$= .6.51x10^4$ ft

Loss through underdrain orifice - assume 10 orifices

$$Q = C° A \quad 2 g h_f$$

A = area of orifice, Q = flow (cfs), C° = discharge coefficient, about .6 for submerged sharp crested orifice

$$A = 2.1 \times 10^{-3} \ ft^2$$

$$Q = 2 \times 10^{-3} \ cfs/orifice$$

$$= >2 \times 10^{-3} \ cfs = .6 \quad 2.1 \times 10^{-3} \quad 644.4 \ h_f$$

$$= >4 \times 10^{-6} = .36 \cdot 4.41 \times 10^{-6} \quad 64.6 \ h_f$$

$$= >3.9 \times 10^{-2} \ ft = h_f$$

Total head loss = .039 + .000651 + .12 + .45 + .25

= .86 ft

APPENDIX 7

Design of Backwash Collection Trough and Trough Collection

1) Backwash Trough

The trough completely surrounded the tank. Therefore it had an inside length of 2.62 ft. The trough was inclined slightly towards a collector but for the purposes of design and grade was assumed to be zero.

The channel was rectangular, therefore,

$$D_c = \frac{Q^2}{g\ Wl^2} \quad 1/3$$

$Wl = 3\ 1/2"$

$Q = .02\ ft^3/sec$

$D_c = .72"$

Accumulative Depth = $(D_u) = 1.73\ D_c = 1.25$ inches

2) Backwash Trough Collector

The trough collector was a column located at the downstream end of the trough.

The flow out of the collector must be .02 ft^3/sec.

Flow $(Q) = C° A \quad 2\ g\ h_f$

where $C°$ = orifice constant

APPENDIX 7 (cont'd.)

A = area of ofifice

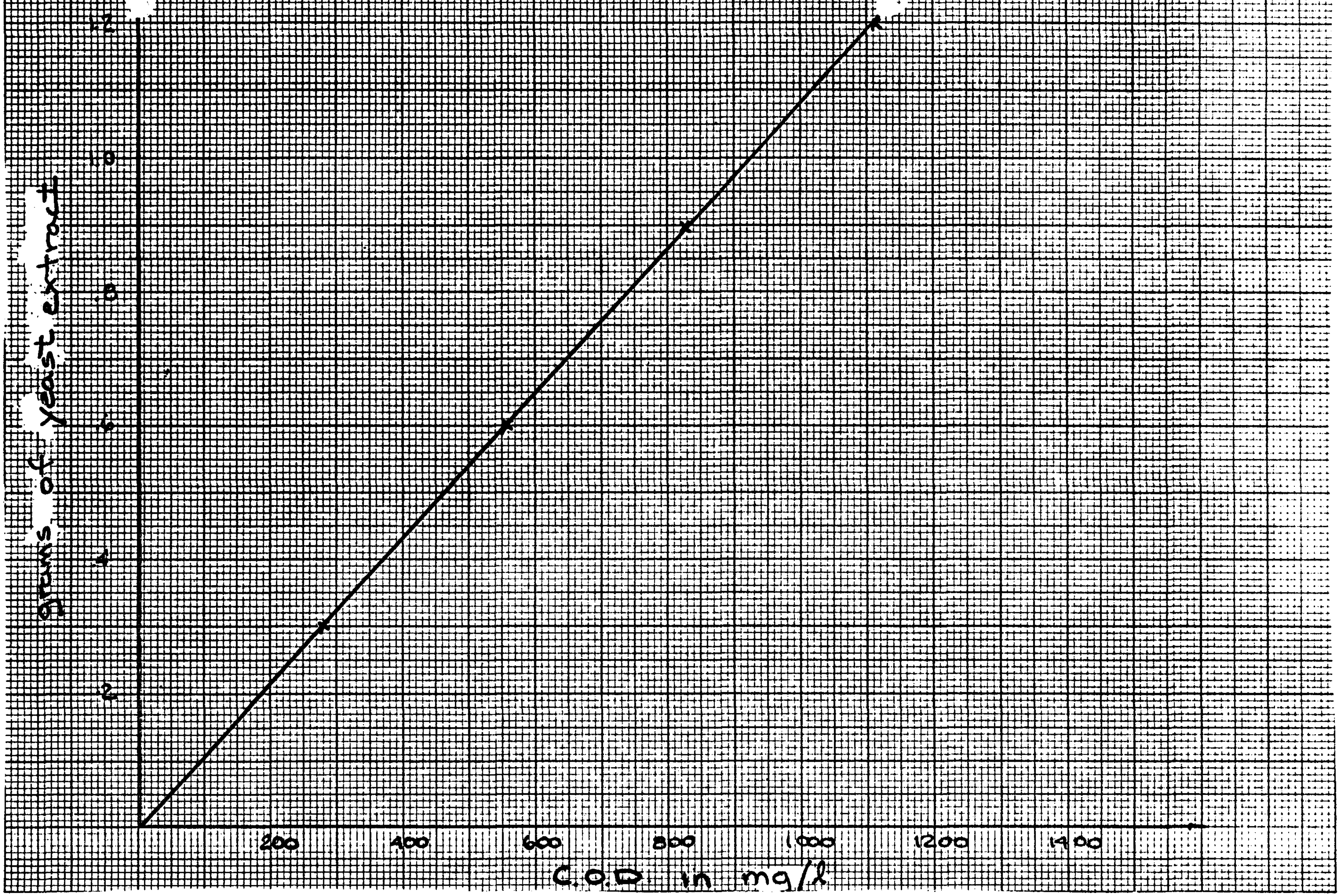
g = acceleration due to gravity

h_f = head loss through orifice

h_F = .58 ft = 7"

Therefore the collection depth of 9" was recommended.

FIGURE A8-1 CALIBRATION OF YEAST EXTRACT TO C.O.D



APPENDIX 9

Residence Time Distribution

Table A9-1 - Recorded Data

Time (min)	Clarifier Reading (Red)	Concentration mg/l NaCl	Overall System Reading (Green)	Concentration mg/l NaCl
0	7	0	6	0
20	9	50	6	0
40	32	450	6	0
60	44	590	6	0
80	47	630	6	0
120	46	620	6	0
300	43.5	600	6	0
600	37.5	520	6	0
780			6	0
800			8.5	100
820			13	200
840			20	330
860			30.5	470
880			34	520
900	32	450	33.5	500
1200	26.5	370	32	480
1800	17	230	23.5	380
2400	13.5	150	18.5	300
3000	11	90	15	240

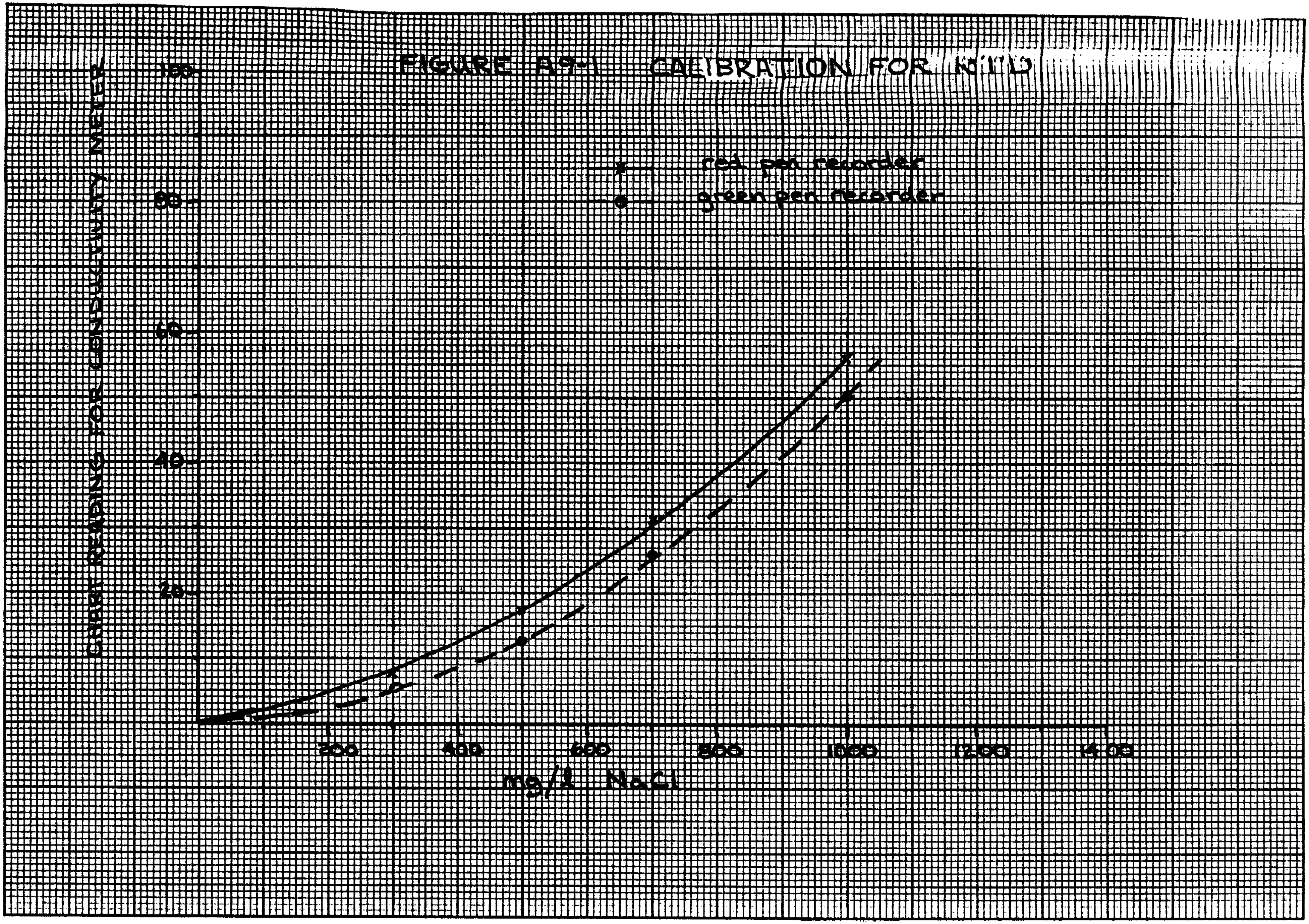


FIGURE A9-1 CALIBRATION FOR NTID

Table A9-2
Adsorption-Filtration Column

Time (min)	Reading	Concentration
0	9	0
1	9	0
1.55	9	0
2 hr	12	70
2 hr 10 min	22	200
2 hr 20 min	26	310
2 hr 30 min	30	370
2 hr 40 min	31	380
3 hr	30	370
3 hr 20 min	38	340
3 hr 40 min	25.5	310
4 hr	22.5	260
4 hr 20 min	20	230
4 hr 40 min	18	190
5 hr	15.5	140

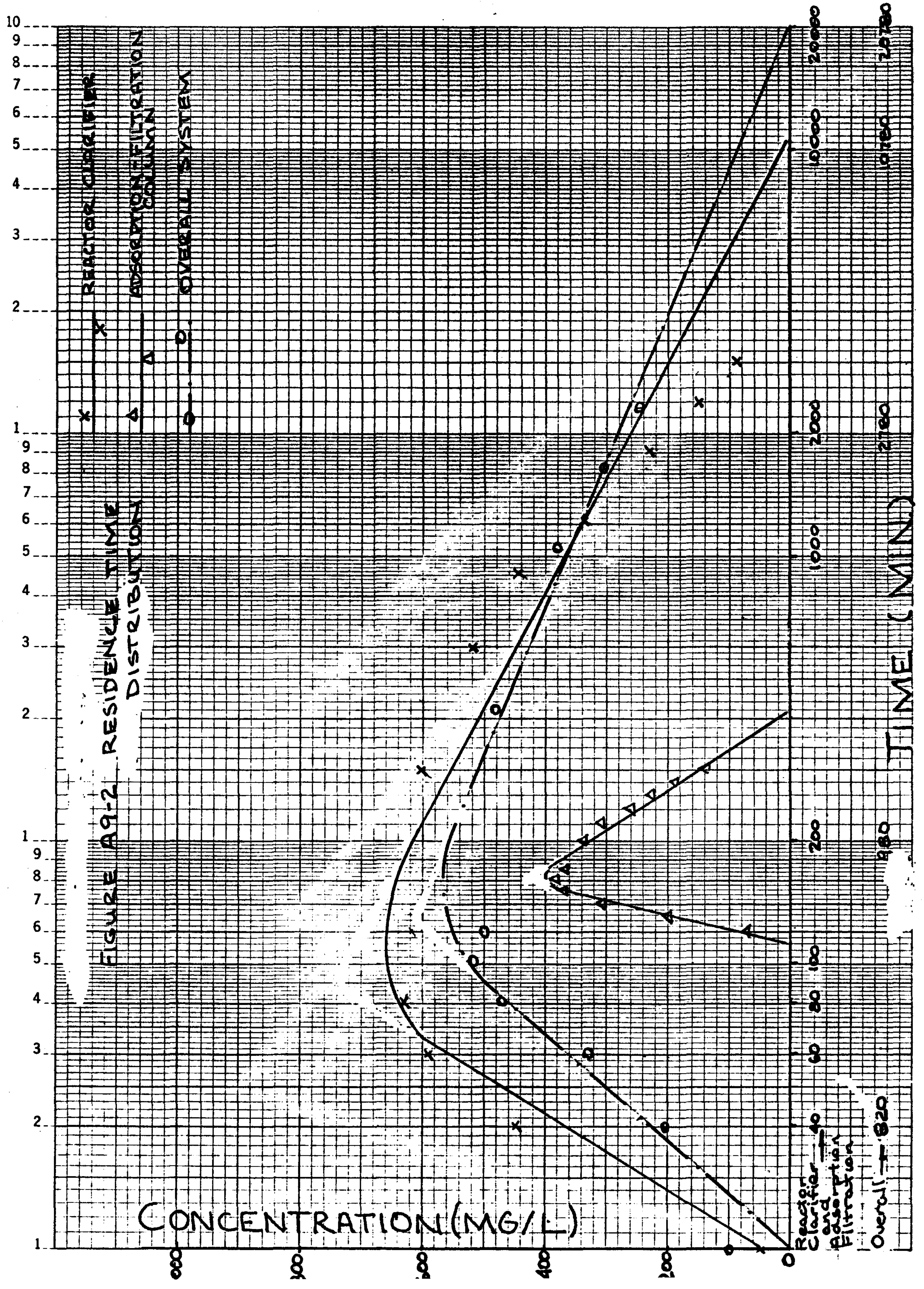


FIGURE A9-2 RESIDENCE TIME DISTRIBUTION

Table A9-3

Overall System Dispersion Number

t (hr)	t^2	Δt	Ci (mg/l)	$Ci\Delta t$	$ti^2 Ci\Delta t$	$tiCi\Delta t$
.33	.11	.33	50	17	2	6
.67	.44	.44	155	51	22	34
1.0	1.0	.33	285	94	94	94
1.33	1.77	.33	410	135	239	180
1.67	2.8	.33	490	162	454	270
2.0	4	.33	570	188	752	1,380
5.0	25	3	555	1,665	41,625	8,325
10.0	100	5	465	2,325	232,500	23,250
20.0	400	10	375	3,750	1,500,000	75,000
30.0	900	10	320	3,200	2,880,000	96,000
40.0	1600	10	280	2,800	4,480,000	112,000
60.0	3600	20	235	4,700	16,920,000	282,000
80.0	6400	20	195	3,900	24,960,000	312,000
100.0	10,000	20	165	3,300	33,000,000	330,000
150.0	22,500	50	125	6,250	140,625,000	937,500
200.0	40,000	50	80	4,000	160,000,000	800,000
250	62,500	50	50	2,500	156,250,000	625,000
300.0	90,000	50	30	1,500	135,000,000	450,000
300.0	108,900	50	10	500	54,450,000	165,000
			4,845	41,037	730,340,686	4,217,039

Table A9-3
(cont'd)

$$t = \frac{\Sigma\ ti\ Ci\ \Delta t}{\Sigma\ Ci\ \Delta t} = 102 \text{ hrs} + 13 = 115 \text{ hrs}$$

$$\text{amt recovered} = \Sigma\ Ci\ \Delta t \times 8.33 \text{ l/hr} = 242{,}000 \text{ mg}$$
$$= 120\%$$

$$\sigma^2 = \frac{\Sigma\ t^2\ Ci\ \Delta t}{\Sigma\ Ci\ \Delta t} = (102)^2$$

$$= 17797 - 10404 = 7393 \text{ hr}^2$$

$$\sigma_\theta{}^2 = \frac{7393}{(102)^2} = .71 = 2\frac{D}{UL} + 8\left(\frac{D}{UL}\right)^2$$

$$\frac{D}{UL} = .2$$

Table A9-4

Reactor-Clarifier Dispersion Number

t (hrs)	t^2	Δt	avg. Ci (mg/l)	Ci Δt	ti^2 Ci Δt	ti Ci Δt
.33	.11	.33	25	8	1	3
.66	.44	.33	210	70	31	46
1.0	1.0	.33	460	152	152	152
1.33	1.77	.33	625	206	365	274
1.67	2.8	.33	705	233	652	389
2.0	4.0	.33	710	234	936	468
5.0	25	3	625	1,875	46,875	9,375
10.0	100	5	500	2,500	250,000	25,000
20.0	400	10	395	3,950	1,580,000	79,000
30.0	900	10	310	3,100	2,790,000	93,000
40.0	1,600	10	260	2,600	4,160,000	104,000
60.0	3,600	20	200	4,000	14,400,000	240,000
80.0	6,400	20	150	3,000	19,200,000	240,000
100.0	10,000	20	105	2,100	21,000,000	210,000
120.0	14,400	20	75	1,500	21,600,000	180,000
140.0	19,600	20	50	1,000	19,600,000	140,000
160.0	25,600	20	30	600	15,360,000	96,000
170.0	28,900	10	10	100	2,890,000	17,000
			5,445	27,228	122,879,012	1,434,707

Table A9-4
(cont'd)

$$t = \frac{\Sigma ti\ Ci\ \Delta t}{Ci\ \Delta t} = \frac{1,434,707}{27,228} = 52.7 \text{ hrs}$$

$$\text{amt recovered} = \Sigma\ Ci\ \Delta t \times 8.33 \text{ l/hr} = 226,900 \text{ mg/l}$$
$$= 113\% \text{ recovery}$$

$$\sigma^2 = \frac{\Sigma\ t^2\ Ci\ \Delta t}{\Sigma\ Ci\ \Delta t} - (52.7)^2$$

$$= 1735$$

$$\sigma_\theta^{\ 2} = \frac{1735}{(52.7)^2} = .62 = 2\ \frac{D}{UL} + 8\left(\frac{D}{UL}\right)^2$$

$$\frac{D}{UL} = .18$$

Table A9-5

Adsorption-Filtration Column
Dispersion Number

t (hrs)	t^2	Δt	avg Ci (mg/l)	Ci Δt	ti^2 Ci Δt	ti Ci Δt
0	0	0	0	0	0	0
1	1.0	1.0	0	0	0	0
1.5	2.25	.5	0	0	0	0
2	4	.5	40	20	80	40
2.17	4.71	.17	140	24	113	52
2.34	5.47	.17	240	41	224	96
2.5	6.25	.16	325	52	325	130
2.67	7.13	.17	390	66	471	176
3	9.0	.33	405	134	1,206	402
3.33	11.1	.33	350	116	1,288	386
3.67	13.5	.34	310	105	1,418	385
4.0	16.0	.33	270	90	1,440	360
4.33	18.75	.33	230	76	1,425	329
4.67	21.8	.34	195	66	1,439	308
5.0	25	.33	165	55	1,375	275
7.0	49	2.0	80	160	7,840	1,120
				1,005	18,644	4,059

$$\bar{t} = \frac{\Sigma\ ti\ Ci\ \Delta t}{\Sigma\ Ci\ \Delta t} = \frac{4059}{1005} = 4.04 \text{ hrs}$$

amt recovered = Σ Ci Δt x 8.33 l/hr = 8,372 mg
= 84% recovery

$$\sigma^2 = \frac{\Sigma\ t^2\ Ci\ \Delta t}{\Sigma\ Ci\ \Delta t} = (4.04)^2 = 2.23$$

$$= \frac{2.23}{(4.04)^2} = .1366 = 2\ \frac{D}{UL} + 8\left(\frac{D}{UL}\right)^2$$

$$\frac{D}{UL} = .055$$

APPENDIX 10

Tabulated Experimental Results

TABLE A10-1 DATA SUMMARY

RUN	SYSTEM INFLUENT (mg/l)			REACTOR-CLARIFIER EFFLUENT (mg/l)			FINAL EFFLUENT (mg/l)			MLSS (mg/l)	% USS	FLOW (l/day)
	SS	COD	DISSOLVED COD	SS	COD	DISSOLVED COD	SS	COD	DISSOLVED COD			
1-1	690	1045		56	332		8	45		0		300
1-2	451	940		54	235		8	50				280
1-3	720	1365		50	368		18	100				250
1-4	684	1297		58	315		18	120		1000	90	330
2-1	304	514		42	100	47	2	2	5	300	97	220
2-2	303	362		20	95	40	3	1	3			210
2-3	270	455	78	20	35	80	9	5	5			90
2-4	219	310		39	17	17	[illegible]	5	5	1100	85	220
3-1	900	1269		92	110	14	7	70	11	1000	82	110
3-2	655	1063		72	112	18	18	50	4			200
3-3	670	950		110	144	76	4	5	4			220
3-4	800	1100	90	101	108	18	14	5	4	3030	85	200
4-1	405	660		78	180	60	10	39	10	1425	79	300
4-2	400	680	100	76	114	34	15	4	10			285
4-3	485	710		56	70	24	18	27	9			220
4-4	305	595		86	101	16	22	23	8	3755	86	320
5-1	380	430		66	108	24	36	40	3	3710	78	30
5-2	300	585		66	123	40	20	32	8			100
5-3	275	450		88	102	24	12	8	1			200
5-4	290	555	85	60	80	24	8	16	8	3890	85	225
6-1	250	500	110	90	256	34	28	86	9	3900	76	230
6-2	200	525		70	156	22	20	40	3			110
6-3	190	515		80	144	28	36	62	8			110
6-4	255	415		58	155	23	17	42	10	3800	77	215
7-1	240	395		78	103	25	12	38	2	1700	71	150
7-2	140	344		96	159	20	8	4	3			110
7-3	275	485		60	95	20	8	10	10			180
7-4	275	500	111	42	72	20	4	7	7	1850	85	230
8-1	195	455		128	230	49	20	132	29	1580	74	220
8-2	220	500	100	78	147	21	10	46	19			150
8-3	330	535		52	90	30	12	40	8			220
8-4	255	410		60	108	25	19	25	8	2236	90	220

TABLE A10-2 OBSERVED YIELDS AND SOLID REDUCTIONS

RUN	Micro-organism produced in R-C (g)	Micro-organism leaving system (g)	Micro-organism leaving R-C (g)	Solids produced in R-C (g)	Solids entering R-C (g)	Solids leaving R-C (g)	Coagulant precipitate produced in R-C	Overall		Reactor Clarifier (R-C)		Adsorption Filtration column solid reduction	Substrate Removed (COD mg/l) system	Substrate Removed (COD mg/l) R-C
								yield	solid redctn	yield	solid redctn			
1	186.65		61.0	207.0	739.21	63.56	19.36g of $Al(OH)_3$ ⇒ 21mg/l of $Al(OH)_3$			* .13	63%			985
2	133.31		20.8	165.6	203.09	23.82	19.36g of $Al(OH)_3$ ⇒ 21mg/l of $Al(OH)_3$			.30	7%			249
3	363.39		63.5	420.21	558.12	66.54	19.36g of $Al(OH)_3$ ⇒ 21 mg/l of $Al(OH)_3$			.30	13%			728
4	403.65		81.4	482.3	440.35	84.90	19.36g of $Al(OH)_3$ ⇒ 21 mg/l of $Al(OH)_3$			.40	-22%			604
5	24.5	33.96	39.68	37.3	161.65	39.68	no alum added	.21	56%	.30	52%	15%	277	232
6	-7.9	41.12	47.0	-20.7	155.23	49.67	8.8g of $Al(OH)_3$ produced during last two days of run	.08	86%	.10	79%	17%	279	193
7	75.66	42.8	42.72	31.1	164.15	42.72	no alum added	.20	55%	.26	55%	0%	291	233
8	174.6	52.8	59.9	135.8	204.6	64.50	7.26 g of $Al(OH)_3$ produced in first two days	.35	8%	.44	2%	18%	333	267

* deceptive value should not be included in average (see sect 5.3.2.3)

Table A10-3

$$\frac{\underline{\text{COD Corresponding to SS}}}{\text{SS}} \left(\frac{\text{COD}_{S_S}}{\text{SS}}\right)$$

$$\frac{\text{COD}_{S_S}}{\text{SS}} = \frac{\text{Total COD - dissolved COD}}{\text{SS}}$$

Run	Influent$^+$	RC Effluent*	RC % Removal	AF Col. Effluent	Col. % Removal$^+$
2-1	1.4	1.26	87	1.39	94
2-2	1.0	2.75	81	1.46	93
2-3	1.38	0	100	0	0 or 100
2-4	1.20	0	100	0	0 or 100
3-1	1.2	1.05	81	8.5	39
3-2	1.33	1.3	89	2.5	52
3-3	1.2	.62	91	0	100
3-4	1.26	.83	90	0	100
4-1	1.33	1.54	78	2.9	76
4-2	1.4	1.05	86	0	100
4-3	1.2	.82	92	1.0	61
4-4	1.6	1.00	83	7.0	82
5-1	.93	1.27	76	1.0	56
5-2	1.6	1.26	83	1.2	71
5-3	1.3	.89	78	.6	83
5-4	1.6	.93	88	1.0	86
6-1	1.56	2.47	43	3.1	66
6-2	2.15	1.92	69	1.9	72
6-3	2.22	1.4	82	1.5	54
6-4	1.33	2.3	61	1.9	76
7-1	1.35	1.0	76	3.0	54
7-2	2.00	1.5	51	0	100
7-3	1.41	1.25	81	0	100
7-4	1.41	1.24	87	0	100

Cont'd./.....

Table A10-3

(cont'd.)

Run	Influent$^+$	RC Effluent*	RC % Removal	AF Col. Effluent	Col. % Removal$^+$
8-1	1.91	1.4	51	5.2	43
8-2	1.86	1.5	72	2.7	72
8-3	1.33	1.2	86	2.7	47
8-4	.29	1.2	78	.87	77

Notes:

$$* \quad \frac{(COD_{Feed} - DCOD_{FEED}) - (COD_{R\text{-}C\ Eff.} - DCOD_{R\text{-}C\ Eff.})}{COD_{Feed} - DCOD_{Feed}}$$

$$+ \quad \frac{(COD_{R\text{-}C\ Eff.} - DCOD_{R\text{-}C\ Eff.}) - (COD_{Col.Eff.} - DCOD_{Col.Eff.})}{COD_{R\text{-}C\ Eff.} - DCOD_{R\text{-}C\ Eff.}}$$

Table A10-4

Mass Loading on Clarifier

MLSS (mg/l)	Flow (l/day)	Mass Loading (g)	% Removal
250	300	75	78
500	280	140	89
750	250	188	93
1000	330	330	94
300	220	66	86
550	210	116	96
800	90	72	97
1100	220	242	96
1000	110	110	91
1700	200	340	96
2400	220	528	95
3030	200	606	97
1425	300	428	95
2225	285	634	96
3025	220	666	98
3755	320	1202	97
3710	30	111	98
3770	100	377	98
3830	200	766	98
3890	225	875	98
3905	230	898	98
3870	110	426	98
3835	110	422	98
3800	215	817	99
1700	150	255	96
1750	110	193	95
1800	180	324	97
1850	230	426	98
1580	220	348	92
1800	150	270	96
2010	220	442	97
2236	220	492	97

Table A10-5

Reactor-Clarifier Mass Loading - Suspended Solids

(1) Low concentration - high and low flow

SS (mg/l)	Flow (l/day)	Mass Loading (g)	% Removal
304	220	66.88	86.0
303	210	63.63	93.4
270	90	24.30	92.6
219	220	48.18	92.2
405	300	121.5	80.7
400	285	114.0	81.0
485	220	106.7	88.0
305	320	97.6	71.2

(2) Low flow - high and low concentration

SS (mg/l)	Flow (l/day)	Mass Loading (g)	% Removal
304	220	66.88	86.0
303	210	63.63	93.4
270	90	24.3	92.6
219	220	48.2	82.2
900	110	99.0	90.0
655	200	131.0	89.0
670	220	147.0	84.0
800	200	160.0	87.0

Table A10-6

Reactor Clarifier Mass Loading - COD

(1) Low concentration - high and low flow

COD (mg/l)	Flow (l/day)	Mass Loading (g)	% Removal
514	220	113	80
362	210	76	74
455	90	41	92
310	220	68	94
660	300	198	73
680	300	198	73
680	285	194	83
710	220	156	90
595	320	190	83

(2) Low flow - high and low concentration

COD (mg/l)	Flow (l/day)	Mass Loading (g)	% Removal
514	220	113	80
362	210	76	74
455	90	41	92
310	220	68	94
1269	110	140	91
1063	200	213	89
950	220	209	85
1100	200	220	90

Table A10-7

Food to Micro-organism Ratio

Run-Day	Micro-organisms (g)	COD Removed	Food Applied Each Day / Micro-organisms Present Each Day
1-1	46.6	213.2	4.58*
1-2	93.2	197.4	2.12*
1-3	139.7	249.1	1.78*
1-4	186.3	325.3	1.74*
2-1	60.2	81.5	1.52
2-2	106.0	56.2	.53
2-3	147.4	37.7	.26
2-4	193.6	64.8	.33
3-1	169.7	127.0	.75
3-2	295.6	191.3	.65
3-3	417.3	177.7	.43
3-4	533.2	200.2	.38
4-1	233.1	144.5	.62
4-2	373.0	161.2	.43
4-3	519.8	140.6	.27
4-4	668.4	158.0	.24
5-1	599.1	9.7	.02
5-2	634.3	46.2	.07
5-3	650.2	70.2	.11
5-4	684.6	107.4	.16
6-1	614.4	56.4	.09
6-2	608.8	40.5	.07
6-3	611.3	40.8	.07
6-4	605.7	56.2	.07
7-1	250.0	43.9	.18
7-2	275.3	21.5	.08
7-3	301.8	70.0	.23
7-4	325.6	99.0	.30
8-1	242.0	49.0	.20
8-2	294.4	53.3	.18
8-3	349.0	97.7	.28
8-4	415.7	66.8	.16

* Deceptive values - should not be included in average; see Section 5.3.2.4

Table A10-8

Sludge Age

Run-Day	Sludge Age
1	0.5
2	1.5
1-3 avg.	2.0
4	2.5
1	2.24
2	3.24
2-3 avg.	3.74
4	4.24
1	3.08
2	4.08
3-3 avg.	4.58
4	5.08
1	3.72
2	4.72
4-3 avg.	5.22
4	5.72
1	9.8
2	10.8
5-3 avg.	11.3
4	11.8
1	16.8
2	17.8
6-3 avg.	18.3
4	18.5
1	19.5
2	20.5
7-3 avg.	21.0
4	21.5
1	21.4
2	22.4
8-3 avg.	22.9
4	23.4

Note: adjustment was made for portion of sludge from previous runs which is wasted in effluent.

Table A10-9

Dissolved Oxygen Levels

Run*	Temp. (°C)	Reactor-Clarifier	Influent to Column	Effluent from Column	Collection Tank
1	7	8.0 ppm	9.0	3.0	3.0
2	6	9.5	9.8	2.7	3.2
3	4	10.2	11.0	1.9	1.9
4	7	6.3	7.1	1.5	3.1
5	4	7.7	9.6	0.4	1.6
6	14	1.5	6.4	0.4	1.9
7	4	7.7	8.2	1.0	1.9
8	6	7.2	8.0	2.0	2.0

*all values taken on day 3 of each run

Table A10-10

Backwash Data

Run	Backwash Number	Volume (gal)	Time	Rate (gpd)	Remarks
5	1	24.2	6 min 17 sec	4 gal/min	did not reach proper expansion level for most of the run
	2	26.4	4 min 15 sec	6.24 gal/min	
6	1	27.5	4 min 10 sec	6.52 gal/min	
	2	28.6	4 min 30 sec	6.4 gal/min	
7	1	24.2	4 min 50 sec	5 gal/min	did not reach proper expansion level for most of the run
	2	28.0	4 min 36 sec	6.15 gal/min	
8	1	25.8	5 min 29 sec	4.7 gal/min	did not reach proper expansion level for most of the run
	2	24.2	4 min 40 sec	5.2 gal/min	did not reach proper expansion level for most of the run

APPENDIX 11

Solids Build-up in Effluent

Average solids reduction in R-C = 33.25%

Average solids removal overall = 97.0%

Average solids removal in R-C = 83.0%

Average solids in feed = 385 mg/l

Average solids in final effluent = 11.6 mg/l

Average solids in backwash = 140 mg/l

Average volume of backwash = 120 l

Number of days between backwash = 2

Solids produced per backwash = 16.8 g

	Feed	Solids Removed in R-C	Solids in Effluent Which is Used As Next Day's Feed	Conc. of SS
Day 1	84.7	70.3	2.5	11.6
Day 2	(84.7 + 2.5) = 87.2	72.38	2.6	11.8
	backwash produces 16.8 g of solids			
Day 3	(84.7 + 2.6 g + 16.8 g) = 104.1 g	86.4	3.12	9.2*
Day 4	87.82	72.89	2.64	12.0
	backwash produces 16.8 g of solids			
Day 5	104.14	86.44	3.12	9.2*
Day 6	87.82	72.89	2.64	12.0

*flow = 340 l/day due to backwash.

APPENDIX 11 (cont'd.)

Method of Calculation

1. Day 1's feed was approximated by the average feed concentration times the expected daily feed of 200 l.
2. Day 1's effluent was calculated by multiplying the average SS in effluent x 200 l.
3. Day 2's feed is day 1's solid + solids in day 1 effluent.
4. Day 2's effluent is average SS removal in overall system x day 2's feed.
5. Day 3's feed is day 3's solids + day 2's effluent solids + backwash solids.
6. Day 3's effluent is calculated in the same manner as day 2's effluent.
7. Etc.

APPENDIX 12

COD Build-up in Effluent

Average COD removal in overall system = 94.26%

Average COD in feed for 8 runs = 660 mg/l

Average COD in final effluent = 36.3 mg/l

Assuming an average COD/SS = 1.43, and assuming no DCOD in backwash, the COD of backwash on the average is approximately 1.43 x average SS = 200 mg/l.

Average volume of backwash = 120 mg/l, which means each recycled backwash contributes 24 g of COD.

	Feed	COD in Effluent	Conc. of COD
Day 1	145.2 g	8.0	36.3 ppm
Day 2	153.2	8.43	38.3 ppm
	backwash added 24 g of COD		
Day 3	(145.2 + 8.43 + 24) = 177.6	9.77	*28.7 ppm
Day 4	(145.2 + 9.77) = 154.97	8.52	38.7 ppm
	backwash added 24 g of COD		
Day 5	177.72	9.78	*28.8 ppm
Day 6	154.98	8.52	38.7 ppm
	backwash added 24 g of COD		
Day 7	177.72	9.78	*28.8 ppm
Day 8	154.98	8.52	38.7 ppm

*flow = 340 l/day due to backwash

APPENDIX 12 (cont'd.)

Method of Calculation

1. Day 1's feed was approximated by the average feed concentration times the expected daily feed of 200 l.
2. Day 1's effluent was the average COD in effluent times 200 l.
3. Day 2's feed is the COD in day 1's feed plus COD in day 1's effluent.
4. Day 2's effluent is average COD removal in overall system times day 2's feed.
5. Day 3's feed is day 1's COD + day 2's effluent COD plus backwash's COD.
6. Day 3's effluent is calculated in the same manner as day 2's effluent.
7. Etc.

APPENDIX 13

Filter Ash Lost Test

When measuring mixed liquor volatile suspended solids (MLVSS) a temperature of 600°C is used to ash the solids. However, to provide an accurate measurement of the ashed solids remaining, it is essential to know the change in weight with the filters and containers used during ashing.

Therefore a test was set up to establish what container (i.e. aluminum cup or crucible) should be employed to hold the filters and the effect of the temperature on the filters (i.e. after ashing). The results indicated that aluminum cups varied greatly in weight before and after ashing, but crucibles, if allowed to cool in a moisture-free environment for 15-20 minutes, had no appreciable weight variation. In addition, within the accuracy of measurement, the .45μ Membran filter had a total weight loss and the fibreglass prefilter, which was used in conjunction with the .45μ filter, had no detectable weight loss.

APPENDIX 14

Coagulant Jar Tests

Jar tests were carried out to establish the optimum alum dosage in the reactor-clarifier.

The testing procedure followed was similar to that recommended by Banci and Benedek (Banci and Benedek, 1976).

Six identical jars contained wastewater similar to that which would be used when testing the system. One jar was used as a control (i.e. no alum was added). To the remaining five jars, alum was added in various concentrations (i.e. 90, 120, 150, 180, 200 mg/l of $Al_2(SO_4)_3.10H_2O$). The optimum dosage was found to be 150 mg/l.

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