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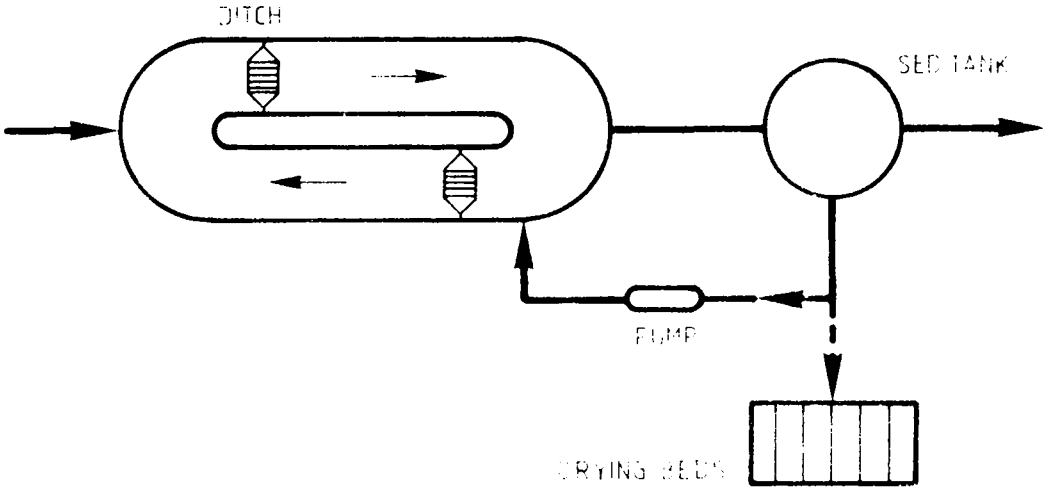
Hadgu Rassu

Temperature Effects on Extended Aeration Activated Sludge Process



Tampere 1984

TEMPERATURE EFFECTS ON EXTENDED AERATION ACTIVATED SLUDGE PROCESS

by

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ACTIVATED SLUDGE PROCESS

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ABSTRACT

Results obtained from this experiment and literature information were analyzed to define and evaluate the effects of temperature variation on the performance efficiency and some fundamental biological kinetic coefficients. The parameters investigated were: substrate removal (both BOD_7 and COD), nitrification, suspended solids (SS) removal, sludge settleability, cell yield coefficient, specific organism decay rate, and sludge production and characteristics. Removal efficiencies of pollutants at different temperatures were calculated and compared. Cell yield coefficient and specific organism decay rate were computed using regression analysis. Settled sludge volumes and sludge volume indices (SVI) were determined to assess sludge settleability.

Variations of BOD removal efficiencies with varying temperature were not discerned. Slight variations in COD removal rates were observed with a maximum removal efficiency at 22°C/13°C (alternate mixed liquor temperatures). The classic progression of nitrification with increasing temperature was observed. SS removal rates were mostly erratic and difficult to assess. SVI values were found to be more affected by F/M ratio than temperature variation. Both cell yield coefficient and specific organism decay rate showed maximum values at room temperature (22°C), with lower values at 11,2° and 32,5°C.

1. INTRODUCTION

Wastewater treatment practice in the developed world mostly relies heavily on high maintenance, mechanically complex systems. Such systems are far from being optimal for the developing world. The cost of these systems, both capital and operational, and the requirement for high-level operator skills limit their feasibility for such applications. Consequently, a serious need exists for reliable, low-maintenance, mechanically simple wastewater treatment systems.

The increased emphasis on a high degree of wastewater treatment, among other things, has increased the popularity of the activated sludge process. However, this process in its use in the last seven decades has not been without its brand of design and operational problems; but the majority of modifications to the process have developed through solving operational problems rather than from research.

The suitability of the conventional activated sludge process and some of its modifications for the conditions of the developing world is questionable due to the level of technical knowhow required for the design, operation and maintenance of these systems.

One of the widely used versions of the activated sludge process is the extended aeration system. Extended aeration activated sludge plants have fulfilled a need for high efficiency treatment of small volume wastes of both domestic and industrial origin. They have been found especially suitable for rural or small communities, hotels, schools, institutions and some industrial installations. Besides producing high quality effluents, plants functioning in this mode have been acclaimed for their simplicity of operation and maintenance, low cost and stability against environmental changes. The oxidation ditch, for example,

which is one form of the extended aeration process, with a closed loop earth channel performs better and needs less maintenance than an oxidation pond which may seem to be a cheaper alternative.

The extended aeration process is actually a completely-mixed activated system. This explains why plants in this mode have worked well and have not been plagued with the operational problems of conventional systems.

In light of the above mentioned features of the extended aeration process, which are observations of years of usage, it would not be presumptive to say that this process could suit well the needs and conditions that exist in most of the developing world, although its power needs could pose problems in some cases.

Temperature effects on activated sludge systems have received little attention in the past, probably because of the lack of an inexpensive method for changing or controlling wastewater temperatures. Most of literature dealing with design or simulation studies characterizes temperature effects on the substrate removal rate for an activated sludge system in terms of the modified Arrhenius Equation. Most of this literature contains insufficient description of process parameters, insufficient data, only qualitative observations, or data observed under widely varying test conditions, which perclude the derivation of meaningful temperature-rate relationships for activated sludge systems in general.

The activated sludge process is basically a biological process; and biological processes are known to be temperature dependent. In a completely mixed activated sludge (CMAS) system, temperature is known to affect the growth kinetics of the bacteria, the settling characteristics of the sludge,

and the oxygen transfer characteristics of the system, all of which must be coupled to determine the overall system performance.

The deficiency of experimental data on the influence of temperature on the activated sludge process performance is even more pronounced when it comes to the advantageous ambient temperature prevailing in the tropics. Because most developing countries are in the tropics, it is important and useful to investigate and ascertain how temperature influences the efficiency of waste treatment processes with the eventual aim of drawing criteria that could enable plant designers for the tropics to allow for beneficial ambient temperature effects.

The reasearch reported in this paper is an effort to provide additional data regarding certain effects of temperature on the extended aeration activated sludge process. Concurrently, some of the earlier works done and reported by researchers have been re-examined to give a background review and to ascertain differences and similarities in the results obtained. The study considered the specific influence of temperature on the performance of a CMAS system, which was simulated by two laboratory scale models, and the effects on some of the biological kinetic parameters. The primary objective was to determine in a preliminary manner, the variation of performance, the cell yield coefficient and the specific organism decay rate with temperature. Performance was determined by effluent quality and sludge settleability, while actual values of the kinetic parameters were computed for the different temperatures investigated.

2. EXTENDED AERATION ACTIVATED SLUDGE PROCESS

2.1 General

Modification of the activated sludge process which has gained popularity in recent years is the extended aeration process. Typical applications of this process are to be found in the Pasveer and Carrousel type oxidation ditches widely used in Europe and elsewhere, in the "package" plants available in the U.S.A. for small installations and mechanically aerated lagoons which can be designed on the extended aeration principle if desired (extended aeration lagoons).

The main features of an extended aeration activated sludge system are 1) an extended period of areation, 2) low organic loading (low F/M ratio) and 3) usually high biological solids concentration (high MLSS concentration). Wastewater treatment plants intended to function in the extended aeration mode are usually desinged for an organic loading ranging from 0,02 to 0,1 kg BOD/kg MLSS·d /21/22/, a hydraulic retention time (aeration time) of 12 - 36 hours /22/2/ and MLSS concentration of 3 000 - 8 000 mg/l /33/.

This process is simpler to construct and operate than the conventional activated sludge plant /2/. Plants employing this process are usually designed without primary sedimentation tanks and the employment of an extended aeration time fulfills the purpose of aerobically digesting or destroying the majority of the biological sludge produced. Since the sludge is sufficiently mineralized and excess quantity does not need any further treatment in a digester before dewatering, the operation of these plants simplifies down to only one type of process, namely, the aerobic type, and general operational control becomes subsequently easier /2/.

Although different modifications of the extended aeration process have been used, the basic features of any plant functioning in this mode are illustrated in figure 1.

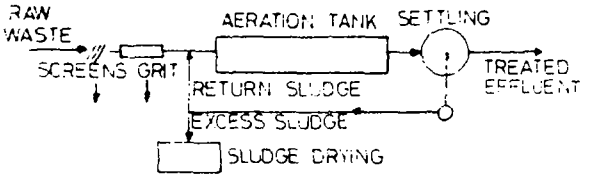


Figure 1. Flowchart of an extended aeration process./2/

Screening or comminuting or both are carried out ahead of the aeration tank in most plants /26/. Supplemental units such as sand filters, stabilization lagoons and even sludge holding tanks have been added to withhold suspended matter from the final effluent /26/.

One of the best-known versions of the extended aeration process is the oxidation ditch. Oxidation ditch usually assumes the shape of an elongated oval, but may be bent at one end, both ends, circular or any other geometrical shape as long as it forms a complete circuit /33/. The original oxidation ditch (developed by Pasveer in the Netherlands) operated on the "fill and draw" method /28/. This ditch comprised a continuous circuit excavation, 1 - 2 m deep and an earth bank having a 45° slope /28/. Horizontal rotors mounted in the ditch provided oxygenation, circulation and mixing of the contents /28/.

Three basic types of oxidation ditches (fig. 2) are now in use: 1) the continuous, 2) split channel, and 3) intermittent types of plant /28/. Of these the continuous system is the most common, and this is the type usually referred to as the Pasveer oxidation ditch /28/. As larger Pasveer ditch plants were constructed it became apparent that, using the original TNO rotors for oxygen transfer, the channel

depth was restricted (~1,8 m was the maximum that could be constructed)/28/. This restriction often resulted in plants with larger surface area, and hence was expensive in the use of land /28/.

This technical limitation led to the development of the Carrousel system and use of vertical aerators. With a vertical aerator, oxygen transfer could take place in a deep section of a channel which contains the aerator. Besides retaining some of the advantages of the original Pasveer ditch, the Carrousel was developed with a more energy-efficient and lower cost system in mind /21/.

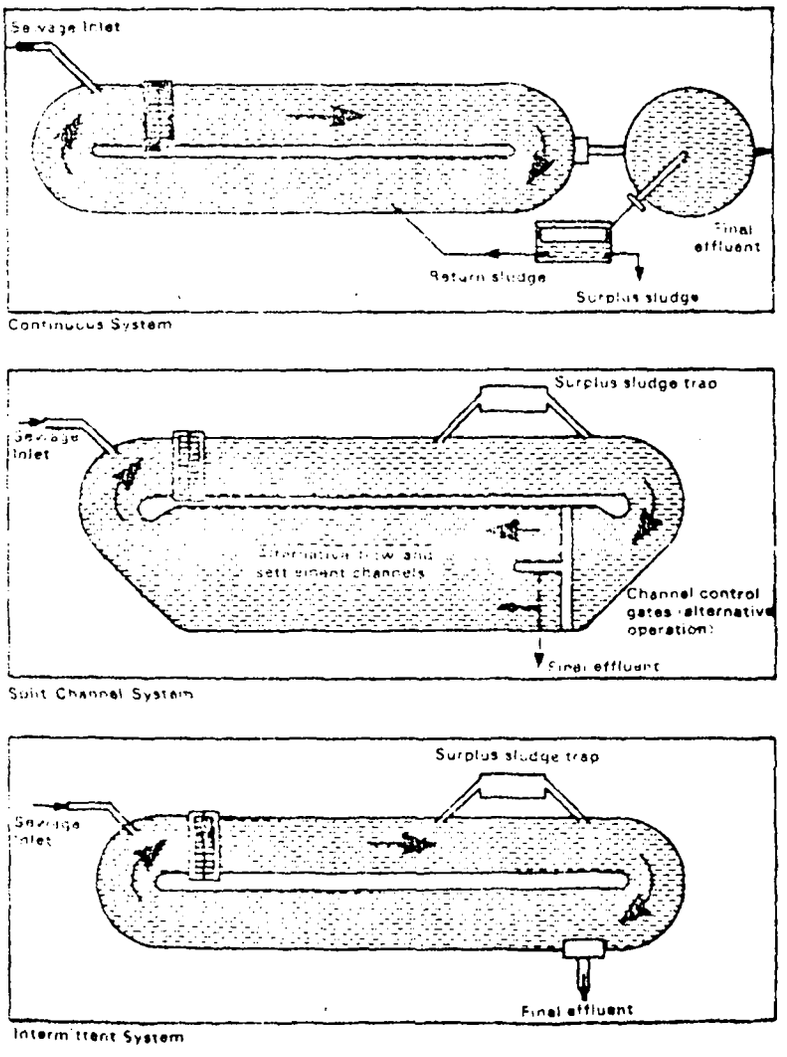


Figure 2. Basic types of oxidation ditch./28/

Since all the parameters required for efficient biological removal of the organic wastes are present in this process (extended aeration), the highest BOD removal rate (95 to 98 %) /2/, compared with any other process is attained. In practice, BOD removal efficiencies of less than the theoretical figures are obtained. However, this is not related to the initial conversion of waste into biological solids, but rather to the discharge of degradable biological solids to the plant effluent /26/.

It must be mentioned here that, the simplification and ease of operation of this process is offset to some extent by the fact that more power (for aeration) is consumed and the suspended solids concentration in the effluent is usually higher than in the conventional activated sludge process /12/.

2.2 Theory of Operation

2.2.1 Process Control Parameters

The overall control of an extended aeration process depends on biomass control and control over dissolved oxygen levels /21/. Biomass control may be obtained in a number of ways. The most common ways of achieving this are /30/:

1) control by maintenance of a constant mixed liquor volatile suspended solids (MLVSS),

2) control by maintenance of a constant food-to-micro-organism (F/M) ratio, and

3) control by maintenance of a constant mean cell residence time (MCRT).

The above parameters may be adjusted as required to meet changes in wastewater characteristics or temperature.

2.2.1.1 Constant mixed liquor volatile suspended solids

With this method, a constant mass of microorganisms is maintained to make use of the incoming food supply. If a certain concentration of MLVSS is found to work effectively at a given plant then this level of concentration will be maintained. If the MLVSS concentration happens to increase above this optimum level, some solids should be wasted till the concentration falls down to the optimum concentration level. If the MLVSS drops below the optimum, less solids should be wasted to allow the MLVSS concentration to increase.

This system of solids control is simple to understand and manage, involves a minimum amount of laboratory work, and can produce good results if the incoming wastewater strength is stable. However, this method has a rather severe limitation in that the important F/M ratio is ignored. If the BOD load of the incoming wastewater is increased over a substantial period of time, the increased solids production from the high loading would be wasted to maintain the constant MLVSS. The result of this action, however, is that the F/M ratio is higher than the previously maintained ratio. The resulting high F/M ratio easily could lead to process inefficiency or failure /30/.

2.2.1.2 Constant food-to-micro-organism ratio

With this method, the MLVSS concentration is increased or decreased to match an increase or decrease in the BOD entering a plant, so as to maintain a certain desired F/M ratio. The use of this method presupposes the knowledge of the organic (BOD) load present and the MLVSS concentration. While the latter parameter could be measured readily at any time, real-time data regarding the former parameter are not available (BOD_7 analysis takes 7 days). This organic load can sometimes be approximated by making use of chemical oxygen demand (COD) or total organic carbon (TOC) measurements combined with BOD/COD or BOD/TOC correlations /21/. As can be imagined this method requires a large amount of laboratory work.

An additional disadvantage of this method is that the MLVSS is not an accurate measure of active mass of micro-organisms. Among other things, dead cells (unlysed once) are registered in volatile solids test. As a result, even with a constant BOD_7/MLVSS ratio, there is no assurance that the F/M ratio is constant /30/.

2.2.1.3 Constant mean cell residence time

The mean cell residence time (MCRT) is the most logical parameter on which to base the operation of an activated sludge plant /16/. The control of an activated sludge plant through MCRT has been practiced since the inception of the treatment process /16/. Activated sludge is wasted from the return sludge flow to maintain a desired concentration of mixed liquor suspended solids (MLVSS). If, by this wasting operation, the MLVSS concentration is kept constant, then the weight of waste sludge solids represents the net sludge growth in the system. Sometimes, a certain MCRT that is required to produce a desired degree of treatment is set (beforehand) and the maintenance of a solids wasting rate that would produce this MCRT is calculated. In doing this there are two obstacles to overcome. The first is that solids are wasted unintentionally in the effluent and secondly, the relationship between MLVSS and return sludge solids is not constant and cannot always be controlled closely by the rate of return sludge pumping since settleability of the activated sludge influences the ratio of MLVSS to return sludge volatile suspended solids (RSVSS) /16/. Hence a return sludge wasting rate, established to produce a given MCRT at a low sludge volume index (SVI), will not be great enough to maintain this same MCRT if the SVI increases significantly. The SVI increase would cause the return sludge solids concentration eventually to decrease. A measurement of RSVSS on a composite sample easily can establish the desired waste flow rate to be used to accomplish the wasting of a certain quantity of sludge solids /16/.

Another way of control of a plant through MCRT is by wasting sludge direct from the aeration tank. This process of wasting mixed liquor rather than return sludge is called hydraulic control of the activated sludge system /30/. This control method is easier yet than the previous one. If a steady state is achieved i.e. solids produced equal those wasted both intentionally and unintentionally, control through MCRT could be achieved by simply wasting a constant percentage of the mixed liquor in the aeration tank (reactor) each day. If 5 % of the mixed liquor in the reactor is wasted daily, a MCRT of 20 days will be obtained. Since the active part of the MLVSS will be wasted approximately as a fraction (within a range) of the total solids wasted, control could be excercised simply by wasting the necessary percentage of total solids within the reactor /21/.

The hydraulic control system is self-regulating. As the concentration of the MLSS increases, the mass of solids wasted also increases, lowering the MLVSS concentration over a period of time. If the MLSS concentration decreases below the level dictated by the selected MCRT, the mass of solids wasted decreases, raising the MLVSS concentration over time. The hydraulic control system lends itself to automation readily and is the simplest system to control /30/. This method of controlling a treatment process results in stable operation of the process because there will be less variation in the F/M ratio and MCRT than the other control methods /30/.

2.2.1.4 Dissolved oxygen control

The other major control parameter mostly applicable to oxidation ditch plants is the dissolved oxygen concentration in the reactor. Almost all oxidation ditch plants are designed to allow some control over oxygen input /21/.

This may be accomplished with variable submergence or speed on mechanical aerators and blower turndown when jet aerators are used /21/. In plants designed for BOD removal and/or nitrification, control is utilized to maintain a dissolved oxygen level greater than or equal to 2 mg/l /21/. Considerable cost savings through reduced aeration power requirements may be obtained through control to prevent unnecessarily high dissolved oxygen levels /21/.

2.2.1.5 Sludge recycle rate

The required rate of recycle of sludge from the settling tank to the reactor is related to the sludge volume index (SVI) and MLSS concentration of the mixed liquor in the reactor . In return the SVI limits the reactor MLSS concentration that can be achieved because it controls the settling tank underflow concentration. For example, if a mixed liquor with MLSS concentration of 5 000 mg/l and SVI of 100 is taken and 1 l of this mixed liquor were settled for half an hour, the solids would settle to a volume of 0,5 l. At least 0,5 l would have to be recycled to the reactor to retain all the sludge. This would correspond to a recycle rate of 1:1 based on the influent flow to the plant /23/.

According to McCarty and Brodersen, the sludge recycle rate may be formulated as follows /23/:

$$\frac{R}{R+1} = \frac{V}{1000} \qquad (2.1)$$

where, R = ratio of sludge recycle rate to influent flow,

V = the volume of concentrated sludge in ml after half an hour settling of mixed liquor in a 1-l graduated cylinder.

Sludge recycle rates of 1:1 or 2:1 are desirable and adequate for most plants /23/. This range of recycle rate gives reasonable felxibility to adjust the MLSS to the desired concentration /7/. However, the ability of a plant to retain a maximum concentration of MLSS depends not only on the recycle rate, but also on the ability of the settling tank to efficiently settle and retain the solids. Because of this, the efficiency of an extended aeration activated sludge plant depends to a large degree on the proper design of the settling tank, which must not only provide for proper settling at average flows, but also must provide for efficient operation in the prescence of any surge flows that might come into the plant /23/. It is often due to such surge flows that large quantities of solids are discharged to the effluent /23/.

2.2.2 Substrate Stabilization

The biological removal and stabilization of organic wastes in an activated sludge process is shown in figure 3. This conversion of organic wastes into end-products can be thought to take place in two different stages, which occur simultaneously in the same tank.

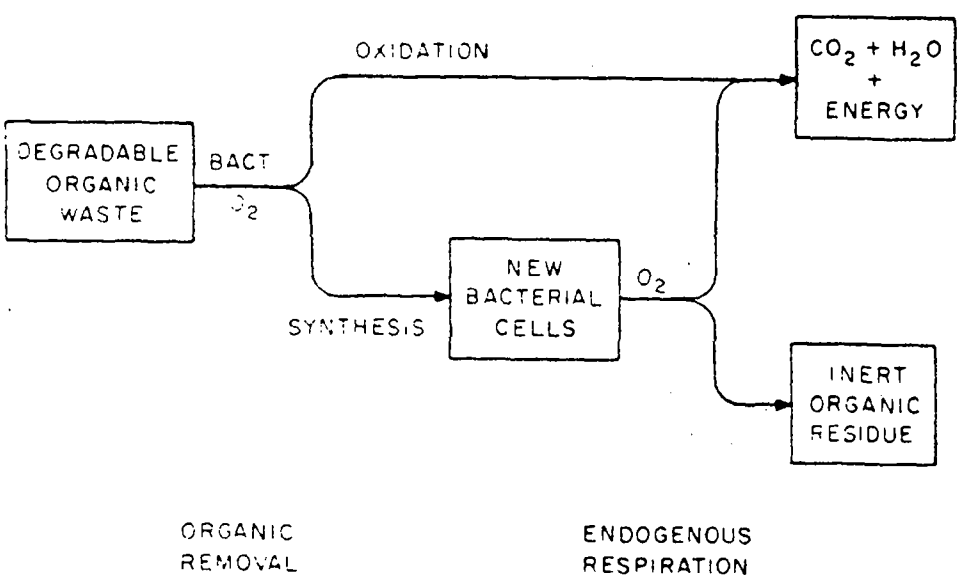


Figure 3. Conversion of organic wastes in activated sludge. /23/

First the degradable organic wastes are metabolized and partially used for energy and partially synthesized into new microbial cells. With continued aeration, in the second stage, the microbial cells formed (their protoplasm) are metabolized for further energy (this so-called endogenous phase takes place when substrate concentration is at a minimum). The active biological solids utilize oxygen while undergoing endogenous respiration to oxidize the degradable portion of their cells to carbon dioxide and water with a resulting decrease in sludge mass /23/. Ideally, with extended aeration this sludge mass would continue to oxidize itself to carbon dioxide and water so that no net sludge accumulation would occur in the system /23/. However, it has been found out that some of the biological solids produced are relatively inert to biological oxidation and so accumulate in the system /23/. As a result, with continuous operation and no sludge wasting intentional or unintentional (controlled sludge wastage or loss of biological solids to the effluent), from an extended aeration system an increase in mixed liquor volatile solids (MLVSS) will result.

The total accumulation of sludge in the system will be equal to the accumulation of biological solids plus the accumulation of biologically undegradable suspended solids which were originally present in the influent waste.

In extended aeration plants where controlled sludge wasting is not practiced, the solids will increase until the system capacity to settle and retain the solids is exceeded /23/. The excess solids produced will then be discharged to the effluent, resulting in an increase in effluent BOD and as a consequence a decrease in overall efficiency /23/.

2.2.3 Nutrients Removal

2.2.3.1 Nutrients Requirements

Domestic wastewater contains more than a sufficient amount of nitrogen and phosphorus, compared with carbon, for the correct balance of nutrients required for bacterial growth /2/. The question of nitrogen and phosphorus utilization must be considered from two view points /2/; 1) minimal requirement and 2) the extent of removal in biological treatment.

To ensure maximum removal of carbon (and therefore BOD) from a given substrate, carbon must be made the limiting nutrient, while nitrogen and phosphorus (and other nutrients also) must be present in at least the minimum required proportion for cell growth /2/.

Nutrient requirements in the conventional activated sludge process are higher than those of the extended aeration process since, in the latter, there is more recycling and less sludge withdrawal /2/. It is suggested that the ratio of BOD_7:N:P should be about 100:5:1 /22/.

2.2.3.2 Phosphorus Removal

Phosphorus removal in biological treatment is due mainly to the removal of microbial solids from the system. At steady state, the surplus microbial solids removed equal those produced per unit time. Thus it is necessary to estimate the net volatile solids produced, and to know the phosphorus content of the solids /2/.

Phosphorus removal from a given wastewater is a function of the MCRT of the system /2/. As MCRT increases, the phosphorus removal rate decreases since less surplus sludge is removed from such a system. Thus due to a higher MCRT, an extended aeration process should give less phosphorus removal than conventional activated sludge. Conversely when treating phosphorus-deficient wastewater, the extended aeration process should require less artificial phosphorus addition than the activated sludge process /2/.

Generally, chemical precipitation is found to be a more efficient method of phosphorus removal than biological treatment.

2.2.3.3 Nitrification

Most of the nitrogen in treated wastewater is in the form of ammonia /25/. Consequently, when wastewater containing ammonia is discharged to the environment, depletion of recipient-water oxygen can occur as the ammonia is oxidized to nitrate. This depletion of oxygen can be avoided if the ammonia is first oxidized to nitrate before it is discharged. Nitrification, which is the bio-oxidation of ammonia to nitrate, is the process used to accomplish this objective. This conversion of ammonia to nitrate is performed by nitrifying micro-organisms (either heterotrophic or autotrophic) which are present in almost all aerobic biological treatment processes, but usually their numbers are limited /12/25/. In spite of the fact that over one hundred species have been cited as forming nitrite (NO_2^-) from ammonia, it is doubtful that significant quantities of nitrate are generated heterotrophically in natural systems /12/. Consequently, most research into nitrification in wastewater treatment systems has concentrated on the autotrophic micro-organisms /12/.

The major nitrifying bacteria are thought to be of the genera Nitrosomonas and Nitrobacter. These organisms obtain their cell carbon from carbon dioxide and energy from the oxidation of inorganic compounds (in this case, ammonia and nitrite)/22/. Nitrosomonas oxidizes ammonia nitrogen (NH_4^+-N) to nitrite (NO_2^--N) through a relatively complex pathway which is not yet completely understood, and remains largely speculative /12/. Nitrobacter oxidizes NO_2^--N to nitrate nitrogen (NO_3^--N) in a single step /12/.

The theoretical stoichiometric equations for the growth of Nitrosomonas and Nitrobacter are /12/:

for Nitrosomonas,

$$55NH_4^+ + 76O_2 + 109HCO_3^- \longrightarrow \overset{\text{cells}}{C_5H_7O_2N} + 54NO_2^- + 57H_2O + 104H_2CO_3 \qquad (2.2)$$

for Nitrobacter,

$$400NO_2^- + NH_4^+ + 4H_2CO_3 + HCO_3^- + 195\,O_2 \longrightarrow \overset{\text{cells}}{C_5H_7O_2N} + 3H_2O + 400NO_3^- \qquad (2.3)$$

These equations are important for two reasons /12/. First, they tell us that a large amount of alkalinity (HCO_3^-) will be utilized during the oxidation of NH_4^+-N: 8,64 mg HCO_3^-/mg NH_4^+-N oxidized /12/. A small part of this will be incorporated into the cell material, but the majority will be used to neutralize the hydrogen ions (pH) released during the oxidation /12/. If the water contains insufficient alkalinity, nitrification will be retarded because of the unavailability of a needed reactant and the deleterious effect of the resulting low pH /12/. Efficient nitrification usually consumes alkalinity and consequently causes pH depression. Secondly, they tell us that approximately

3,22 mg O_2 will be required for each mg NH_4^+-N oxidized to NO_2^--N, and 1,11 mg O_2 will be needed for each mg of NO_2^--N oxidized to NO_3^--N for a total of 4,33 mg O_2 per mg of NH_4^+-N oxidized all the way to NO_3^--N /12/.

The nitrifying organisms are sensitive to their environment, and even under favourable conditions they are slower growing than the usual BOD-removing micro-organisms (heterotrophs)/2/. The former have a doubling time of 1 - 2 days while the latter have typical doubling time of 0,25 - 1,5 hours /22/. Since the growth rate of Nitrosomonas is lower than that of Nitrobacter, it is the one that determines the rate of nitrification /20/.

It has been found out that the following factors have a significant effect on the nitrification process /25/: ammonia concentration, BOD/Total Kjeldahl Nitrogen (TKN) ratio (TKN = NH_4^--N + $N_{organic}$), dissolved oxygen concentration, pH, temperature and MCRT.

Ammonia concentration in a reactor affects the growth rate of the nitrifying organisms. It has been found out that the reaction that determines the growth of the organisms proceeds at a constant rate (zero-order reaction) till the concentration of ammonia reduces to about 1,5 to 2 mg/liter, below which the nitrification rate is reported to drop off rapidly /2/. When the wastewater to be treated contains a very high concentration of NH_4^--N, the concentrations of NO_2^--N and NO_3^--N formed are likely to be high enough to inhibit the organisms so that complete nitrification cannot occur, even at very long MCRT's /12/. Jenkins and Garrison report that previous work has shown that the rate of NH_4^--N oxidation by Nitrosomonas is not limited by NH_4^+-N concentration above 1 to 3 mg/l, the exact value depending on the temperature /16/.

The choice of MCRT for an activated sludge treatment plant must take into account the actual growth rate of nitrifiers at field conditions. Unless the MCRT is somewhat greater than the cell doubling time, cell washout would occur /2/. Thus,

$$\text{MCRT} > \frac{1}{\mu} \qquad (2.4)$$

where μ is the growth rate for Nitrosomonas at the worst operating temperature and other conditions /2/.

Activated sludge plants in cold climates designed for MCRT less than 10 days may show relatively poorer nitrification in winter /2/. Extended aeration plants generally designed for MCRT greater than 10 days may show nitrification uniformly over all seasons /2/. Even activated sludge plants in warm climates may show nitrification uniformly, provided sufficient aeration capacity is available to meet the total oxygen demand and maintain required high dissolved oxygen levels in mixed liquors /2/.

Contact time between incoming ammonia and micro-organisms in the mixed liquor must also be sufficient, besides appropriate MCRT, for successful nitrification /2/.

Nitrifying bacteria are very sensitive to pH /12/. The effects of pH on nitrifiers in mixed cultures, such as activated sludge, are not quite as severe /12/. A wide range of pH optima have been reported, but most researchers agree that as the pH moves to the acid range the rate of ammonia oxidation declines /12/. pH ranges of 7,8 to 9,2 and 8,5 to 9,2 are thought to be optimum for Nitrosomonas and Nitrobacter respectively /2/.

Temperature has a strong influence on the nitrification process. It affects the growth rate of nitrifying bacteria just as it does heterotrophs. Several workers have shown that the effect of temperature on the maximum specific growth rate (μ_m, refer to 3.2.1 for detail) fits an Arrhenius -type equation over the physiological range /12/. It is reported that Wong-Chong and Loehr found that deactivation of Nitrobacter occurred at lower temepratures than did deactivation of Nitrosomonas and that the temperature dependency of both genera was a function of pH /12/.

Dissolved oxygen concentration (DO) in a reactor is one of the parameters that could determine the extent of nitrification. Jenkins and Garrison reported that work by Garret and a group at the British Water Pollution Research Laboratory showed that the rate of NH_4^+-N oxidation by Nitrosomonas was dependent on the DO below levels of about 3,0 mg/l /16/. It is assumed that a DO concentration greater or equal to 1 - 2 mg O_2/l is required for the nitrification process /20/.

It should be mentioned here that nitrification could be a major problem in extended aeration plants. Due to the large excess of air and low organic loadings normally used in extended aeration plants, the condition becomes favourable for the growth of nitrifying bacteria. pH depression (mentioned earlier) in aeration tanks and rising sludge in settling tanks are frequent problems.

Another problem of importance is the effect of nitrification on measurement of effluent BOD and evaluation of efficiency of operation /23/. Effluent samples taken from highly nitrifying activated sludge plants exhibit higher BOD values than would be caused by the carbonaceous demand alone /23/. This has been shown to be due to the nitrifying organisms present in these effluents, which caused nitrification to take place immediately in the BOD test /23/.

2.2.4 Dissolved Oxygen

2.2.4.1 Dissolved Oxygen Concentration

In the activated sludge treatment process, oxygen is used in those reactions required to degrade the substrate to produce the high energy compounds required for cell synthesis and for respiration. For long MCRT systems, as in the extended aeration process, the oxygen needed for cell maintenance can be of the same order of magnitude as substrate metabolism /7/. Usually a minimum residual dissolved oxygen (DO) is maintained in a reactor basin to prevent oxygen deficiencies from limiting the rate of substrate removal. In the literature different values are given for the minimum DO concentration, ranging from 0,5 to 2 mg/l /7/12/. However, it must be recognized that high DO concentrations could allow deeper and more rapid penetration of oxygen into flocs, thereby removing an oxygen limitation on the inner cells /12/.

Since DO concentration and mixing energy can affect the apparent kinetic characteristics of the sludge, careful consideration should be given to them /12/. If this is not done the resulting parameters will not adequately reflect the true nature of the system.

Several "rules of thumb" for determining oxygen requirements have been developed over the years. For diffused air aeration plants, 3,7 to 15 m^3 air/m^3 wastewater was considered adequate, but later, 30 to 55 m^3 air/kg BOD_5 was used /7/. Values ranging from 43 to 123 m^3 of air/kg of BOD_5 treated have been used /24/. For extended aeration systems, the air supplied may range up to 125 m^3/kg BOD_5 removed in those cases where the effects of endogenous respiration and nitrification are significant /7/. It should be noted that these "rules of thumb" are generally conservative in that

they include the air needed for mixing as well as for biological needs and also presume a fixed transfer efficiency of the aeration device /7/.

2.2.4.2 Oxygen Transfer and Aeration Systems

Oxygen transfer to the activated sludge floc is usually accomplished by:

a) oxygen absorbed from diffused bubbles of air entrained in the mixed liquor by subsurface injection and

b) mechanical or natural surface agitation in which oxygen is absorbed from the air above the reactor surface.

The rate of oxygen transfer for a given aeration system depends largely on the contact time between the bubble and the liquid, the size of the bubble, and the turbulence in the liquid /30/. Good transfer efficiencies will result if the contact time is as long as possible, the bubble size is maintained as small as possible and the turbulence is such that the bubble is held in the liquid as long as possible before reaching the surface /30/.

The two most common types of aeration systems are subsurface diffusion and mechanical aeration /30/. In the diffused air system, compressed air is introduced at the bottom of the reactor near one side (see fig. 4). This causes the reactor's contents to be circulated by the air-lift effect. Many different patterns of diffuser placement are being used. Systems that allow longer and more complete contact between the air and the liquid are preferred. Coarse bubble diffusion devices produce larger bubbles than do fine bubble diffusors.

Since oxygen-transfer is a function of mass transfer across a gas-liquid interface, an increase in bubble size will decrease the amount of surface area in the gas-liquid interface. Essentially, if bubble size is doubled then the gas-liquid interfacial area is halved. However, fine bubble diffusers are easily clogged resulting in injection inefficiency.

There are several types of mechanical aeration devices. The floating or fixed bridge aerators are quite common /30/. Some use a blade to agitate the surface of the reactor and dispense air bubbles into the mixed liquor /30/. Others circulate the mixed liquor by an updraft or downdraft pump or turbine (see fig. 5). This action produces surface and subsurface turbulence, at the same time diffusing air through the liquid mass /30/.

Rotating brushes or blades partially submerged are used in oxidation ditch plants (see fig. 6). The turbulence produced entrains the air bubbles and keeps the mixed liquor in motion.

Turbulence of the mixed liquor, as mentioned above, is one of the main factors that determines the transfer of oxygen in the aforementioned aeration systems. However, excessive turbulence (mixing of the mixed liquor) can damage the floc particles of the mixed liquor, causing poor sedimentation and loss of solids in the effluent /24/.

Figure 4.

Diffused aeration system./22/

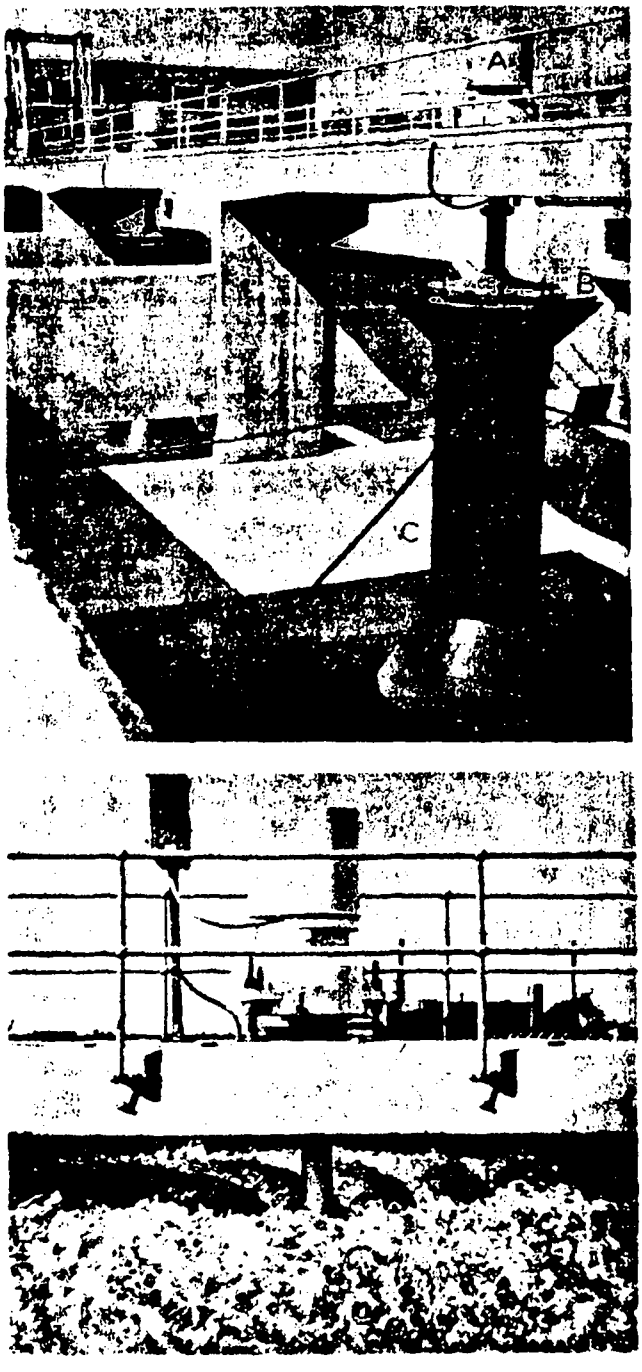


Figure 5.

Surface aerator. Above, at rest; below in action.

A, drive motor
B, cone
C, draft tube.
/22/

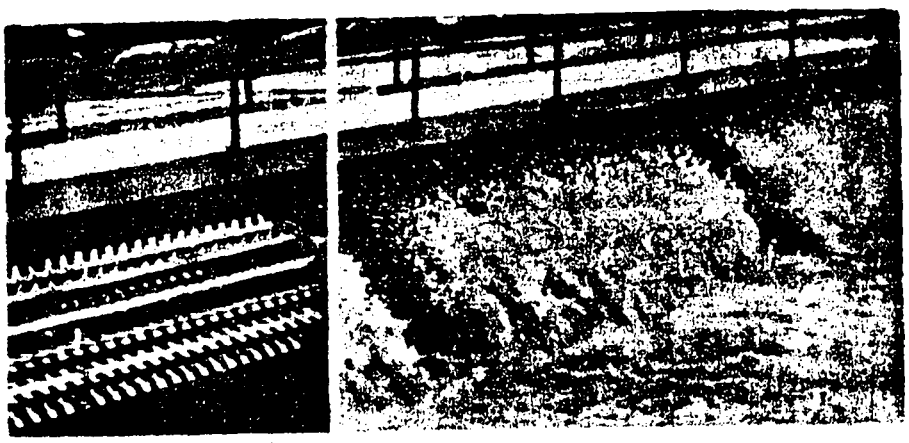


Figure 6. Brush aerator./22/

2.2.5 pH

In biochemical oxidations the enzymes play an important role. The rate of enzyme catalyzed reaction is pH dependent and in a certain pH region the activity of the enzymes is maximum /27/. For this reason, pH will have some influence on the rate of BOD removal of domestic and industrial wastewaters and better BOD reduction may be achieved in the optimum pH range.

Hiidenheimo mentions that, according to Keefer and Meisel, the best performance of the activated sludge process is at pH values of 7,0 to 7,5 /15/.

Generally, the optimum pH for the growth of micro-organisms lies between 6,5 and 7,5 and most organisms cannot tolerate pH levels above 9,5 or below 4,0 /25/.

BOD removal is, however, almost as good at pH values ranging from 6,0 to 9,0 as at values ranging from 7,0 to 7,5 /15/. If the pH is 4,0, the BOD removal drops to only 43 % of the BOD removal at a neutral pH value and 54 % when pH is 10,0, using the same organic loadings /15/.

In the activated sludge process, CO_2 is produced due to respiration of bacteria. If the wastewater being treated happens to be alkaline, the CO_2 will react with the base

and neutralize the mixed liquor /15/. If the pH value of the wastewater happens to be very high, the CO_2 produced cannot neutralize enough of the mixed liquor, pH stays high and the rate of substrate utilization becomes low /15/. Other than CO_2, organic acids and nitrification (see 2.2.4.3) also lower the pH of a mixed liquor /20/.

Some wastes do not contain sufficient buffer capacity to hold the pH within or around about the optimum range during biological metabolism. In this case chemicals must be added to provide buffer.

3. PROCESS KINETICS OF THE EXTENDED AERATION SYSTEM

3.1 General

Literature on the theory of continuous-flow bacteriological systems has grown since the original development of the theory in the 1940's. The process has been applied to diverse areas of study, including aerobic and anaerobic stabilization of wastes. Three fundamental relationships /26/, 1) growth rate, 2) a relationship between an essential nutrient and growth rate, and 3) growth yield applied in conjunction with material balances, allow the development of kinetic equations for continuous-flow systems.

3.2 Microbial Growth

Multiplication of microbial cells is by binary fission, one cell yielding two indentical cells. A cell has to receive its nutrition through its outer surface and as a result there is a limit to the amount of food that can diffuse into the cell. As the cell grows in size, its nutritional requirements also increase and the cell has to divide itself into two to provide more surface area for diffusion of food.

Binary fission of cells can occur every few minutes under favourable conditions /2/. The growth of microbial cells is a function of many factors, among which are; available nutrition, energy, temperature and changes produced in the microenvironment by the microbes themselves /2/. In the growth of microbial populations four principal phases are often recognized as shown in figure 7.

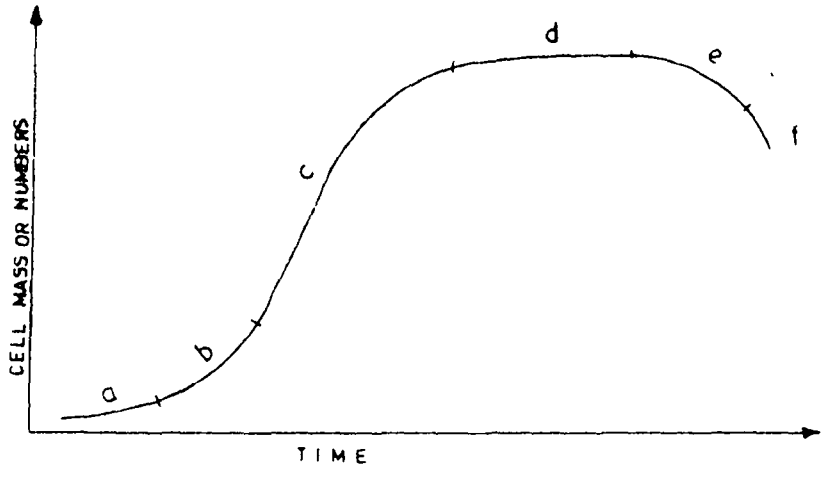


Figure 7. Typical microbial growth pattern./2/

The first phase designated (a) is the lag phase and takes place when the micro-organisms in a given medium have not yet adapted themselves to the local environment. The second phase is the log-growth phase, made up of logarithmic and first-order growth (b + c) in which the growth rate reaches the constant maximal value. This takes place when there is an abundance of the necessary nutrients for growth. The third phase, the stationary population phase (d), occurs when limiting conditions exist (i.e. nutrient limitation) and cell production rate equals cell death rate. The last, the endogenous (death) phase (e + f) is also logarithmic. Here the available nutrition is at a minimum, and micro-organism metabolize part of their own protoplasm in order to get energy. In this phase the death rate of organisms is faster than the production rate, and the numbers decline with time /2/.

3.2.1 Specific Growth Rate

Since micro-organisms multiply by binary fission, their growth is a function of their numbers (or mass) at any given time /2/. In growth kinetics of continuous cultures, the instantaneous growth rate per unit organism per unit

time, i.e. the specific growth rate μ is employed, where during constant growth phase (b+c in fig. 7),

$$\mu = \frac{dx/dt}{x} \qquad (3.1)$$

where,

μ = specific growth rate, time^{-1}
x = organism concentration, mass of organism/ volume, and
t = time.

When all nutrients are present and available to the micro-organisms in excess, the value of the specific growth rate remains constant for the particular set of nutrients, environmental conditions, and species of micro-organisms. When the concentration of one of the nutrients becomes growth-limiting, however, the value of the specific growth rate declines.

A simple relationship between the growth rate and the concentration of an essential (limiting) nutrient was proposed by Monod as a hyperbolic function similar to the equations used to describe enzyme-substrate interaction /26/. Although this relationship is empirical, many natural processes fit this general category of reaction /26/. The relationship is /26/:

$$\mu = \mu_{max} \frac{S}{K_s + S} \qquad (3.2)$$

where,

μ_{max} = maximum growth rate, time^{-1}
S = limiting nutrient concentration, mass/ volume, and
K_s = saturation constant, equal to nutrient concentration at one-half the maximum growth rate, mass/volume.

The above relationship is illustrated in figure 8. When S, the limiting nutrient concentration, is large compared to K_s, μ will approach μ_{max} and growth becomes a zero-order (constant rate)/36/ reaction with respect to S. In other words the growth rate will be independent of S. However, when the concentration of the limiting nutrient becomes small compared with the value of K_s (the saturation constant) growth becomes a fist-order reaction /36/ and is proportional to the concentration of the limiting nutrient.

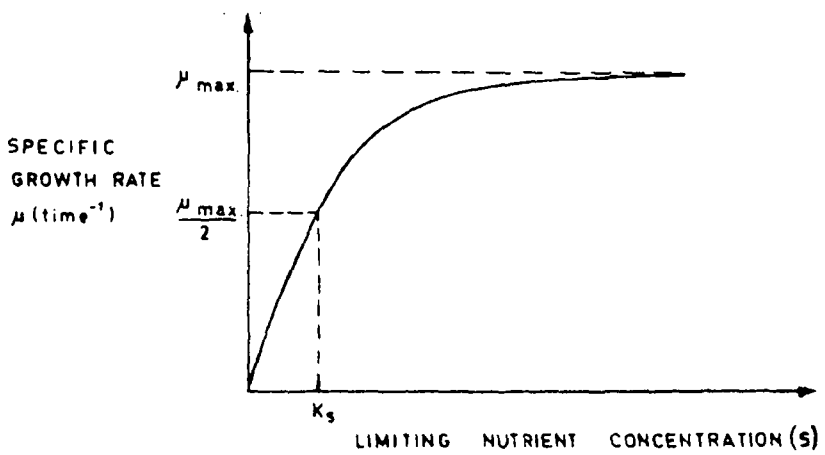


Figure 8. Specific growth rate as a function of the limiting nutrient concentration./36/

The above expression (Eqn. 3.2) has been used very successfully, but it must be remembered that it relates the effect of a single nutrient on the growth rate of a pure culture. Various substances can act as limiting nutrient and many have been shown to affect the growth rate /26/. In situations involving complex media and mixed cultures the expression may apply only over limited ranges and environmental factors also influence the growth rate /26/.

3.2.2 Specific Organism Decay Rate

In applying equation 3.1 to a given population of micro-organisms, it is assumed that all of the organisms are viable or capable of reproduction. This is a reasonably good assumption for homogenous cultures /26/. But it has been found out that, for organisms maintained at a low growth rate (as in extended aeration process), losses due to death of the micro-organisms may be significant /26/. The dead cells (micro-organisms) in extended aeration systems probably are represented by the biological inert solids which accumulate in the system /26/.

When the death rate is significant, the equation for growth must be modified to include it. If the terms in equation 3.1 are rearranged to give:

$$\frac{dx}{dt} = \mu x \qquad (3.3)$$

The effect of death of organisms on the growth rate could be considered by inserting K_d, which represents the specific organism decay rate, in equation 3.3. This yields:

$$\frac{dx}{dt} = (\mu - K_d)x \qquad (3.4)$$

where,

K_d = specific organism decay rate, $time^{-1}$

The decay (death) of organisms is due to endogenous respiration, which is the utilization of cellular material to furnish energy required for the replacement of protoplasm and cellular death with subsequent lysis /26/. Therefore, the specific organism decay rate, K_d in equation 3.4 represents the overall decrease in cellular mass.

The value of K_d appears to vary with the type of organism, substrate concentration and time /26/. Its value can be estimated from graphical plots /36/. Such a procedure is not always too precise and, as a result, a wide range of values can be obtained under similar experimental conditions /36/. The value of K_d as measured under continuous flow conditions has been found to vary with solids age (sludge age) /36/. At temperatures upto 20°C, it can be calculated from /36/:

$$K_d = 0,48\ t_s^{-0,415}(1,05)^{T-20} \qquad (3.5)$$

where,

t_s = mean solids age, days
T = temperature, °C

In the activated sludge process, the mean solids age can be equated to the mean cell residence time. A plot of equation 3.5 for different temperatures is given in figure 9.

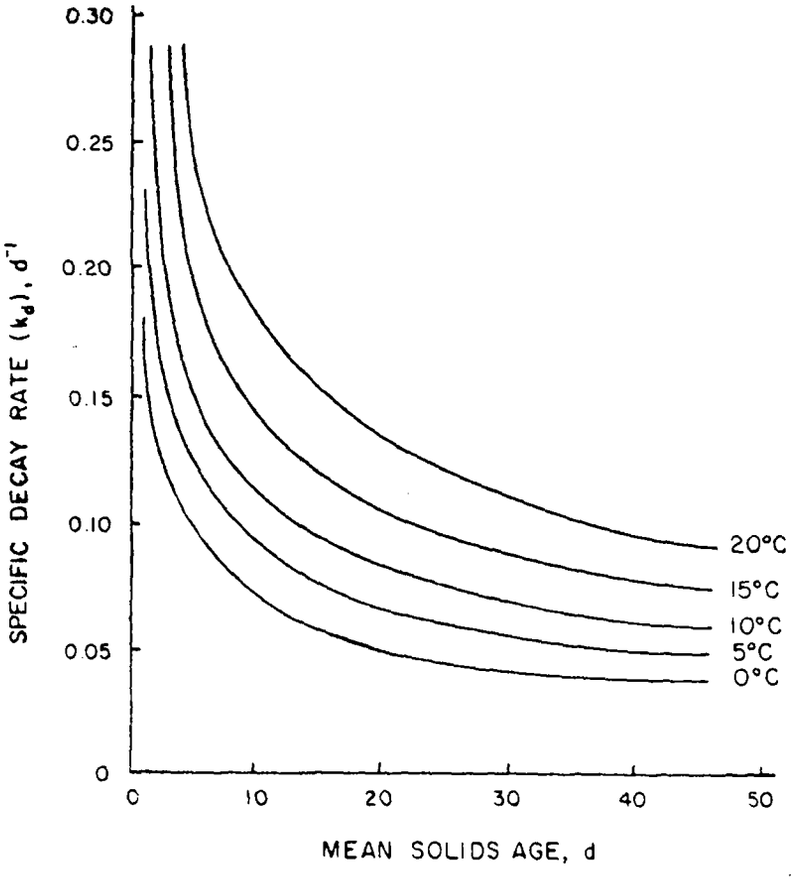


Figure 9.
Specific organism decay rate as a function of MCRT and Temp. /36/

It was mentioned above, that the value of K_d, among other things, varies with substrate concentration. Middlebrooks and Garland report that Stewart and Ludwig have indicated that the specific organism decay rate decreases with a reduction in food for the micro-organisms /26/. They were working with extended aeration systems and based their calculations of the specific organism decay rate on the volatile solids present. They did not consider that these solids may be composed of an accumulation of biologically inert materials and dead cells which do not contribute to organism decay /26/. It is also reported that, in a later publication, Stewart indicated that the specific rate of organism decay decreases with a reduction in the F/M ratio /26/.

Although there is sufficient evidence against the notion of a constant K_d, the concept has been found satisfactory when applied over a limited range of MCRT and F/M ratio /26/.

3.2.3 Growth Yield Coefficient

The relation between new cell production and soluble substrate consumption can be stated as /2/:

$$\frac{dx}{dt} = Y\left[\frac{d(S_o-S_1)}{dt}\right] \qquad (3.6)$$

where,

S_o = influent substrate concentration, mass/volume

S_1 = effluent substrate concentration, mass/volume

S_o-S_1 = substrate removed (consumed), mass/volume

Y = true growth yield coefficient, mass/mass (mass of microbial cells produced/unit mass of substrate utilized).

Monod established that, for a given organism and essential nutrient under the same environmental conditions, the weight of microbial cells produced per weight of nutrient (substrate) utilized is constant /26/. This relationship is expressed as:

$$Y = \frac{\text{weight of organisms produced}}{\text{weight of substrate utilized}} \qquad (3.7)$$

The value of Y is virtually constant for a wide variety of substrates treated aerobically /2/. This is due to the fact that the build-up of protoplasm involves so many transformations that the overall energy requirements are virtually the same regardless of the substrate being metabolized /2/.

Middlebrooks and Garland report that Moser, Herbert, McKinney and many others have expressed the growth yield as a differential /26/:

$$\frac{dx}{dt} = -Y\frac{ds}{dt} \qquad (3.8)$$

which reduces to:

$$\frac{dx}{ds} = -Y \qquad (3.9)$$

Exceptions to a constant growth yield have been observed /26/. These observations were made in experiments performed under nitrogen limiting conditions and the variation was attributed to the storage of substances in the cell /26/.

3.3 Mathematical Model Development

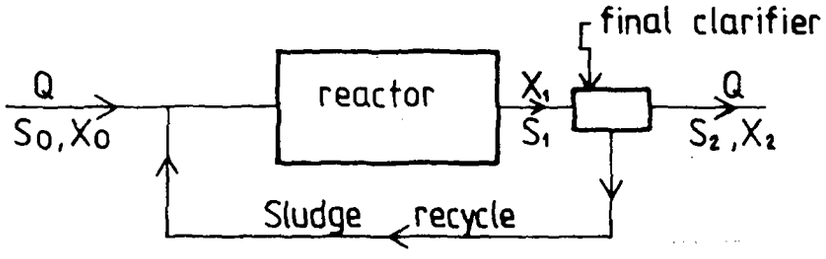


Figure 10. Flowchart of continuous flow system of CMAS process.

A single reactor with its contents completely mixed, so that the composition of the effluent is essentially the some as that of the reactor contents, along with sludge-recycle system is depicted in figure 10. By applying material balances to the relationships presented in equations 3.1, 3.2 and 3.9 one can develop a mathematical model for a completely mixed single-stage reactor with sludge recycle /26/. The following development is a modification by Middlebrooks and Garland of a presentation by Herbert /26/. The following assumptions were made in developing the model, to simplify the mathematics:

1) Monod's equation can be used to express the relationship between the growth rate and limiting nutrient concentration.

2) A constant proportion of the organisms is viable.

3) The yield of organisms is constant.

The mass balance for the organisms is /26/:

$$\left[\begin{matrix}\text{organism change}\\ \text{in reactor}\end{matrix}\right] = \left[\begin{matrix}\text{organisms in}\\ \text{influent}\end{matrix}\right] + \left[\begin{matrix}\text{growth of}\\ \text{organisms}\end{matrix}\right] - \left[\begin{matrix}\text{loss of organisms}\\ \text{in effluent}\end{matrix}\right]$$

$$- \left[\begin{matrix}\text{loss due}\\ \text{to decay}\end{matrix}\right] \qquad (3.10)$$

and for the designations given in figure 10., equation 3.10 becomes,

$$V(dX_1)_{net} = x_o Q dt + (dX_1)_g V - X_2 Q dt - K_d X_1 V dt \qquad (3.11)$$

where,

V = reactor volume

$(dX_1)_{net}$ = net change in organism concentration in the reactor

x_o = organism concentration in the influent, mass/volume

Q = flow rate through reactor, volume/time

t = time

X_1 = organism concentration in the reactor, mass/volume

X_2 = organism concentration in the system effluent, mass/volume

$(dX_1)_g = \mu X_1 dt$

K_d = specific organism decay rate, $time^{-1}$.

The hydraulic detention time, H_t, can be expressed as follows:

$$H_t = \frac{V}{Q} \qquad (3.12)$$

The concentration of organisms in the influent is insignificant when compared with the concentration of organisms in the mixed liquor /26/. Therefore, dividing equation 3.11 by Vdt and neglecting the organisms in the influent yields /26/,

$$\left(\frac{dX_1}{dt}\right)_{net} = X_1 \mu - K_d X_1 - \frac{X_2}{H_t} \qquad (3.13)$$

At steady-state conditions

$$\left(\frac{dX_1}{dt}\right)_{net} = 0$$

If the ratio $\frac{X_1}{X_2} = b$, where b is a concentration factor which represents the ratio of the concentration of organisms in the reactor to the concentration of organisms in the system effluent, equation 3.13 reduces to /26/:

$$\mu = K_d + \frac{1}{bH_t} \qquad (3.14)$$

The ratio $b \geq 1,0$

Similarly a materials balance for the substrate is /26/:

$$\begin{bmatrix}\text{substrate}\\ \text{change in}\\ \text{reactor}\end{bmatrix} = \begin{bmatrix}\text{substrate}\\ \text{in}\\ \text{influent}\end{bmatrix} - \begin{bmatrix}\text{consumption}\\ \text{by}\\ \text{organisms}\end{bmatrix} - \begin{bmatrix}\text{loss of}\\ \text{substrate}\\ \text{in effluent}\end{bmatrix} \qquad (3.15)$$

Using the designations in figure 10, equation 3.15 becomes:

$$V(dS_1)_{net} = S_oQdt - V(dS_1)_g - S_2Qdt \qquad (3.16)$$

where,

S_o = substrate concentration in the influent, mass/volume

S_1 = substrate concentration in the reactor, mass/volume

S_2 = substrate concentration in the effluent, mass/volume

$(dS_1)_{net}$ = net change in substrate concentration in the reactor, and

$(dS_1)_g$ = change in substrate concentration due to growth.

The term $(dS_1)_g$ in equation 3.16 could be put in another form, i.e.

$$(dS_1)_g = \frac{dx}{Y}$$

But $dx = \mu X dt$

Therefore,

$$(dS_1)_g = -\frac{\mu X_1 dt}{Y}$$

Dividing equation 3.16 by Vdt and substituting the above expression yields /26/:

$$(\frac{dS_1}{dt})_{net} = \frac{S_o}{H_t} - \frac{\mu X_1}{Y} - \frac{S_2}{H_t} \qquad (3.17)$$

It is assumed that the effluent from the reactor is the same as the effluent from the final clarifier, or:

$$S_1 = S_2$$

At steady state conditions,

$$(dS_1/dt)_{net} = 0$$

since,

$$\mu = K_d + \frac{1}{bH_t}$$

equation 3.17 reduces to /26/:

$$X_1 = \frac{bY(S_o - S_1)}{(K_d bH_t + 1)} \qquad (3.18)$$

Combining equations 3.2 and 3.14 results in /26/:

$$\mu = \frac{1}{bH_t} + K_d$$

$$= \frac{\mu_{max} S_1}{K_s + S_1}$$

$$= \frac{1+bH_t K_d}{bH_t} \qquad (3.19)$$

Rearranging equation 3.19 and solving for S_1 gives

$$S_1 = \frac{K_s [1+bH_t+K_d]}{\mu_{max} bH_t-(1+bH_t K_d)} \qquad (3.20)$$

Equations 3.18 and 3.20 can be utilized to calculate the micro-organism and substrate concentration in the reactor effluent for any given hydraulic detention time, H_t, or influent substrate concentration after determining the values of the kinetic parameters Y, μ_{max}, K_d and K_s.

To show compliance with, or deviation from, the theoretical model, it is necessary to calculate the above kinetic parameters making use of some forms of the aforementioned equations. Dividing equation 3.18 by (S_o-S_1), gives

$$\frac{X_1}{(S_o-S_1)} = \frac{bY}{K_d bH_t+1} \quad ,$$

inverting the above equation multiplying all terms by b and rearranging it results in /26/:

$$b\left(\frac{S_o-S_1}{X_1}\right) = \left(\frac{K_d}{Y}\right)bH_t + \frac{1}{Y} \qquad (3.21)$$

The above expression (eqn. 3.21) is a straight-line equation, where $\frac{K_d}{Y}$ is the slope and $\frac{1}{Y}$ is the y-intercept of the resulting straight line if plotted, thus determining the cell growth yield coefficient Y and the specific organism decay rate K_d.

Inverting equation 3.19 gives,

$$\frac{K_s+S_1}{\mu_{max}S_1} = \frac{bH_t}{1+bH_tK_d}$$

Rearranging the above expression,

$$\frac{bH_t}{1+bH_tK_d} = \left(\frac{K_s}{\mu_{max}}\right)\frac{1}{S_1} + \frac{1}{\mu_{max}} \qquad (3.22)$$

This equation 3.22 again, if plotted will give a straight line with a slope of K_s/μ_{max} and a y-intercept of $\frac{1}{\mu_{max}}$, thus determining the parameters K_s and μ_{max}.

4. THE EFFECT OF TEMPERATURE ON ACTIVATED SLUDGE PERFORMANCE

4.1 General

Effects of temperature on the performance of the activated sludge process have been observed and discussed since the inception of the process. Most work reported to date is of the type that does not provide comprehensive kinetic or stoichiometric information and, therefore, cannot give enough predictive ability in relation to temperature effects on the treatability of different wastewaters by the activated sludge process /37/. Another point is that regarding the information currently available, there are disagreements on the reason for, and magnitude of, temperature effects.

4.2 The Arrhenius Equation

It is an agreed upon fact that the rate of any simple chemical reaction is increased when the temperature is elevated, provided that the higher temperature does not produce alternations in the reactants or the catalyst /35/. Arrhenius proposed that the effect of temperature on the reaction-rate constant in a chemical reaction may be described by the following equation /35/:

$$\frac{d(\ln K)}{dT} = \frac{E_a}{R}\,\frac{1}{T^2} \qquad (4.1)$$

where K = reaction rate constant
E_a = activation energy, cal/mole
R = ideal gas constant, 1,98 cal/mole-degree
T = reaction temperature, °K

when integrated between limits, equation 4.1 gives the following:

$$\ln\left(\frac{K_2}{K_1}\right) = \frac{E_a}{R}\,\frac{T_2-T_1}{T_2T_1} \qquad (4.2)$$

using equation 4.2, if K_1 is known for T_1, it is possible to compute K_2 for the temperature T_2, if the activation energy for the reaction is known.

It is generally assumed that the same relationship also holds for biochemical reactions over the temperature range that is compatible with the enzymes catalyzing the reaction of interest /35/.

4.3 The Streeter-Phelps Empirical Modification of the Arrhenius Equation

It is commonly assumed that the quantity E_a/RT_2T_1 in equation 4.2 is a constant for biological processes, and subsequently equation 4.2 is modified as follows /35/:

$$\ln\left(\frac{K_2}{K_1}\right) = \text{constant}\ (T_2-T_1)$$

or,
$$K_2/K_1 = e^{\text{constant}\ (T_2-T_1)} \qquad (4.3)$$

If a temperature characteristic term, θ, which has a value equal to that given by e^{constant} is introduced in equation 4.3, it reduces to /35/:

$$K_2/K_1 = \theta^{T_2-T_1} = \theta^{\Delta T} \quad /3/ \qquad (4.4)$$

where,

T_1, T_2 = any two temperatures within the considered temperature range °C

K_1, K_2 = reaction rate coefficients at temperatures T_1 and T_2, respectively, consistent units.

θ = the Streeter-Phelps temperature sensitivity coefficient, dimensionless

ΔT = a temperature differential, $T_2 - T_1$, where $T_2 > T_1$

Equation 4.4 is known as the Streeter-Phelps empirical modification of the Arrhenius Equation (Law). This equation has been widely used to describe temperature effects on the reaction rate constants involved in the biological treatment of wastewaters; however, recent findings have shown that the modified Arrhenius Equation cannot be used to describe such temperature effects /37/.

The generalized applicability of the modified Arrhenius Equation is limited because θ, the temperature activity coefficient, is a variable that depends on many parameters among which are temperature range, type of substrate, F/M ratio and the gross bacterial culture adaptation /29/37/.

Pure culture studies of aerobacter aerogenes have shown that cellular growth and substrate removal reaction rates do follow the above relationship (eqn. 4.4) over the temperature range from 25°to 40°C /35/. Studies have also shown that mixed microbial cultures, such as activated sludge, will obey the relationship for the same two reactions over the temperature range from 5°to 20°C /35/. Consequently in the design of biological processes, it is widely assumed that the reaction rate constant of interest can be determined at 20°C, and then adjusted to any desired operating temperature over the range from 5°to 45°C,

using the same modified Arrhenius Equation and a single value of Θ developed for that particular biological process /35/. This means the modified Arrhenius Equation could be written as:

$$K_T = K_{20}\Theta^{(T-20)} \qquad (4.5)$$

where,

K_T = the unknown value of the reaction rate at required temperature of T°C

K_{20} = a known value of the reaction rate at a temperature of 20°C and

Θ = the same as in equation 4.4

The value of Θ has been found to vary from 1,000, indicating that the reaction rate of a given system is independent of temperature ($K_1 = K_2$, in eqn. 4.4), to about 1,250 for various bio-oxidation processes /3/. Most values reported are between 1,000 and 1,100 /3/.

In general, it seems that the Θ value determined by some researchers depends on either the substrate concentration or the food-to-micro-organism ratio /29/. It also seems that low-cell systems are more temperature-sensitive than processes where high organism levels are maintained /29/.

4.4 Temperature Effects on Micro-organisms

4.4.1 Acclimation of Micro-organisms

When organisms are subjected to a temperature change within the biologically active range, the response of the organisms at the new temperature depends on their new ability to adapt or acclimate to the new environment. The exact processes by which temperature acclimation occurs are not well established for pure culture systems, and even less is known about this phenomenon for mixed culture

systems /4/. The temperature acclimation of mixed cultures is complicated by the fact that gross culture adaptation may depend on shifts in population (different species) as well as on the adaptation of specific organisms within the culture /4/.

Although acclimation is generally considered to be significant in determining the temperature response of mixed culture systems, the length of time required for adaptation is not well established /4/. For research purposes the time is usually set arbitrarily. The time required seems to be related to the magnitude and rate of the temperature change among other things /4/.

Benedict and Carlson reported that at least one investigator has concluded that temperature acclimation is not important in determining the treatment efficiency of completely mixed, activated sludge systems /4/. The same aforementioned authors also reported that Adamse found that there was no significant difference in the bacterial compositions of two mixed cultures maintained at 8° to 12°C and 15° to 20°C, respectively, suggesting that shifts in bacterial population with changes in temperature within this range may not affect the temperature adaptation phenomenon /4/.

4.4.2 Effects on Growth of Micro-organisms

Since growth processes are dependent on chemical reactions, different species of bacteria have their specific temperatures that are optimum for their growth. For any given species of bacteria there is a minimum temperature below which growth does not occur. From this value, as the temeprature is increased, a point is reached where the rate of growth is maximum or optimum /35/. Increases in temperature beyond this point result in the denaturation of the heat sensitive cell components such as enzymes and the growth rate drops rapidly /35/.

The growth rate of the bacteria would not be expected to follow the Arrhenius relationship beyond the optimum temperature because the net effect would be the combination of two reactions, the normal reaction and denaturation /35/. For a mixed culture process to follow the relationship, growth dominance would have to constantly shift to species that are below optimum temperature as the previously dominant organisms exceed their optimums /35/. In actuality, it is more likely that growth gaps in the shifts would occur wherein the dominant reaction would be that of a group beyond its optimum temperature for an interval before a sub-optimum group assumes dominance /35/.

Based on the temperature range within which they can proliferate, bacteria can be classified as psychrophilic, mesophilic, or thermophilic. The acceptable temperature range for each class of bacteria is shown in figure 11./35/. The hatched portion within each range indicates an approximate optimum temperature that allows for the most rapid growth during a short period of time. Facultative thermophiles and facultative psychrophiles are bacteria which have optimum temperatures that extend into the mesophilic range. Optimum temperatures for obligate thermophiles and obligate psychrophiles lie outside the mesophilic range.

Figure 11. shows that the species dominating growth from 0°to 25°C are likely to be fairly uniform and the rates would probably follow the Arrhenius relationship /35/. A potential gap occurse between 20°and 25°C where a decrease in overall reaction rate might take place in a mixed culture because of a shift in the dominant species /35/. Further, changes in dominant species are likely to occur at 30°C, at 45°C, and at 55°C. Of course, if members of some of the groups are not present in the original mixed culture, more frequent and larger changes in the overall rate of reaction could occur as the temperature of the reactor is increased /35/.

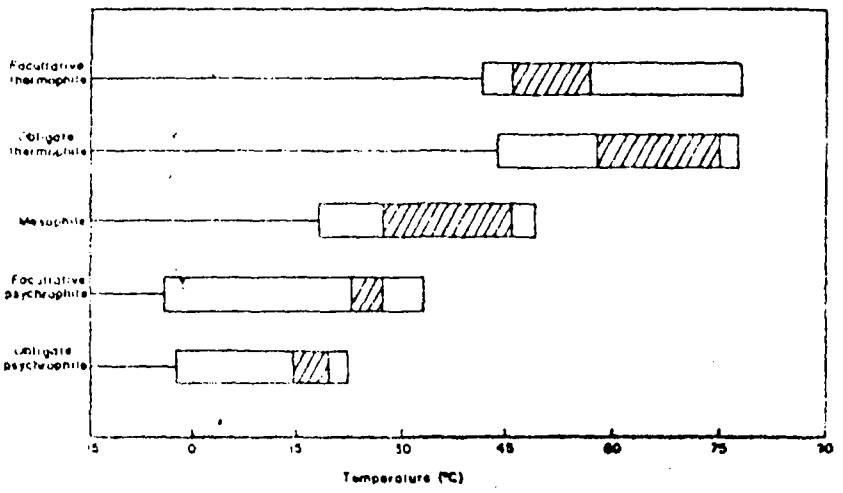


Figure 11. Temperature ranges for reproduction of psychrophilic, mesophilic and thermophilic bacteria./35/

Of chief concern in biochemical operations are mesophilic organisms which grow well over the range of 10 - 35°C. The two other groups (psychrophilic and thermophilic) have ranges on either side and find use under special conditions /12/. Hiidenheimo reports that Viehl (1964) has pointed out that the adapatability of activated sludge is good at the temperatures 12 - 30°C /15/. At lower temperature (actually 2°C), however, he noted considerable retardation in the purification process /15/.

4.5 Temperature-Substrate Interactions

If the specific growth rate μ for a given organism and substrate is assumed to follow a general equation in the form,

$$\mu = f(\text{composition}) \cdot f(\text{temperature}) \qquad (4.6)$$

then proper evaluation of the composition and temperature functions should produce an equation that describes growth rate and the interrelation between substrate and temperature /29/.

Novak reported that an empirical equation similar to the Monod Equation has been used by Lawrence and McCarty to describe the substrate uptake rate as a function of the limiting nutrient concentration /29/:

$$\frac{dS}{dt} = \frac{ksx}{K_s+S} \qquad (4.7)$$

where,

dS/dt = rate of decrease in substrate (nutrient) concentration, mass/volume-time

k = the maximum rate of substrate degradation time^{-1}

K_s = the saturation constant; the substrate concentration when dS/dt = 1/2 k, mass/volume.

The specific utilization rate, U, or rate of substrate uptake per unit mass of organisms may then be defined as /29/:

$$\frac{dS/dt}{x} = U = \frac{kS}{K_s+S} \qquad (4.8)$$

which is the same general form as the Monod Equation, except that the Monod Equation describes the micro-organism specific growth rate while equation 4.8 describes the rate of substrate uptake per unit mass of micro-organism /29/.

Although equation 4.8 is useful for describing substrate utilization as a function of substrate concentration, it does not consider the temperature effects on the utilization rate. The temperature effects must then be applied to this equation in such a way that the temperature correction is substrate-dependent.

Most temperature modifications have been applied in the same manner as in equation 4.5.

$$U = k_{20}\theta^{(T-20)} \frac{S}{K_s+S} \qquad (4.9)$$

Such corrections do not account for a substrate-temperature interaction because this correction modifies only k, a substrate independent term /29/. Since K_s modifies the specific utilization rate for substrate variations, K_s may be a function of temperature along with k /29/.

Making use of equation 4.3 it is postulated that the rate of substrate degradation, k, and saturation constant, K_s, at a given temperature could be predicted from /29/:

$$k_2/k_1 = e^{C_1\Delta T} \Rightarrow k_2 = k_1 e^{C_1\Delta T} \qquad (4.10)$$

and

$$\frac{K_{s2}}{K_{s1}} = e^{C_2\Delta T} \Rightarrow K_{s2} = K_{s1} e^{C_2\Delta T} \qquad (4.11)$$

where,

k_2 = k at a reaction temperature, T_2
k_1 = k at a reference temperature, T_1
ΔT = the difference between T_2 and T_1
C_1 = a constant equal to the slope of log k versus temperature line

K_{s2} = K_s at a reaction temperature, T_2
K_{s1} = K_s at a reference temperature, T_1, and
C_2 = a constant equal to the slope of log K_s versus temperature line.

Thus substituting for k and K_s in equation 4.9, the specific substrate utilization rate becomes /29/:

$$U = \frac{k_1 e^{C_1 \Delta T} S}{K_{s1} e^{C_2 \Delta T} + S} \qquad (4.12)$$

This equation, describing substrate utilization, should account for variations in temperature response with either organic loading or organism concentration /29/. When incorporated into kinetic models for various treatment processes, this same equation should allow prediction of process performance under a variety of organic loadings and operating temperatures /29/.

The kinetic parameters k and K_s along with the specific utilization rate, U, can be determined experimentally from data for the degradation of organic substrates by the activated sludge process. Novak reported that Sanders, using batch cultures containing varying concentrations of substrate and seeded with activated sludge biomass, was able to obtain estimates of the specific utilization rate of certain organic substrates /29/. Novak reported further, that he (Sanders) measured substrate as soluble organic carbon and biomass as volatile suspended solids in the batch reactors /29/. For each initial substrate concentration an initial rate of substrate removal was determined and this rate divided by the initial biomass concentration to yield a specific utilization rate, as shown in figure 12.

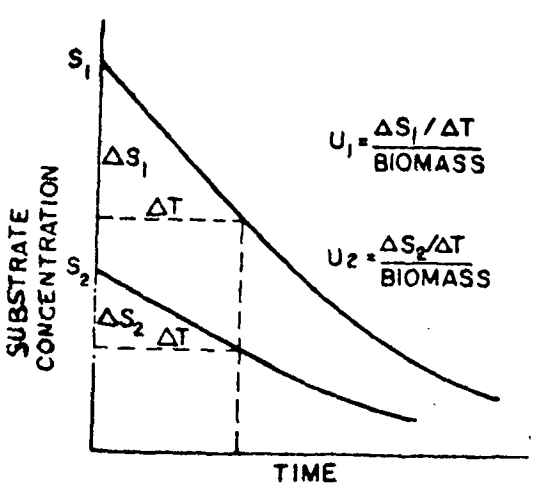


Figure 12.
Determination of specific utilization rate from a batch culture /29/.

If equation 4.8 is inverted and rearranged it would assume an expression:

$$\frac{1}{U} = \left(\frac{K_s}{k}\right) \frac{1}{S_1} \left(\frac{1}{k}\right) \qquad (4.13)$$

The plotting of $\frac{1}{U}$ versus $\frac{1}{S_1}$ would enable one to determine the kinetic constants k and K_s, where K_s/k is the slope and $\frac{1}{k}$ the y-intercept of the resulting straight line.

From the relationship of food-to-micro-organism (F/M) ratio to the specific substrate utilization rate U /25/,

$$U = \frac{(F/M)E}{100} \qquad (4.14)$$

(Here the F/M ratio is based on MLVSS concentration.)

where,

E = process efficiency

$$= \left(\frac{S_o - S_1}{S_o}\right) 100$$

If the loading (F/M ratio) and process efficiency (E) are already known (calcualted) for a given experiment, then the term 1/U could readily be determined and plotted against $1/S_1$ to determine k and K_s.

5. EXPERIMENT

5.1 Objective

The objective of the experimentation reported herein is an evaluation of the effects of temperature on the process kinetics parameters and the performance efficiency of the extended aeration activated sludge process treating domestic wastewater.

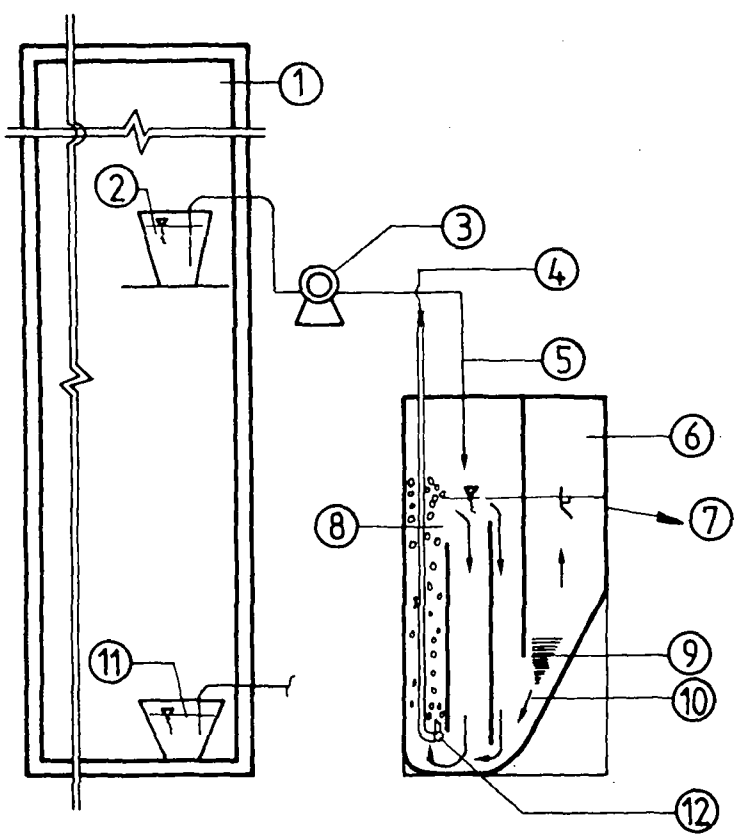
The operational parameters that were evaluated include cell yield coefficient, organism decay coefficient, performance efficiencies, regarding substrate removal, nitrification and suspended solids removal, the settling characteristics of sludge in terms of settled sludge volume and the sludge volume index (SVI), and solids production and accumulation.

The values of standard design kinetic parameters and removal efficiencies of pollutional parameters as reported in the literature are used as the measure of the effects observed under the different working temperature values selected.

5.2 Apparatus

A schematic view of the laboratory bench-scale unit used for the experiment work is shown in figure 13. Two continuous-flow, complete-mix activated sludge models were used. Each unit (model), which was made of polyacryle, has an aeration chamber and clarifier volume of 7,4 l and 2,1 l respectively. The aeration chamber was divided into three compartments by baffles which run vertically down, short of touching the bottom surface of the chamber, thus allowing free-flow between compartments. The clarifier was separated from the aeration chamber also by means of a vertical baffle open at the bottom end.

Sludge return was accomplished by means of a mild sludge roll from the clarifier, which entered (mixed with) the continually agitated mixed liquor by gravity and cavitation. Influent wastewater was fed continuously using a peristaltic pump. Air was introduced through a multipored nozzle made of polyacryle (see figure 14). The treated effluent was withdrawn through an overflow effluent pipe located in the clarifier.



LEGEND
1. Fridge at 4°C (also 10°C for sometime)
2. Influent container
3. Pump (peristaltic)
4. Air supply
5. Influent inflow
6. Clarifier (2,1l)
7. Effluent outlet
8. Aeration chamber (7,4l)
9. Settled sludge
10. Return sludge flow by gravity
11. Effluent container
12. Aeration nozzle

Figure 13. Schematic of experiment system.

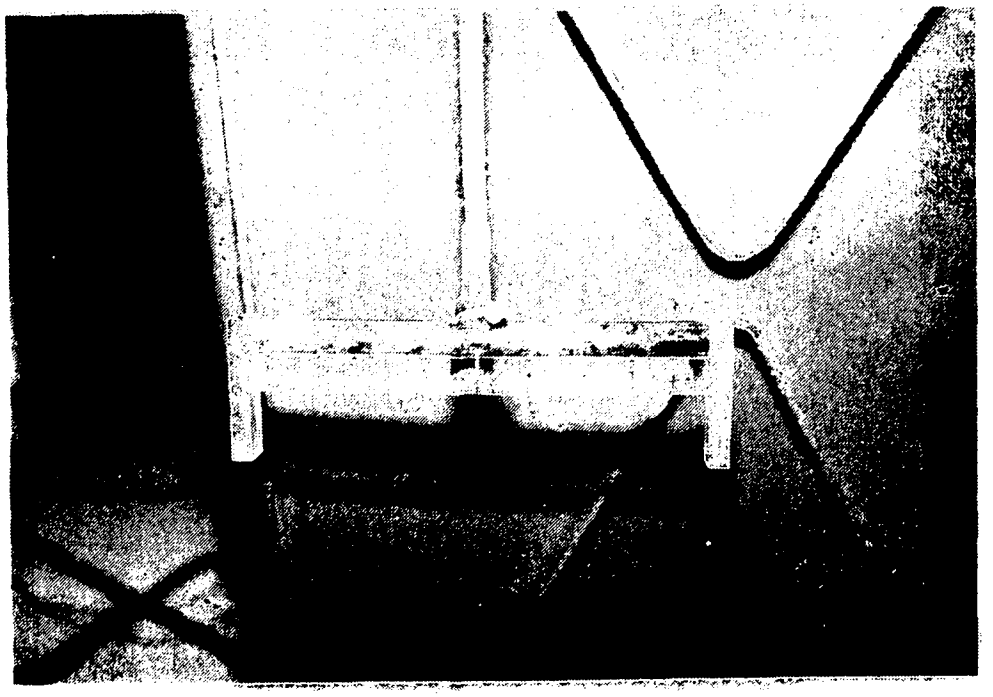


Figure 14. Multi-pored aeration nozzle.

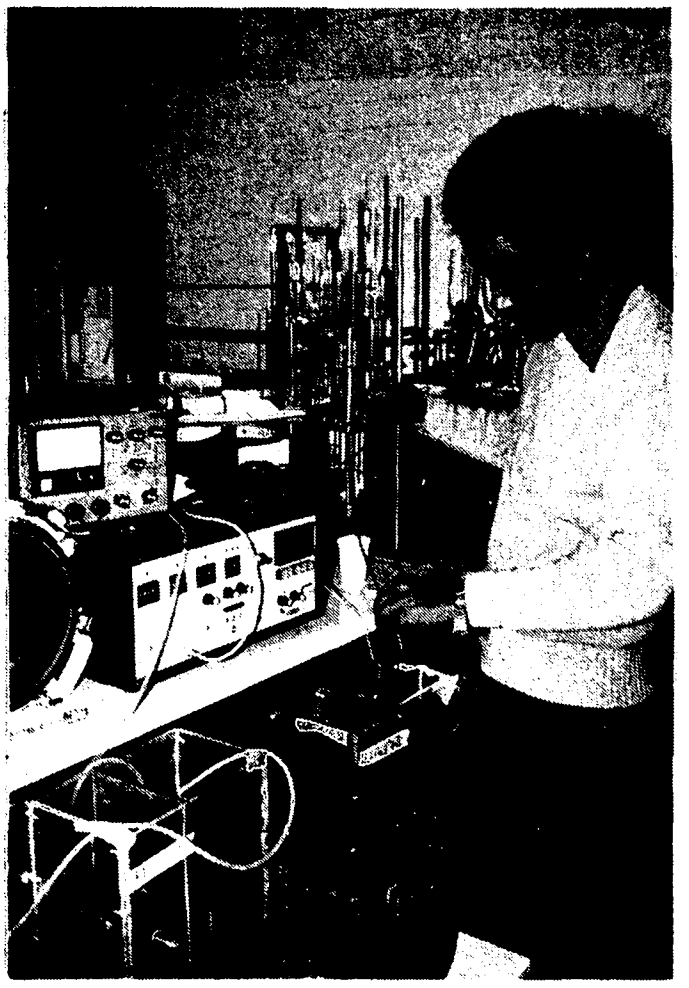


Figure 15

The two models being run parallel.

5.3 Operation of Models

The experiment was carried out in three phases using two identical reactors as shown in figure 15, and lasted about fourteen weeks. As is shown in table 1, phase I of the experiment comprised of running the two reactors parallel at approximately room temperature. In phase II, reactor I (RI) was left running at room temperature, while reactor II (RII) was alternately put in a regultable refrigerator, set at 10°C, for seven hours and put back at room temperature for the rest seventeen hours. This part of the experiment was meant to simulate warm daytime and colder night temperatures in some parts of the tropics during certain seasons. During the final phase (phase III), RI was moved to a heated room of temperature around 38°C and RII placed in the same fridge used in phase II, with the temperature set at 10°C.

Table 1. Experiment schedule.

Phase number	Reactor number	Date	Mean mixed liquor Temperature, °C
I	R I R II	22/9/83 ....25/10/83 26/9/83 ....25/10/83	22,2 22,4
II	R I R II	26/10/83....29/11/83 26/10/83....23/11/83	22,2 22,0/13
III	R I R II	29/11/83....2/1/84 28/11/83....2/1/84	32,5 11,2

During the experiment the maximum mixed liquor temperature deviation from the mean values was about +2,1°C, while the standard deviation for all temperature values remained below $\pm$ 0,7°C (see table 2).

Table 2. Mixed liquor temperature during experiment.

Phase number	Reactor number	Mixed liquor temperature °C			
		Maximum	Mean	Minimum	Standard deviation
I	R I	23,2	22,2	21,1	± 0,41
I	R II	22,8	22,4	21,8	± 0,37
II	R I	23,2	22,2	21,1	± 0,41
II	R II	22,6/14	22,0/13	21/11,8	± 0,42/± 0,57
III	R I	34,6	32,5	31,4	± 0,69
III	R II	12,0	11,2	10,3	± 0,41

Domestic wastewater was continuously fed to the two reactors from a common storage bucket kept at 4°C, during phase I. In phase II the storage temperature was raised to 10°C (since RII in this phase had to be kept at 10°C and there was only one fridge available for use). This same temperature was maintained in phase III for RII for the same reason mentioned above, while for RI a different storage at 4°C was used.

The wastewater, mainly from a residential area, was collected from a nearby manhole on Fridays and Tuesdays. It was collected in the morning hours in an attempt to obtain a substrate of constant strength (later, in the last half of the experiment, the collection was done in the afternoons). The total biochemical oxygen demand (BOD_7) during the whole experiment ranged from 90 - 373 mg/l.

The culture was developed with activated sludge from a nearby sewage works (located at Sahalahti), which mainly treated wastewater from a food-processing plant. An acclimation period of about ten days was allowed before actual evaluation of the performance of the models was started.

Complete mixing was achieved by diffused air aeration, which was provided by a small compressor through a multi-pored nozzle diffuser. Air also provided the necessary dissolved oxygen (DO) concentration to maintain aerobic conditions. Careful positioning of the diffusers at the bottom corner of the aeration chambers was necessary to maintain an adequate and constant aeration level. Dissolved oxygen concentration was measured once a day by means of the dissolved oxygen meter of YSI Model 57 with probe Model 5739. The concentration ranged from 1 - 8 mg/l, with the higher concentration values measured in the reactor which was kept in a fridge at 10°C (RII).

Temperature measurements of reactor contents (mixed liquor) were taken once a day except for RII in phase II, where measurements had to be taken twice a day once when the reactor was inside and another when it was outside the fridge.

Although it would have been much easier to employ a constant wastage of solids from the reactors as an operational control parameter, the food-to-micro-organism (F/M) ratio was used instead due to a low rate of solids production throughout the experiment. In the middle of phase III, when the solids concentration reached the 7000 - 8000 mg/l level, wastage of solids (250 ml a day from each reactor) was started not as a control parameter but rather to limit the solids concentration and lower them to those values achieved in the previous phases of the experiment. Solids concentration (MLSS) ranged from a minimum of 1860 mg/l to a maximum of 8924 mg/l, whereas the volatile part of the solids (MLVSS) ranged from 1000 - 4740 mg/l (see table 3).

The food-to-micro-organism ratio (F/M) was kept within the range for extended aeration (0,15 kg BOD_7/kg MLSS.d) throughout the experiment by adjusting influent feed rate and MLSS concentration. The actual range of F/M ratio employed was 0,0054 - 0,1492 kg BOD_7/kg MLSS.d (0,0115 - 0,3547 kg BOD_7/kg MLVSS.d), with the mean ratios ranging from 0,0351 - 0,0861 kg BOD_7/kg MLSS.d (0,0617 - 0,121 kg BOD_7/kg MLVSS.d). See table 3 for detail.

pH of the mixed liquor was monitored every other day and was maintained between 6 - 7, although there was a period of depression for sometime, manifested in the consumption of alkalinity and lowering of effluent pH down to about 4. Lime was added for sometime to overcome this problem. An electrochemical device was used to measure pH.

Hydraulic detention time (H_t) ranged from a minimum of 14 hours to a maximum of 68 hours with mean values of 26-42 hours (see table 3). It was difficult to maintain a constant detention time due to influent flow variations. The influent flow had to be varied to achieve a desired range of F/M ratio and also the pumps used did not allow much the maintenance of a constant flow, thereby simulating conditions encountered at operational sewage works.

Table 3. Maintained values of some parameters during experiment.

Phase No.	Reactor No.	MLSS max mg/l	MLSS mean mg/l	MLSS min mg/l	MLVSS max mg/l	MLVSS mean mg/l	MLVSS min mg/l	F/M max $kgBOD_7$ / kgMLSS·d	F/M mean $kgBOD_7$ / kgMLSS·d	F/M min $kgBOD_7$ / kgMLSS·d	F/M max $kgBOD_7$ / kgMLVSSd	F/M mean $kgBOD_7$ / kgMLVSSd	F/M min $kgBOD_7$ / kgMLVSSd	H_t max h	H_t mean h	H_t min h
I	RI	8516	4374,7	1860	2956	2117,1	1376	0,1492	0,0623	0,0232	0,1880	0,1056	0,0594	46,48	31,47	14,03
	RII	3240	2970,9	2748	2756	2394,9	2212	0,1123	0,0861	0,0571	0,1215	0,1075	0,0672	35,53	26,32	14,8
II	RI	8516	4374,7	1860	2956	2117,1	1376	0,1492	0,0623	0,0232	0,1880	0,1056	0,0594	46,48	31,47	14,03
	RII	8924	5267,1	3036	3420	2628,2	2000	0,0698	0,0429	0,0223	0,1453	0,0800	0,0498	42,5	33,68	19,89
III	RI	7096	4565,5	2680	3976	1938,5	1000	0,1295	0,0489	0,0101	0,3547	0,121	0,0265	68,26	39,43	17,8
	RII	8368	5756,7	3108	4740	3090,6	1992	0,0994	0,0351	0,0054	0,1686	0,0617	0,0115	50,54	42,58	19,44

5.4 Sampling Procedures

Equal effluent volumes were collected each day of the week, except on Tuesdays and Fridays, and stored in a fridge at 4°C to minimize decompostion. Those collected on Wednesdays and Thursdays were mixed and analyzed on Fridays, whereas those collected on Saturdays, Sundays and Mondays were analyzed on Tuesdays after mixing.

On Tuesdays and Fridays, since the feed was replenished, effluent samples were taken only after an allowance of a detention time of 24 hours (determined arbitrarily) for the previous feed to flow out of the models completely. That was why effluent samples were not collected on Tuesdays and Fridays.

5.5 Analyses and Methods

Analyses were made for suspended solids (SS) and volatile suspended solids (VSS) of influent and effluent. Total BOD_7 and chemical oxygen demand (COD) of influent, COD and BOD_7 with ATU of effluent (ATU was used to inhibit nitrification) and ammonium nitrogen concentration and alkalinity of both influent and effluent. As was mentioned above in section 5.3, measurements of pH, DO concentration and temperature (of mixed liquor) were also taken.

All effluent analyses, except for SS and VSS, were carried out from filtered samples for two reasons;

1) to avoid the interference of escaped biological solids in the BOD_7 test of the effluent load, and
2) to avoid interference from biological solids in nitrification tests in the effluent /35/.

To evaluate the real performance efficiency of the models, the influent samples (except those for SS, VSS, BOD_7 total and total COD) had to be filtered also along with the effluent samples.

All sample filtrations were carried using GF/C glass fiber filters, except those for SS and VSS analyses which were filtered by ashless filter S&S 597.

The standards and methods used for the different analyses are presented in table 4.

Table 4. Standards and methods used for analyses.

Analysis	Standards and methods employed
BOD_7	The Finnish Standard SFS 3019 Determination of biochemical oxygen demand (BOD) of water. Dilution method.
COD	Hach Chemical Company The micro COD digestion procedure. Oxidation with dichromate = COD_{cr}
NH_4-N	Standard methods for the examination of water and wastewater. Nesslerization method.
SS&VSS	Water treatment handbook (Degrémont) Measurement of suspended solids No.406.By filtration (- sample filtered on ashless filter S&S 597 - filter rinsed with distilled water - dried at 105°C for two hours -SS determination - muffled at 550°C for one hour -VSS ")
Alkalinity	The Finnish Standard SFS 3005. Alkalinity and acidity in water. Potentiometric filtration.
pH	The Finnish Standard SFS 3021. Determination of pH-value of water.

Apart from the analyses of influent and effluent samples, MLSS and MLVSS concentrations in the aeration chambers (mixed liquor) were measured every other day along with pH of mixed liquor, settled sludge volume and sludge volume index (SVI). MLSS and MLVSS analyses of samples taken from a total mixture of aeration chamber and clarifier contents were done once every week to determine solids production. Settled sludge volume was measured by filling a 100 ml graduated cylinder with mixed liquor and letting the solids settle quiescently for half an hour. The height of the interface between supernatant and settled solids was recorded every 5 minutes. SVI was calculated by the formula:

$$\frac{\text{settled MLSS (ml/100 ml) after 1/2 h} \cdot 10}{\text{MLSS (mg/l)}}$$

At the end of each phase of the experiment, oxygen uptake rate (OUR) tests were conducted to help determine the characteristics (in this case endogenous respiration rates) of the activated sludge. Samples were taken direct from aeration chambers and the OUR tests carried out using the same oxygen meter used for measuring DO concentrations and a chart recorder.

5.6 Problems

Although the overall running of the experiment was accomplished without major difficulties, there were some persistent and temporary but minor problems worth mentioning here.

Foaming in the aeration chambers was a nuisance on some occasions as in many treatment plants. Mixed liquor suspended solids were deposited on the sides of the models above the water line and dried out, becoming difficult to disperse in the mixed liquor when washed down during cleanup.

During phases of the experiment other than at room temperatures, there was a blanket of floating sludge in both clarifiers but thicker in clarifier II. The blanket was stirred to break up the floating masses of sludge and free any accumulated gases (may be released due to denitrification) in an attempt to control the loss of solids in the effluent.

As was mentioned earlier in section 5.3, there was a depression of pH in both reactors in the middle of phase I of the experiment. Lime was added (1,258 g in each reactor) daily in both reactors for 9 days. Still the pH was being depressed and the dosage of lime had to be doubled (2,516 g in each reactor). This dosage was maintained for about one month and the pH during this time was raised to the 6-7 range and remained stable at this level.

6. RESULTS AND DISCUSSION

6.1 Substrate Removal

Not considering the results in the first few weeks at the beginning of the experiment (to allow for bacterial adaptation), it is observed from figure 16 that almost 100 percent of the time the effluent BOD_7 was less than 10 mg/l. With the results of the first few weeks included, the maximum effluent BOD_7 was less than 20 mg/l (at room temperature) with almost no BOD_7 at one stage at 11,2°C. Here it must be mentioned that the effluent BOD_7 figures represent soluble BOD (see section 5.5).

It is evident, again from figure 16, that sequential fluctuations in BOD_7 load affected performance more than mixed liquor temperature changes. It is also seen that at higher BOD_7 loadings (again excluding the first few weeks' results), effluent quality was better than at lower loadings.

During phase I of the experiment, eventhough the two models were operating at the same temperature (22,2° and 22,4°C), it is seen that there was a slight variation in their performance. This was possibly associated with the variation in micro-organism populations, which were never the same in both models at least in population concentrations.

The percent removals of BOD_7 and COD achieved at different mixed liquor temperatures are summarized in tables 5 - 7. The percentage of BOD_7 removal in the whole experiment ranged from a minimum of 95 percent at 22,2°C to a maximum of 99 percent at 22°C/13°C (see tables 5 and 6). The mean removal percentage for the different mixed liquor temperatures was between 98 percent and 99 percent (see table 7).

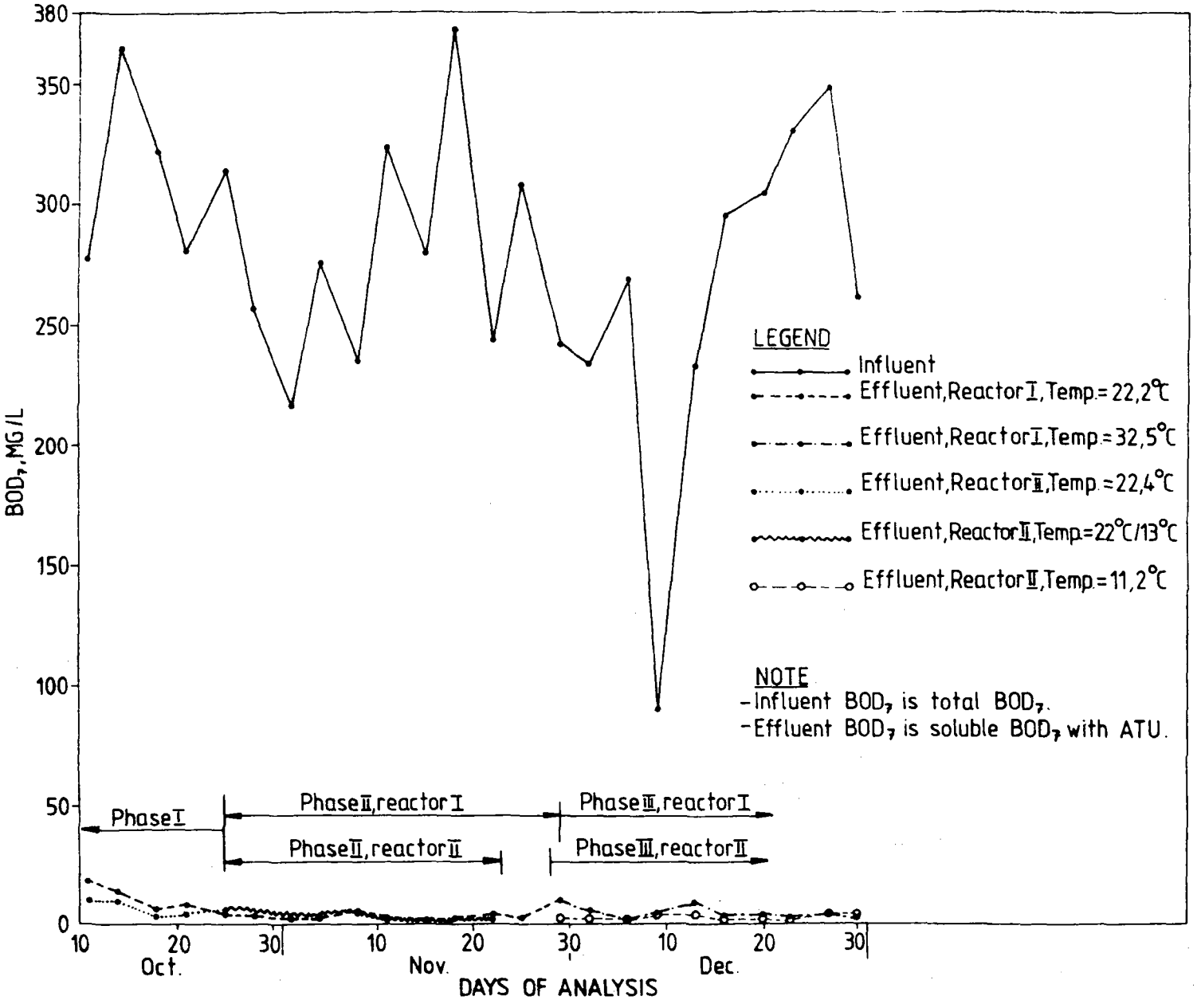


Figure 16. Variation of BOD_7 of influent and effluent with time and temperature respectively.

Table 5. Summary of weekly mean removal rates of pollutants, Reactor I.

Week	Phase No.	BOD_7 removal	COD removal	SS removal	Nitrification
No.		%	%	%	%
41		94,85	79,7	73,15	53,15
42		97,65	86,35	88,6	63,9
43	I	98,4	94,0	59,55	96,85
44	&	98,75	89,65	90,4	95,45
45	II	98,3	90,85	59,0	97,55
46	(22,2°C)	99,4	94,7	56,05	98,10
47		98,9	95,1	93,4	98,35
48		96,65	90,2	66,55	97,8
49		96,85	90,9	91,0	98,45
50	III	97,33	93,2	74,55	98,3
51	(32,5°C)	98,9	86,9	69,85	98,25
52		99,05	90,95	63,3	98,8*

* The result of the other analysis in the same week not included due to a lower nitrification rate caused by a relatively higher loading.

Table 6. Summary of weekly meand removal rates of pollutants, Reactor II.

Week	Phase No.	BOD_7 removal	COD removal	SS removal	Nitrification
No.		%	%	%	%
41		96,85	84,70	72,45	56,95
42	I	98,50	87,75	93,15	65,1
43	(22,4°C)	98,10		43,2	98,0
43		98,40	93,60	39,7	96,7
44	II	98,40	89,65	77,8	95,45
45	22°C/13°C	98,65	92,40	51,1	97,55
46		99,40	93,90	58,8	97,8
47		99,10	95,10	90,45	98,1
48		98,95	93,9	69,2	95,7
49	III	97,60	92,9	54,0	90,0
50	(11,2°C)	98,65	93,2	73,1**	86,8
51		99,35	89,85	70,25	86,3
52		98,60	91,60	59,95	86,85

** Result of analysis on 16.12.1983 only. The result of analysis done on 13.12.1983 was unusual in that the effluent SS concentration was higher than that of the influent, therefore it is not considered here.

From figure 17 (or table 7), it is seen that the highest mean percentage of BOD_7 removal, 99 percent, was obtained at 22°C/13°C and 11,2°C. The weighted average of the 22°C/13°C value was taken (see Appendix B).

Table 7. Summary of mean removal rates of pollutants over weeks.

Week	Temperature	BOD_7 removal	COD removal	SS removal	Nitrification
from - to	°C	%	%	%	%
41 - 47	22,2	98,0	90	74	97*
41 - 43	22,4	98	86	70	98,0*
43 - 46	22/13	99	92	57	97
48 - 52	11,2	99	92	65	89
48 - 52	32,5	98	90	73	98

* Lower rate of nitrification in weeks 41 and 42 (due to nitrifying bacteria still growing). Therefore not entered in the mean calculation.

At mixed liquor temperature of 32,5°C, the removal percentage was 98 percent (1 percent less than at 11,2°C).

Although not in a convincing way, figure 17 shows that BOD_7 removal dropped with increasing temperature. The mean F/M ratio, which could have had an influence on the BOD_7 removal, was not that much varied. It was, for example, for the mixed liquor temperatures of 22°C/13°C and 32,5°C, 0,043 and 0,049 kg BOD_7/kg MLSS.d respectively.

The above result is both supported and contradicted by other findings reported in the literature. Lee et al /17/ experimenting with bleached kraft effluents in a continuous flow activated sludge unit found that BOD_5 removal efficiency decreases with increasing temperature. For the tested F/M

ratio range of 0,1 - 2,4 kg BOD_5/kg MLVSS•d, the BOD_5 removal efficiency they found averaged 91, 89, 85 and 83 percent at 16, 30, 36 and 40°C respectively. Ludzack et al /18/ also operated continuous-flow activated sludge units at temperatures of 5° and 30°C with a variety of substrates, but got different results. Their system required about 2 weeks to approach equilibrium after significant temperature changes. They concluded that BOD and COD removal rates were about 30 percent higher at 30°C than at 5°C.

Keefer /10/ conducted a statistical study on BOD removal/ temperature relationships for a wastewater treatment plant (serving a town) with data collected over a 20-year period. He found that the BOD removal efficiency was 85 percent at 12,2°C, 90 percent at 18,9°C, and 91 percent at 23,4°C.

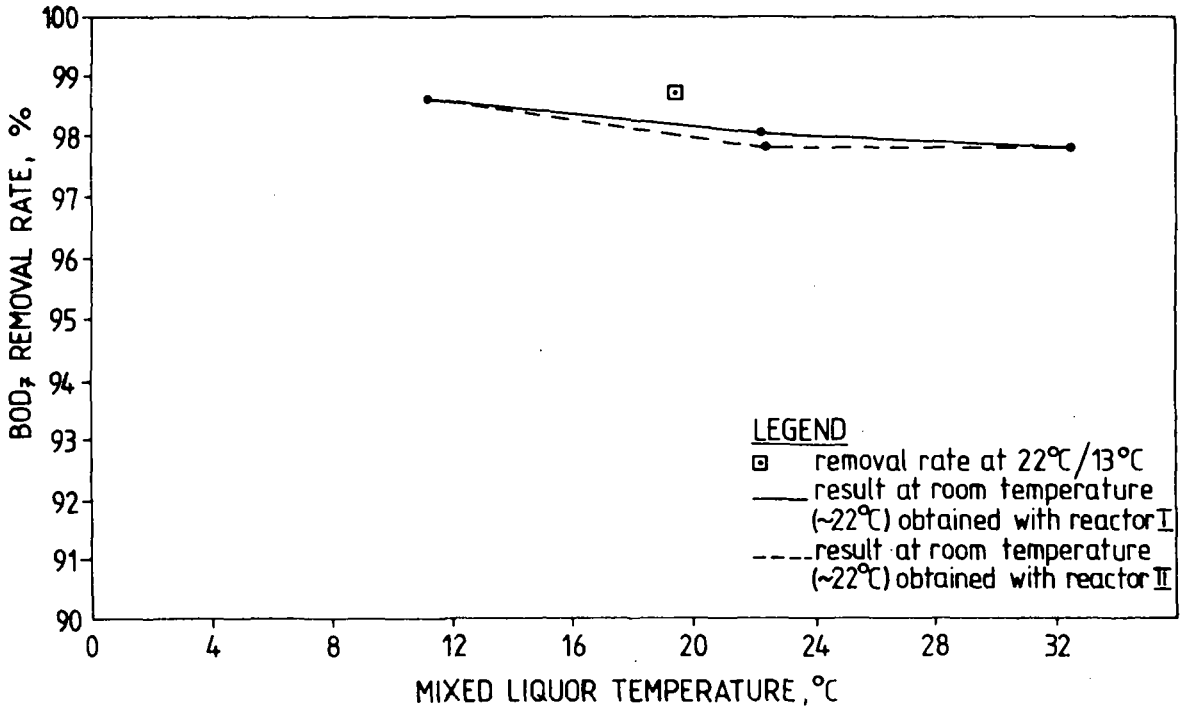


Figure 17. BOD_7 removal rates at different temperatures.

The removal of soluble substrate by activated sludge has been found to be relatively insensitive to changes in temperature /12/. This is probably due more to the long MCRT's employed than to any insensitivity of the kinetic parameters themselves /12/.

Eckenfelder /2/ reported that in domestic wastewater, BOD is mainly in suspended and colloidal form so that removal of bioflocs is largely physical and relatively independent of temperature, but in the case of soluble industrial wastes, temperature effects may be higher.

Gehm /10/, using data from the literature, found little difference in the efficiency of BOD removal for an activated sludge system operating at either 32°C or 49°C. Okun /10/, after reviewing the literature also concluded that the rates of carbonaceous oxidations were not affected by temperature changes.

It is clear that findings of temperature effects on BOD removal, as reported in the literature, are both inconsistent and contradictory, which makes it all the more difficult to reach a conclusive remark regarding the results obtained in the experiment.

Notwithstanding this, if the effects of experimental error, both in analyses and operation, are allowed for; it can safely be postulated that figure 17 shows that temperature effects on BOD_7 removal of low-loaded systems are not discernable. This could be possibly explained by the facts that:

1. In low-loaded systems the substrate concentration is so low that the bacteria are almost in a constant state of "hunger" all the time and consequently make use of all the available food under varying environmental conditions (including temperature).

2. Activated sludge comprises mixed cultures, and significant shifts in the dominant species and the relative numbers of each species are likely to occur with varying temperature thus dampening its effects.

Effluent soluble COD ranged from 20 - 120 mg/l. As is seen from figure 18, the COD removal in the first few weeks at room temperature was relatively low, most probably due to bacterial adaptation. The percentage removal was highest, 92 percent, at mixed liquor temperature of 22°C/13°C and lowest, 80 percent, at room temperature with R II (see table 7). Again here as in the case of BOD_7 removal, the two models running at the same temperature (room temperature) showed a slight variation in COD removal.

Figure 19 shows that COD removal was lowest at room temperature while an increment in efficiency is observed at temperatures below and above it. Streebin and Phillips |10|, using batch cultures, found that COD removals were constant (96 percent) at temperatures between 20° and 40°C

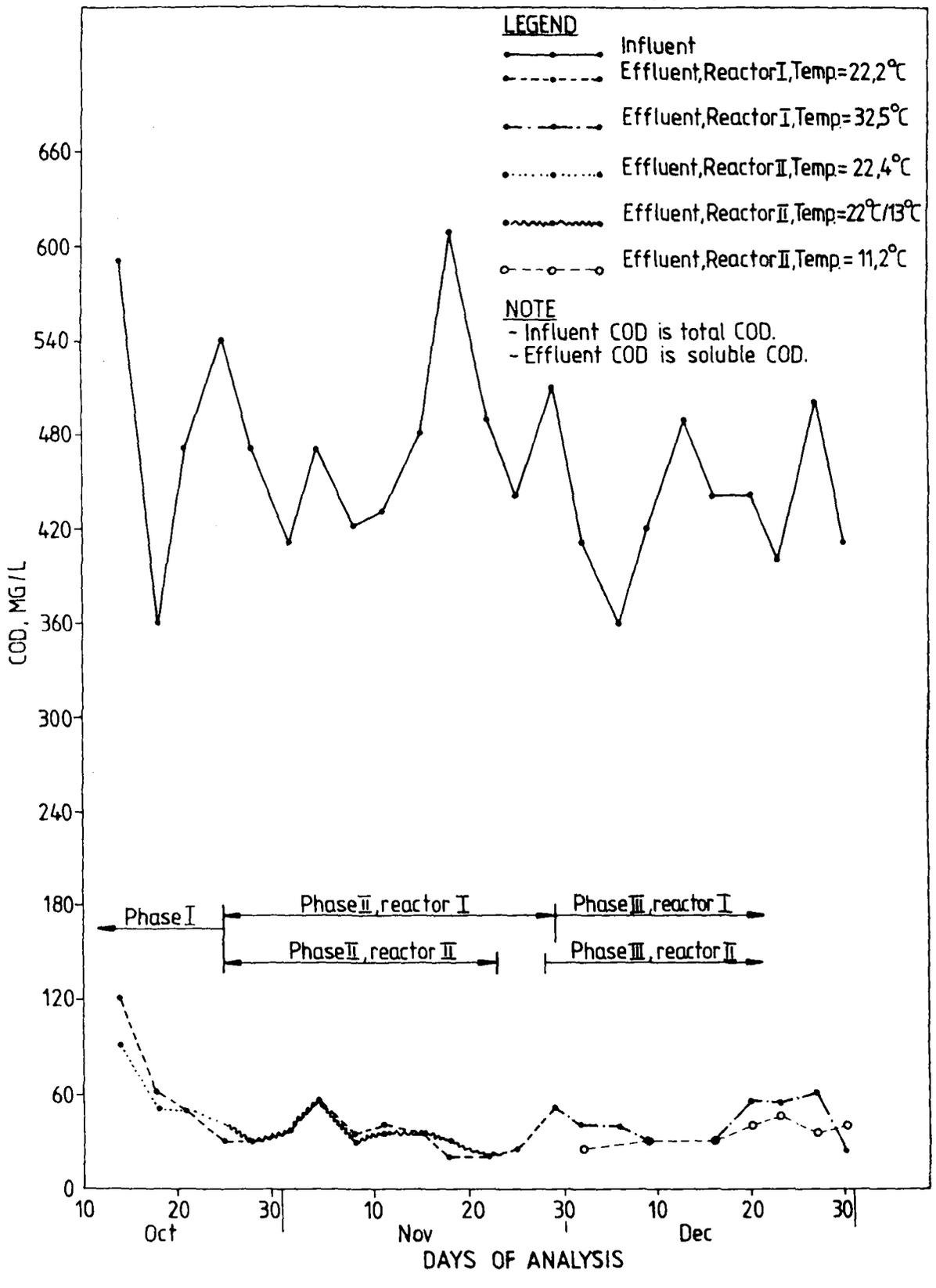


Figure 18. Variation of COD of influent and effluent with time and temperature respectively.

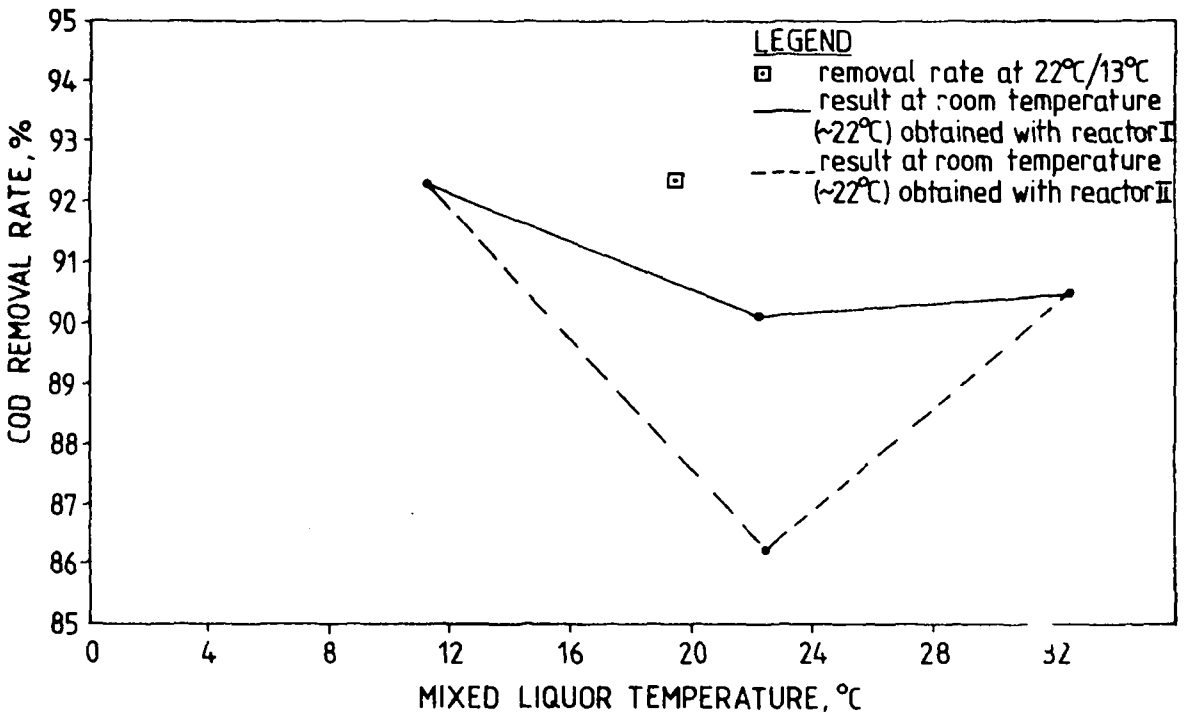


Figure 19. COD removal rates at different temperatures.

6.2 Nitrification

The results shown in figure 20, illustrate that the heterotrophs began metabolizing efficiently (see figure 16) earlier than the nitrifiers. This proves the fact that the nitrifiers are slower growing than the BOD-removing heterotrophs. Since the time both reactors (models) were started to run, the effluent ammonia nitrogen (NH_4-N) concentration kept on dropping from around 33 mg/l till it stabilized at concentrations of less than 2 mg/l in about one months time. This implies that, assuming parameters affecting nitrification (DO concentration, temperature and MCRT) remained constant, then the nitrifiers took about one month to multiply and reach their maximum number.

As is seen from the same figure, effluent NH_4-N concentration ranged from 0,3 - 33,2 mg/l. Most activated sludge plants which nitrify successfully are capable of producing effluents with less than 1 or 2 mg/l NH_4-N /40/. Except for a small bump (see figure 20) in the beginning of November, the effluent NH_4-N concentration, during the whole duration of phase II does not seem to have varied much in both reactors. Since the DO concentration ($\geq$ 2 mg/l) did not drop below the usually maintained range during the experiment and the temperature was kept constant, the sudden increase in effluent NH_4-N concentration (the above mentioned bump) during phase II is not easy to explain.

It is clearly seen that, while the effluent NH_4-N concentration in R II increased when the mixed liquor temperature was changed from 22°C/13°C to 11,2°C (phase III), that in R I decreased for a temperature change from 22,2°C to 32,5°C (phase III). The slight increment of NH_4-N concentration in both reactors towards the end of the phase (phase III) is due to a lowering of DO concentration caused by a higher loading (influent flow was increased to raise the F/M ratio).

Here it is noticed that unlike BOD_7 and COD removal rates, nitrification at mixed liquor temperature of 22°C/13°C is lower than at 32,5°C or even at room temperature, but still higher than at 11,2°C.

Results tabulated in table 7 show that nitrification rate was highest, 98,0 percent, at mixed liquor temperature of 22,4°C and 32,5°C, and lowest, 89 percent, at 11,2°C. Figure 21 shows the classic progression of nitrification rate as temperature was increased. The rate of increment of nitrification between 11,2°C and 22,2°C is faster than that between 22,2°C and 32,5°C. This could probably be due to a more significant micro-organism population change in the former temperature range than in the latter /13/.

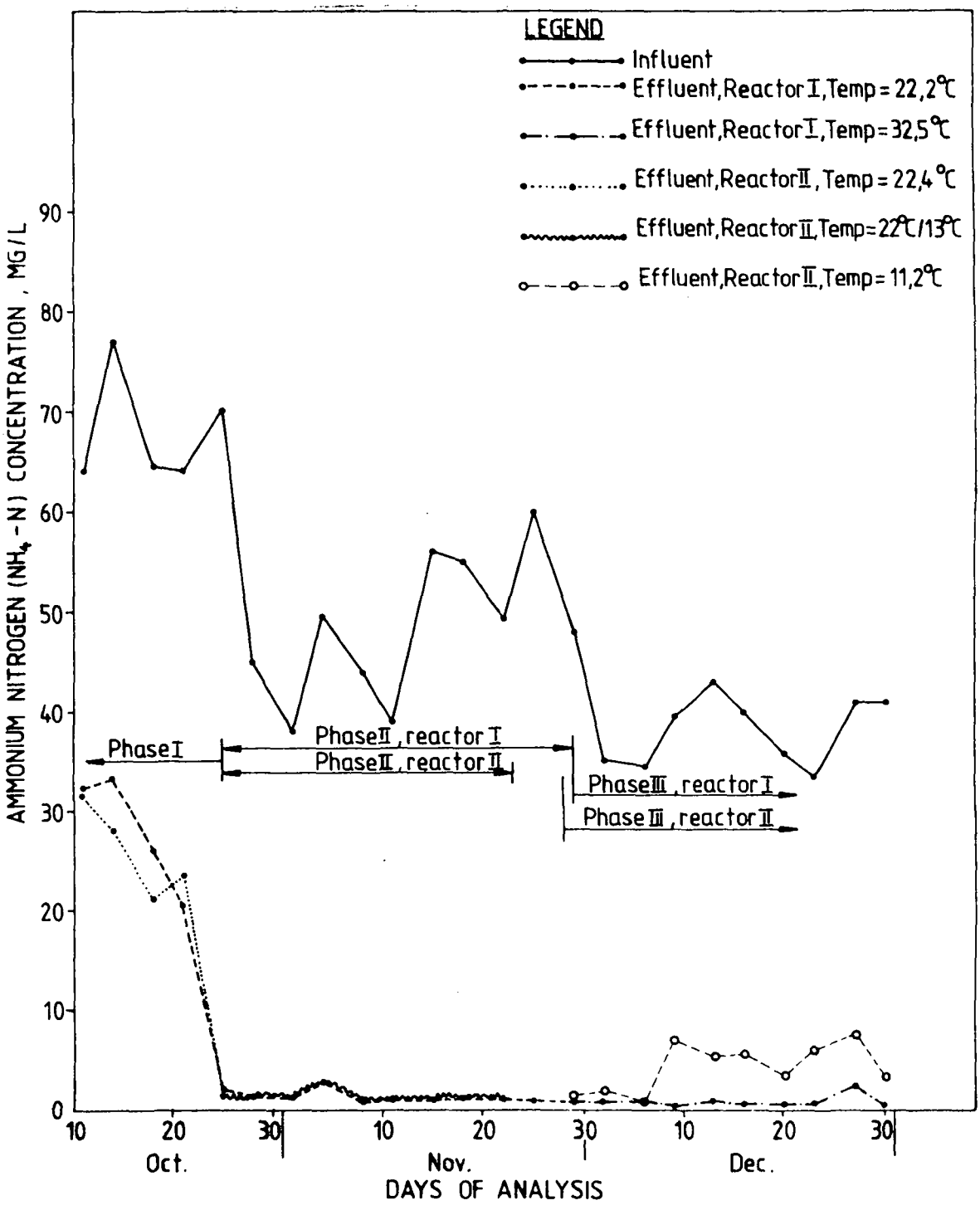


Figure 20. Variation of ammonium nitrogen (NH_4-N) concentration in influent and effluent with time and temperature respectively.

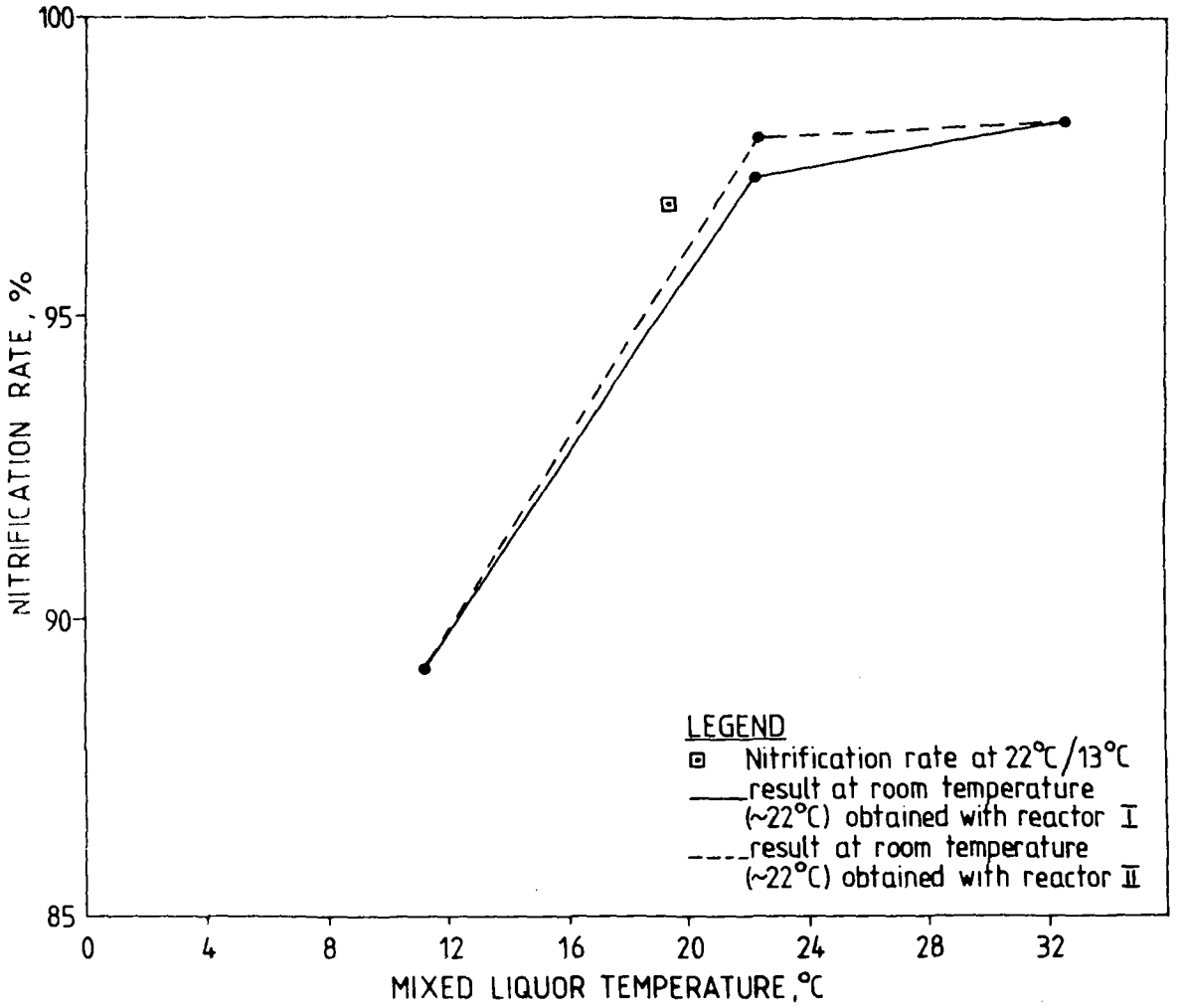


Figure 21. Nitrification rate at different temperatures.

Although the rate of nitrification seems to slow down between room temperature and 32,5°C, from the results of this experiment, 30°C has been reported as the optimum temperature for growth of nitrifying bacteria in pure culture /34/. The rate of enzyme reaction has been shown to double approximately with 10°C rise in temperature up to 35°C with an optimum growth temperature of 42°C for nitrobacter although some enzymes undergo denaturation at temperatures above 35°C /34/.

From the literature review, it is apparent that the DO concentration requirements for nitrification are not well defined /40/. The maximum growth rate of both nitrification reactions (see section 2.2.4.3) are reported to be affected by DO concentration over the range of 0,3 mg/l to as much as 4,0 mg/l /40/.

In some instances, it has been reported that a DO concentration in excess of 4,0 mg/l is required to achieve maximum nitrification rates, while other investigators have found that only 0,5 to 1,0 mg/l is required /40/. Most of the times the DO concentration in both reactors was above 2 mg/l (actual range was 1 - 8 mg/l). In the last phase of the experiment, almost always, the DO concentration in R II, at 11,2°C, was higher (with mean concentrations of 6,2-6,8 mg/l) than that in R I at 32,5°C (with mean concentrations of 3,8 - 4,2 mg/l). Still the nitrification was higher in R I.

It has been well established that successful activated sludge nitrification must be accomplished at MCRT greater than the wash-out MCRT of the nitrifying organisms /40/. At low MCRT, it is possible to nitrify as efficiently as at higher MCRT, although the required DO concentration is much greater.

Between October 10th and December 16, there was no sludge wasted at all from both reactors intentionally, except for analysis purposes (which was very little). Therefore the MCRT was long enough for nitrifiers to grow and multiply to their maximum number. This is supported by the theoretical MCRT values which were calculated (see Appendix A).

6.3 Suspended Solids Removal

Effluent suspended solids (SS) concentrations ranged from a maximum of 71 mg/l at 22,2°C to a minimum of zero at 11,2°C, 22,2°C, and 32,5°C (see figure 22). In some cases, even higher effluent SS concentrations than in the influent were observed at temperatures of 22,4°C, 11,2°C and 22,2°C.

Generally low effluent SS concentrations corresponded to low influent SS concentrations at all phases of the experiment (again see figure 22). Tabulated values in tables 5 and 6 show that results were erratic, i.e. SS removal rates could not be related to mixed liquor temperature except for that of 22°C/13°C where the weekly removal percentages were relatively lower than at other temperatures.

In an activated sludge plant, elimination of the primary clarifier tank results in a proportionately increased organic load on the aeration tank as well as increased solids on the final clarifier tank. Not only will there be a greater quantity of biological solids, but all inert organic and inorganic settleable solids will pass on to the final clarifier tank.

Loss of high concentration of solids in the effluent of extended aeration plant is attributed, among other things, on the following: i) flow variation, ii) excessive solids in the mixed liquor, iii) nonflocculent solids, and iv) denitrification resulting in solids floation due to nitrogen gas entrapped in the sludge blanket /26/. Most of the times, the models were operated within a constant range of flow rate and air supply, and thus the degree of nitrification of the effluent would be expected to be relatively constant, for a given temperature, during periods when the loading rate was stable. This was true as shown by the effluent NH_4-N concentrations in figure 20.

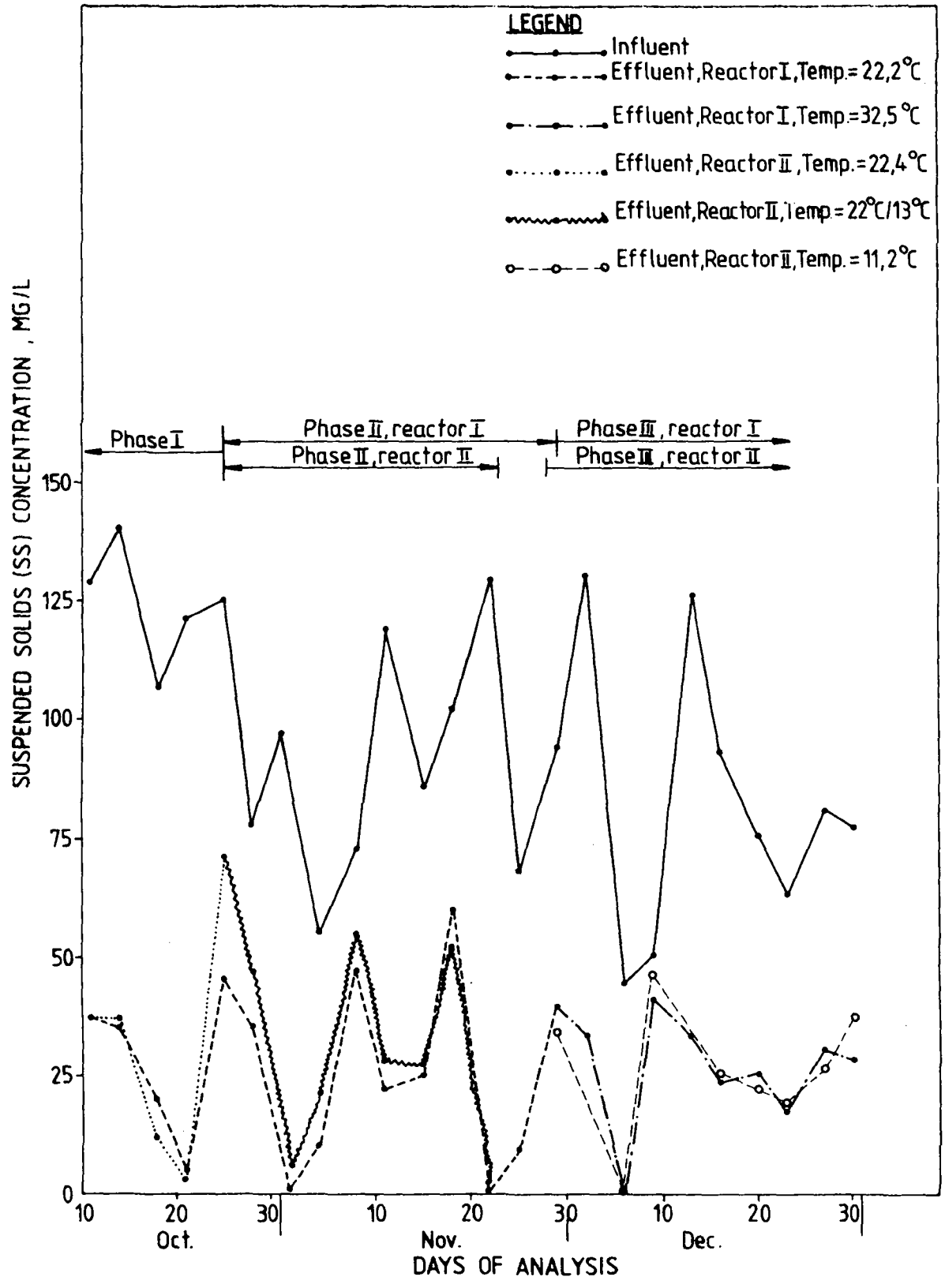


Figure 22. Variation of suspended solids concentration in influent and effluent with time and temperature respectively.

Large losses of solids in the effluent were caused most probably by the inability of the clarifiers to return high concentrations of MLSS to the aeration chambers, which may have resulted in denitrification (although occasional measurements of DO concentrations within the clarifiers indicated that there was some DO, $\geq$ 0,5 mg/l) and floating solids. Another cause for the loss of solids could have been the configuration of the clarifiers in the models used. However, solids losses have also occured at low MLSS concentrations, and these losses may have been due to an accumulation of nonflocculent solids. Nonflocculent solids were also observed with high MLSS concentrations and probably contributed significantly to the loss of solids in the effluent. The nonflocculent solids were lost gradually in the effluent and were present on the surface of the liquid in the clarifiers in small masses.

In figure 23, SS removal rates are plotted against temperature. The solid line shows SS removal increasing from 11,2°C to 22,2°C, but decreasing as the temperature was raised to 32,5°C. The dashed line shows an increasing removal rate from 11,2°C through 22,4°C up to 32,5°C. Here it is observed that the performance efficiency, regarding SS removal, at 22°C/13°C drastically falls down from that of substrate removal or nitrification at the same temperature.

Temperature affects SS concentration through its influence on the kinetic parameters and the water viscosity /6/. Collins et al /6/, after experimenting with domestic sewage in two different models of completely mixed activated sludge, found that there was a decrease in effluent SS concentration with increasing temperature less or equal to 20°C, while for increasing temperature greater or equal to 20°C, there was a decrease in one model and an increase in the other. In both models they attributed the decrease in effluent SS concentration for increasing temperature below or equal to 20°C to the decrease in water viscosity.

For increasing temperture greater or equal to 20°C, the effluent SS reduction was attributed to, in one model, both reductions in water viscosity and MLSS concentrations in the aeration chamber. In the other model, the increase in SS concentration was said to be due to the reduction in MLSS concentrations off-setting the effect of reduced viscosity.

Lee et al /17/ found that the amount of SS in clarified effluents increased drastically with increasing temperature. SS as a function of increasing temperature showed similar increases at all F/M ratios, with significant increases in the range of 30 - 36°C. They postulated that, apparently important changes had occured in the flocculation and settling characteristics of the biological sludges.

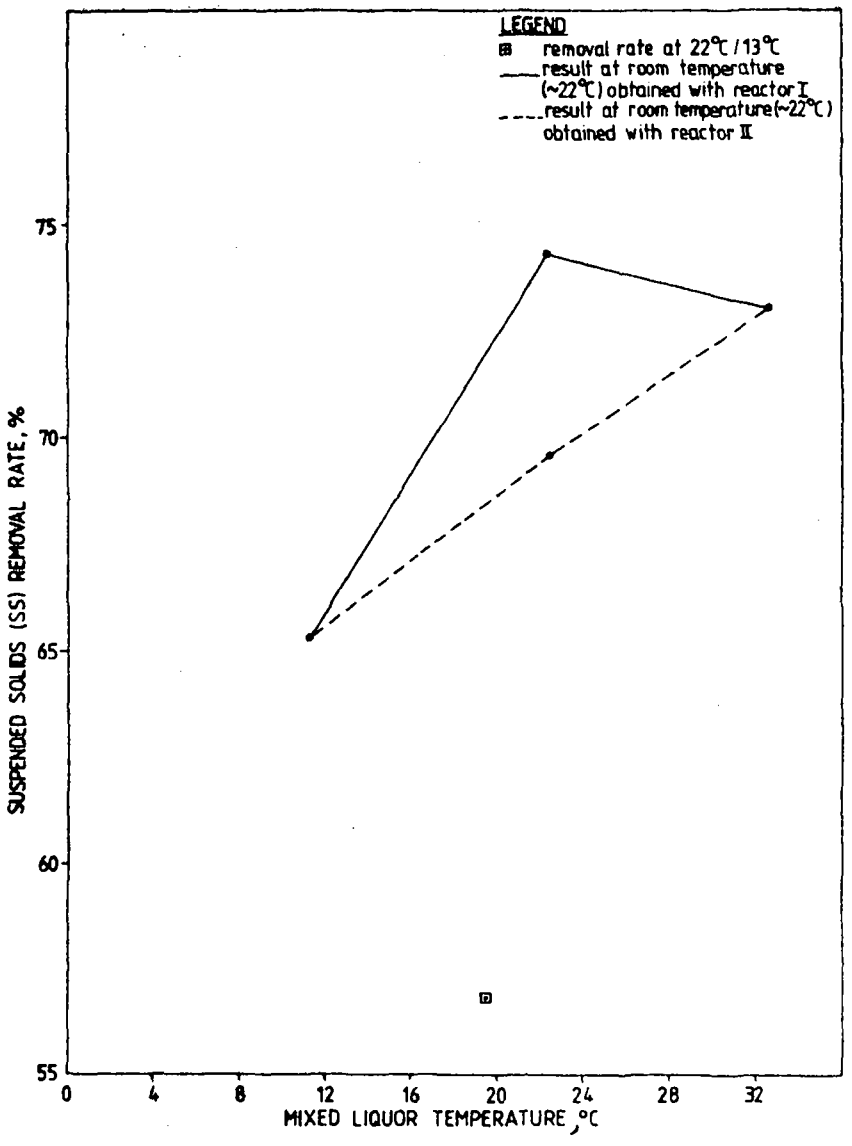


Figure 23. Suspended solids (SS) removal rates at different temperatures.

6.4 Sludge Settleability

Settled sludge volume and sludge volume index (SVI) were used as measures of sludge settleability. The results of the rate of quiescent settlings of mixed liquor solids in a 100 ml graduated cylinder within half an hour are presented for the different mixed liquor temperatures in figure 24 (and table 8).

TABLE 8 settle ability of mixed liquor in ml/100ml.

TIME min	MIXED LIQUOR TEMP. °C 22,2	22,4	22/13	11,2	32,5
0	100	100	100	100	100
5	48	95	71	50	43
10	37	89	51	38	32
15	32	84	41	33	28
20	29	78	37	30	25
25	28	74	33	28	24
30	26	67	31	27	23

+ Numbers in parentheses are MLSS and SVI values respectively.

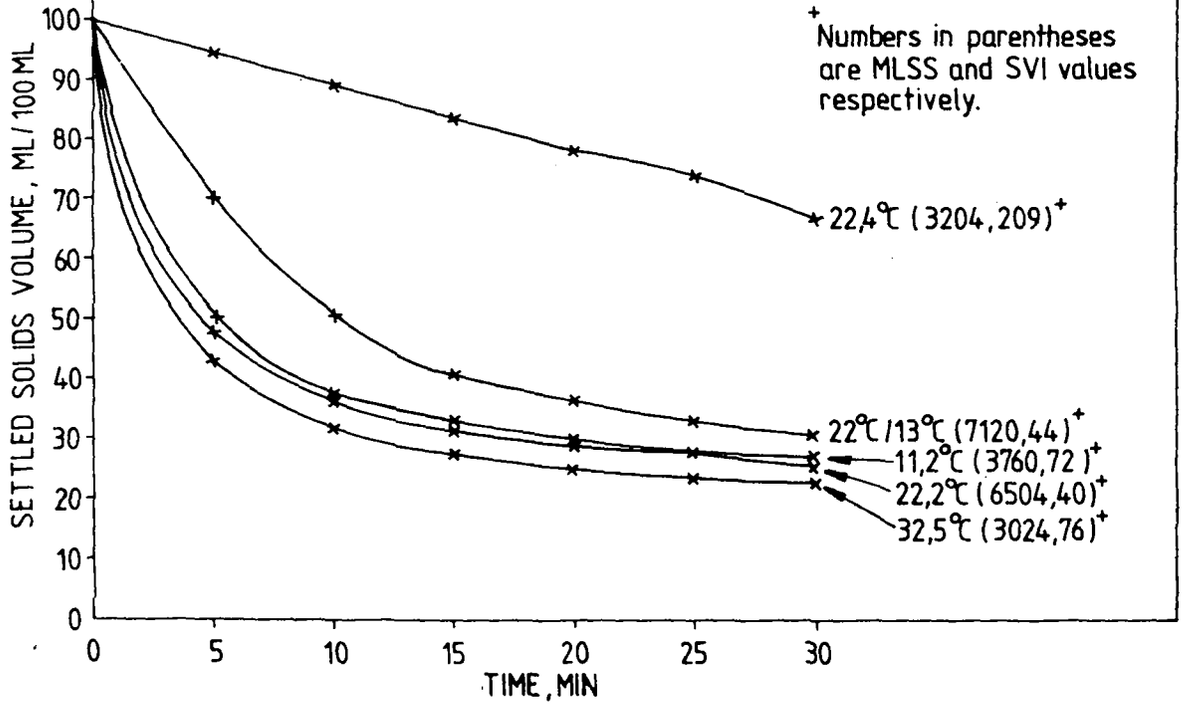


Figure 24. Settleability of mixed liquor solids with time at different temperatures.

The figure shows values obtained for tests done at the end of each phase of the experiment for each reactor, i.e., when the activated sludge was supposed to have acclimated well to the particular temperature it was subjected to. At a first glance the results may seem to indicate apparent bulking. In this case, with the exception of the curve for 22,4°C, it is observed that an increase in settled sludge volume and consequent slower solids settlement rate are related to a build-up of MLSS rather than temperature (see also table 8).

Considering R I, it had an MLSS concentration of 6504 mg/l with an SVI of 40 ml/g at 22,2°C. When the MLSS concentration went down to 3024 mg/l at 32,5°C, the SVI increased to 76 ml/g, and as is seen from the figure, the rate of solids settlement was relatively faster. With R II, the picture is the same. The MLSS concentrations of 7120, 3760 and 3204 mg/l corresponded to sludge volume indices of 44, 72, and 209 ml/g respectively. Regarding the rate of solids settlement, the more the MLSS concentration the slower the settlement rate (with the exception of the results obtained at 22,4°C, where the MLSS concentration was relatively low but still the sludge settlement rate remained the slowest).

An activated sludge that settles very rapidly (SVI of less than approximately 70 ml/g) can leave behind a turbid supernatant, the condition known as "pin-point floc" /32/. This phenomenon is usually associated with plants operating at the low end of the loading range, i.e., operating in the extended aeration mode /7/.

The above phenomenon could be the reason for the visually observed more turbid supernatant at 32,5°C, during settleability tests in a 100 ml graduated cylinder, than at other temperatures. It was also observed that tiny solids were adhering to the walls of the clarifier throughout the phase of the experiment at this same temperature (32,5°C).

Benedict et al /4/ experimenting with domestic wastewater in a continuous flow activated sludge system at different temperatures noticed that there was some culture dispersion at 32°C, which supports the above observations in the experiment.

It is seen from table 9 that the sludge volume indices in the first few weeks of the experiment were high compared to later values. This may have been, among other things, due to the characteristics of the seed activated sludge which was brought from a plant mainly treating wastewater from a food-processing factory.

Table 9. Weekly mean MLSS, MLVSS, SVI and F/M ratio values.

Week No.	Phase No.	MLSS mg/l	MLVSS mg/l	$\frac{\text{MLVSS}}{\text{MLSS}}$ %	SVI ml/g	F/M $\frac{\text{kgBOD}_7}{\text{kgMLSS}\cdot\text{d}}$
			Reactor I			
41		2549	2118	83	208,5	0,086
42		2271	1679	74	184,9	0,116
43	I	2343	1708	73	117,8	0,095
44	&	3637	1907	52	82,8	0,066
45	II	4816	2082	43	56,2	0,046
46	(22,2°C)	6617	2517	38	42,2	0,036
47		7172	2545	36	37,7	0,025
48		6676	3427	51	38,5	0,022
49		5098	1899	37	39,7	0,019
50	III	5634	2310	41	46,6	0,030
51	(32,5°C)	3902	1626	42	50,9	0,037
52		3068	1208	39	63,4	0,081
			Reactor II			
41		2924	2415	83	284,3	0,081
42	I	3033	2368	78	256,1	0,093
43	(22,4°C)	3200	2591	81	153,9	0,072
44	II	3731	2221	60	92,0	0,054
45		5029	2501	50	64,2	0,044
46	22°C/13°C	6289	2769	44	51,7	0,04
47		7187	2948	41		
48		7868	3985	51	65,1	0,016
49	III	7299	3500	48	66,3	0,014
50		6580	3388	52	62,7	0,02
51	(11,2°C)	5023	2926	58	60,2	0,031
52		4069	2381	59	60,4	0,064

Figure 25 (and table 9) shows that the weekly mean sludge volume indices for mixed liquor temperatures of 11,2°, 22°/13° and 32,5°C were under 100 ml/g within the same range of F/M ratio values (0,014 - 0,081) kg BOD_7/kg MLSS·d). The SVI values for temperatures of 22,2° and 22,4°C (room temperature) were much higher, mostly lying between 82 and 284 ml/g inclusive; but it should be noticed that the range of F/M ratio employed at room temperature was higher than for the other temperatures (0,025 - 0,116 kg BOD_7/kg MLSS·d). The interesting point here is the SVI values at room temperature, but within the same F/M ratio range as the other temperatures, are also below 100 ml/g. This most probably could imply that SVI is much more affected by F/M ratio rather than by temperature at least for the tested temperatures and used F/M ratios.

This is contradicted by results that were obtained by Lee et al /17/. They reported that in the 16-36°C temperature range, SVI of less than 100 ml/g were obtained regardless of the F/M ratio, and added that values of SVI did not rise to the 100-150 range until 40°C and F/M ratios of 1,4-2,4 kg BOD_5/kg MLVSS·d were reached. On the other hand Chao et al /5/ experimenting with a synthetic feed (which they claimed produced sludge with similar characteristics to those expected for sludges derived from domestic and industrial activated sludge wastewater treatment systems) in a continuous flow activated sludge system noted that there were two process loading intensity (PLI) ranges for which the sludge had relatively low SVI levels. The lower range corresponded to the design window employed for the extended aeration and conventional modes of operation of the activated sludge process. Magara et al /19/ studying the settling characteristics of activated sludge with synthetic wastewater as feed found out that SVI increases and settling velocity decreases with an increase in organic load (F/M ratio).

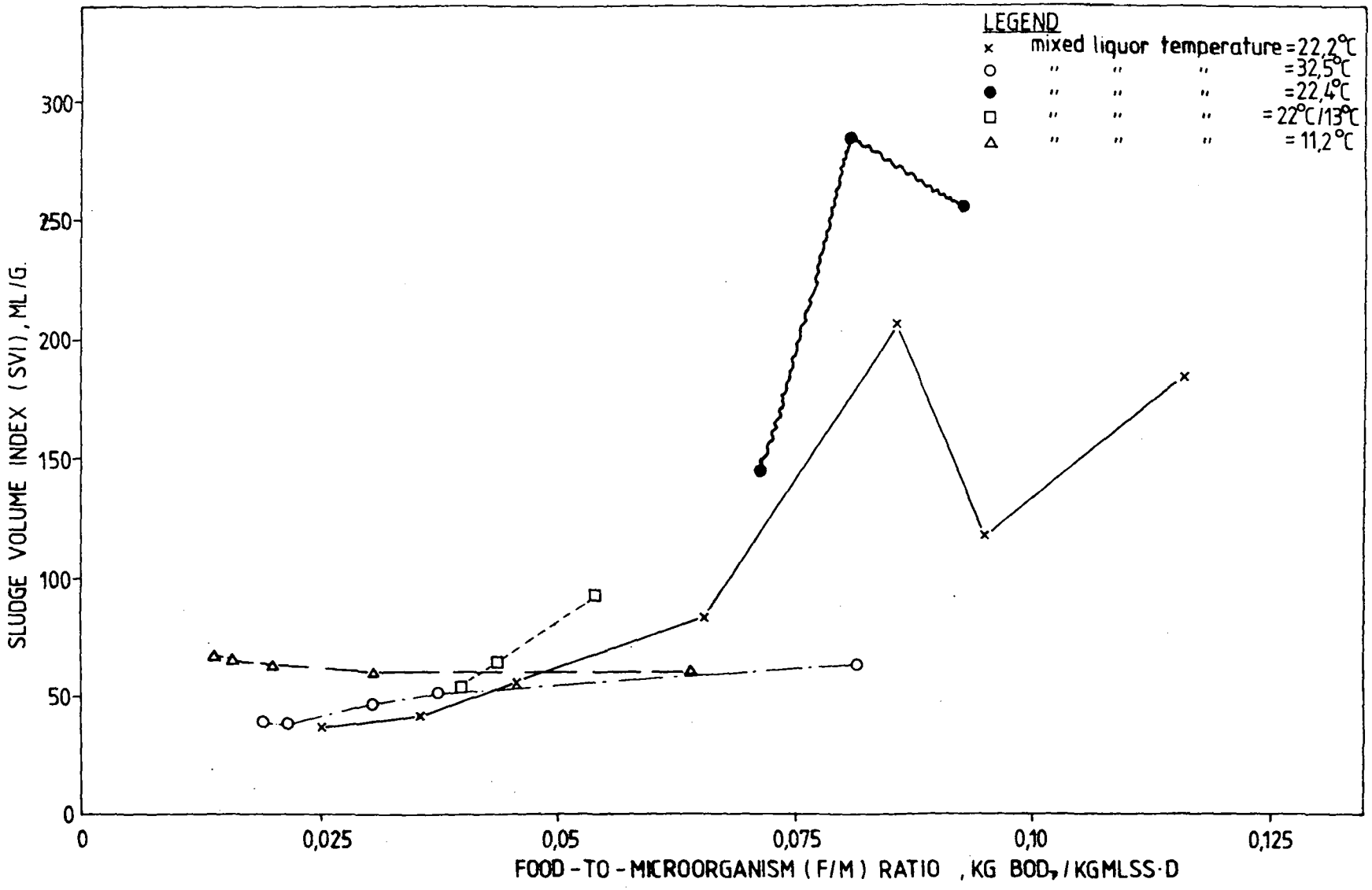


Figure 25. Relationship between sludge volume index (SVI) and food-to-micro-organism (F/M) ratio at different mixed liquor temperatures.

Solids of good settlement are favored by the endogenous phase of growth normally associated with the low organic loading of extended aeration /11/. At no time during the operation of the reactors in the experiment did the sludge exhibit the characteristics of a bulking, filamentous sludge.

While it is understood that absolute values of SVI may not be conclusive in determining the existence of activated sludge bulking, changes in SVI measurements with time for a given process are useful in detecting changes in sludge-settling characteristics /9/. Different investigators have given different SVI values as the threshold of bulking sludge. Palm et al /32/ classify 150 ml/g as bulking sludge. Sezgin /38/ reported (referring Vesilind, 1979) that a bulking sludge is characterized to yield an SVI of greater than 100 ml/g. Sayigh et al /37/ consider sludge with an SVI of equal to or less than 200 ml/g (over temperature range of 4° to 20°C) to be a good settling sludge. It is also suggested that most plants which operate well will have an SVI of 80 to 120 ml/g /7/.

In light of the above mentioned SVI threshold and range values for good settling sludge, it is difficult to evaluate the settleability of the sludges at different temperatures in the experiment according to the SVI values recorded. Generally it could be said that the weekly mean SVI values were mostly below 210 ml/g except for a couple of points at room temperature as is seen from figure 25. The reason behind the higher SVI values at room temperature (22,2° and 22,4°C) is difficult to ascertain. It is interesting to note though here that the sludge with the slowest settlement rate at 22,4°C (figure 24) corresponds to the highest SVI range in figure 25.

Some researchers have reported the adverse effects of temperature on settleability of sludge. Dougherty et al /10/, working with citrus wastes in a laboratory-scale activated sludge system, found that rapid temperature changes caused temporary bulking and poor quality effluent. Lee et al /17/ also found that at temperatures greater than 30°C, sludge settleability deteriorated, i.e. part of the biological floc was so fine it remained suspended after 1/2 h quiescent settling.

6.5 Biological Kinetic Parameters

The biological kinetic parameters, cell yield coefficient (Y), and specific organism decay rate (K_d) along with the correlation coefficients estimated from this study results were computed by regression analysis.

Both, cell yield coefficient and specific organism decay rate were determined by plotting equation 3.21 (section 3.3) for each set of results obtained at different mixed liquor temperatures. MLVSS concentrations measured at least every other day, effluent volatile suspended solids (VSS), the 7-day BOD of the raw influent and effluent (total BOD_7 and filtered BOD_7 with ATU respectively) analyzed twice a week, and daily hydraulic detention time (H_t) were used in the computation of the kinetic parameters. The MLVSS was used as the concentration of viable organisms, x_1, in the aeration chambers. Effluent VSS concentrations were substituted for the concentrations of viable organisms in the effluent, x_2. S_o, the influent substrate concentrations, were assumed to be equal to the 7-day BOD of the raw wastewater, and the effluent 7-day BOD's (with ATU) were used as the substrate escaping in the effluent, S_1. All the above data are tabulated in tables 10-14 for the different mixed liquor temperatures.

The kinetic parameters resulting from data obtained for mixed liquor temperatures of 22,2°, 32,5°, 22,4°, 22°/13°, and 11,2°C are shown in figures 26, 27, 28, 29 and 30 respectively and summarized in table 15. The correlation coefficients of the regression equations used to determine the kinetic parameters are also tabulated in the same table.

Table 10. Data for plotting equation 3.21
(Phase I and II, Reactor I, Temp. = 22,2°C)

Date	x_1	x_2	$b=\frac{x_1}{x_2}$	H_t	S_o-S_1	bH_t	$\frac{b(S_o-S_1)}{x_1}$
	mg/l	mg/l	—	h	mg/l	h	--
11/10/83	2684	29	92,55	34,74	260	3215	8,965
13/10/83	1968	29	67,86	46,48	260	3154	8,965
14/10/83	1904	26	73,23	30,92	352	2264	13,538
16/10/83	1916	26	73,69	32,57	352	2400	13,538
18/10/83	1584	12	132,00	25,46	316	3361	26,333
20/10/83	1492	12	124,33	45,67	316	5678	26,333
24/10/83	1960	0		23,63	273		
26/10/83	1672	36	46,44	31,80	309	1477	8,583
28/10/83	1376	30	45,87	31,96	253	1466	8,434
30/10/83	2076	30	69,20	29,05	253	2010	8,433
1/11/83	2068	0		26,53	214		
4/11/83	1856	3	618,67	29,41	273	18195*	91,000*
7/11/83	1796	3	598,67	31,69	273	18972*	91,001*
8/11/83	1964	37	53,08	23,70	230	1258	6,216
10/11/83	1908	37	51,67	32,12	230	1656	6,217
11/11/83	2112	11	192,00	32,12	321	6167	29,182
13/11/83	2344	11	213,09	33,94	321	7232	29,182
15/11/83	2284	16	142,75	33,66	278	4805	17,375
17/11/83	2484	16	155,25	30,05	278	4665	17,375
18/11/83	2576	52	49,54	34,54	371	1711	7,135
20/11/83	2724	52	52,38	33,52	371	1755	7,134
22/11/83	2892	0		33,52	240		
24/11/83	2332	0		38,64	240		
25/11/83	2000	8	250,00	38,64	306	9660	38,250
27/11/83	2956	8	369,50	36,24	306	13391	38,250
29/11/83	2724	24	113,50	33,66	232	3820	9,667

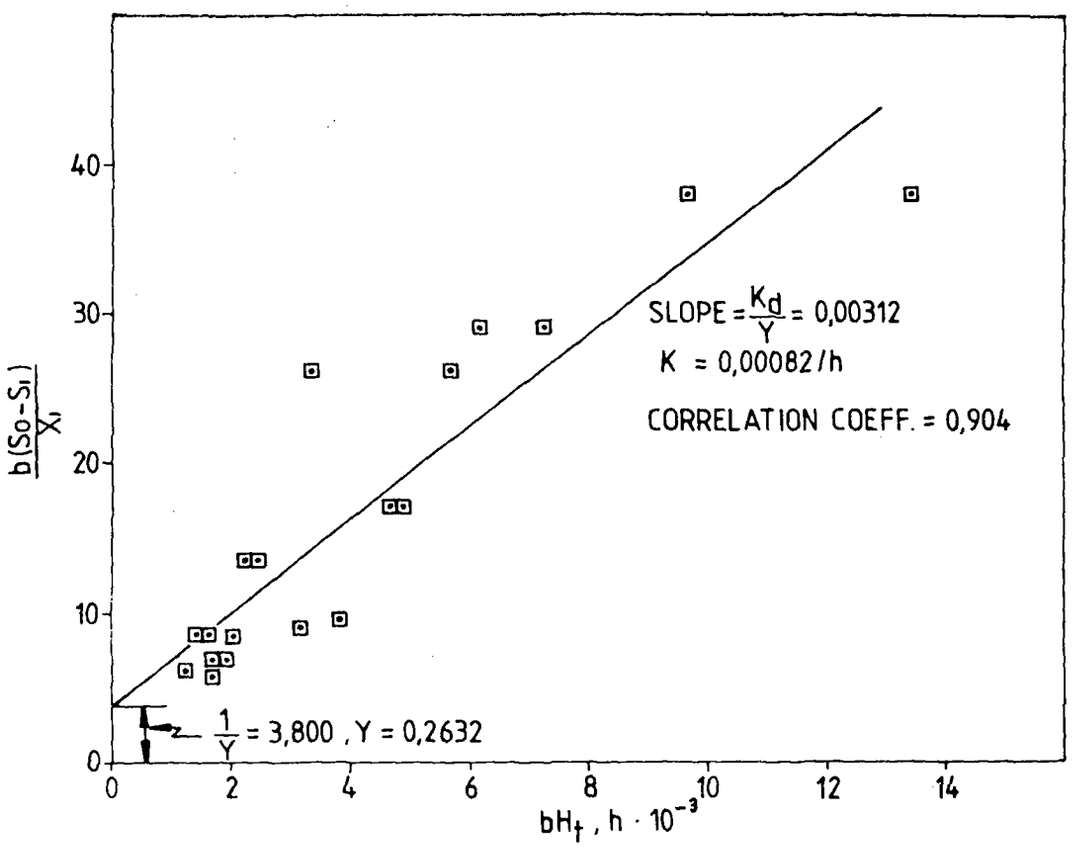


Figure 26. Determination of yield coefficient (Y) and specific organism decay rate (K_d) for temperature of 22,2°C.

Table 11. Data for plotting equation 3.21
(Phase III, Reactor I, Temp.= 32,5°C)

Date	x_1	x_2	$b=\frac{x_1}{x_2}$	H_t	S_o-S_1	bH_t	$\frac{b(S_o-S_1)}{x_1}$
	mg/l	mg/l	—	h	mg/l	h	—
1/12/83	3976	24	165,67	129,64[a]	232		
4/12/83	3580	21	170,48	45,29	228	7721[c]	10,857[c]
6/12/83	2012	0		45,29	267		
8/12/83	1668	0		45,29	267		
9/12/83	1608	29	55,45	45,29	85	2511	2,931
11/12/83	2308	29	79,29	33,34	85	2654	2,931
13/12/83	2444	26	94,00	33,34	224	3134	8,615
15/12/83	2336	26	89,85	33,34	224	2996	8,615
16/12/83	2076[b]	14	148,29	33,34	291	4944[b]	20,786[b]
18/12/83	2384[b]	14	170,29	47,87	291	8152[b]	20,786[b]
20/12/83	2800[b]	17	164,71	47,87	300	7885[b]	17,647[b]
22/12/83	1200	17	70,59	47,87	300	3379	17,647
23/12/83	1448	10	144,80	47,87	327	6932	32,700
25/12/83	1232	10	123,20	68,26	327	8410	32,700
26/12/83	1448	10	144,80	68,26	327	884	32,700
27/12/83	1140	24	47,50	68,26	344	3242	14,333
28/12/83	1192	24	49,67	23,45	344	1165	14,333
29/12/83	1000	24	41,67	23,27	344	970	14,333
30/12/83	1500	18	83,33	18,22	239	1518	13,278
1/1/84	1384	18	76,89	17,80	239	1369	13,278
2/1/84	1188	18	66,00	22,46	239	1482	13,278

a very high H_t due to a very low influent flow, not considered for calculations

b high MLVSS values after wasting of sludge was started, therefore derived points not considered in plotting fig.27

c data before 9/12/83 not considered in plotting fig 27. The time from 29/11/83 to 8/12/83 taken as an acclimation period for micro-organisms.

Table 12. Data for plotting equation 3.21
(Phase I, Reactor II, Temp.= 22,4°C)

Date	x_1	x_2	$b=\frac{x_1}{x_2}$	H_t	S_o-S_1	bH_t	$\frac{b(S_o-S_1)}{x_1}$
	mg/l	mg/l	—	h	mg/l	h	—
11/10/83	2756	31	88,90	34,74	268	3088	8,645
13/10/83	2396	31	77,29	35,53	268	2746	8,645
14/10/83	2268	31	73,16	33,17	356	2427	11,484
16/10/83	2240	31	72,26	27,69	356	2001	11,484
18/10/83	2432	5	486,40	30,39	318	14782	63,600
20/10/83	2212	5	442,40	25,89	318	11454	63,600
24/10/83	2460	0		22,16	276		

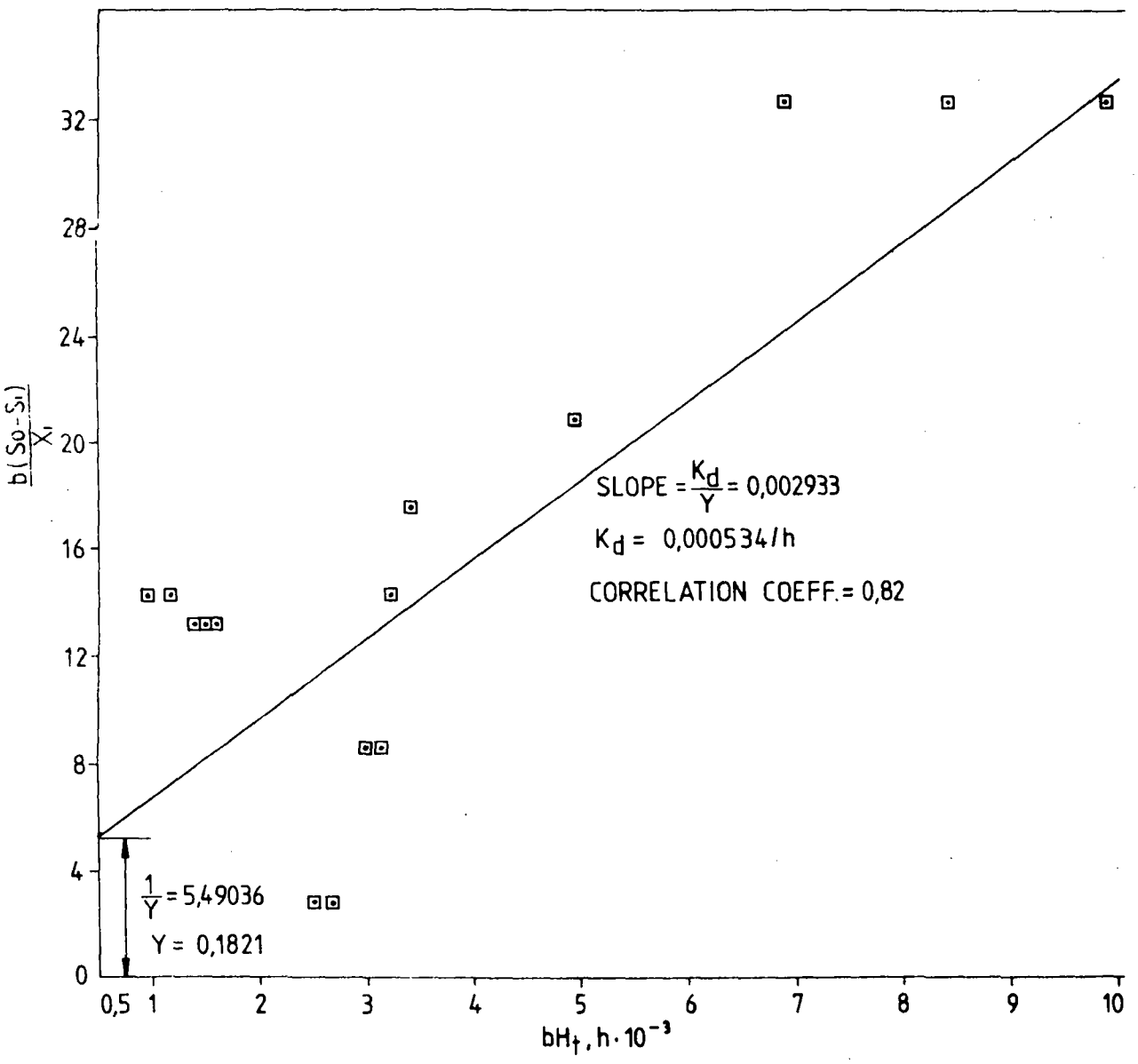


Figure 27. Determination of yield coefficient (Y) and specific organism decay rate (K_d) for temperature of 32,5°C.

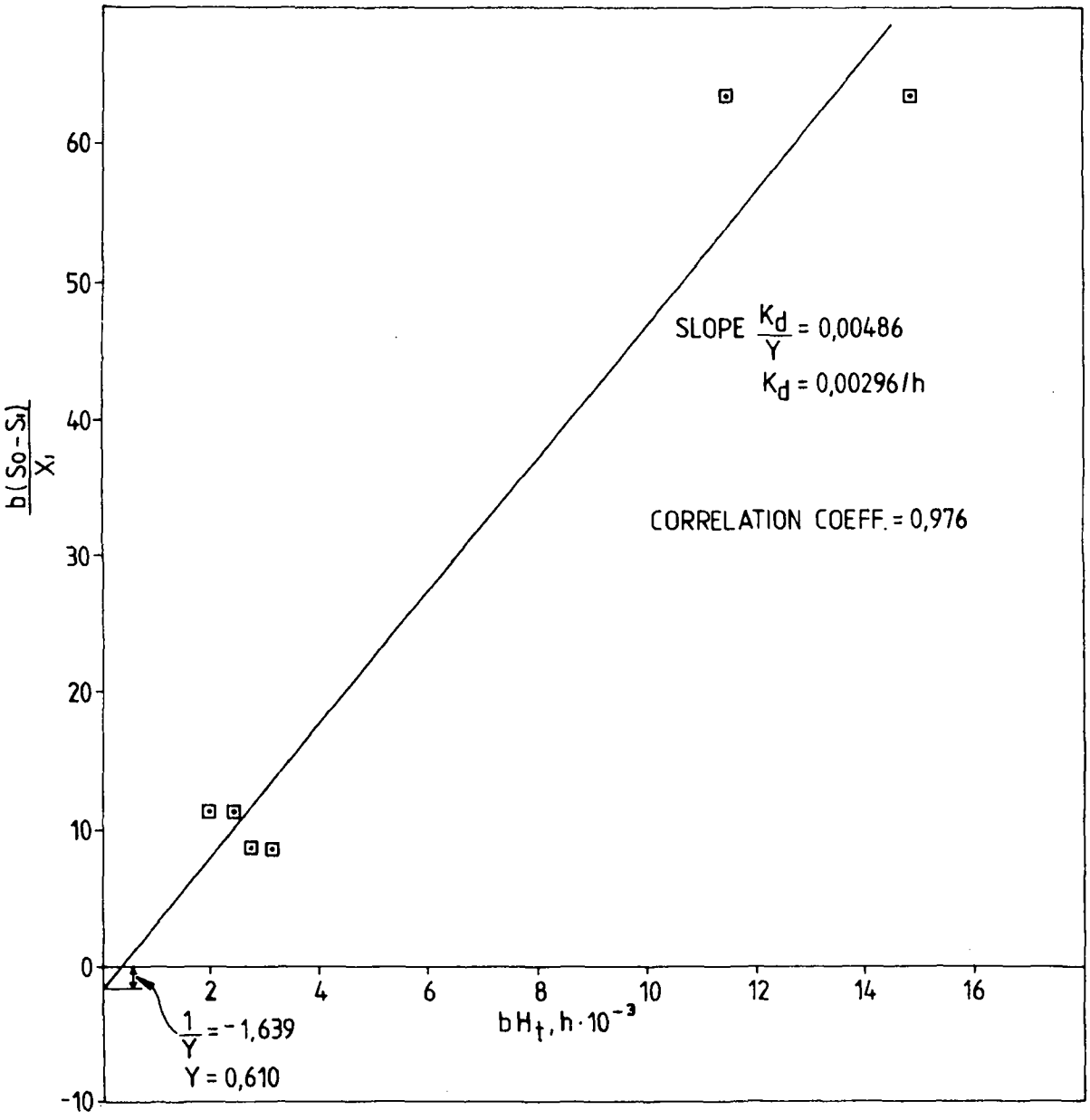


Figure 28. Determination of yield coefficient (Y) and specific organism decay rate (K_d) for temperature of 22,4°C.

Table 13. Data for plotting equation 3.21
(Phase II, Reactor II, Temp.= 22°C/13°C)

Date	x_1	x_2	$b=\frac{x_1}{x_2}$	H_t	S_o-S_1	bH_t	$\frac{b(S_o-S_1)}{x_1}$
	mg/l	mg/l	—	h	mg/l	h	—
26/10/83	2648	58	45,66	33,62	308	1535	5,311
28/10/83	2484	42	59,14	31,67	253	1873	6,024
30/10/83	2640	42	62,86	29,54	253	1857	6,024
1/11/83	2404	0		36,05	213		
4/11/83	2000	4	500,00	32,84	272	16420	68,000
7/11/83	2260	4	565,00	19,89	272	11238	68,000
8/11/83	2424	41	59,12	32,91	231	1946	5,634
10/11/83	2500	41	60,98	36,12	231	2203	5,634
11/11/83	2520	16	157,50	36,12	322	5689	20,125
13/11/83	2560	16	160,00	33,95	322	5427	20,125
15/11/83	2920	11	265,45	32,96	278	8749	25,273
17/11/83	2648	11	240,73	28,35	278	6825	25,273
18/11/83	2520	28	90,00	31,01	371	2791	13,250
20/11/83	2988	28	106,71	34,11	371	3640	13,250
22/11/83	3420	0		34,11	242		
24/11/83	2616	0		42,50	242		
25/11/83	2336	4	584,00	42,50	305	24820*	76,250*
27/11/83	3420	4	855,00	36,40	305	31122*	76,250*

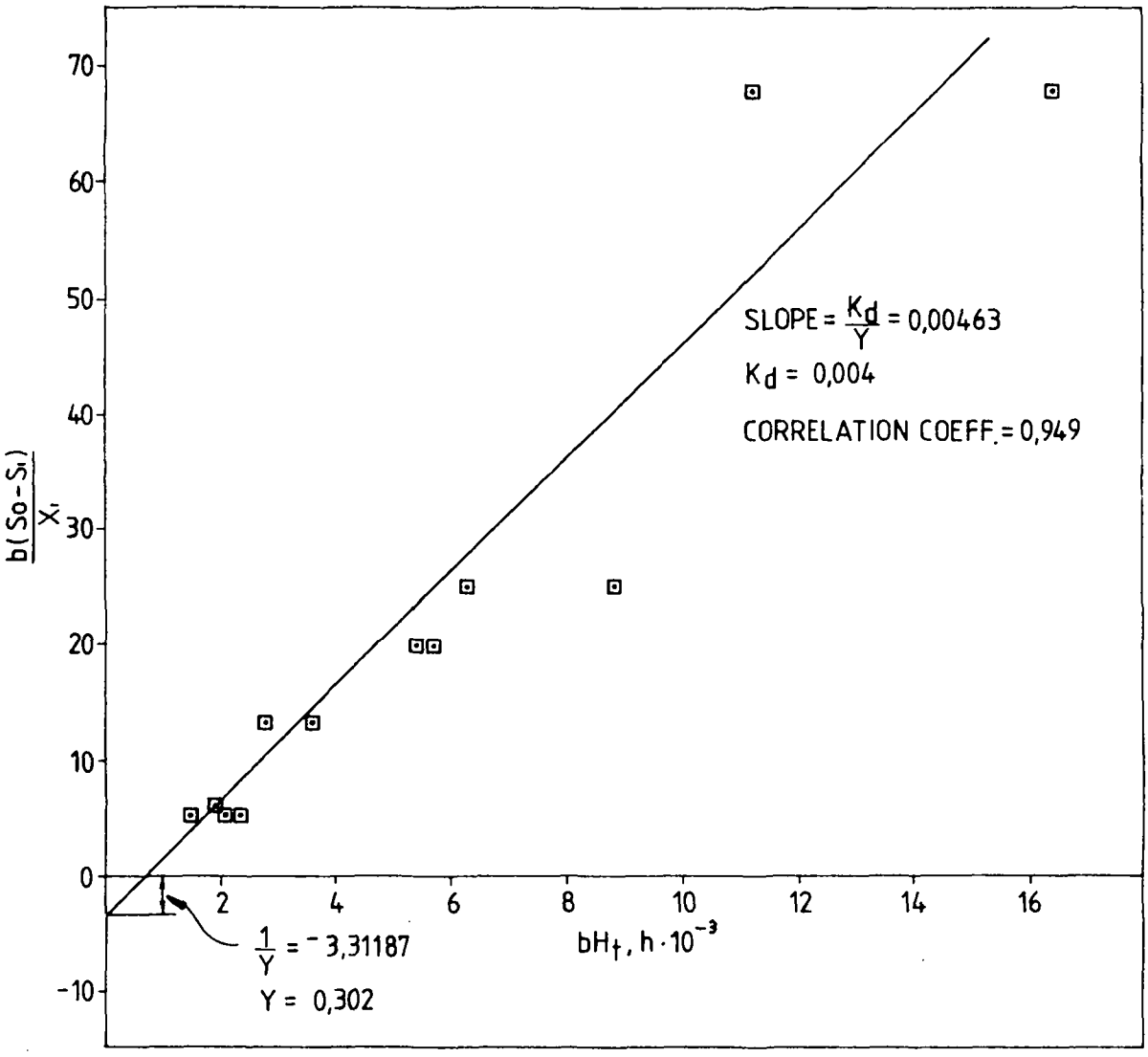


Figure 29. Determination of yield coefficient (Y) and specific organism decay rate (K_d) for temperature of 22°C/13°C.

Table 14. Data for plotting equation 3.21
(Phase III, Reactor II, Temp.= 11,2°C)

Date	x_1	x_2	$b=\frac{x_1}{x_2}$	H_t	S_o-S_1	bH_t	$\frac{b(S_c-S_1)}{x_1}$
	mg/l	mg/l	—	h	mg/l	h	—
29/11/83	3200	24	133,33	49,72	239	6629	9,958
1/12/83	4740	24	197,50	49,72	239	9820	9,958
4/12/83	4016	26	154,46	39,46	232	6095	8,923
6/12/83	3696	0		39,46	> 268		
8/12/83	3296	0		42,51	> 268		
9/12/83	3284	35	93,83	42,51	86	3989	2,457
11/12/83	3724	35	106,40	48,04	86	5111	2,457
13/12/83	3648	226	16,14	48,04	229	775	1,013
15/12/83	3288	226	14,55	48,04	229	699	1,013
16/12/83	3316	14	236,86	48,04	292	11379	20,857
18/12/83	3300	14	235,71	48,04	292	11324	20,857
20/12/83	3928	16	245,50	48,04	301	11794	18,813
21/12/83	4040	16	252,50	48,04	301	12130	18,813
22/12/83	2712	16	169,50	48,04	301	8143	18,813
23/12/83	2220	3	740,00	48,04	329	35550*	109,667*
25/12/83	2320	3	773,33	50,54	329	39084*	109,667*
26/12/83	2336	3	778,67	50,54	329	39354*	109,667*
27/12/83	2364	17	139,06	50,54	344	7028	20,235
28/12/83	2164	17	127,29	28,30	344	3602**	20,235
29/12/83	2292	17	134,82	21,37	344	2881**	20,235
30/12/83	2704	28	96,57	19,44	237	1877	8,464
1/1/84	2504	28	89,43	19,83	237	1773	8,464
2/1/84	1992	28	71,14	24,01	237	1708	8,464

For tables 10 - 14

** Low H_t, combined with comparable BOD_7 reduction with the other data, gave rise to a low bH_t value, the $\frac{b(S_o-S_1)}{x_1}$ term remaining unaffected. If these points are plotted, their scatter will be very much noticeable.

* very low x_2 (compared with the other entries) gave rise to greater values of both terms i.e. bH_t & $\frac{b(S_o-S_1)}{x_1}$.

These points will also spread far away from the other points if plotted.

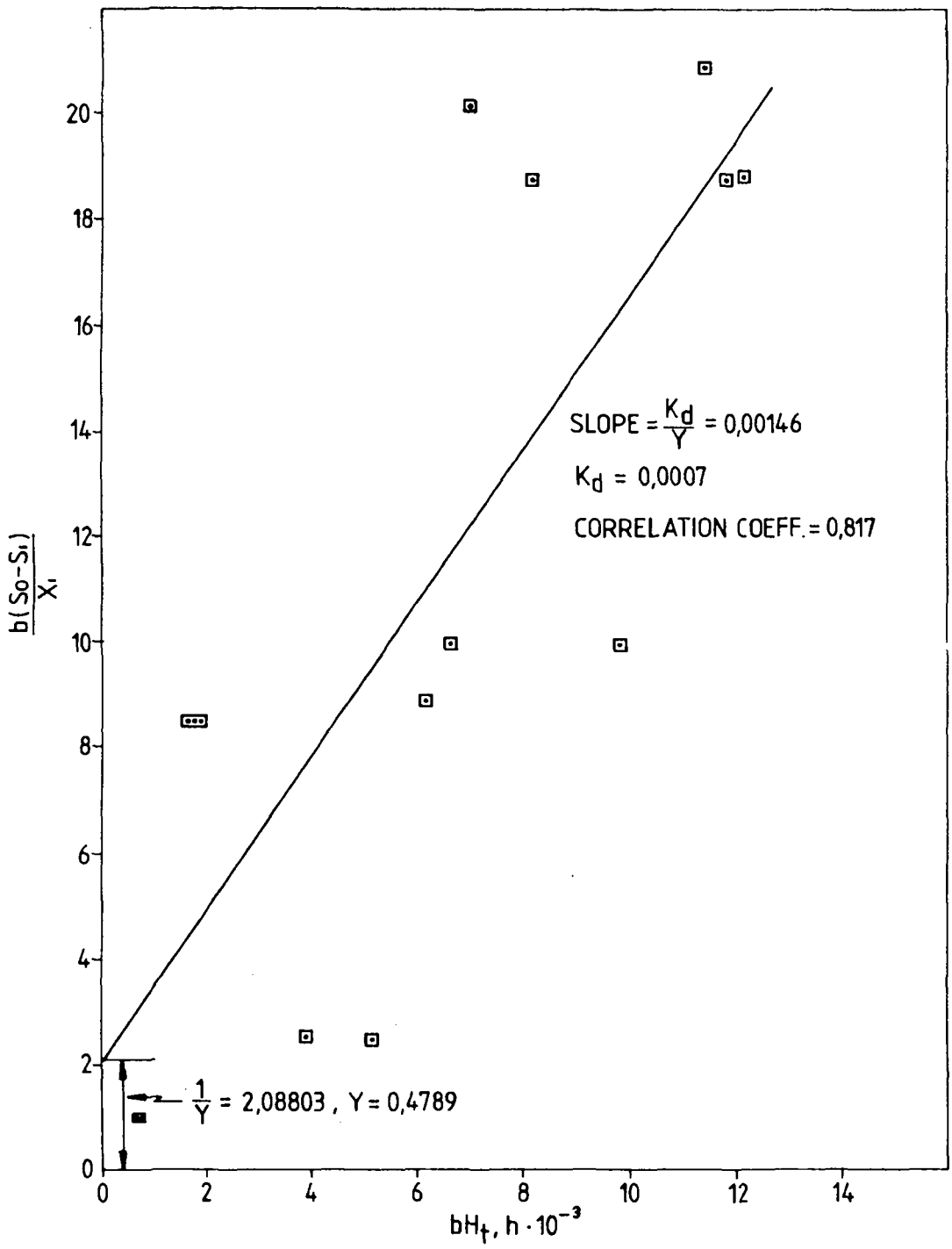


Figure 30. Determination of yield coefficient (Y) and specific organism decay rate (K_d) for temperature of 11,2°C.

Table 15. Cell yield coefficient and specific organism decay rate at different mixed liquor temperatures.

Phase	Reactor	Mixed liquor temperature	Cell yield (Y)		Organism decay rate (K_d)		Correlation coefficient of regression line
			Experimental	Literature	Experimental	Literature	
No.	No.	°C			h^{-1}	h^{-1}	
I	R I	22,2	0,2632	0,44(20°C)[a]	0,00082	0,0040(20°-21°C)[b]	0,904
	R II	22,4	0,6100	0,70(20°-21°C)[b]	0,00296	0,0025(20°C)[c]	0,976
II	R I	22,2	0,2632	0,5(20°C)[c]	0,00082	0,00058(20°-21°C)[b]	0,904
	R II	22/13	0,3020		0,00400		0,949
III	R I	32,5	0,1821	0,38(30°C)[a]	0,00053		0,820
	R II	11,2	0,4789	0,53(10°C)[a]	0,00070		0,817

a obtained from reference /17/
b " " " /26/
c " " " /6/

3.5.1 Cell Yield Coefficient

The yield coefficients computed ranged from a minimum of 0,1821 at 32,5°C to a maximum 0,610 at 22,4°C. Yield values calculated using equation 3.7 (see Appendix A) are presented in table 16. It is seen that allowing for experimental error, the yield values for the same temperature but using two different methods match reasonably with the exception of the values for the temperature of 22°C/13°C. Here there is a wide difference between 0,302 computed by repression analysis and 0,985 calculated using equation 3.7. Due to lack of previous work reported in the literature on this same temperature, it is difficult to discuss the results obtained.

Table 16. Cell yield coefficients computed using equation 3.7

Time span	Mixed liquor temperature °C	Mean Y
8/10 - 3/12/83	22,2	0,2950
11/12 - 31/12/83	32,5	0,1751[d]
8/10 - 29/10/83	22,4	0,4291
30/10 - 26/11/83	22/13	0,9850
4/12 - 10/12/83	11,2	0,6346

[d] Y = 1,41124, obtained for the period 4/12-10/12 was skipped in the mean calculation.

Remembering the highest BOD_7 removal rate (99 %) attained at this temperature, one can reasonably assume a higher yield value if the substrate consumed by the micro-organisms is used more for growth rather than intercellular storage. The mean yield values computed using equation 3.7 for this particular temperature were mostly greater than 1 (table A4, Appendix A). A yield coefficient greater than one is impossible if the growth rate and yield are dependent on the influent BOD /26/. These high yield values are probably attributable to experimental error which might have been compounded by temperature shock effects on the micro-organisms while the reactor was being placed alternately at room temperature and 10°C daily. Therefore the mean yield value of 0,985 is unreliable.

Literature information concerning temperature effects on the yield coefficient, is both confusing and conflicting. The yield coefficient along with other kinetic parameters is generally considered to be constant for a given substrate and biological population /41/. It is also commonly assumed that the yield coefficient does not vary with temperature /35/. This does not seem reasonable for a mixed biological population if significant shifts in dominant organisms occur /35/.

Randall et al /35/ investigated the effect of temperature on kinetic parameters using a laboratory-scale open respirometer techniques with a soluble mixture of glucose and mineral as substrate. They found that the theoretical (true) yield coefficient, Y_T, was relatively constant with an average value of 0,574 (COD used in the study) for the temperature from 15° to 24,5°C. It then increased to an average value of 0,603 for the temperature of 31,5°C, indicating a probable shift in the species dominating the reaction.

Ludzack et al /10/ reported that calculation based on Garrett and Sawyers data indicate yield coefficients of 0,53, 0,44 and 0,38 at temperatures of 10°, 20° and 30°C, respectively (see figure 31).

The effects of temperature on Y_T as determined by a number of investigators are illustrated in figure 31 /6/.

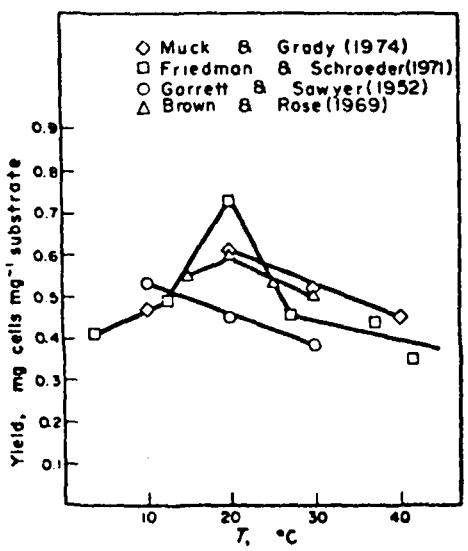


Figure 31. Variation of the true cell yield coefficient with temperature /6/.

It should be noted that the yields of Friedman and Schroeder, and Garrett and Sawyer are not true yields (Y_T), but observed yields (Y) /6/. Y is related to Y_T by a function containing the specific organism decay rate and MCRT /6/. Because of the way that organism decay rate varies with temperature it can be shown that if Y displays a maximum as a function of temperature so will Y_T /6/. The yield coefficients computed (tables 15 and 16) from data of this study are observed yields.

Three of the four sets of data in figure 31 suggest a maximum in Y_T at approximately 20°C /6/. Collins et al /6/ in the same experiment mentioned in section 6.3, found the temperature dependence of Y_T selected for their study as tabulated in table 17. Those values (of temperature) were chosen to give a curve of the general shape shown by the three data sets in figure 31 which exhibit a maximum, and to give a value at 20°C of 0,50 (based on COD), which is a typical value for cells grown on domestic sewage /6/.

Table 17. Variation of the true cell yield coefficient with temperature /6/.

Temperature (°C)	Y_t (mg cells/mg substrate)
10	0.392
15	0.446
20	0.500
25	0.465
30	0.431
35	0.401

The cell yield values as computed for the different mixed liquor temperatures are plotted in figure 32. The solid line shows a decrease in yield value with increasing temperature (yield value at 22,2°C obtained with R I). On the other hand the dashed line shows almost the same trend as the curve obtained with the data of Friedman and Schroeder in figure 31. A maximum yield value at 22,4°C with the values increasing and decreasing below and above this temperature respectively. Benedict /10/ reported that Wineberger's studies resulted with a maximum yield occurring at a temperature of 20°C. Friedman et al /10/ attributed this maximum value of cell yield to the maximum utilization of substrate for synthesis purposes by the mesophilic organisms at temperatures near 20°C.

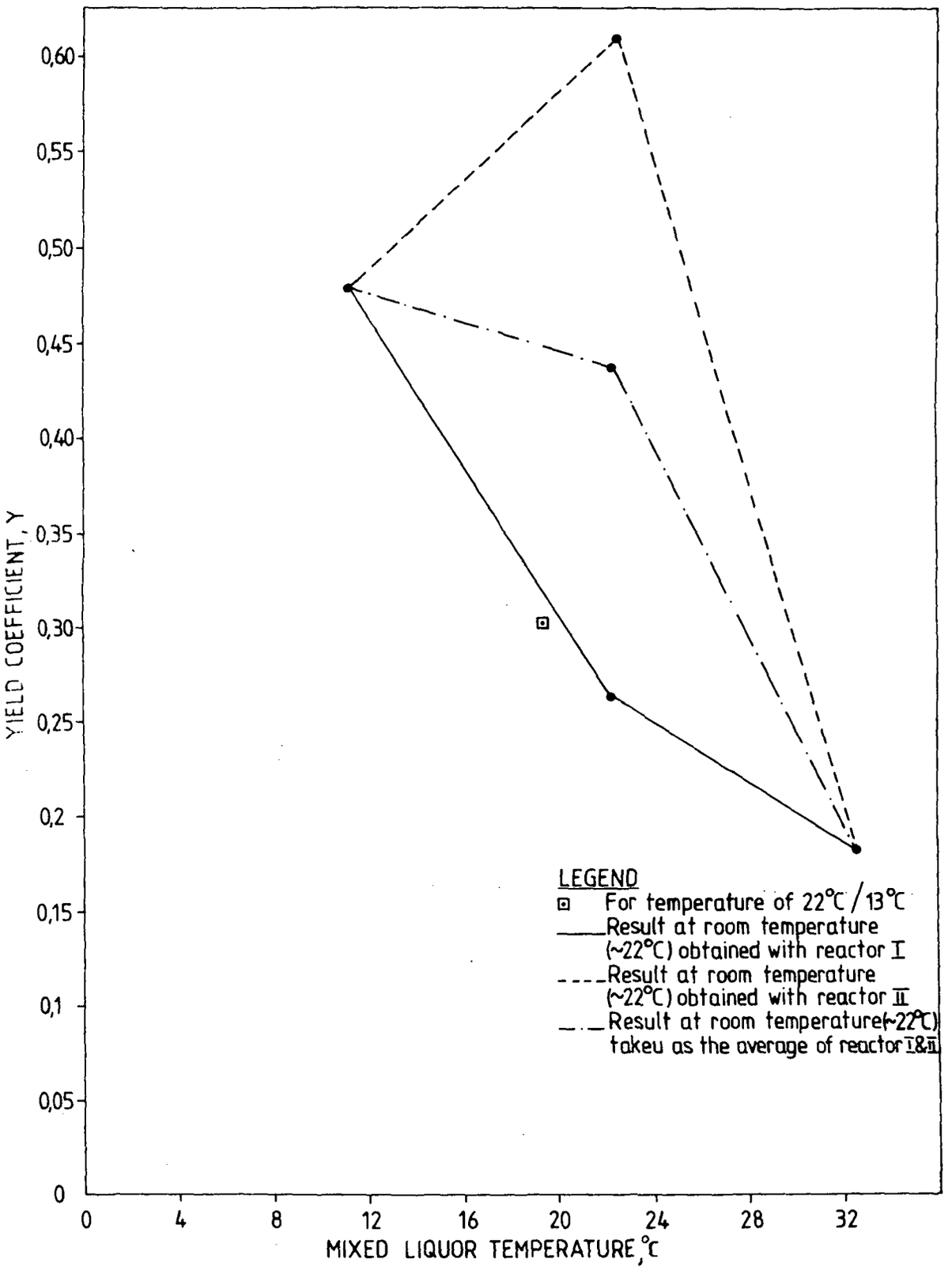


Figure 32. Cell yield coefficient at different temperatures.

In the same review Benedict reported the findings of Wineberger, he also mentioned that Sawyer and Nichols concluded that maximum yield was found at 15°C for activated sludge studies in the 10° to 25°C. In their laboratory study, Ludzack et al /10/ concluded from continuous flow experiments that "solids accumulation per unit weight of BOD input was substantially greater at 5° than 30°C". However, Benedict found that yield coefficient was equal to 0,42 and 0,62 at 4° and 19°C, respectively, for continuous-flow experiments /10/.

Sayigh et al /37/ studying the effects of temperature in a completely mixed continuous-flow model with domestic waste-water as feed obtained results that indicate temperature variations having a minimal effect on the yield coefficient in the 4° to 20°C range. At 31°C the value of the yield dropped appreciably. They explained that the acclimation of the predominant bacterial species seemed to be the reason behind the independence of the cell yield coefficient from temperature variations in the range of 4° to 20°C. They reasoned that such acclimation must result essentially in the same balanced bacterial population predominating at the different temperatures, removing soluble substrate and resulting in the same cell yield. On the other hand, they attributed the sudden drop in the value of the yield coefficient at a temperature of 31°C to a new and different and appreciably more active bacterial species predominating after acclimation at 31°C, and that their yield was much less for the same mass of soluble COD removed.

Some researchers have found the yield coefficient to be dependent on organic loading rather than temperature. Helmers, as reported by Ludzack et al /10/, concluded that solids production varied with BOD removal and was relatively independent of temperature. Pipes /10/, utilizing several substrates, found a linear relationship between cell growth

counts and the COD removed for activated sludge. But on the contrary, Streebin and Phillips /10/, operating batch systems at temperatures from 20° to 60°C, found large variations in yield as a function of temperature accompanied by small changes in COD removal.

A constant yield value with varying loading range has also been observed by some investigators. The data (field data) of Hopwood and Downing /10/, as shown by Eckhoff and Jenkins gave a constant yield value (0,97) over a large range of loading and MLSS concentrations. Eckhoff and Jenkins also found a constant yield value of 0,46 under a wide range of organic loading for laboratory studies and 0,33 for field studies. Presumably, these studies which resulted in a constant yield value and the studies of Pipes were conducted under relatively constant temperature conditions /10/.

6.5.2 Specific Organism Decay Rate

The specific organism decay rates (K_d) calcuted for this study ranged from 0,00053 h^{-1} at 32,5°C to 0,00400 h^{-1} at 22°C/13°C (see figures 26-30 and table 15). The values are far from being constant indicating most probably the effects of temperature. However, even for a fixed mixed liquor temperature, the specific organism decay rate varies widely, and the variation generally is attributed to the operating conditions in the plant or model. Most researchers have reported a decline in the decay rate as the influent substrate concentration increases, but Stewart /26/ has shown that the decay rate increases with an increase in substrate concentration. Maintained mean substrate concentrations at different mixed liquor temperatures during the experiment are tabulated against computed decay rates in table 18. Although not strongly convincing, the decay rates obtained at room temperature (22,4° and 22,2°C) support the findings by Stewart.

Table 18. Mean substrate concentrations and organism decay rate at different temperatures.

Mixed liquor temperature	Mean substrate concentration	Specific organism decay rate
°C	BOD_7, mg/l	h^{-1}
22,4	314,3	0,00296
22,2	294,8	0,00082
22/13	284,9	0,00400
11,2	258,1	0,00070
32,5	256,5	0,00053

Middlebrooks and Garland /26/ working with laboratory scale extended aeration models (with domestic wastewater as feed) found organism decay rates ranging from 0,0018 to 0,0040 h^{-1} for a mixed liquor temperature of 20°-21°C. They also computed a decay rate from field data collected by Morris and obtained a value of 0,00058 h^{-1}. Therefore, at least, the range of values computed for the organism decay rate in this experiment agree with those of Middlebrooks and Garland's.

Yang et al /42/ employing a laboratory-scale extended aeration pilot plant treated different soluble organic industrial wastewaters at ambient air temperature of 20°-30°C. The specific organism decay rates computed from data of the above study ranged from 0,000419 to 0,001306 h^{-1}.

An interesting point is that the mixed liquor temperature, 22°/13°C, at which the highest BOD removal rate (99 %) was obtained was also the temperature for which the maximum organism decay rate, 0,00400 h^{-1}, was computed. The decay rate is primarily a measure of the reduction in mass of the volatile organisms (VSS) due to endogenous respiration. The more substrate oxidized (the more BOD removed) by an organism the greater is the need for maintenance of the cellular

protoplasm /26/. This increased maintenance could account easily for the highest decay rate obtained at a temperature for the highest BOD removal.

Sayigh et al /37/ in the same study mentioned in section 6.5.1 found results that indicate that the organism decay rate could be considered as a constant in the 4° to 20°C range with an average value of 0,00833 h^{-1}. Laboratory-scale batch aerobic digestion of waste activated sludge was performed by Randall et al /35/ at temperatures of 5, 10, 20, 30, 35 and 45°C. Figure 33 shows the results. They found the organism decay rate was a maximum at 20°C. Also shown in the figure are results obtained by Bishop and LePage for aerobic digestion of 11 days. The values of the decay rate at 25° and 30°C were substantially less than the value at 20°C. However they also observed a 15°C value that was substantially less than the 10°C value.

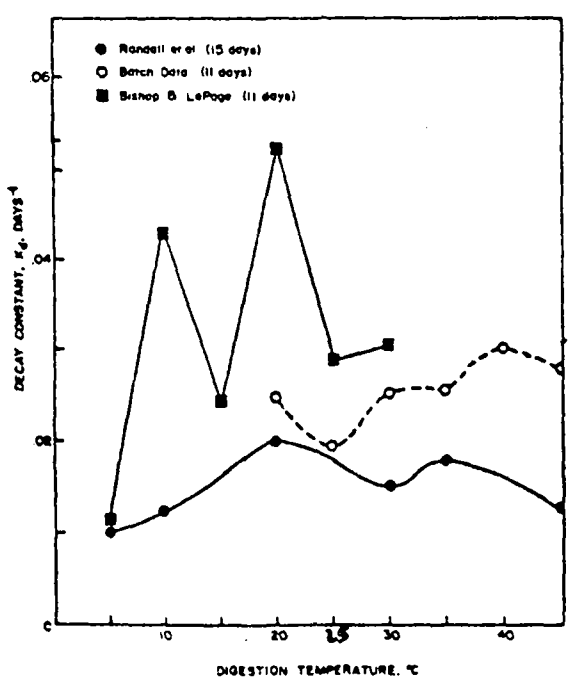


Figure 33. Variation in the specific organism decay rate with temperature./35/

In figure 34, the computed specific organism decay rates are plotted against temperature. In all the three curves, the maximum decay rate (excluding that for the temperature of 22°C/13°C) is at room temperature (22,2° and 22,4°C). This agrees with some of the results of other researchers as mentioned above. The trend of the dashed curve in this figure closely resembles that of Bishop and LePage's between temperatures of 15° and 30°C in figure 33.

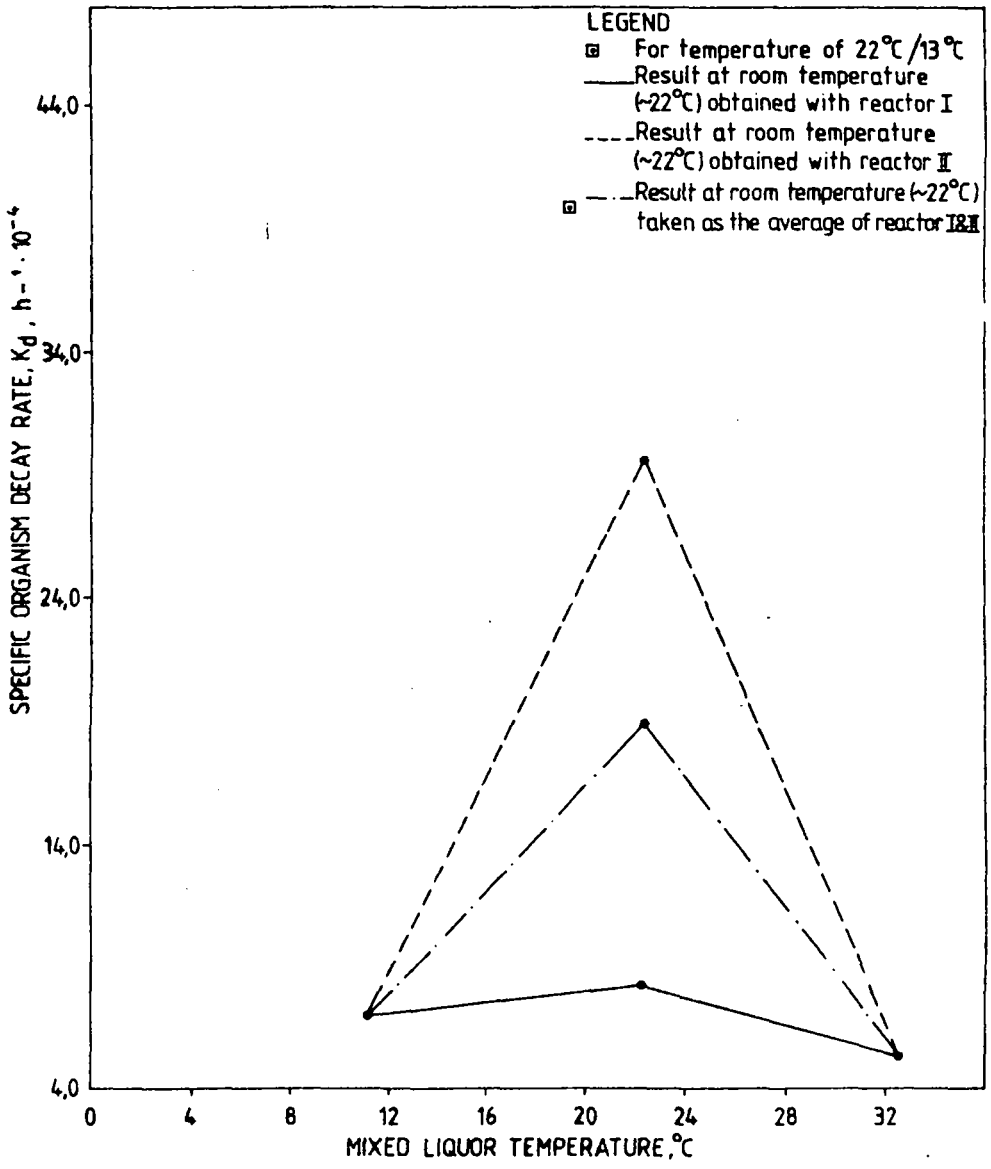


Figure 34. Specific organism decay rate at different temperatures.

6.6 Sludge Production and Characteristics

The exact computation of sludge production is very difficult due to analysis errors in the determination of SS and volatile suspended solids (VSS) concentration and the scale of the experimentation. Since the models employed were of the continuous-flow type, small errors in SS and VSS analysis would be magnified in the final calculations.

The smallness of the scale of the models causes difficulties in accounting for sludge lost or gained in that there are always small amounts of sludge escaping or entering the system which are difficult or even impossible to quantify, i.e. sampels, cleaning of reactors etc.

Solids production (sludge production) for activated sludge systems generally refers to the net difference in the mass of solids formed by anabolism and that destroyed by catabolism /18/. Anabolism is subject to wide rate variations depending on such factors as food supply and population suitability /18/. Catabolism is a low rate process that occurs simultaneously; it is a major fraction of the total respiration rate at low loading /18/.

Many reports on activated sludge operation show that solids accumulation depends on the net difference between growth and decay /18/. A large increase in solids occurs when growth predominates at high loading; a net decrease may occur when decay becomes significant at low loading. Wuhrman /18/ reported that influent oxygen demand, volatile solids percentage of sludge and MCRT are paramount in estimating unit solids gain.

With low MLSS concentration, a given amount of organic material removed from sewage gives rise to a greater sludge accumulation because oxidation is limited by the disparity between the food and the number of organisms and limited

oxidation would result in a relatively greater amount of biological growth /14/. In other words, the relatively fewer organisms present under this condition multiply to a greater degree to carry on the limited oxidation and hence produce more growth and the sludge formed is higher in protoplasmic growth /14/. When a higher MLSS concentration is maintained there is not so much food per unit of organism. As a result, the oxidation is carried on further, giving a smaller sludge accumulation, and the protoplasmic growth is relatively less because there are more organisms per unit amount of food and hence less multiplication per organism /14/. In the former case, the sludge produced is not so fully oxidized and contains proportionally a greater amount of protoplasmic growth; it may be considered as an overgrowth of organisms as a result of disparity between organisms and food /14/. The volatile solids part of the MLSS, should, therefore, decrease with increasing MLSS concentrations, as is confirmed by the relationship shown in figure 35.

There is a small (at 22,2°C very big) but significant decrease in the MLVSS/MLSS ratio with increasing MLSS concentration in all the phases except at temperature of 22,4°C, where again the data available are not adequate. The fact that the MLVSS concentrations at the different temperatures of the experiment are not the same is not so pertinent as the general decrease with increasing MLSS concentration /14/. Heukelekian et al /14/ experimenting with laboratory-scale activated sludge units performing on a batch basis (at room temperature) have observed the same trend.

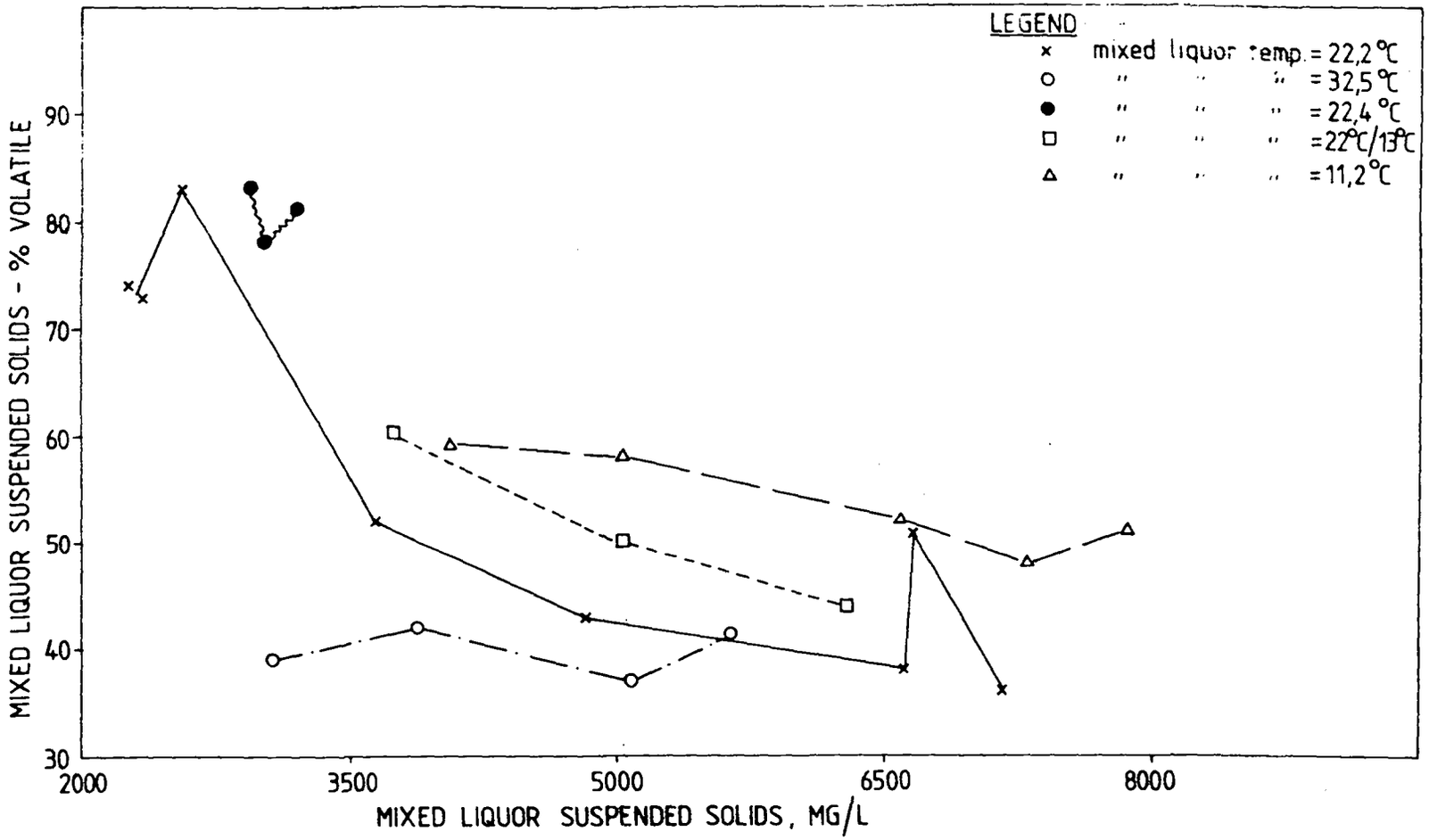


Figure 35. Relationship between the percentage volatile solids in mixed liquor and mixed liquor suspended solids concentration at different mixed liquor temperatures.

Oxygen uptake rate (OUR) or respiration rate is a simple and reliable indicator for the degree of stabilization of sludge /31/. Oxygen uptake rates have been shown to be independent of DO concentrations and directly related to nitrification /8/. Stenström and Andrews /1/, by mathematical modeling and computer simulation established the effectiveness of utilizing a specific OUR as an indicator of process performance.

Endogenous OUR (endogenous respiration rate) tests were conducted at the end of each phase (except at 22,4°C) of the experiment to determine the degree of stabilization of the sludges and to compare activity of the micro-organisms.

It is seen from table 19 that the endogenous respiration rate ranged from a minimum of 5,530 mg O_2/h/g MLVSS at 22°C/13° C to 14,700 mg O_2/h/g MLVSS at 32,5°C. Figure 36 shows that the endogenous respiration rate increases with increasing temperature with the rate of increment being faster from 22,2°C to 32,5°C than from 11,2°C to 22,2°C. Although the interpretation of oxygen uptake results is difficult /24/, the low endogenous respiration rate at 22°/13°C supports the earlier observation of very high BOD removal, 99 percent, (see section 6.1) at this same temperature. At 20°C an aerobically well-stabilized sludge has a respiration rate of 2-4 mg O_2/h/g MLVSS /31/ compared to 8,354 mg O_2/h/g MLVSS at 22,2°C found in this experiment.

Table 19. Endogenous respiration rate of micro-organisms - test results (oxygen uptake rate of activated sludge, OUR)

Phase No. of experiment	Date of OUR test	Reactor No.	Temp of MLSS	Portion of OUR line*	O U R	Average OUR	Final average OUR
			°C		$\frac{mgO_2}{h \cdot g \cdot MLVSS}$	$\frac{mg\ O_2}{h \cdot g \cdot MLVSS}$	$\frac{mg\ O_2}{h \cdot g \cdot MLVSS}$
I&II	24/11/83	I	22,2	I II III IV	10,979 9,130 11,793 6,160	9,516	
I & II	27/11/83	I	22,2	I II	7,660 6,721	7,191	
							8,354
III	4/1/84	I	32,5	I II III	17,993 13,760 12,348	14,700	
							14,700
II	24/11/83	II	22/13	I II III IV	5,467 6,423 5,161 6,896	5,987	
II	27/11/83	II	22/13	I II III IV V	8,225 5,086 4,530 3,778 3,739	5,072	
							5,530
III	3/1/84	II	11,2	I II III IV V	7,593 8,414 7,151 6,670 8,266	7,619	
							7,619

* The OUR lines drawn by the recorder on charts were divided into portions having the same slope.

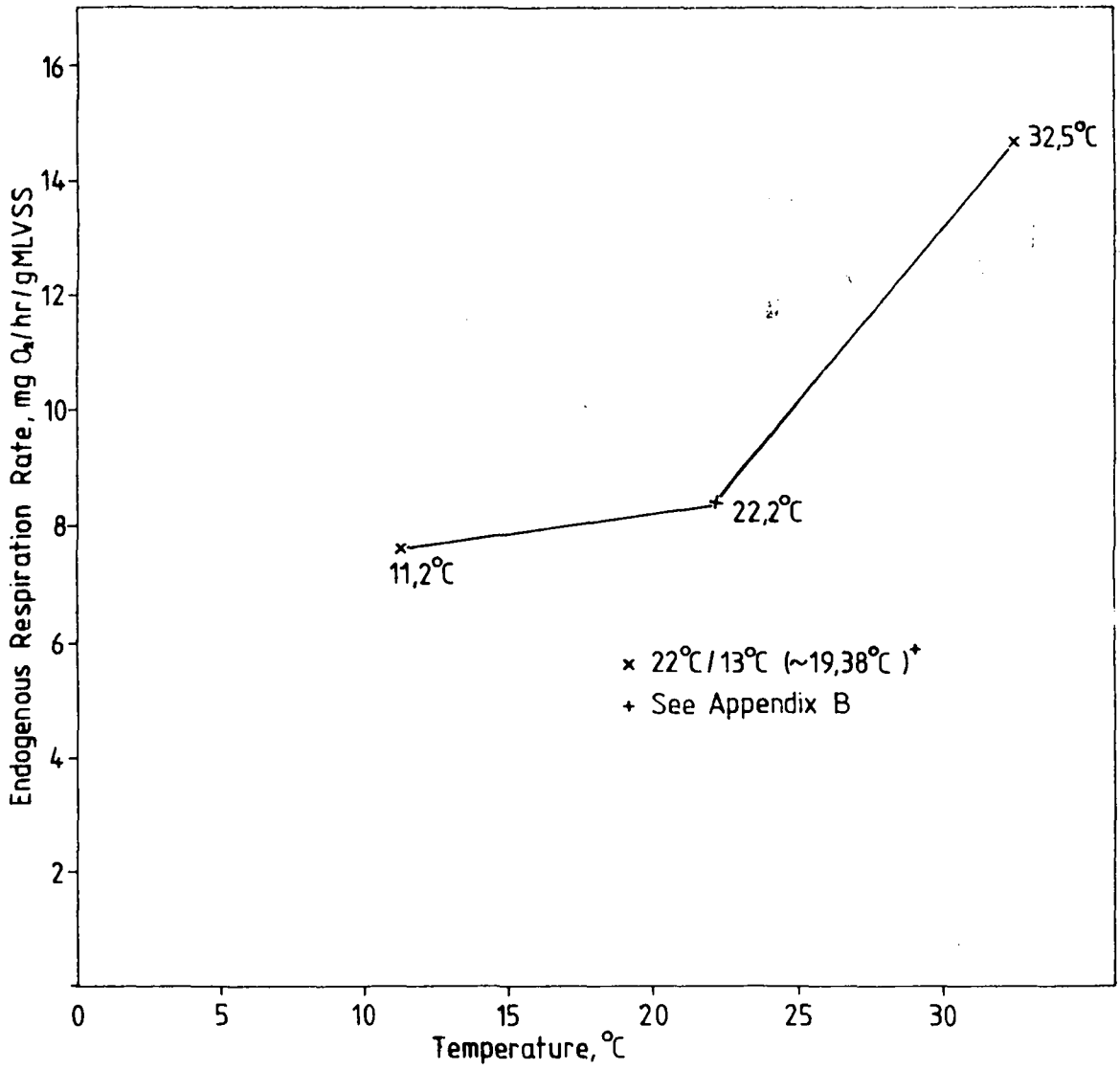


Figure 36. Endogenous respiration rate of sludge.

7. CONCLUSIONS

The following conclusions are based on the results of the research herein:

1. Temperature effects on BOD_7 removal are not discernable in very low-loaded activated sludge systems (mean F/M ratio of 0,04 - 0,09 kg BOD_7/kg MLSS·d).

2. COD removal seems to be highest at mixed liquor temperature of 22°C/13°C, followed by removal rate at 11,2°C. Removal rates at 22,2°C and 32,5°C seem to be lower than that at 22°C/13°C or 11,2°C.

3. Nitrification increases with increasing temperature. The rate of nitrification increases more from 11,2°C to 22,2°C (or 22,4°C) than from 22,2°C (or 22,4°C) to 32,5°C.

4. Suspended solids (SS) removal rates at different mixed liquor temperatures were difficult to evaluate due to the erratic results obtained mainly due to the configuration of the clarifiers in the models used. Generally, it can be postulated that SS removal rate is higher at room temperature (22,2° and 22,4°C) than at 11,2°C.

5. For mixed liquor temperatures of 11,2°, 22°/13° and 32,5°C and F/M ratios of 0,01 - 0,08 kg BOD_7/kg MLSS·d, the sludge volume indices are below 100 ml/g. At room temperature (22,2° and 22,4°C) and F/M ratios of 0,02-0,12 kg BOD_7/kg MLSS·d, the SVI values ranged from 38 - 284 mg/l. It seems that in a low-loaded system, SVI values are affected more by F/M ratios than by temperature for the above range of temperature.

6. Cell growth yield coefficient is a maximum at around 22,4°C with values increasing from 11,2° to 22,4°C and decreasing from 22.4° to 32,5°C. The yield calculated at mixed liquor temperature of 22°C/13°C is greater than that calculated at 32,5°C.

7. Specific organism decay rate is a maximum at room temperature, i.e. 22,2° and 22,4°C. It increases from 11,2°C to room temperature and decreases from room temperature to 32,5°C. The data obtained at mixed liquor temperature of 22°C/13°C show the highest decay rate of 0,004 h^{-1}.

REFERENCES

1. Alleman, James E., and Sherrard, Joseph H.,
Activated Sludge
Journal Water Pollution Control Federation
vol. 52, No. 6, 1069-1854,
1980

2. Arceivala, S.J.,
Wastewater Treatment and Disposal
Marcel Dekker, Inc., New York, USA
1981

3. Benedict, Arthur H., and Carlson, Dale A.,
Rational Assessment of the Streeter-Phelps Temperature Coefficient
Journal Water Pollution Control Federation
vol. 50, No. 4, 678-687,
1974

4. Benedict, Arthur H., and Carlson, Dale A.,
Temperature Acclimation in Aerobic Bio-Oxidation Systems
Journal Water Pollution Control Federation
vol. 45, No. 1, 11-24
1973

5. Chao, Allen C., and Keinath, Thomas M.,
Influence of Process Loading Intensity on Sludge Clarification and Thickening Characteristics
Journal Water Research,
vol. 13, No. 12, 1213-1223,
1979

6. Collins, Caeli E., and Incropera, F.P.,
The Effect of Temperature Control on Biological Wastewater Treatment Process
Journal Water Research,
vol. 12, No. 8, 547-554
1978

7. Design of Wastewater Treatment Plants,
A Manual of Practice, No. 36
Water Pollution Control Federation,
Lancaster Press Inc., New York,
1977

8. Edwards, Gary L., and Sherrard, Joseph H.,
Measurement and Validity of Oxygen Uptake as an Activated Sludge Process Control Parameter,
Journal Water Pollution Control Federation,
vol. 54, No. 12, 1546-1552,
1982

9. Farquhar, G.J., Boyle, W.C.,
Control of Thiothrix in Activated Sludge,
Journal Water Pollution Control Federation,
vol. 44, No. 1, 14-23,
1972

10. Friedman, A.A., and Schroeder, E.D.,
Temperature Effects on Growth and Yield of Activated Sludge
Journal Water Pollution Control Federation,
vol. 44, No. 7, 1433-1442,
1972

11. Goronszy, Mervyn C.,
Intermittent Operation of the Extended Aeration Process for Small Systems,
Journal Water Pollution Control Federation,
vol. 51, No. 2, 274-286,
1979

12. Grady, Jr., C.P. Leslie, and Lim, Henry C.,
Biological Wastewater Treatment,
Marcel Dekker, Inc.,
New York and Basel, USA,
1980

13. Heide, B.A.,
Combined Nitrogen and Phosphorus Removal in a Low-Loaded Activated Sludge System Operating on Oxidation Ditch Principles,
Journal Water Research,
vol. 8, No. 4/5, 615-622,
1977

14. Heukelekian, H., Orford, H.E., and Manganelli, R.,
Factors Affecting the Quantity of Sludge Production in the Activated Sludge Process,
Journal Sewage and Industrial Wastes,
vol. 23, No. 8, 945-957,
1951

15. Hiidenheimo, Heikki,
Investigation of the Biological Treatment of Pulp Mill Effluents by the Use of Extended Aeration,
Soil and Hydrotechnical Investigations,
No. 13, Helsinki, Finland,
1969

16. Jenkins, David and Garrison, Walter E.,
Control of Activated Sludge by Mean Cell Residence Time,
Journal Water Pollution Control Federation,
vol. 40, No. 11, 1831-1980,
1968

17. Lee, E. G-H., Mueller, J.C., and Walden, C.C.,
Effect of Temperature and Sludge Loading on BOD_5 Removal and Sludge Settleability in Activated Sludge Systems Treating Bleached Kraft Effluents,
TAPPI, vol. 58, No. 6, 100-103,
1975

18. Ludzack, F.J., Schaffer, R.B., and Ettinger, M.B.,
Temperature and Feed as Variables in Activated Sludge Performance,
Journal Water Pollution Control Federation,
vol. 33, No. 2, 141-156,
1961

19. Magara, Yasumoto, Nambu, Shoichi, and Utosawa, Kohken,
Biochemical and Physical Properties of Activated Sludge on Settling Characteristics,
Journal Water Research,
vol. 10, No. 1, 71-77,
1976

20. Mäkelä, Markku,
Water Quality and Unit Operations,
Wastewater Treatment I, II, and III
Lecture notes, Tampere University of Technology
1983

21. Mandt, Mikkel G., and Bell, Bruce A.,
Oxidation Ditches in Wastewater Treatment,
Ann Arbor Science Publishers, USA,
1982

22. Mara, Duncan,
Sewage Treatment in Hot Climates,
John Wiley and Sons Ltd.,
Great Britain, 1978

23. McCarty, Perry L., and Brodersen, C.F.,
Theory of Extended Aeration Activated Sludge,
Journal Water Pollution Control Federation,
vol. 34, No. 11, 1095-1103,
1962

24. McKinney, Ross E., and O'Brien, Walter J.,
Activated Sludge - Basic Design Concepts,
Journal Water Pollution Control Federation,
vol. 40, No. 11, 1831-1980,
1968

25. Metcalf and Eddy, Inc,
Wastewater Engineering Treatment/Disposal/Reuse,
2nd edn.,
McGraw-Hill Book Company, USA,
1979

26. Middlebrooks, E.J., and Garland, C.F.,
Kinetics of Model and Field Extended Aeration
Wastewater Treatment Units,
Journal Water Pollution Control Federation,
vol. 40, No. 4, 586-612,
1968

27. Mukherjee, S.K., Chatterji, A.K., and Saraswat, I.P.,
Effect of pH on the Rate of BOD of Wastewater,
Journal Water Pollution Control Federation,
vol. 40, No. 11, 1831-1980,
1968

28. Mulready, C.R., Payne, D.C., and Watkins, D.W.,
A Comparison of the Carrousel and Pasveer Ditch
Activated-Sludge Plant,
Journal Water Pollution Control, Great Britain,
vol. 81, 297-305,
1982

29. Novak, John T.,
Temperature-Substrate Interactions in Biological
Treatment,
Journal Water Pollution Control Federation,
vol. 46, No. 8, 1984-1994,
1974

30. Operation of Wastewater Treatment Plants,
A Manual of Practice, No. 11
Water Pollution Control Federation,
Lancaster Press, Lancaster, Pa, USA,
1976

31. Oxidation Ditch Technology,
International Conference at Amsterdam, October 1982
CEP Consultants Ltd., Edinburgh, U.K.,
1982

32. Palm, Jonathan C., Jenkins, David, and Parker, Denny,S.,
Relationship Between Organic Loading, Dissolved
Oxygen Concentration and Sludge Settleability in the
Completely-Mixed Activated Sludge Process,
Journal Water Pollution Control Federation,
vol. 52, No. 10, 2417-2616,
1980

33. Parker, Homer W.,
Wastewater Systems Engineering,
Prentice-Hall, Inc., Englewood Cliffs,
New Jersey, USA,
1975

34. Pickford, John, and Anakwa, A.O.,
Some Effects of Temperature on BOD Removal and
Nitrification in the Activated Sludge Process,
Environmental Pollution Management,
vol. 14, No. 2, 74-78
1974

35. Randall, C.W., Benefield, L.D., and Buth, D.,
The Effects of Temperature on the Biochemical Reaction
Rates of the Activated Sludge Process,
Journal Water Science and Technology,
vol. 14, 413-430,
1982

36. Rich, Linvil G.,
Low-Maintenance, Mechanically Simple Wastewater
Treatment Systems,
McGraw-Hill Book Company, USA,
1980

37. Sayigh, Basim Aziz, and Molina, Jr., Joseph F.,
Temperature Effects on the Activated Sludge Process
Journal Water Pollution Control Federation,
vol. 50, No. 4, 678-687,
1978

38. Sezgin, Mesut,
Variation of Sludge Volume Index With Activated
Sludge Characteristics,
Journal Water Research,
vol. 16, No. 1, 83-88,
1982

39. Sherrard, James H.,
Activated Sludge,
Journal Water Pollution Control Federation,
vol. 52, No. 6, 1069-1854,
1980

40. Stenstrom, Michael K., and Poduska, Richard A.,
The Effect of Dissolved Oxygen Concentration on
Nitrification,
Journal Water Research,
vol. 14, No. 6, 643-649,
1980

41. Washington, Donald R., and Hetling, Leo J.,
Volatile Sludge Accumulation in Activated Sludge
Plants,
Journal Water Pollution Control Federation,
vol. 37, No. 4, 499-507,
1965

42. Yang, P.Y., and Chen, Y.K.,
Operational Characteristics and Biological Kinetics
Constants of Extended Aeration Process,
Journal Water Pollution Control Federation,
4, vol. 49, 525-724,
1977

Table A1. Weekly mean values of parameters - reactor I (temp = 22,2° and 32,5°C)

Date	Weeks No.	x_1 mg/l	x_2 mg/l	Q l/d	F/M $\frac{kg\ BOD_7}{kgMLVSS \cdot d}$	K_d $(h)^{-1}$	x_o mg/l	(S_o-S_1) mg/l	WAS+ mg
8/10 - 15/10/83	1,143	2206,0	29,38	6,195	0,0929	0,00082	118,875	304,375	220,60
16/10 - 21/10/83	0,857	1664,0	14,67	7,58	0,1449	"	97,667	320,833	166,40
22/10 - 29/10/83	1,143	1669,3	21,00	6,360	0,1397	"	92,125	281,5	166,93
30/10 - 5/11/83	1	2000,0	9,43	6,331	0,1048	"	65,571	242,0	200,00
6/11 - 12/11/83	1	1945,0	19,86	5,878	0,1094	"	59,286	268,286	243,125
13/11 - 19/11/83	1	2422,0	24,86	5,35	0,0933	"	74,286	316,857	544,95
20/11 - 26/11/83	1	2487,0	17,14	4,941	0,079	"	83,000	296,286	310,875
27/11 - 3/12/83	1	3218,7	18,57	4,659	0,0647	0,00053[a]	75,714	252,0	321,87
4/12 - 10/12/83	1	2217,0	14,29	4,122	0,0542	0,00053	53,429	203,857	277,125
11/12 - 17/12/83	1	2291,0	23,43	5,096	0,0656	"	77,143	203,429	286,375
18/12 - 24/12/83	1	1958,0	14,14	3,710	0,0731	"	63,857	305,143	293,70
25/12 - 31/12/83	1	1252,0	18,29	6,072	0,1898	"	58,000	309,143	219,1

[a] The period 27/11 - 3/12/83 falls within both phase II and III of the experiment run with reactor I. Phase III of the experiment was started on 29/11/83, only 2 days after the period (week) 27/11/ - 3/12/83 started. Therefore the K_d value of 0,00053 h^{-1}, which actually was calculated for phase III is also assumed for this period (27/11/ - 3/12/83).

+ WAS = amount of sludge wasted for analysis purposes within the given period.
= no. of MLVSS analysis • 0,025 l • mean MLVSS concentration within the period.

CALCULATION OF THEORETICAL MCRT

Earlier in secion 3.3, the mass balance for micro-organisms was given by equation 3.11, i.e.

$$V(dX_1)_{net} = X_o Qdt + (dX_1)_g V - X_2 Qdt - K_d X_1 Vdt$$

the term on the left handside is the difference in the MLVSS concentrations (after mixing clarifier and aeration chamber contents) in a week's interval in the experiment. The term $(dX_1)_g V$ gives the mass of micro-organisms produced in the reactor. In section 3.2.3, the cell yield coefficient (Y) was given by equation 3.7 as:

$$Y = \frac{\text{weight of organisms produced}}{\text{weight pf substrate utilized}}$$

But,

$$\text{Weight of organisms produced} = (dX_1)_g V$$

where,

$$(dX_1)_g V^* = V(dX_1)_{net} + X_2 Qdt + K_d X_1 Vdt - X_o Qdt$$

Therefore,

$$Y = \frac{V(dX_1)_{net} + X_2 Qdt + K_d X_1 Vdt - X_o Qdt}{\text{weight of substrate utilized}}$$

Mean cell residence time, MCRT (θ_c) could be calculated from,

$$\frac{1}{\theta_c} = Y\,(F/M) - K_d \qquad /2/$$

In the above equation; Y is calculated from equation 3.7, F/M ratio is taken as the weekly mean and K_d obtained from the regression equation (equation 3.21) figures plotted earlier in section 6,5.

* activated sludge used for analysis purposes and therefore wasted, and intentionally wasted sludge should be added here (although not mentioned in the equation).

Table A2. Summary of theoretical MCRT (θ_c) calculations, reactor I (temp = 22,2 and 32,5°C)

Date	dt	$V^b(dX_1)_{net}$	X_2Qdt	$K_dX_1V^b dt \cdot 24$	X_oQdt	$(dX_1)_gV^b$	$(S_o-S_1)Q \cdot dt$	$Y=\frac{(dX_1)_gV}{(S_o-S_1)Qdt}$	$\frac{1}{\theta_c}=Y(F/m)-K_d$	θ_c
	d	mg	mg	mg	mg	mg	mg	—	d^{-1}	d
8/10 - 15/10/83	8	- 651,2	1456,073	2570,114	5891,445	-2295,858	15084,825	0,1522	0,01332	75,08
16/10 - 21/10/83	6	4765,0	667,192	1453,99	4441,895	2611,287	14591,485	0,17896	0,02511	39,82
22/10 - 29/10/83	8	-3167,2	1068,48	1944,828	4687,32	-4674,282	14322,720	0,32635	0,04477	22,34
30/10 - 5/11/83	7	1450,4	417,909	2038,848	2905,910	1201,247	10724,714	0,11201	0,01092	91,58
6/11 - 12/11/83	7	1065,6	817,16	1982,78	2439,382	1669,283	11038,896	0,15122	0,01572	63,61
13/11 - 19/11/83	7	680,8	931,007	2469,045	2782,011	1843,791	11866,295	0,15538	0,01368	73,1
20/11 - 26/11/83	7	4144	593,859	2535,307	2870,721	4713,32	10247,644	0,45994	0,03552	28,15
27/11 - 3/12/83	7	5712,8	603,623	2120,789	2469,261	6774,626[c]	8218,476	0,82432[d]	0,05280	18,94
4/12 - 10/12/83	7	-8909,6	412,324	1460,772	1541,640	-8301,019	5862.09	1,41124	0,05367	18,63
11/12 - 17/12/83	7	-1480	835,795	1509,531	2751,845	1027,394[e]	7256,719	0,14158	0,00876	114,16
18/12 - 24/12/83	7	-42992	367,216	1290,118	1658,366	- 572,832[e]	7924,564	0,07229	0,00475	210,53
25/12 - 31/12/83	7	2545,6	777,398	824,938	2465,232	4092,804[e]	13139,814	0,31148	0,05859	17,07

[b] V = volume of reactor = 7.4 l

[c] 150 ml of wasted sludge added i.e.; 0,150 l • 3218,7 mg/l = 482,805 mg added.

[d] Cell yield coefficient, Y, taken as = 1,0, since $Y \leq 1$ if bacterial growth depends on influent BOD only, which was the case in the experiment conducted.

[e] 250 ml of removed (wasted) sludge each day starting from 16/12/83 is added here,
i.e. for period 11/12 - 17/12/83; 0,25 l•2 days • 2291 mg/l = 1145,5 mg added
" " 18/12 - 24/12/83; 0,25 l•7 days • 1958 mg/l = 3426,5 mg added
" " 25/12 - 31/12/83; 0,25 l•7 days • 1252 mg/l = 2191,0 mg added

Table A3. Weekly mean values of parameters - reactor II (temp = 22, ° 22°/13° and 11,2 °C)

Date	Weeks No.	x_1 mg/l	x_2 mg/l	Q l/d	F/M $\frac{kg\ BOD_7}{kgMLVSS \cdot d}$	K_d $(h)^{-1}$	x_o mg/l	(S_o-S_1) mg/l	WAS+ mg
8/10 - 15/10/83	1.143	2496,0	29,88	6,433	0,0855	0,00296	118,875	304,5	249,60
16/10 - 21/10/83	0,857	2294,7	12,83	7,73	0,1248	0,00296	97,667	323,667	229,47
22/10 - 29/10/83	1,143	2530,7	32,25	6,544	0,0939	0,00348[f]	92,125	282,25	253,07
30/10 - 5/11/83	1	2348,0	13,14	5,288	0,0788	0,004	65,571	241,286	234,80
6/11 - 12/11/83	1	2426,0	23,29	5,824	0,0903	0,004	59,286	268,714	303,25
13/11 - 19/11/83	1	2662,0	17,29	5,554	0,0903	0,004	74,286	317,143	598,95
20/11 - 26/11/83	1	2840,0	9,14	4,722	0,0658	0,004	83,000	296,857	355,00
27/11 - 3/12/83	1	3786,7	18,86	3,891	0,0499	0,007	75,714	255,857	378,67
4/12 - 10/12/83	1	3573,0	17,43	4,316	0,035	0,007	53,429	205,714	446,63
11/12 - 17/12/83	1	3494,0	110,86	3,697	0,0304	0,007	77,143	206,143	436,75
18/12 - 24/12/83	1	3240,0	11,71	3,697	0,0498	0,007	63,857	306,429	486,00
25/12 - 31/12/83	1	2363,3	16,14	6,199	0,1025	0,007	58,000	309,143	413,58

f The period 22/10 - 29/10/83 falls between phases I and II with K_d values of 0,00296 h^{-1} and 0,004 h^{-1} respectively. Therefore for the above period the mean of the two values was taken.

Table A4. Summary of theoretical MCRT (θ_c) calculations, reactor I (temp = 22,4°, 22°/13° and 11,2°C)

Date	dt	$V^b(dX_1)_{net}$	X_2Qdt	$K_{d-1}X_1V^b dt \cdot 24$	X_oQdt	$(dX_1)_g V^b$	$(S_o-S_1)Q\,dt$	$Y=\frac{(dX_1)_g V}{(S_o-S_1)Qdt}$	$\frac{1}{c}=Y(F/M)-K_d$	θ_c
	d	mg	mg	mg	mg	mg	mg		d^{-1}	d
8/10 - 15/10/83	8	-444,0	1537,744	10497,098	6117,783	5722,659	15670,788	0,36518	0,02826	35,39
16/10 - 21/10/83	6	352,0	595,055	7237,888	4529,795	3884,618	15011,575	0,25877	0,02933	34,09
22/10 - 29/10/83	8	172,0	1688,352	12512,753	4822,928	9803,247	14776,352	0,66344	0,05882	17,00
30/10 - 5/11/83	7	-456,0	486,390	11676,134	2427,176	9514,148	8931,443	1,06524[d]	0,0748	13,37
6/11 - 12/11/83	7	204,0	949,487	12064,013	2416,972	11103,778	10954,932	1.01359[d]	0,0863	11,59
13/11 - 19/11/83	7	-32,0	672,201	13237,594	2888,091	11588,654	12329,886	0,93988[d]	0,08087	12,37
20/11 - 26/11/83	7	608,0	302,114	14122,752	2743,482	12644,384	9812,311	1,28862[d]	0,0618	16,18
27/11 - 3/12/83	7	1200,0	513,69	3295,338	2062,222	3325,476[h]	6968,777	0,4772	0,02311	43,27
4/12 - 10/12/83	7	-952,0	526,595	3109,368	1614,197	1980,886[g]	6215,031	0,31873[d]	0,01046	95,60
11/12 - 17/12/83	7	212,0	2868,946	3040,619	1996,384	6308,931[g]	5334,775	1,18261[d]	0,0297	33,67
18/12 - 24/12/83	7	-804,0	303,043	2819,578	1652,555	6822,066[g]	7930,076	0,86028	0,04214	23,73
25/12 - 31/12/83	7	32,0	700,363	2056,638	2516,794	4821,562[g]	13414,642	0,35943	0,03614	27,67

[g] 250 ml of removed (wasted) sludge each day starting from 16/12/83 is added here.
i.e. for period 11/12 - 17/12/83; 0,25 l • 2 days • 3494,0 mg/l = 1747 mg added
" " 18/12 - 24/12/83; 0,25 l • 7 days • 3240,0 mg/l = 5670 mg added
" " 25/12 - 31/12/83; 0,25 l • 7 days • 2363,3 mg/l = 4135,775 mg added

[h] 130 ml of wasted (unintentional) sludge on 10/12/83 is added here.
i.e. 0,13 l • 3573,0 mg/l = 464,49 mg added

APPENDIX B

WEIGHTED AVERAGE TEMPERATURE OF MIXED LIQUOR IN REACTOR II , PHASE II

The reactor was kept at room temperature i.e. mixed liquor mean temperature of 22°C for 17 h and at mean temperature of 13°C (inside fridge) for 7 hours. In order to compare the performance results of this reactor at 22°C/13°C with results at other temperatures, a respresentative temperature value has to be assumed. One such value was assumed to be the weighted average of the two temperature values (22°C and 13°C).

Weighted average of temperature,

$$= \frac{22°C \cdot 17\ h + 13°C \cdot 7\ h}{24\ h}$$

$$= \underline{\underline{19,38}}\ °C$$

WASTEWATER ANALYSIS RESULTS

DATE		INFLUENT											MIXED LIQUOR							EFFLUENT											DATE		REMARK
		SS	VSS	BOD_5 settl	BOD_5 filt	COD settl	COD filt	TKN filt	NH_4-N filt	P_{tot} filtered	pH	Alkal	MLSS	MLVSS	Settled sludge volume (SV)	SVI	pH	TEMP.	Dissolved Oxygen	SS	VSS	BOD_5 filt	BOD_5 (with ATU) filt	COD filt	TKN filt	NH_4-N filt	P_{tot} filtered	pH	Alkal			
WEEK	DAY	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l		m mol/l	mg/l	mg/l	ml/l	ml/g		°C	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l		m mol/l	WEEK	DAY	
		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28			

Format used for tabulating measurement and analysis data.

APPENDIX C

MONTH
REACTOR NO.
TEMPERATURE

Operational Parameters

DATE		INFLUENT FLOW	BOD_7		BOD LOAD	MLSS	MLVSS	F/M	F/M	AERATION T_t	REMARK
WEEK	DAY	(Q)	INFLUENT BOD_7 tot (S_0)	EFFLUENT BOD_7* FILT (S_1)	$\frac{(S_0-S_1)Q}{106}$			$\frac{(S_0-S_1)Q}{MLSS\cdot 7,4l}$	$\frac{(S_0-S_1)Q}{MLVSS\cdot 7,4l}$	(H_t)	
–	–	l/d	mg/l	mg/l	kg/d	mg/l	mg/l	$\frac{kg\ BOD\ 7}{kg\ MLSS\cdot d}$	$\frac{kg/BOD\ 7}{kg\ MLVSS\cdot d}$	h	

* BOD_7 with ATU (to inhibit nitrification)

Tabular form used to record parameters for the computation of F/M ratio and hydraulic detention time along with the computed values.