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MİNERİLİZASYON VE NİTRİFİKASYON İÇİN
YENİ OKSİJEN MODELİ**

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**A NEW OXYGEN MODEL FOR MINERALIZATION
AND NITRIFICATION IN AEROBIC SUSPENDED
CULTURE SYSTEMS**

Ph. D THESIS

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N O T A T I O N L I S T

Symbol	Indice	Dimension	Description
A		m^2	interfacial area for transfer
A_f		m^2	surface area of the sludge floc
a		m^2/m^3	interfacial surface area
a_i		m^2	surface area of inner sphere of the floc
a_f		m^2/m^3	specific surface area of the floc
a_s		m^2	specific surface area of the floc
a_t		m^2	total surface area of the floc
α		dimensionless	alpha correction factor
α^-		dimensionless	alpha bar, correction factor
β		dimensionless	beta factor
β_n		mg/l	empirical concentration constant for transition from mineralization to nitrification
C^*_{α}		mg/l	saturation dissolved oxygen concentration for wastewater and tap water
C_f		kg/m^3	oxygen concentration in the floc
C_L		kg/m^3	oxygen concentration in bulk liquid
C_{Lf}		kg/m^3	oxygen concentration at the surface of the floc-liquid interface

C_L^k		kg/m ³	critical oxygen concentration in bulk liquid in mineralization and nitrification
C_i		kg/m ³	oxygen or substrate concentration around the innermost cell of the aggregate
C_s		kg/m ³	saturation concentration at the surface of the aggregate
C_{sc}		kg/m ³	effective saturation concentration
d		m	diameter of the floc
D	f_{O_2}	m ² /sec	molecular diffusion coefficient of oxygen in the floc
D_s		m ² /sec	diffusion coefficient of substrate in the floc
D_s		m ² /sec	overall diffusion coefficient of diffusing species through the membrane template assembly
D_{sf}		m ² /sec	diffusivity of a given substrate through the floc matrix
D_L		m ² /sec	diffusion coefficient of oxygen in the liquid
D_f		m ² /sec	diffusion coefficient of oxygen in the floc
ε		dimensionless	porosity of floc
F		kg/m ² sec	diffusion flux
F/M	BOD, NH ₄ ⁺ -N	day ⁻¹	food to microorganism ratio for mineralization and nitrification

k_d		day^{-1}	decay or maintenance coefficient
K_s	1,2	day^{-1}	half saturation concentration of substrates
K_L		m/sec	liquid film coefficient
$K_L a$	ww	$1/\text{sec}$	volumetric mass transfer coefficient, for wastewater
k_o		$\text{kg/m}^3 \text{ sec}$	reaction rate constant of oxygen in zero order reaction in the floc
k	$\text{BOD}, \text{NH}_4^+ \text{-N}$	$\text{kg/m}^3 \text{ sec}$	reaction rate constant in zero order, for mineralization and nitrification
k_n		dimensionless	empirical constant for transition from mineralization to nitrification
M_{O_2}		kg/m	overall mass transfer of oxygen
X	MLSS, f	gr/cm^3	solid concentration of the activated sludge and floc
N_o		kg/m^3	influent ammonia nitrogen concentration
N_e		kg/m^3	effluent ammonia nitrogen concentration
P_w		kWh	wire power
R		m	radius of sphere floc
r		m	distance in the sludge floc
r_f		$\text{kg/m}^3 \text{ sec}$	oxygen consumption rate in the sludge floc

r_s		kg/m ³ sec	specific consumption rate of substrate by the cell of the floc
r_{O_2}		kg/m ³ sec	specific consumption rate of oxygen by the cell of the floc
r_L		kg/m ³ sec	oxygen consumption rate in the sludge suspension
R_p		m	mean pore radius
r_{max}		kg/m ³ sec	maximum oxygen uptake rate based on dry weight of the floc
r_{red}		kg/m ³ sec	reduced oxygen uptake rate based on dry weight of the floc
r_{wet}		kg/m ³ sec	maximum oxygen uptake rate based on wet volume of the floc
R_c		m	radius of anoxic core
X_c		m/m	ratio of radius of the anoxic core, to total floc radius
X_{MLSS}		kg/m ³	mixed liquor suspended solid concentration in mineralization and nitrification
S	1,2	kg/m ³	limiting substrate concentration
S_g		m ² /g	surface area per unit mass of porous media
SVI		m ³ /kg	sludge volume index
S_o		kg/m ³	influent BOD concentration
S_e		kg/m ³	effluent BOD concentration
S_h		dimensionless	sheerwood number
μ		day ⁻¹	net specific growth rate

$\mu_{\max}$		day ⁻¹	maximum specific growth rate
θ		day	hydraulic retention time
θ_c		day	sludge age
θ_A		dimensionless	arithmetic temperature correction factor
θ_G		dimensionless	geometric temperature correction factor
ρ_f		g/cm ³	bulk density of the floc
ρ_s		g/cm ³	density of dried sludge
τ		dimensionless	tortuosity factor
ξ	BOD, NH ₄ ⁺ -N	%	percent removal efficiency
ν	N	kg/kg	stoichiometric oxygen consumption ratio in mineralization and nitrification
W		kg	amount of BOD removed
Ψ		kg/kg	oxygen yield factor, amount of oxygen consumed per unit amount of BOD removed
V_i		m ³	volume of inner sphere
V_w		m ³	wet volume of the floc
V_f		m ³	volume of the sludge floc

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SUMMARY

In this study, critical oxygen concentrations, C_L^k for various floc diameters in spherical form in the aeration tank, with respect to the design and operational parameters such as organic loading F/M , removal efficiency ξ , and activated sludge concentration X_{MLSS} in mineralization and nitrification processes is given in the frame of the developed model.

In the first chapter, the basic principles leading to the description of the subject of concern are outlined and the scope and objectives of this study are defined.

In the second chapter, emphasis is placed upon studies in the literature dealing with oxygen requirement and supply for aerobic suspended culture systems. Special attention was directed towards oxygen transfer mechanisms, diffusivity of oxygen through activated sludge and flocs.

In the third chapter, the oxygen diffusion models on microbial aggregates are reviewed in great extend. The main approaches in the literature were to use floc diameter in spherical form for oxygen diffusion models.

In the fourth chapter, A New Oxygen Model for Mineralization and Nitrification in Aerobic Suspended Culture Systems is presented and given respectively as follows:

$$C_L^k = \left[\frac{1}{3D_L} + \frac{1}{6D_f} \right] \frac{d^2}{4} [v(\xi)_{BOD} (F/M)_{BOD}] X_{MLSS}$$

$$C_L^k = \left[\frac{1}{3D_L} + \frac{1}{6D_f} \right] \frac{d^2}{4} \left[\left(\frac{v(S_o - S_e)}{v_N(N_o - N_e)} \right) + 1 \right] X_{MLSS}$$

In the fifth chapter, the graphical representation is given.

In the last chapter, results and discussion are presented.

HAVALI ASKIDA KÜLTÜR SİSTEMLERDE MINERİLİZASYON VE NİTRİFİKASYON İÇİN YENİ OKSİJEN MODELİ

ÖZET

Evsel ve endüstriyel atıksuların mineralizasyon ve nitrifikasyonunda genel olarak havalı askıda askıda kültür sistemleri uygulanır. Havalı biyolojik arıtma tesislerinin ana kısmını havalandırma oluşturmakta ve genelde enerjinin en fazla kullanıldığı ünitelerdir. Mekanik havalandırıcılar veya hava üfleyiciler yardımı ile havalandırma tankı içine verilen hava içindeki oksijen veya doğrudan verilen oksijen organik substratların biyooksidasyonu, hücrelerin içsel solunumu ve amonyak azotunun nitrifikasyonu için kullanılmaktadır.

Havalı biyolojik arıtma proseslerinin tipik konfigürasyonları biri ile diğeri arasında farklılık göstermesine rağmen, her bir prosesin performans değerlendirmesinde prosesin geniş aralıklarda değişen işletme şartları altında enerji kullanımı ile organik madde giderim verimi arasında kantitatif ilişki kurmak gerekir.

Mineralizasyon ve nitrifikasyon için havalı askıda kültür sistemler kullanan biyolojik atıksu arıtma tesisinin dizaynı sırasında, proses seçimi kadar havalandırma sisteminin konfigürasyonunda önemlidir. Havalandırma sisteminin konfigürasyonu için belirleyici ana etkenler oksijen transferinin gerçekleşeceği havalandırma tankı reaktör tipi ve oksijen transferini gerçekleştirecek havalandırıcı ekipmanların fiziksel boyutları ile işletme sırasındaki gerçek verimleridir.

Minerilizasyon için havalı askıda kültür sistemler kullanan konvan siyonel aktif çamur prosesi ile dizayn edilen biyolojik bir atıksu arıtma tesisinin tam karışimli reaktör modelindeki havalandırma tankı, dizayn sırasında uygun boyutlarda gözükmemesine rağmen, tank içindeki sıvı faza oksijen transferi sağlayacak mekanik havalandırıcıların ve/veya tank tabanına döşenecek difüzörlerin toplam sayısı, dizayn edilen su derinliği için yetersiz veya fazla olabilmektedir. Bu durum, dizayn sırasında havalandırıcı ekipmanlarının muhafazakar şekilde dizaynı ile aşılmaya çalışılmakta, ancak tesisin işletilmesi sırasında, öngörülen çıkış suyu kalitesi standartları sağlanmakla birlikte, birim kullanılan enerji başına giderilen organik madde miktarı düşük, bu nedenle tesis ekonomik işletme olmaktan uzaklaşmaktadır.

Minerilizasyon için havalı askıda kültür sistemler kullanan konvansiyonel aktif çamur prosesi ile dizayn edilen bir biyolojik atıksu arıtma tesisinin piston akımlı reaktör modelindeki havalandırma tankı, dizayn sırasında uygun boyutlarda gözükmese bile oksijen transferini sağlayacak ekipmanların fiziksel boyutları ve/veya sayısı, dizayn edilen su derinliğindeki tank yüzey alanı veya tank taban alanına sığmamaktadır. Diğer taraftan, havalandırma tankı girişinde birim mikroorganizma kütlesi başına organik madde yüklemesinin fazla olması nedeniyle, havalandırma tankının bu kısımlarında, sıvı fazdaki çözünmüş oksijen konsantrasyonları hızlı oksijen kullanımı nedeniyle kritik hale gelip havasız ortamların oluşmasına neden olmaktadır. Böyle durumlar, tesisin işletilmesinde kontrol zorlukları getirmesi yanında tesis de verim düşüklüğü, çıkış su kalitesi standartlarının sağlanamaması şeklinde kendini göstermektedir.

Rasyonel olarak dizayn edilmiş bir arıtma tesisinden işletme sırasında meydana gelebilecek dalgalanmalara cevap vermesi beklenir. Bu dalgalanmalar atıksu akımı ve bu akım içindeki kitletici konsantrasyonların da değişimler şeklinde ortaya çıkmaktadır. Atıksu akımındaki dalgalanmalar gerekli savak yapıları veya kontrol ekipmanları ile önlenilebilmekte ancak akım içindeki kirletici konsantrasyonlarındaki artış veya azalmalar doğrudan prosesi ve proses kontrol parametrelerini olumsuz yönde etkilemektedir. Bu etkilerden biri, havalandırma tankına oksijen sağlayan ekipmanların yetersiz kalması veya gereğinden fazla enerji kullanmaları şeklinde sonuçlanmaktadır.

Havalandırma tankı girişinde dizayna esas alınan atıksu kirletici konsantrasyonun yüksek, oksijen transferini sağlayan ekipmanların yetersiz kalarak sıvı faz çözünmüş oksijen konsantrasyonunun kritik değer altına düşmesi halinde ise; sistem çıkışında artılmış su kalitesi öngörülenden yüksek, dolayısıyla giderim verimi düşmekte, buna bağlı etkileşim çöktürme tankında çamur yükselmesi şeklinde ortaya çıkmaktadır. Havalandırma tankı girişinde dizayna esas kirletici konsantrasyonunun düşük, oksijen temininin gerekenden fazla olması halinde ise gereksiz şekilde fazladan enerji kullanılmış olmakta, bu gibi durumlar megawatt/saat mertebesinde enerji kullanan arıtma tesisleri için büyük önem taşımaktadır.

Atıksuların konvansiyonel veya ileri arıtımı için oluşan maliyetlerin önemi; bir tarafta su kalite kriterleri düzenli olarak sağlanırken, diğer tarafta her türlü anlamdaki maliyetlerin azaltılmasına yönelik tedbirlerin alınması için de belirleyici olmasıdır. Günümüzde biyolojik atıksu arıtma proseslerine optimum enerji kullanımı ile ilgili getirilen geniş kavram, birim enerji kullanımına karşı giderilen BOD miktarı olarak tanımlanmaktadır.

Havali Askıda Kùltür Sistemlerde Minerilizasyon ve Nitrifikasyon için Yeni Oksijen Modeli adli bu alıřmada, mineralizasyon ve nitrifikasyon prosesi sırasında organik madde yùklemesi F/M, organik madde giderim verimi ξ , aktif amur konsantrasyonu X_{MLSS} gibi tesisin dizayn ve iřletme parametrelerine baėlı olarak, havalandırma tankı içinde kùresel formdaki deėiřik flok apları için sıvı faz daki kritik özünmùř oksijen konsantrasyonları C_L^k geliřtirilen model erevesinde verilmiřtir.

Birinci bölümde alıřmanın ama ve kapsamı, anlam ve önemi açıklanmıřtır.

İkinci bölümde havali askıda kùltür sistemler için oksijen gereksinimi ve oksijen temini üzerinde literatürde daha önce yapılmıř alıřmalar verilmiřtir. Özellikle, oksijen transfer mekanizması, oksijenin aktif amur ve floklar içinde difüzyonu, mineralizasyon ve nitrifikasyon üzerinde özünmùř oksijenin etkileri ile ayrıca, oksijen temini, oksijen kùtle transferi, karıřtırma ve havalandırma prosesi için oksijen transfer düzeltme faktörleri ve girişimler ayrıntılı olarak verilmiřtir.

Üçüncü bölümde, mikrobiyolojik yapılar için literatürde daha önce yapılmıř difüzyon modelleri tarihi geliřimi içinde incelenmiřtir. Literatürdeki modelleme alıřmalarında yapılan genel kabuller, biyolojik yapıların deėiřik geometrik řekilleri için biyolojik yapının dıř yüzeyinden iç kısmına oksijen ve/veya substrat transferinin konsantrasyon farkı ve difüzyon ile gerekleřtirdiklerini, havali askıda kùltür sistemler ile alıřan aktif amur proseslerinde mikrobiyolojik yapıların genelde kùresel formda floklar olarak ele alındıėı gözlenmiřtir. Model alıřmalarında, difüzyon mekanizması altında oksijenin önce sıvı faz sonra sıvı fazdan flokların içine olduėu, oksijen ve deėiřik sübstratlar için deneysel alıřmalardan elde edilen difüzyon katsayıları arasında farklılıklar bulunduėunu göstermiřtir. İncelenen model alıřmalarından birinde, aktif amurun yoğunluk, porosite, spesifik yüzey alanı, ortalama pore apı ile ilgili yapılmıř deneysel alıřma sonucuna göre aktif amur ile kalınlařtırılmıř aktif amur için aynı diffüsitivite katsayısının verildiėi görölmüřtür.

Dördüncü bölümde, bu alıřmanın konusu olan havali askıda kùltür sistemlerde mineralizasyon ve nitrifikasyon için yeni oksijen modeli ve modelin dayandıėı esaslar verilmiřtir. Modelleme alıřmasında kullanılan tek bir flok'un geometrik gösterimi, yapılan kabuller ve sınır řartları açıklanmıřtır. Havali askıda sistemlerde mineralizasyon ne nitrifikasyon için yeni oksijen modelinde, mineralizasyon ve nitrifikasyon için 0. mertebe reaksiyonda reaksiyon hızları tanımı yapılmıřtır. Flok içinde oksijenin reaksiyon hız sabiti k_o , atıksu kirletici parametrelerinden kimyasal oksijen ihtiyacı (COD) ve biyokimyasal oksijen ihtiyacı (BOD) ve amonyak azotu için ayrı ayrı verilmiřtir. Oksijen transfer mekanizmasında ana rol oynayan flokların biyolojik mikroorganizma kùtlesinden oluřtukları hesaba katılarak,

bunun doğal sonucu oksijen reaksiyon hız sabiti k_0 ile olan doğrudan ilgisi, organik madde yüklemesi F/M , substratın giderim verimi ξ , ve aktif çamur konsantrasyonu X_{MLSS} ile BOD ve azot için oksijenin stokiyometrik kullanım oranlar v terimleri birlikte ele alınarak 0. derece reaksiyon mertebesinde, aktif çamur floklarının oksijen kullanım hızı r_f mineralizasyon için;

$$r_f = [v(\xi)(F/M)]_{BOD} X_{MLSS}$$

nitrifikasyon için,

$$r_f = [[v(\xi)(F/M)]_{BOD} + [v_N(\xi)(F/M)]_{NH_4^+-N}] X_{MLSS}$$

şeklinde verilmiştir.

Havalı askıda kültür sistemlerde mineralizasyon ve nitrifikasyon için biyolojik atıksu arıtma tesisi havalandırma tankı içinde sıvı fazdaki kritik çözülmüş oksijen konsantrasyonları oksijen transferinin difüzyon mekanizması altındaki flok çapları, proses dizayn ve kontrol parametreleri ile olan ilgili model denklemleri ise mineralizasyon için ;

$$C_L^k = \left[\frac{1}{3D_L} + \frac{1}{6D_f} \right] \frac{d^2}{4} [v(\xi)(F/M)]_{BOD} X_{MLSS}$$

nitrifikasyon için,

$$C_L^k = \left[\frac{1}{3D_L} + \frac{1}{6D_f} \right] \frac{d^2}{4} [(v(\xi)(F/M))_{BOD} + (v(\xi)(F/M))_{NH_4^+-N}] X_{MLSS}$$

şeklinde verilmiştir.

Beşinci bölümde, mineralizasyon ve nitrifikasyonda kullanılan model sabitleri literatür çalışmaları dikkate alınarak tablo halinde verilmiştir.

Model içinde kullanılan sabitler; mineralizasyon ve nitrifikasyonda oksijenin stokiyometrik kullanım oranları $v = 0.84 \text{ kgO}_2/\text{kgBOD}$, $v_N = 4.20 \text{ kgO}_2/\text{kgN}$, aktif çamur karışımında oksijenin difüzyon katsayısı $2.1 \times 10^{-9} \text{ m}^2/\text{san}$, flok içinde oksijenin difüzyon katsayısı $0.74 \cdot 10^{-9} \text{ m}^2/\text{san}$ alınmıştır.

Mineralizasyon sırasında biyolojik atıksu arıtma tesisi havalandırma tankı aktif çamur ortamında, dizayn ve işletme parametrelerinin değişik değerlerine karşı hesaplanan kritik çözünmüş oksijen konsantrasyonunun aldığı değerler tablo ve şekillerle verilmiştir. Sistemde mineralizasyondan nitrifikasyona geçiş şartları ile değişik BOD konsantrasyonlarına karşı hesaplanan amonyak azotu değerleri ayrı tablo ve şekil ile verilmiştir.

Bu bölümde son olarak, model çerçevesinde öngörülen mineralizasyondan nitrifikasyona geçiş ile nitrifikasyon prosesinde belirli işletme şartları altında belirli flok çapı için hesaplanan kritik oksijen konsantrasyonunun aldığı değerler konunun önemini vurgulamak için ayrı bir tablo ve şekil ile gösterilmiştir.

Son bölümde, sonuçlar ve tartışma sunulmaktadır.

CHAPTER 1. INTRODUCTION

1.1 Aim and Scope

In order obtain high quality effluent from biological wastewater treatment plants, considerable amount of energy is required. Although the configurations of typical modern aerobic treatment processes differ from one to the other, the necessary supply of oxygen to the microorganisms, which are decomposing and assimilating organic matter in the wastewater, is the main cause of energy consumption.

Recently, owing to the increased need for energy conservation, a method to decrease requisite energy, while maintaining a sufficiently high quality of effluent wastewater from the treatment plant, is of great concern. The correct selection of the process and rational design of the configuration and the dimensions of the plant, necessary to meet the quantity and quality of the raw wastewater, are essential in order to realize the optimal practices which minimize the energy required.

In order to assess the performance of each process, information is required about the quantitative relationships between performance of organic removal and the energy consumption under widely varied operating conditions of the process. For these reasons, the application of mathematical models is essential to depict the characteristics of the process as a function of operating parameters.

1.2 Importancy and Meaningful

The importance of costs incurred for the conventional or advanced treatment of wastewater fully justifies that a particular attention be given to any means of reducing the impact of these costs, while properly meeting the water quality requirements.

Costs savings in wastewater treatment plant can be achieved mainly in two ways:

- (1) by optimizing the design of the treatment plant
- (2) by optimizing the operation of the plant

Ideally, the design phase should account for the expected operation characteristics of the plant during its lifespan, whenever the latter are predictable: this can yield additional savings as recent studies indicate. Also it is meaningful to optimize the operation of a previously designed plant. This in turn can be effected in two manners :

- (1) by looking at the seasonal or monthly variations of the plant influent and the characteristics of the receiving water body
- (2) by concentrating on the daily or hourly variations and attempting to cope with these through automatic control

In the first case a mathematical model based on cost minimization will indicate how to readjust the operational variables, while in the second case cost savings will result from actions whose main purpose is the minimization of shock loads, plant failures or energy consumption.

The concept of power economy was first introduced for oxygen supply in the activated sludge wastewater treatment process and was defined as the amount of oxygen dissolved into the water in the aeration tank per unit power consumption. It has been used as an indicator in evaluating the performance of an air-liquid contacting facility. Today one of the extended concept of power economy to wastewater treatment process is defined as the amount of BOD removed per unit power consumption.

In the activated sludge process, the amount of oxygen consumed can not be readily determined, while the amount of BOD removed can be directly observed. By using the power economy concept based on the amount of BOD removed, the performance of different biological wastewater treatment processes, with respect to their power consumption, can be easily compared.

Under steady state operating conditions with constant inflow rate of wastewater Q , influent BOD concentration S_0 , effluent BOD concentration S_e , and fraction of BOD removal ξ_{BOD} , the amount of BOD removed W in a wastewater treatment unit is given as

$$W = Q (S_0 - S_e) = Q S_0 \xi_{\text{BOD}} \quad (1.1)$$

if the power required is denoted by P_w , the power economy is evaluated by

$$W / P_w = Q S_0 \xi_{\text{BOD}} / P_w \quad (1.2)$$

In cases, where there exists daily fluctuations in Q and S_0 the daily average power economy is given by

$$W / P_w = \frac{\int_0^{24} Q (S_0 - S_e) d\theta}{\int_0^{24} P_w d\theta} = \frac{\int_0^{24} Q S_0 \xi_{\text{BOD}} d\theta}{\int_0^{24} P_w d\theta} \quad (1.3)$$

If the oxygen consumption rate is evaluated in the wastewater treatment unit, the power economy can also be evaluated by

$$W/P_w = Q_{\text{Oxygen}} / \Psi P_w \quad (1.4)$$

where Ψ is the overall oxygen yield factor, which is defined as the amount of oxygen consumed per unit amount of BOD removed.

In aerobic biological wastewater treatment processes, oxygen is consumed by the bio-oxidation of organic substrate, the endogenous respiration of cells, and the nitrification of ammonia. Oxygen transfer in aeration tank in activated sludge process energy is required mainly to supply oxygen to the aeration tanks. An increase in aeration efficiency is essential to save the cost of energy necessary to remove the BOD of wastewater in both mineralization and nitrification.

In activated sludge aeration tanks, equipped with air diffusers as aeration devices, the air diffusers are located along the side walls near the bottom of the tank. Air bubbles generated from the diffusers exist in only a part of the tank, where oxygen is transferred from the air bubbles into the sludge floc. The oxygen transfers into the sludge floc via bulk liquid for various floc diameters under the operational conditions of the suspended culture system in aerobic treatment process was seemed to be widely studied in large perspective for the definition of meaningful critical oxygen concentration in mineralization and nitrification which is a subject of this study.

CHAPTER 2. LITERATURE REVIEW

2.1 Oxygen Requirement in Aerobic Suspended Culture Systems

Aeration is an essential part of almost all wastewater treatment systems and usually the major energy consuming process. The reported beneficial effects of high dissolved oxygen concentration in aeration tanks are often explained by the possible formation of dissolved oxygen deficits in the center of the floc. Thus an assessment of diffusion resistance of oxygen within a floc has become closely related with the explanations of the role of dissolved oxygen in the activated sludge process. Before any diffusion resistance can be calculated, data on the diffusion of oxygen in activated sludge must be obtained. This requires an understanding of oxygen transfer mechanisms within activated sludge flocs, and diffusion in heterogeneous solid is closely related.

Several researchers have attempted to measure diffusion coefficient in activated sludge or trickling filter slimes. Pasveer (1954) suggested that diffusion of oxygen in the floc is important in the performance of activated sludge especially at low levels of turbulence when the floc size is large. In his mathematical analysis of Fickian diffusion he assumes a value of diffusivity equal to oxygen in pure water $2.1 \cdot 10^{-9} \text{ m}^2/\text{sec}$.

Wuhrmann (1960) has calculated the oxygen concentration gradient in activated sludge flocs based on assumed value of the diffusivity of oxygen in the floc of $5 \cdot 10^{-10} \text{ m}^2/\text{sec}$ that was one-quarter of the diffusivity in pure water.

Tomlinson and Snaddon (1966) have calculated the diffusivity of oxygen in a bacterial slime grown on the inner surface of a rotating

tube that was partially submerged in sewage. The critical depth at which nitrification was stopped by the thickness of slime was measured experimentally by allowing a bacterial film to grow over a performed layer of nitrifying bacteria and using the extension of ammonia oxidation or nitrite production as a measure of virtual absence of dissolved oxygen. The rate of oxygen uptake was calculated from growth rate measurements. The results gave diffusion coefficients in the range of $1.18 \cdot 10^{-9}$ to $1.72 \cdot 10^{-9} \text{ m}^2/\text{sec}$.

Mueller *et al.*, (1967) have studied oxygen transfer through a pure culture of *Zooglea ramigera* I-16 M. At 30°C they reported an average value for diffusivity of $2.1 \cdot 10^{-10} \text{ m}^2/\text{sec}$ with $\text{SD} = 0.95 \cdot 10^{-10} \text{ m}^2/\text{sec}$. At 20°C , the average value $1.84 \cdot 10^{-10} \text{ m}^2/\text{sec}$ with $\text{SD} = 0.73 \cdot 10^{-10} \text{ m}^2/\text{sec}$. However, the technique based on a static floc sizing technique, i.e., the flocs were stationary on a microscope slide and the appearance of the flocs may bear little resemblance to the dynamic situation in aeration tank. Depending on how Mueller analyzed and interpreted the floc geometry, they obtained totally different results varied by a factor of 10 has been concluded by Smith and Coackley (1984).

Bungay *et al.*, (1969) have used a microprobe dissolved oxygen electrode to determine dissolved oxygen profiles through a bacterial slime. The diffusion coefficient then calculated assuming steady state diffusion and a homogenous respiration in the slime. The calculated value of diffusion coefficient from this method reported to vary from $9.3 \cdot 10^{-10}$ to $4.0 \cdot 10^{-11} \text{ m}^2/\text{sec}$.

Williamson and McCarty (1976) have measured the diffusivity of ammonia, nitrate and oxygen for the film. Two mixed chambers were separated by the biological film and various reactants allowed to diffuse through the film. Experiments with nitrate and nitrite were carried out with no oxygen in the film and hence no metabolism by the bacteria. Oxygen diffusivity was determined when no nitrogen species were

present. The diffusivity results were 80-100% of their corresponding values through pure water.

Matson and Characklis (1976) have measured the diffusivity of oxygen through activated sludge in an apparatus of consisting of two well-mixed chambers separated by a sandwich with a hollow metal disc containing sludge an oxygen permeable membranes on either side of the discs. The sludge was prepared by concentration in a centrifuge at low speed, followed by deactivation by mercuric chloride or ultraviolet radiation. Prior experiments were carried on by Matson had examined the difference between diffusivity of non-metabolizable salts through live and deactivated sludge. No difference was found indicating that deactivation had no effect on diffusion measurement. The results were indicated an average diffusivity for oxygen through activated sludge of $1.27 \cdot 10^{-9} \text{ m}^2/\text{sec}$.

Many models have been proposed for nitrifying suspended growth systems (Downing et al., 1964; Lijkema, 1973; Poduska and Andrews, 1975; Stenstrom and Poduska, 1980; Beck, 1981; Batchelor, 1982; Henze et al., 1987), but some of them deal specifically with inter particle diffusion. Because microbial floc and liquid exist as seperate phases, inter particle diffusion limitation may be significant if the systems are operated under low dissolved oxygen and/or low turbulence levels. Under such conditions, oxygen depletion due to its diffusion resistance inside the floc may adversely affect nitrification efficiency due to the partial ability of the floc to nitrify and may also enhance de nitrification due to the existence of anoxic zone inside the floc.

A number of investigators including La Motta and Shieh (1979), Benefield and Molz (1983,1984) have attempted to show through various mathematical models, the significance of diffusion resistance in floc on the rates of substrate removal in nitrifying growth systems. The diffusion model of La Motta and Shieh (1979) is a steady state model and hence is limited in utility because the inputs to most

wastewater treatment systems are time-variant. Furthermore, the model is applicable only to excess oxygen availability and a single limiting substrate (ammonia). Benefield and Molz (1984) developed non-steady state diffusion models but neglected the possibility of nitrite accumulation and the ability of the floc to denitrify under oxygen-limited conditions.

In most studies on intraparticle diffusion limitation including those involving non- nitrifying systems (Mueller et al., 1967 ; Baillod and Boyle, 1970 ; Matson and Characklis, 1976 ; Lau et al., 1984), model application relied on experimental evaluations of floc density, floc size and/or substrate diffusivities within floc. As noted by several investigators including Mueller et al.(1967), Kossen (1978) and from the experimental determination of Bakti and Dick (1992) reliable experimental determination of these parameters is found extremely difficult.

Several investigators including Mueller et al. (1967), Baillod and Boyle (1970) and La Motta and Shieh (1979), while determining diffusivity from theory, they have obtained the values of floc size and floc density from the experiments.

In the study of Bakti and Dick (1992), an alternative approach to the experimental evaluation of the floc, floc size and substrate diffusivities was developed. The basis of their model is the consideration that the floc behavior is related to the bulk liquid characteristics. When a floc is formed from aggregation of organisms, metabolic reactions occur concurrently, gradients vary with changes in the bulk liquid characteristics. Because of the concentration gradients needed for mass transfer, the overall reaction rates in a flocculent system will not be equal to those in a dispersed system.

According to Bakti and Dick (1992) the occurrence of nitrification near the external surface of floc and denitrification toward

the center of the floc as depicted in Figure 2.1 may be realized under low DO and /or turbulence levels, concentrations inside and outside the floc may be nearly equal and the intraparticle diffusion effect may be insignificant.

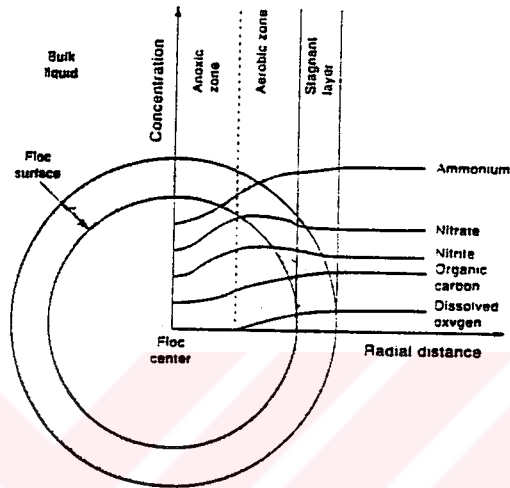


Figure 2.1 Illustration of substrate concentration profiles within a microbial floc showing simultaneous nitrification and denitrification (Bakti and Dick, 1992).

According to the one of the latest model and, experimental study for fixed film system on RBD investigated by Gönenç and Harremoes (Gönenç and Harremoes, 1985; 1990) mineralization and nitrification can take place in the aerobic part of the biofilm as a “zero-order” reaction in Rotating Biological Disc (RBD). The inlet section of a RBC is characterized by high organic matter concentration, high rates of oxygen utilization due to mineralization, therefore low but above zero oxygen concentrations. In this section, the biofilm grows so fast and oxygen penetrates so little that the slowly growing nitrifiers can not keep up. Furthermore, the slow growth rate of the nitrifiers to become only a small, insignificant fraction of biomass. Accordingly, nitrification becomes insignificant.

For an oxygen concentration in the order of 4 mg/l to achieve the nitrification requires the soluble BOD concentration to be less than 20 mg/l, if the oxygen concentration decreases soluble BOD concentration should be decreased accordingly has been concluded by Gönenç and Harremoes (Gönenç and Harremoes, 1990).

From the literature review, apparently that the DO concentration requirements for nitrification are not well defined and that a broad range of DO half saturation coefficients have been reported. There are several factors which could produce the variability in the reported effects of DO concentration within the flocs may be considerably less than the bulk liquid concentration. This phenomenon was originally investigated by Wuhrmann(1963) and has more recently been studied by Kossen (1978).

The concentration of DO in the biological floc is dependent upon a number of factors. Most prominent among these are bulk DO concentration, size and shapes of the flocs, bulk mixing, turbulence and intensity and the oxygen uptake rates of both heterotrophic nitrifying organisms in the floc. Therefore, it would be expected that the results of pure culture investigations could be different from the results of investigations using mixed or activated sludge cultures.

The growth rate dependence of the nitrifying bacteria on inorganic nitrogen substrates and the intensity of mixing has been discussed by Poduska (1975). Based on work by Lilly and Sharp (1968) on enzyme kinetics in continuous flow stirred reactors, it was shown that the half-saturation coefficient was inversely proportional to mixing intensity and mathematically related to the diffusion boundary layer thickness surrounding the enzyme particles.

Powell (1967) has suggested that diffusion of substrates to cell can reduce the availability of those substrates at low concentrations. He mathematically demonstrated that the effect of diffusion would be to

increase the apparent half-saturation coefficient. Therefore, the variations reported for the oxygen half-saturation coefficient for nitrification could in part be due to difference in the intensity of each study.

A second possible source of variability in reported results is the dynamic or steady-state nature of the measuring techniques for half-saturation coefficients. Several of the values for half-saturation coefficients given in Table 2.1 were calculated from batch tests using dissolved oxygen exhaustion rates. Other values were calculated from steady-state operation at several values of liquid phase dissolved concentrations.

According to Stenstrom and Poduska (1980), very little is known about the dynamic behavior of nitrification under oxygen limited conditions, and only one dynamic model (Laudelout et al., 1974) includes the effects of oxygen limiting kinetics. The results of this model appear to confirm the possibility of difference between dynamic and steady-state measuring techniques to determine needed DO concentrations.

A third possible source of variability in Table 2.1 is the likelihood of double substrate kinetics. Normally, micro-biological cultures are thought to be limited by only one substrate. Most mathematical models for biological growth account for only one substrate, as shown in equation (2.1) since experimental studies are usually performed with all other nutrients in excess.

$$\mu = \mu_{\max} \frac{S}{K_s + S} \quad (2.1)$$

where

$$\mu : \text{growth rate} \quad (T^{-1})$$

$\mu_{\max}$	: maximum specific growth rate	(T^{-1})
S	: limiting substrate concentration	(ML^{-3})
K_s	: half-saturation coefficient	(ML^{-3})

Ryder and Sinclair (1972) and Sykes (1973) have recently investigated the possibilities of multiple, but not simultaneous substrate limiting kinetics. Both investigators show that one operating point or condition, substrate A limiting, while at another operating point or condition substrate B may be limiting. Both investigator propose to determine maximum growth rate by evaluating the microbial yields associated with growth rate, and then selecting the expression that is limiting.

Bader et al.(1975) and Bader (1978) have showed that the analysis proposed by Ryder and Sinclair (1972) and Sykes (1973) are not general and have discussed simultaneous multiple substrate limiting kinetics, using growth rate contours, similar to that shown in Figure 2.2.

The growth rate contours in Figure 2.2 were calculated from a double Monod function with endogenous or decay coefficient by Stenstrom and Poduska (Stenstrom and Poduska,1980).

$$\mu = \mu_{\max} \left[\frac{S_1}{K_{s1} + S_1} \cdot \frac{S_2}{K_{s2} + S_2} \right] - k_d \quad (2.2)$$

where

μ	: net specific growth rate	(T^{-1})
$\mu_{\max}$	: maximum specific growth rate	(T^{-1})
S_1	: substrate 1 concentration	(ML^{-3})
S_2	: substrate 2 concentration	(ML^{-3})

- K_{S_1} : half-saturation concentration for substrate 1 (ML⁻³)
 K_{S_2} : half-saturation concentration for substrate 2 (ML⁻³)
 k_d : decay or maintenance coefficient (T⁻¹)

To illustrate the effects of double substrate limiting kinetics, the net growth rate has been calculated and plotted as a contour graph shown in Figure 2.2. A value of 0.5 mg DO/l was selected for the DO half-saturation coefficient. The values are chosen as a representative of the ranges of the coefficients is shown in Table 2.1. It is interesting that Laudelout et al. (1974) have also used a value of 0.5 mg DO/l in their study. The analysis shown in Figure 2.2 is also applicable for other values of half-saturation coefficients. The figure shows regions where the growth rate can be limited by both substrates, and regions where only one substrate is limiting. Most activated sludge plants that nitrify successfully are capable of producing effluents with less than 1 or 2 mg/l of ammonia nitrogen. The results shown in Figure 2.2 indicate that only way to achieve this effluent nitrogen concentration at a net growth rate of 0.005 hr⁻¹ (8.3 days mean cell retention time) is to operate at a DO concentration in excess of 2.5 mg/l.

A contour graph of effluent ammonia nitrogen concentration as a function of MCRT and DO concentration has been prepared by Stenstrom and Poduska (1980) is shown in Figure 2.3 and a perspective diagram is shown in Figure 2.4. In their opinion from these two figures, the hypothetical effects of DO concentration upon nitrification at any given MCRT are clearly demonstrated that, at low MCRT, it is possible to nitrify as efficiently as at higher MCRT, although the required DO concentration is much greater. In addition, from Figure 2.2 it becomes obvious why numerous investigators have reported different values of DO concentration required for nitrification. For example, it is possible to explain why Wuhrmann (1963) observed poorly to zero nitrification at 4.0 mg/l DO oxygen concentration and excellent nitrification at 7.0 mg/l DO concentration. Wuhrmann does not report MCRT for this

observation, but he reports the hydraulic loading rate and mixed-liquor solid's concentration, which can be used to estimate a food to mass ratio, F/M or MCRT. The estimated MCRT for this observation is low and borders upon washout. Therefore, it would be expected that a high DO concentration was needed for Wuhrmann's observation as Figure 2.3 shows.

It has been argued by Ryder and Sinclair (1972) and Sykes (1973) that stoichiometric considerations will make only one of multiple substrates limiting. Undoubtedly, for many cases they may correct but for the case for nitrification in wastewater treatment plants, the DO concentration can be independent from MCRT, or for concentration of other substrates. According to Stenstrom and Poduska (1980), this situation exists because many treatment plants are provided with variable aeration rates. The treatment plant operators can maintain constant DO concentration over a range of growth rates or substrate loading rates. As treatment plants are modernized and energy becomes more expensive, more treatment plants will install real time DO control systems. The results of a recent survey by Genthe et al., (1978) indicated that more automatic DO control systems are being installed, and concluded that the trend will continue. Moreover, this survey shows that DO control can result in improvements in process efficiency. It also appears from work performed by Sezgin et al.,(1978) that DO controls may be used to improve process performance with respect to sludge bulking.

From the previous discussion has been made by Stenstrom and Poduska (1980), it was obvious that there exists no clearly defined DO concentration for optimum nitrification. It appeared that at higher MCRT, nitrification can be achieved at DO concentration in the range of 0.5 to 1.0 mg/l and at lower MCRT higher DO concentrations are needed. Based on the analysis of their study, the lowest DO concentration at which nitrification occurs appears to be approximately 0.3 mg/l.

According to Stenstrom and Poduska (1980), there appear to be several mechanisms which may explain the wide distribution of reported half saturation coefficients and inhibitory DO concentrations. It is concluded that one possible reason is the effect of oxygen diffusion with the flocs, a second mechanism is double substrate limiting kinetics, which can explain the wide variation of reported values.

Table 2.1 Summary of the Effects of DO on Nitrification
(Stenstrom and Poduska, 1980)

Organisms	Dissolved Oxygen Parameter		Conditions and Remarks	Reference
	$K_s(\text{DO})$ mg / l	(DO) mg / l		
<i>Nitrosomonas</i>	0.3		20 °C Pure culture	Loveless & Painter (1968)
	0.25		Pure culture	Peter et al. (1969)
		1.0	Zero-order ¹ , pure culture	Schoberl & Engel (1964)
		2.0	Zero-order	Knowles et al. (1965)
<i>Nitrobacteria</i>	1.84	2.46	Pure culture, by respirometric	Peter et al. (1969) and
	0.83		amperometric and micro	Laudelout et al. (1976)
			calorimetric technique	
	0.34, 0.48		at 20, 29, 35 °C,	
	0.72		respectively, Calculated from	
			Boon & Laudelout (1962)	
<i>Nitrocyclic</i>		2.0	Zero-order	Schoberl & Engel (1964)
		4.0	Zero order	Knowles et al. (1965)
		7.5	Zero order	Gundersen (1966)
<i>Oceanous</i>	0.5-0.7		Inhibitory	Forster (1974)
<i>Nitrifiers</i>	1.0-1.5		Proper operation in ASP ²	Jenkins (1966), Balakrishnan & Eckenfelder (1969), Wild et al. (1971), Wuhrmann, (1964)
	0.8		In soil	by Shah & Coulman (1978)
		0.5	Inhibitory ³ in ASP	Bragstad & Bradney (1937)
		0.3	Inhibition ⁴ in ASP	Downing and Boon (1963)
		0.2	Inhibitory in ASP	Downing et al. (1964)
		0.3-1.0	Proper operation in ASP	Downing et al. (1964)
		4.0	Max. rate of nitrification	Wuhrmann (1963)
	0.42		In pure oxygen ASP	Stankewich (1972)
	2.0		In ASP	Nagel & Haworth (1969)

1 Zero- order minimum DO concentration for zero order kinetics

2 ASP- activated sludge process

3 Inhibitory - no reaction

4 Inhibition- reaction depends upon concentration

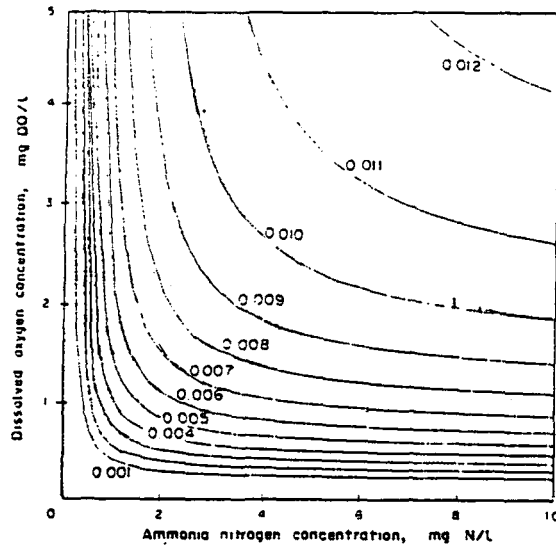


Figure 2.2 Net growth rate contours for nitrification : ammonia nitrogen vs. dissolved oxygen concentration (numbers on curve indicate growth rate, μ , hr^{-1}) (Stenstrom and Poduska, 1980)

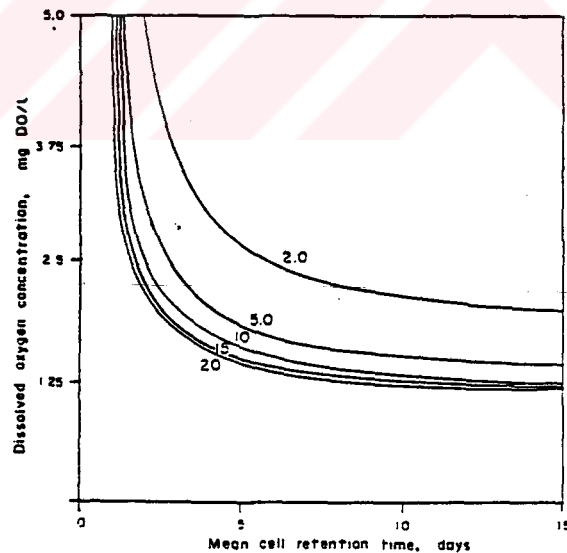


Figure 2.3 Effluent ammonia-nitrogen contours: cell retention times vs. dissolved oxygen (numbers on curves indicate effluent ammonia nitrogen concentration). (Stenstrom and Poduska, 1980)

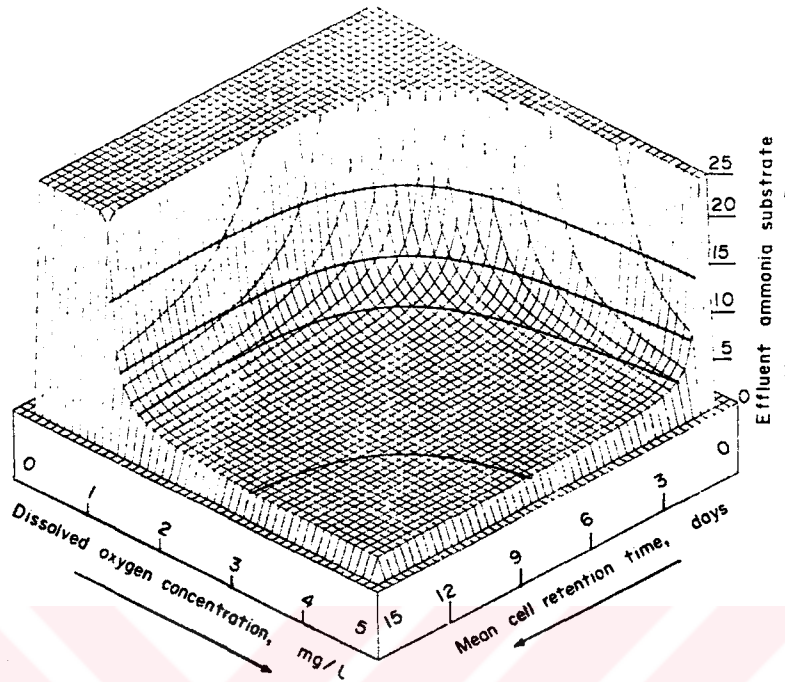


Figure 2.4 Ammonia nitrogen substrate perspective diagram (Stenstrom and Poduska, 1980)

$$\begin{aligned}\mu_{\max} &= 0.02 \text{ hr}^{-1} \\ K_s &= 1.0 \text{ mg NH}_4^+ - \text{N/l}\end{aligned}$$

$$\begin{aligned}K_{s \text{ DO}} &= 0.5 \text{ mg DO l}^{-1} \\ k_d &= 0.005 \text{ hr}^{-1}\end{aligned}$$

2.2 Oxygen Supply for Aerobic Suspended Culture Systems

2.2.1 Submerged Aeration Systems

Submerged aeration systems have been used in the treatment of wastewater since the discovery of the activated sludge process by Ardern and Lockett in 1914. The first aeration systems were coarse bubble diffusers, but interest in reducing power and compressor requirements led to the development of fine bubble diffusers. The fine bubble diffusers are capable of transferring more oxygen per unit of power consumed because of the increased bubble surface area created

by the smaller bubbles; however, the small bubble systems require highly filtered air and are troubled by inorganic deposits and biological growths on the diffuser openings.

In submerged aeration, oxygen transfer takes place in three phases: first, there is a rapid transfer across the expanding bubble surface during formation; second, oxygen is transferred to the bulk liquid as the bubble rises; and third, the turbulence created by the bubble breaking the surface promotes surface aeration. The amount transferred during the first phase of bubble formation depends upon the type of diffuser unit being employed. Fine bubble diffusers provide a great deal of oxygen transfer because of the frequency of bubble formation and surface area per unit volume of air. Coarse bubbles provide less interfacial area. Sparger units and open holes in air pipes do not provide large amounts of oxygen transfer in this phase because they do not form individual bubbles on the diffuser, but instead issue an air stream that separates into bubbles after leaving the diffuser. As the bubbles rise after formation they are initially experiencing a high driving force due to the large value of the oxygen is rapid at this point and the oxygen content within the bubble decreases. This transfer process rate decreases as the bubble rises until at same point near the surface a condition may occur where oxygen is stripped from the liquid back into the air bubble. This phenomenon usually occurs only in low oxygen deficit periods and is due to oxygen enriched water near the surface that contains dissolved oxygen in excess of the saturation concentrate on at that shallow depth. The amount of transfer in the rising phase is enhanced by small bubbles (large interfacial area per unit volume), high turbulence (rapid surface renewal),deep submergence (increased driving force due to larger saturation concentration),and long contact time. The surface turbulence created by the bursting bubbles exposes new surfaces of liquid to the atmosphere that allows oxygen transfer especially when the dissolved oxygen concentration is low.

There is no question that submerged aeration in its many forms has a place in wastewater treatment today. Increased land costs and, in some cases, scarcity of land has forced treatment plants to go deeper and be placed closer to living areas. Submerged systems offer the advantage of increased oxygen transfer as the air is released from greater depths and they do not create airborne particles of mixed liquor mist. This mist is particularly objectionable where it may be blown over land adjoining the treatment facility.

2.2.2 Oxygen Mass Transfer

Oxygen in a gaseous state disperses itself through a body of liquid by the process of diffusion which tends to produce a stable state of uniform concentration. Mass transfer by diffusion occurs between the gaseous phase and the liquid phase when driving force is created by a departure from equilibrium. This driving force is a partial pressure gradient in the gas phase and a concentration gradient in the liquid phase. The rate of mass transfer of a dissolved gas in a liquid is dependent on a characteristic of the gas and the liquid, the temperature, the concentration gradient and the cross-sectional area through which diffusion occurs.

Numerous theories have been put forward to interpret the mass transfer of a sparingly soluble gas such as oxygen to water. However, all these theories are approximations of the physical process as none completely accounts for all observed phenomena. Of the three reviewed, the first the Whitman, two-film theory is the most widely accepted and used in aeration process today.

The film theory is used based on a model where two films exist at the gas/liquid interface. The resistance to gaseous molecules transferring from the bulk-air to the bulk-liquid state occurs in the films. In considering the transfer of gas molecules from the air to the liquid, soluble gases encounter major resistance from the gaseous film and

slightly soluble gases such as oxygen, find the major resistance from the liquid film.

This theory first proposed by Whitman (1923) and Lewis and Whitman (1924) pictures a stagnant film of thickness, Z at the surface of the liquid next to the gas. While the rest of the liquid is kept uniform in composition by agitation, the concentration in the film falls from C_s at the gas interface to C at the liquid surface (there is no convection in the film) and the dissolved gas crosses it by molecular diffusion alone. This relation is known by Fick's Law as shown mathematically,

$$\frac{dm}{dt} = D (C_s - C) / Z = K_L (C_s - C) \quad (2.3)$$

where

dm/dt is flux ($M L^{-2} T^{-1}$)

C_s is the saturation concentration of the dissolved gas ($M L^{-3}$)

C is the average bulk liquid concentration ($M L^{-3}$)

K_L is the liquid film coefficient ($L T^{-1}$)

D is the molecular diffusion coefficient ($L^2 T^{-1}$)

The liquid film coefficient, K_L is defined as the diffusivity, D divided by the hypothetical film thickness, Z and in general K_L can be related to resistance from the liquid interface renewal and turbulence. In practice it is not possible to measure K_L directly, so the new term $K_L a$, the overall mass transfer coefficient is measured. $K_L a$ is the product of the liquid film coefficient, K_L and the interfacial surface area (a), generated per unit of volume of liquid (V) and defined as follows;

$$\frac{1}{V} \frac{dm}{dt} = \frac{dc}{dt} = K_L (A / V) (C_s - C) = K_L a (C_s - C) \quad (2.4)$$

where

A : interfacial area for transfer (L^2)

V : unit volume of liquid	(L ³)
a : interfacial surface area	(L ² L ⁻³)

This theory is not very realistic as it assumes a layer of uniform thickness. Nevertheless, this theory incorporates an essential feature in the real system, namely that the gas must get into the liquid by dissolution and molecular diffusion before it can be transported by convection. The Whitman two film theory, despite its deficiencies, is the theory most accepted and used in commercial aeration analysis today (Beak Consultants Limited, 1977).

Instead of the discontinuity implicit in the film theory, there may be a progressive transition from purely molecular transport to predominantly convective transport as the distance from the gas interface increases. King (1964) has considered this possibility and assumed that transport is the combined result of molecular and eddy diffusivity. He showed that the transfer coefficient may be proportional to any power of diffusivity between 0 and 1 and not necessarily 1 as given by the film model. He also provides an expression for K_L , $K_L = D/Z$ that contains two parameters related to the hydrodynamic conditions, whereas the film model uses only one i.e. Z .

Since the liquid film coefficient (K_L) and the interfacial surface area (a) can not be determined separately, in evaluating the performance of commercial aeration equipment it has been a general practice to consider the aeration process in terms of the overall oxygen transfer coefficient ($K_L a$).

The mass transfer data shown in literature is in general based on the Whitman two-film theory and the corresponding general expression of oxygen transfer,

$$dc/dt = K_L a (C_{se} - C) \quad (2.5)$$

where

dc/dt : Rate of oxygen transfer, change of oxygen concentration (C) with respect to time (t), ($ML^{-3} T^{-1}$)

K_L : The liquid film coefficient ($L T^{-1}$)

a : A/V , the interfacial surface area of the air (A) through which diffusion can occur generated by the particular aeration system per unit volume (V) of water.

$K_L a$: Overall oxygen transfer coefficient, (T^{-1})

C_{se} : Effective saturation value of dissolved oxygen concentration at given temperature and atmospheric conditions, (ML^{-3})

C : Average concentration of dissolved oxygen in liquid (ML^{-3})

Finally the total mass transfer of oxygen into water

$$M_{O_2} = K_L a (C_{se} - C) V \quad (2.6)$$

where

M_{O_2} : Overall mass transfer of oxygen ($M T^{-1}$)

V : Total volume of water (L^3)

2.2.3 Fluid Mixing By Aeration

When an aeration system is required to be effective, it must not only transfer the required amounts of dissolved oxygen to the water in the aeration basin, but it must also disperse the oxygen throughout the basin and provide sufficient turbulence to maintain solids in suspension. Aeration systems differ in the method in which mixing energy is applied. Mechanical aerators are point sources of very intense energy that must penetrate the basin volume. Submerged compressed air

systems vary from multiple point sources of turbulence as typified by the static air-gun units, to rather uniform mixing inputs in the diffused air units that occupy a large portion of the basin bottom. However, no matter what system is employed the mixing functions it must provide include :

- a) the mixing of the basin biomass content with the incoming wastewater,
- b) the bulk liquid mixing of the basin contents to the extent of ' complete mixing ' desired,
- c) transport of oxygenated water away from the aerator unit and the cycling back of deoxygenated water,
- d) mixing turbulence by large scale for the suspension of biological solids
- e) mixing turbulence by small scale to promote the transfer of dissolved oxygen and organic into the bacterial cell and to carry bacterial wastes away from the cell.

The first four mixing functions can be accomplished by large scale eddy turbulence macromixing (Price et al., 1973, Kalinske, 1968).

The fifth mixing functions requires mixing turbulence that is characterized by eddies smaller than the bacterial floc particles, so called micro mixing (Price et al., 1973, Kalinske, 1968, Tsai et al., 1969).

The controversy over the macro mixing abilities of the various types of aeration equipment is largely centered around how much mixing input is required and which system can provide it most efficiently.

Kalinske (1960) has presented a thorough analysis on the fundamental principles of fluid mechanics as they apply to mixing. The analysis of mixing energy reveals that large scale eddy turbulence which

is important in the diffusion process is relatively unimportant so far as energy dissipation is concerned; it is the viscosity of liquid and the small eddy turbulence which are the controlling factors for fluid mixing. Fluids rapidly dissipate energy into heat when small scale, intense eddies are produced. Orlob (1959) has proposed the following equation for relating the rate of energy dissipation to eddy turbulence,

$$D_t = K E^{1/3} L^{4/3} \quad (2.7)$$

where

D_t : Eddy diffusion coefficient

E : Rate of energy dissipation per unit of mass

L : Scale of eddies participating in turbulence diffusion

K : Constant of proportionality

This equation suggests the following: turbulence, diffusion or mixing can be increased by adding energy to the fluid mass in the form of large scale eddies ; if small scale eddies are produced rapid energy dissipation into heat will occur. Because it is normally impractical and uneconomical to fill a basin with mixing devices, turbulence and mixing energy is provided by point sources. These point sources provide a localized origin of relatively high intensity turbulence which is directed to penetrate the mass of liquid. The intensity and scale of turbulence produced in the volume adjacent to the unit will be proportional to the velocity of the stream and its relative size.

2.2.4 Oxygen Transfer Correction Factors and Interferences

Many factors influence oxygen transfer mechanisms in wastewater treatment process. Wastewater contaminants, temperature, dissolved oxygen concentration (driving force), type of aeration device, turbulence, and basin geometry all affect oxygen transfer rate. Since

these factors make field application of an aeration device unique, it has become standard practice to specify aeration equipment at "standard conditions" and to develop additional techniques for adjusting rates at standard conditions to rates at field conditions. Standard conditions in the United States are considered to be 20 °C, 760 mm Hg barometric pressure, tap water, and zero dissolved oxygen concentration. In Europe, standard conditions are slightly different and 10°C is the standard temperature. Additionally, in the United Kingdom, a surfactant (commercial detergent) is used to change the properties of wastewater, and to mask the effects of trace contaminants in tap water (Stenstrom and Gilbert, 1981) .

Field oxygen transfer rates are calculated from standard transfer rates through the use of alpha (α), beta (β), and teta (θ) factors. These factors are defined as follows;

$$(K_L a)_{ww} = \alpha K_L a_{TP} \quad (2.8)$$

$$C^*_{\alpha ww} = \beta C^*_{\alpha TP} \quad (2.9)$$

$$(K_L a)_T = (K_L a)_{20} \theta^{T-20} \quad (2.10)$$

where

$K_L a$: volumetric mass transfer coefficient (T^{-1})

C^*_{α} : saturation dissolved oxygen concentration ($M L^{-3}$)

ww : subscript indicating wastewater

TP : subscript indicating tap water

T : temperature ($^{\circ}C$)

$(K_L a)_T$: K_L at temperature T

$K_L a_{20}$: K_L at 20 °C

Field oxygen transfer rates (OTR) can be easily calculated from standard oxygen transfer rate as follows :

$$\text{OTR} = \alpha \left(\frac{\beta C_{\infty \text{TP}}^* - C_L}{C_{\infty 20}^*} \right) \theta^{T-20} \text{SOTR} \quad (2.11)$$

where

C_L : desired value of oxygen concentration under normal operation

SOTR : oxygen transfer rate at standard conditions

The importance of properly determining the alpha, beta and teta factors can not be overestimated. For example, a field transfer rate is only 52% of the standard transfer rate when $C_L = 2$, $\alpha = 0.8$, $\beta = 0.9$, $T = 18^\circ\text{C}$, when $\theta = 1.024$. Drastic over or under designed aeration system can result if inaccurate correction factors are used. Figure 2.5 shows a profile of errors in oxygen transfer rate for incorrect alpha and beta factors prepared by Stenstrom and Gilbert (Stenstrom and Gilbert, 1981).

2.2.4.1 Alpha Factor

There are several technical problem associated with measuring the alpha factor as discussed by Gilbert (Gilbert, 1979). The alpha factor varies with many processes conditions including wastewater quality, intensity of mixing or turbulence, suspended solid's concentration, method of aeration, scale, and other factors. The effect of aeration methods is particularly important with respect to the alpha factor.

The recent increases in energy costs have generated new interest in aeration efficiency and fine bubble aeration systems. Kiiskinen (Kiiskinen, 1979) has examined the efficiency of seven types of diffusers in clean water, and sites evidence to show that the alpha

factors for fine bubble systems range from 0.4 to 0.5. Houck (Houck, 1980) presented the results of a survey of 19 plants using fine bubble diffusers, and tabulates process efficiencies, as well as operating experience with fouling and plugging problems, along with mitigating and maintenance techniques. He also raises the question of bubble coalescence on fine bubble diffusers due to biological growths on the liquid side of the diffusers.

The effects of surfactants on α have been investigated for various types of surface aerators. Downing et al., (1960) have investigated α factors for the Searle aerator (a modified brush type aerator) and Simplex Cone aerator (vertically rotating low speed surface aerator with draft tube). The α factor was approx. 2.0 for the Searle aerator in the presence of 10 mg/l Alkil Benzen Sulfonat. For the simplex Cone aerator, they concluded that an increase in oxygen transfer rate of 10- 15% could be expected with surfactant addition, which corresponds to an α factor of 1.1-1.15. They also report that anti-foaming agents reduced the α factor. Similar results were reported by Eckenfelder and Ford (1968) and von der Emde (1968).

The mechanisms of surfactant interference have been investigated by Mancy and Okun (1960), McKeown and Okun (1963), and Mancy and Barlage (1968), among others Mancy and Barlage (1968) have showed that the effects of surfactants on aeration can be divided into two categories: the effect of the adsorbed surfactant film which increases the resistance to transfer; and changes in hydrodynamic behavior of the air- liquid interface, which results from changes in surface tension produced by the surfactant. They used three experimental designs in their investigation; a 50 cm bubble column, surface aerator, and laminar jet orifice to observe the effect of surfactants on the molecular diffusion coefficient independently of changes in bubble characteristics and the hydrodynamic conditions, of which large part are created by the intensity of mixing, power input, and unique characteristics of each aeration device, will greatly affect

oxygen transfer rate. Their results, which have been supported by Mueller (Mueller, 1976).

The study of Mancy and Barlage (Mancy and Barlage, 1968) also shows that the rate of surfactant adsorption (a function of the type of surfactant) to the gas - liquid interface also affect oxygen transfer.

The alpha factor also changes substantially with wastewater characteristics. For example, Bass and Shell (Bass and Shell, 1977) reported alpha factors fluctuation for 0.5-1.0 for various wastewaters and Eckenfelder (Eckenfelder, 1959) has reported the alpha factor for an industrial wastewater to vary between 0.3 and 0.8 . The variance alpha factor can largely be attributed to the changing of the raw wastewater with time of day or day of week.

Kalinske (Kalinske, 1968) and Pfeffer (Pfeffer et al., 1968) have reported that the alpha factor also varies with degree of treatment and with location of sampling point in activated sludge aeration basins. Generally, alpha factor is lowest for the influent wastewater, and increases to maximum for the effluent, Wheatland and Boon (Wheatland and Boon, 1979) have reported data showing the change in alpha factor with increasing levels of treatment, for fine bubble diffuser aeration systems. They have related the alpha factor change to total oxygen consumed during treatment, finding an alpha factor ranging from approx. 0.3-0.4 for 100 mg O₂ absorbed to 0.7-0.8 for 300 mg O₂ absorbed. Their findings have particular significance for the "plug flow" type aeration systems, where one would expect the greatest spatial variability in alpha factors. Moreover, the use of tapered aeration amplifies this effect, since a simple arithmetic average of alpha factor may not accurately reflect the true average.

Downing (Downing, 1960) has reported on the range of suspended solids from 0 to 3000 mg/l and found that alpha decreased with increasing solids concentration.

Turbulence has a very strong effect on alpha factors. Stenstrom and Hwang (Stenstrom and Hwang, 1979) have showed that a range of alpha factors can be measured for a specific wastewater (tap water containing synthetic detergent), depending upon the level of turbulence, as indicated by power input per unit volume. Eckenfelder and Ford (Eckenfelder and Ford, 1968) have noted similar effects.

There is convincing evidence to indicate those alpha factors for fine bubble diffusers are lower than alpha factors for coarse bubble diffusers, and mechanical aerators.

It is the opinion of a number of researchers, including Kayser (1979), and Zlokarnik (1979 b), those alpha factors simply can not be measured reliably on small scale devices.

Alpha testing should be performed with generic devices identical to those proposed for full scale installation. For diffuser systems, the bubble diameter is the most characteristic of the device. For surface aerators, the level of turbulence appears to be the most important characteristic.

The scale-up of the test equipment seems to dramatically affect the determination of alpha factors. Using small scale laboratory equipment, it is easy to test using high levels of turbulence, or power input per unit volume. In general, these conditions will not be representative of full scale conditions.

The general question of scale-up procedures has been addressed by Ellis and Stanbury (1980), Horvath (1966,1979), Schmidtke and Horvath (1977), Harremoes (1979), and Zlokarnik (1979a). They have showed that scale-up for certain devices, such as surface aerators, can be performed with acceptable precision and accuracy ; however, none have addressed the problem of scale-up with reduced or changing surface tension, or reduced or changing molecular

diffusivity of oxygen. Another scale-up problem that must be overcome before diffused aeration devices can be properly scaled, is the influence of changing oxygen transfer mechanisms in a rising bubble plume. There exist three distinct transfer "regions" for diffused aerators. These are the region of bubble formation, where surface renewal and transfer are very high; regions of rising bubbles, where reduced transfer rates occur, and surface region, where the surface "boil" produces transfer. Several investigators, including Morgan and Bewtra (1960), Aiba et al. (1963), Ippen and Carver (1954), and others have demonstrated the effects of depth in oxygen transfer. According to Stenstrom and Gilbert (1981), ideally, alpha factor testing should be performed using the actual fluids to be aerated. For activated sludge plants, this means that the mixed liquor, containing the biologically active suspended solids, should be used for testing. This may create problems if the treatment plant has not been built, or if it is not possible to accurately measure the oxygen uptake rate of the biological solids. In general, the influent wastewater is a poor substitute for mixed-liquor, and will generally produce lower alpha factors, although occasionally, the opposite effects have been noted. Special problems exist for "plug flow" type aeration systems, especially those with tapered aeration. Wheatland and Boon (1979) have presented evidence to indicate that alpha factors, for certain specific conditions, may change from as low as 0.3 at the influent end of an aeration tank to as 0.8 at the effluent end. Bubble coalescence can strongly affect alpha factors. Most wastewater treatment systems operate with coalescing fluid properties; however, with addition of high concentrations of salt, the fluid properties can be changed to non-coalescence. Under these conditions the alpha factors can be many times greater, and Zlokarnik (1979 b) has provided an analysis of these effects. Gas side oxygen depletion affects alpha factors in a way that is analogous to its effect on transfer coefficients. The magnitude of these effects is usually small, and much less than the experimental error associated with alpha factor testing. However, Baillod and Brown (1980) have developed procedures for correction.

The previous discussion has shown that it is extremely difficult to determine meaningful alpha factors. An alternate approach has suggested by Stenstrom and Gilbert (1981) is given by below equation;

$$\alpha^- = \frac{K_L a_{ww}}{K_L a_{TP+S}} \quad (2.12)$$

where

- α^- : alpha bar
- $K_L a_{ww}$: mass transfer coefficient in wastewater
- $K_L a_{TP+S}$: mass transfer coefficient in tap water with a surfactant added.

2.2.4.2 Beta Factor

The beta factor β , is defined as the ratio of the saturation dissolved oxygen concentration in wastewater and in clean water or tap water as follows:

$$\beta = \frac{C^*_{\alpha ww}}{C^*_{\alpha TP}} \quad (2.13)$$

The saturation value of dissolved oxygen is affected by a large number of variables and process conditions, including barometric pressure, temperature, suspended solids, dissolved organic, and dissolved solids. Wastewaters contain sufficiently high dissolved solids to significantly reduce the saturation dissolved oxygen concentration.

The beta factor has been found to vary over a broad range, although variations are generally less than those reported for the alpha factor. Eckenfelder et al. (1956) have reported that the beta factor for

domestic wastewater is generally about 0.95 and that it can vary over a much broader range for industrial wastewater.

Technical problems exist with measuring beta factors. The most common wet chemical analysis for dissolved oxygen, the Winkler Analysis (as specified by APHA, 1975) is subject to interferences that can make the test ineffective for some types of wastewater. Interferences have been reported by Kalinske et al. (1973), among others has been found that the amount of manganese sulfate added to the sample for analysis changes the measured values of DO concentration and hypothesized that organic matter in wastewater chelates metal ions, such as manganese or cobalt, which reduces their activity in solution and changes results. DO probe be used to determine beta values. This alternative is attractive since the galvanic cell in a DO probe is isolated from the wastewater constituents which interfere with the Winkler test. DO probe cannot be used to measure beta factors since the probe responds to the activity of molecular oxygen, and not concentration. This phenomenon has been discussed extensively by many investigators (Gilbert and Stenstrom, 1981). An alternative method for measuring beta factors, which have been proposed by Bass and Shell (1977), among others, is to use an analytical correction factor, based upon total dissolved solids concentration.

2.2.4.3 Theta Factor

Temperature strongly affects aeration systems in a variety of ways. Perhaps the greatest effects on saturation dissolved oxygen concentration, which is well documented. The effects of saturation concentration are not included in the theta factor since they can be handled in the field transfer equations.

The theta factor has normally used to relate mass transfer coefficients as shown in the following equations:

$$K_L a (T) = K_L a (20^\circ) \theta_g^{T-20^\circ} \quad (2.14)$$

$$K_L a (T) = K_L a (20^\circ) + \theta_A (T-20) \quad (2.15)$$

where

$K_L a (T)$: mass transfer coefficient at temperature T

$K_L a (20^\circ)$: mass transfer coefficient at temperature = 20°C

θ_g : geometric temperature correction coefficient

θ_A : arithmetic temperature correction coefficient

T : temperature, degrees Celsius.

The geometric technique, in equation (2.14) is more commonly used. This technique of correcting for the effects of temperature on oxygen transfer is empirical and attempts to lump all possible factors, such as changes in viscosity, surface tension, diffusivity of oxygen, etc. This empirical approach has produced a great variety of correction factors. Some investigators reported that geometric model gives better correction; others reported that an arithmetic model is preferred.

Figure 2.6 shows the error that can result from using incorrect theta factors using a geometric correction technique. It is observed that the deviation from unity increases rapidly as the range of temperature correction increases. For example, using a theta factor of 1.032 instead of 1.024 with 20°C as standard temperature, the oxygen transfer at 30°C rate will be overestimated by over 10%. (Stenstrom and Gilbert, 1981).

According to Stenstrom and Gilbert (Stenstrom and Gilbert, 1981) there is no consensus for a temperature correction factor. The results of a review of the literature indicate that geometric temperature correction factors can range from 1.008 to 1.047, and that some

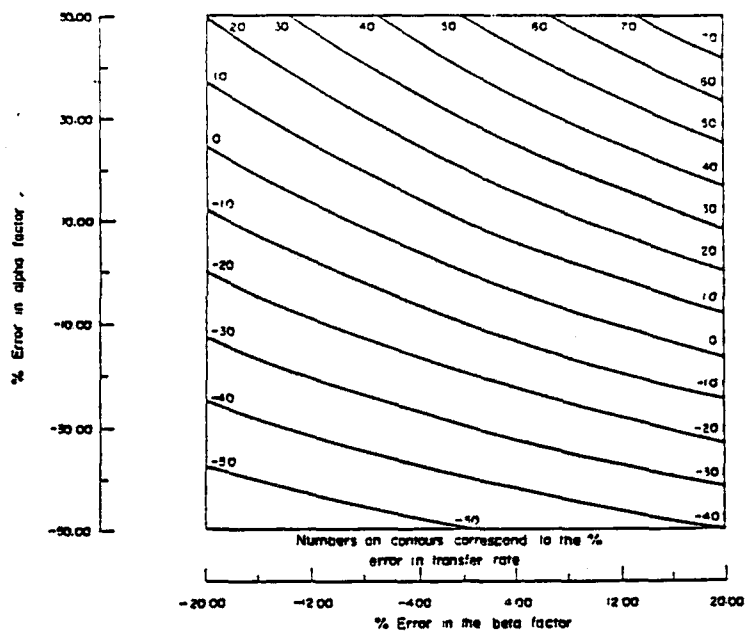


Figure 2.5 Errors in oxygen transfer rate as a function of errors in measurement of alpha and beta factors, when $\alpha=0.8$ and $\beta = 0.9$.(Stenstrom and Gilbert , 1981)

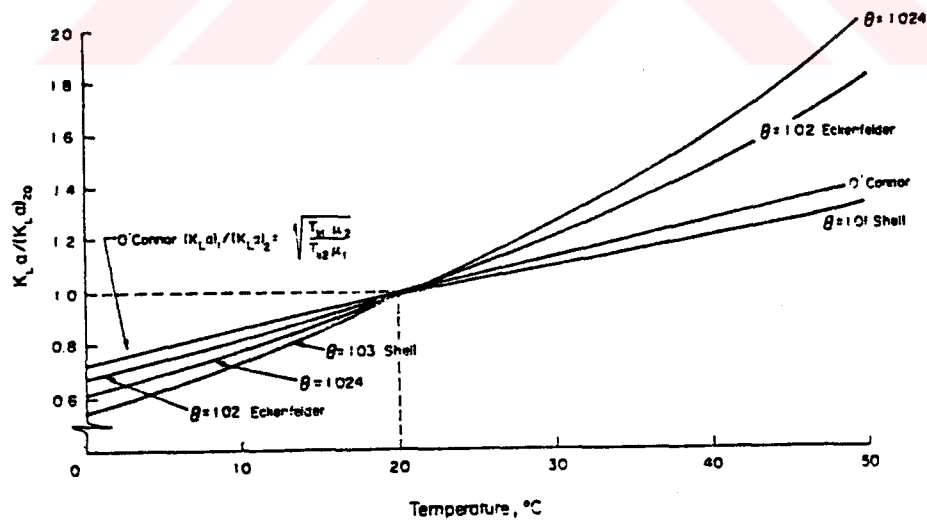


Figure 2.6 The effect of theta value on temperature correction (Stenstrom and Gilbert, 1981)

aeration systems have mass transfer coefficients which are linearly related to temperature (as opposed to the geometric relation). Furthermore, it has been shown that the temperature correction factor is dependent upon turbulence.

It appeared that a theta factor of 1.024 should be used unless it is known that a different geometric theta factor is more suitable (Stenstrom and Gilbert, 1981).



CHAPTER 3. OXYGEN DIFFUSION MODELS ON MICROBIAL AGGREGATES

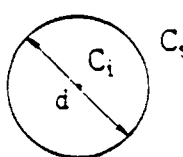
3.1 Wuhrmann Model, 1960

Although the rate of oxygen absorption by aerobic microorganisms in relation to the oxygen tension in the surrounding medium was fairly well known, the role of oxygen tension as an operational parameter in biological treatment plants was a matter of discussion, and according to Wuhrmann (Wuhrmann,1960) ,there were three distinct problems to be considered for a better understanding of the interactions between oxygen tension and the oxidation performance of a mixed population of microorganisms, namely:

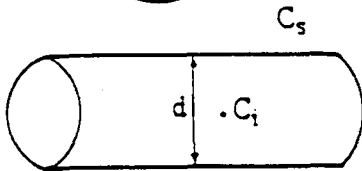
- (1) The oxygen supply of the cells living in aggregates, such as flocs of activated sludge or pellicules of fixed film slimes.
- (2) The dependency of the species composition of the biocoenosis in a biological treatment plant on oxygen tension.
- (3) The effect of oxygen deficiencies on organisms in treatment plants.

Wuhrmann (Wuhrmann,1960) assumed a concentration gradient ($C_s - C_i$) from the surface of a cell aggregate to a cell in the interior of the aggregate, the quantity of transported molecules per time unit is a function of the diffusion coefficient (D), the distance of travel (d), and the rate of consumption (r), of the diffusing molecules by the cell situated between the floc surface and any cell in the floc center. It was suggested that this loss (r), can be calculated for per unit floc volume and time.

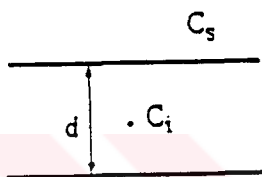
As an application of Fick's law of diffusion, the following set of equations were derived for different shapes of cell aggregates when resolve for the square of diameter (thickness) of floc or pellicule:



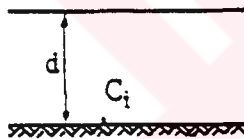
$$d^2 = (C_s - C_i) \frac{24 D}{r} \quad (3.1)$$



$$d^2 = (C_s - C_i) \frac{16 D}{r} \quad (3.2)$$



$$d^2 = (C_s - C_i) \frac{8 D}{r} \quad (3.3)$$



$$d^2 = (C_s - C_i) \frac{2 D}{r} \quad (3.4)$$

Figure 3.1 Diffusion of metabolites into aggregates of living cells (Wuhrmann, 1960)

where

D : diffusion coefficient (cm^2/sec)

C_s : concentration at the surface of the aggregate (g/cm^3)
(assumed to be identical in the surrounding medium-bulk liquid)

C_i : concentration around the innermost cell of the aggregate (g/cm^3)

r : specific consumption of the diffusing molecules by the cell of the floc ($\text{g}/\text{cm}^3 \text{ sec}$)

d : diameter or thickness of the aggregate (cm)

Practical applications of these equations require some numerical values of C_i , the oxygen concentration that must be maintained within a floc for aerobic metabolism of all the cell.

Critical oxygen tensions for a number of bacteria have been published by several authors from which Wuhrmann (Wuhrmann,1960) deduce those oxygen concentrations in the center of a floc about $0.1\text{mgO}_2/\text{l}$ may be adequate for the micro-aerophilic organisms predominant in sewage treatment plants. The estimated diffusion coefficient of oxygen in cell material was about $5 \cdot 10^{-6} \text{ cm}^2/\text{sec}$ at 15°C . It has been stated that the values for the specific rate of consumption rate r , for the diffusing molecules are difficult task to obtain. Considering respiration, the oxygen consumption, from manometric measurements, led to the following most probable figures at 15°C in activated sludge :

$$\text{minimum } r_{\text{O}_2} = 1 \cdot 10^{-4} \text{ mg O}_2/\text{cm}^3 \text{ sec}$$

$$\text{maximum } r_{\text{O}_2} = 5 \cdot 10^{-3} \text{ mg O}_2/\text{cm}^3 \text{ sec}$$

On the basis of the above numerical values, the tolerable floc diameter (d), or pellicule thickness may be calculated when the innermost cell if gets enough oxygen at any time for maintaining an undisturbed respiration rate, assuming that the surrounding medium of the flocs or pellicule contains a certain oxygen concentration (C_s). The results of such estimate by Wuhrmann (Wuhrmann,1960) are shown in Figure 3.2.

According to Wuhrmann (Wuhrmann,1960) the equations (3.1) to (3.4) are valid for oxygen as well as for any other molecules present which are metabolized by the micro-organisms, provided diffusion was the only transport mechanism for their dislocation from the floc surface to the floc center. The activity of biological sludge in treatment plants being not only a function of the supply, but depending as well as on the rate of supply with metabolic substrates. It is briefly

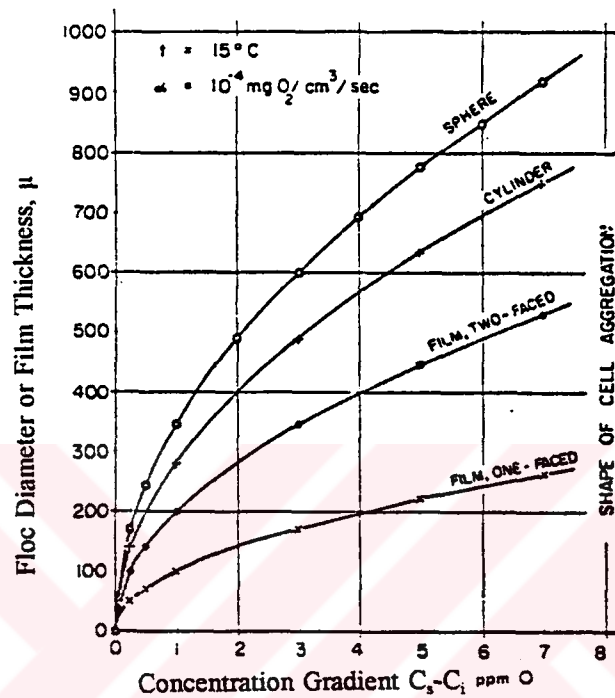


Figure 3.2 Oxygen supply to cell aggregates. Relation between floc size and oxygen supply to the innermost cells at different oxygen concentration gradients (Wuhrmann, 1960).

discussed that the popular assertion that the oxygenation rate of a plant acts as a limiting factor for the oxidation performance.

The relative rate of oxygen (O_2) and substrate (S) supplies to a cell in a spherical floc derived from equation (3.1) by Wuhrmann (Wuhrmann, 1960) as follows:

For spherical floc $d = 2R$, from the same floc,

$$(2R)^2 = (C_s - C_i) \frac{24 D}{r} \quad (3.5)$$

For oxygen, $R^2 = \Delta C_{O_2} \frac{24 D}{r_{O_2}} \quad (3.6)$

For substrates, $R^2 = \Delta C_S \frac{24 D}{r_S} \quad (3.7)$

dividing equation (3.6) and (3.7) side by side,

$$\Delta C_S = \Delta C_{O_2} \frac{D_{O_2}}{D_S} * \frac{r_S}{r_{O_2}} \quad (3.8)$$

where

- R : Floc radius (cm)
- r_{O_2} : Specific consumption of oxygen ($\text{mg}/\text{cm}^3/\text{sec}$)
- r_S : Specific consumption of substrate ($\text{mg}/\text{cm}^3/\text{sec}$)
- D_S : Diffusion coefficient for substrate in the floc material (cm^2/sec)
- D_{O_2} : Diffusion coefficient for oxygen in the floc material (cm^2/sec)
- ΔC_S : Concentration gradient of substrate molecule from the floc surface to the floc center (mg/cm^3)
- ΔC_{O_2} : Concentration gradient of oxygen molecule from the floc surface to the floc center (mg/cm^3).

As an example is given by Wuhrmann (Wuhrmann,1960), assuming glucose to be the principal polluting material within a waste, with using of estimated diffusion coefficient of oxygen, $D_{O_2} = 5.10^{-6}$ cm²/sec and glucose, $D_S = 1.5 \cdot 10^{-6}$ cm²/sec

$$\Delta C_{\text{glucose}} = \Delta C_{\text{oxygen}} \frac{D_{\text{oxygen}}}{D_{\text{glucose}}} * \frac{r_{\text{glucose}}}{r_{\text{oxygen}}} \quad (3.9)$$

$$\Delta C_{\text{glucose}} = \Delta C_{\text{oxygen}} (3.3) (1.9)$$

$$\Delta C_{\text{glucose}} = 6.2 \Delta C_{\text{oxygen}} \quad (3.10)$$

Above numerical solution showed that at any floc diameter, the glucose concentration in the suspending medium has to be approximately six times the oxygen concentration for a steady state of glucose respiration in accordance with the available oxygen supply. The estimate made it clear that even at rather small oxygen concentrations in trickling filter fluids or suspensions of activated sludge, an adequate oxygen supply for the organisms may be maintained in view of generally low concentrations of fermentable material in sewage, in addition, this example has been calculated with one compound only and the sum of all respiration substrates should be taken into account in mixed media stated by Wuhrmann (Wuhrmann,1960).

3.2 Mueller Model, 1967

In order to ascertain the importance of oxygen concentration to floc particles, a fundamental knowledge of the mechanisms involved in the supply of oxygen to the organisms within floc particles was found necessary by Mueller (Mueller et al.,1967). His study was aimed to provide such knowledge, and in particular to determine the role of

oxygen diffusion through the floc particles. The study confined entirely to the movement and consumption of oxygen from the liquid immediately adjacent to the floc surface and into its interior. It is stated that to insure elimination of gas transfer in the experiments, bubble aeration not used during kinetic studies. Rather, all oxygen during the studies was supplied from the initial dissolved oxygen inventory of the solution. Diffusion system analyzed by Mueller is shown in Figure 3.3.

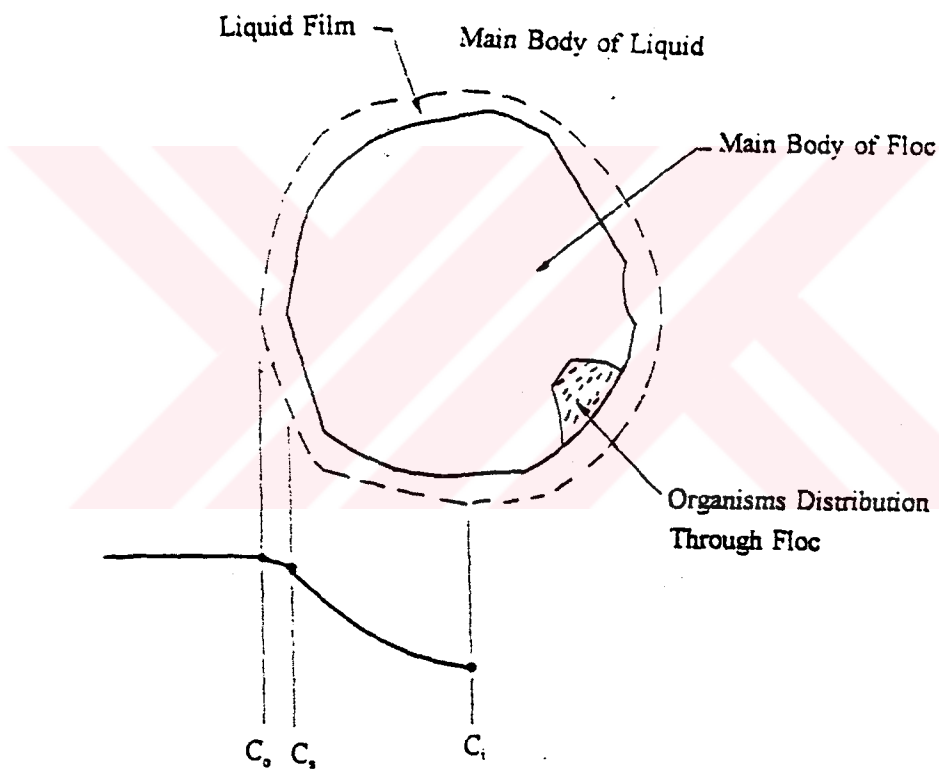


Figure 3.3 Diffusion System Analyzed by Mueller (Mueller et al., 1967)

The continuity equation for oxygen given in the simple form as follows,

$$\frac{\partial C}{\partial t} = D \nabla^2 C - r_f \quad (3.11)$$

where

C is the oxygen concentration at any point in the floc

D is the oxygen diffusivity

r_f is the homogenous oxygen uptake rate of the floc

t is time

By integrating equation (3.11) for suitable geometric shapes and concentrations, the effective diffusivity through floc material could be determined by fitting the integrated equation to kinetic data. To do this however requires techniques for controlled growth of floc particles under sterile conditions and the measurement of the sizes and oxygen uptake rates of the floc. The organisms used in his study was a pure culture floc-forming bacterium identified as *Zooglea ramigera*, strain *I-16 M*. Three methods were used to measure floc size. One method was consisting of microscopically measuring the floc lengths and breaths, but not used for the diffusivity determinations. The other two methods used for the diffusivity determinations were the nominal diameter and specific surface floc sizing methods.

The measured wet density of the *I-16 M* floc 1.09 g/cm^3 and dry density of the floc 1.29 g/cm^3 were used. These densities indicated that the floc contained 63% water used for the nominal diameter determinations. It has been pointed out by Mueller (Mueller et al., 1967) that the nominal diameter yielded reproducible results but measures the maximum possible diffusion dimension of a floc.

An estimate of the average diffusion path of floc particles from the floc surface given as follows:

$$a_s = \frac{1}{\lambda} = \frac{a_t}{v_w} \quad (12)$$

where

a_s : specific surface area of the floc

a_t : total surface area of the floc

v_w : wet volume of the floc

λ : direct function of the diffusion path through the floc

The three curves given in Figure 3.4 represents the maximum, average and minimum values of oxygen uptake rate and specific spherical floc diameter obtained from the study. Under the maximum conditions without agitation, the oxygen concentration drop through the liquid film was only 0.36 mg/l.

It has been concluded that the oxygen concentration drop across the liquid film surrounding the *I-16 M* floc was usually less than 10 percent of the oxygen concentration drop through the floc. This was indicating that the resistance to oxygen transfer caused by the liquid boundary layer around a floc was small when compared to the resistance of the floc. Hence, the assumption of negligible liquid film resistance in the following derivation of the diffusion equations is valid (Mueller et al., 1967).

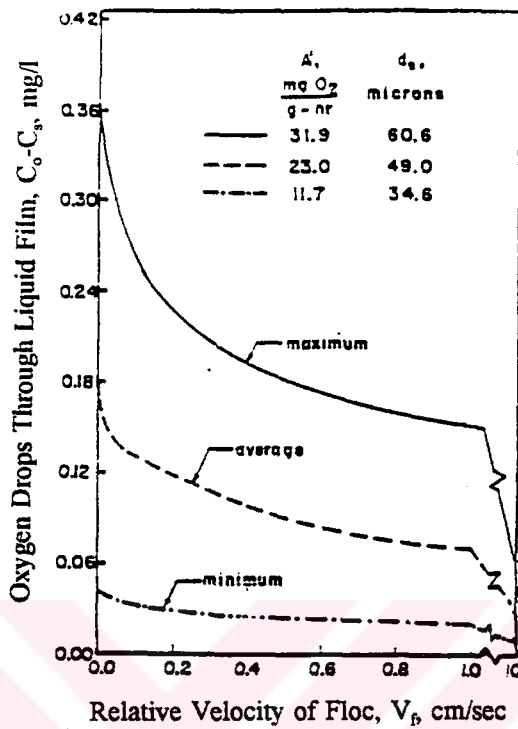


Figure 3.4 Relation between oxygen drops through liquid film and velocity of floc relative to the liquid (Mueller et al., 1967)

The model assumes that the critical oxygen concentration at the surface of the organisms within the floc is zero. For a spherical floc, the anoxic core equations were used to estimate oxygen diffusivity for the nominal diameter, (d_n) floc measurement is given by,

$$\left(\frac{r_{\text{red}}}{r_{\text{max}}} \right)_{d.s} = 1 - (X_c)^3 \quad 0 \leq X_c \leq 1 \quad (3.15)$$

$$\frac{6 D C_s}{(r_{\text{max}})_{\text{wet}} R^2} = 1 - 3X_c^2 + 2X_c^3 \quad 0 \leq X_c \leq 1 \quad (3.16)$$

where

- R : radius of sphere, $R = d_n / 2$
- $r_{\max}$: max. oxygen uptake rate based on dry weight of the floc.
- r_{red} : reduced oxygen uptake rate based on dry weight of the floc.
- r_{wet} : max. oxygen uptake rate based on wet volume of the floc.
- D : diffusivity of oxygen in the floc.
- C_s : oxygen concentration at floc surface assumed equal to the oxygen concentration in the bulk liquid.
- X_c : ratio of radius of the anoxic core to the total floc radius, $X_c = R_c / R$
- R_c : radius of anoxic core

From the study on *I-16 M* organisms, it is stated that, diffusion of oxygen through floc material was the mechanism controlling the oxygen uptake rate of *I-16 M* floc when oxygen becomes limiting. The effect of oxygen concentration on the oxygen uptake rate of biological floc was found to be similar to that shown in Figure 3.5. The curves depicted in this figure for various sizes spherical flocs were calculated by anoxic core equations for the highest oxygen uptake rate obtained in the study. The larger floc particles need a much higher oxygen concentration at their surface than the smaller particles, if the maximum uptake rate was to be maintained. It also concluded that if a system containing biological flocs to be operated at low dissolved oxygen levels, the degree of agitation in the system must be great enough to yield small floc particles (Mueller et al., 1967).

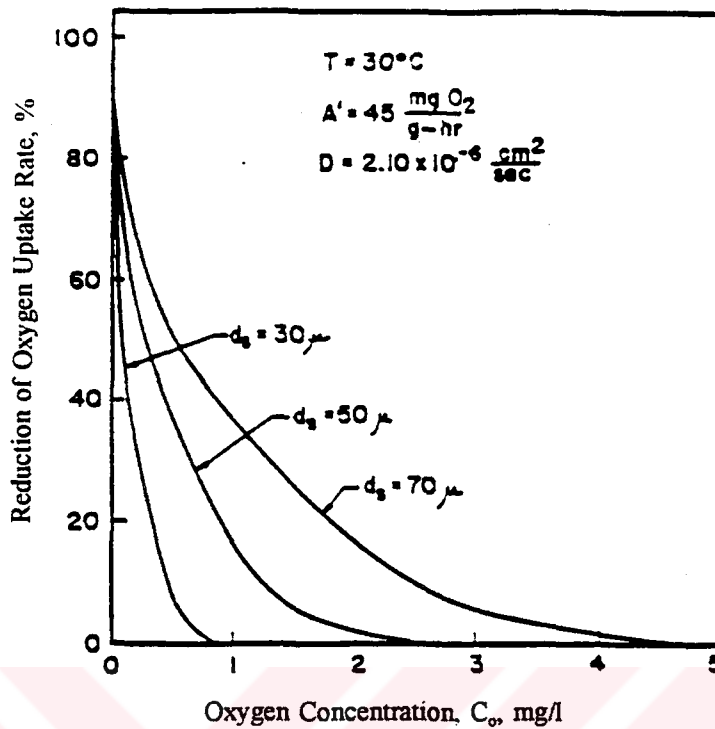


Figure 3.5 Effect of oxygen concentration on oxygen uptake rate (Mueller et al., 1967)

According to Mueller (Mueller et al., 1967) the determinations of the nominal diameter of activated sludge floc permit the estimation of the limiting oxygen concentrations of these flocs. Since oxygen uptake rate decreases with the time of aeration of activated sludge, they have calculated the limiting oxygen concentrations for their nominal diameter measurements as a function of the time of aeration. In their calculations, they used a diffusivity value of $0.5 \cdot 10^{-5} \text{ cm}^2/\text{sec}$ and a critical oxygen concentration of 0.1 mg/l . It is also concluded that, if these activated sludge flocs have the same oxygen transfer resistance as *I-16 M* floc, a value of $0.5 \cdot 10^{-5} \text{ cm}^2/\text{sec}$ seems reasonable to use for the smaller nominal diameters. However, for the nominal diameter of 115 microns , a value of $1.2 \cdot 10^{-5} \text{ cm}^2/\text{sec}$ was recommended as a better value to use. The dashed line in Figure 3.6 shows that using the higher diffusivity value for the larger floc reduces the effect of size on the limiting oxygen concentration (Mueller et al., 1967).

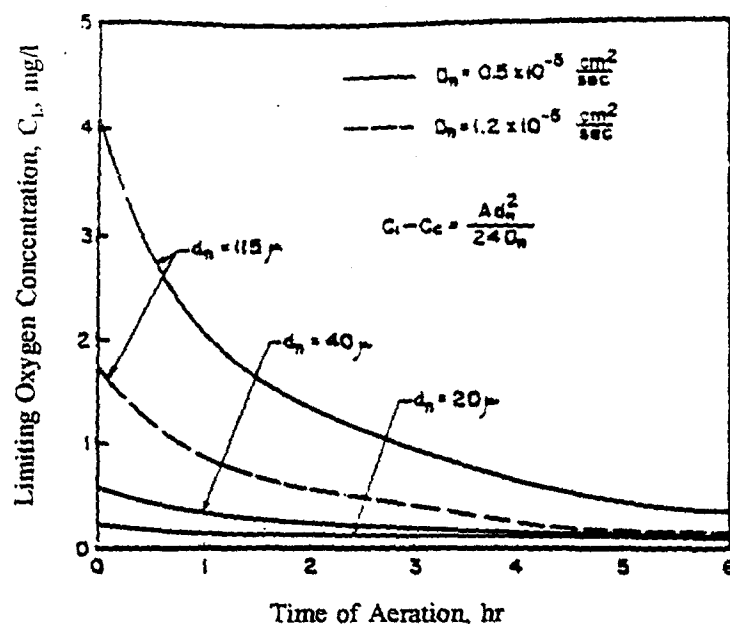


Figure 3.6 Variation of limiting oxygen concentration of activated sludge floc with the time of aeration for plug-flow systems (Mueller et al., 1967)

From Figure 3.6 the degree of agitation normally encountered in activated sludge plant yields floc small enough to allow them to exhibit a maximum oxygen uptake rate at dissolved oxygen concentrations less than 2.0 mg/l in addition, 75 percent of the treatment plants sampled by was containing the floc with nominal diameters less than 43 microns. Thus the oxygen concentration necessary to maintain a maximum oxygen uptake rate in these plants was probably be less than 0.6 mg/l (Mueller et al., 1967).

3.3 Matson and Characklis Model, 1976

To test the significance of sludge age and the growth media conditions on reactant diffusivity and reaction rate in fluidized biological aggregates primarily emphasized in the study of Matson and Characklis (Matson and Characklis, 1976).

The theory applied in the study was based on the experimental apparatus consisting of two well-mixed reservoirs separated by the template sandwich through which substrate diffusing. The membrane on each side of the template was preventing connective molecular transport. The floc in the template sandwich was deactivated, thus no reaction was occurring. A mass balance established over either reservoir is given as follows,

Diffusion flux $\times$ Surface area = Change in reservoir concentration

$$D_s (dC / dX)_s A = V (dC / dT) \quad (3.17)$$

where

A : effective area of transfer in the template sandwich

D_s : overall diffusion coefficient of diffusing species through the template assembly

$(dC/dX)_s$: overall concentration drop across the assembly

V : reservoir volume

C : species concentrations

C_o : initial species

T : time

In the experimental study the two variables at three levels with two different reactants, oxygen and glucose were investigated. The mass fluxes of the reactants through biological material measured in an apparatus consisting of two well-mixed chambers separated by a membrane assembly, a kind of modified filtration apparatus called as Membrane Template Assembly (MTA). The result of the diffusion experiments of the investigators is given in Table 3.1.

Table 3.1 Diffusion Coefficient of Oxygen and Glucose in
Floc from various Biological Wastewater
Treatment Plants (Matson and Characklis, 1976)

Source of Floc Type of Plant	Diffusion Coefficient 10 ⁻⁵ cm ² /sec		Sludge Age (days)
	Oxygen	Glucose	
Municipal (domestic)	1.37 ± 2 %	0.25 ± 20 %	8 - 10
Municipal (Dom + Ind)	1.17 ± 21 %	0.35 ± 10 %	5
Industrial (petrochemicals)	0.85 ± 25 %	0.18 ± 14 %	20

As a result of the experimental study and the analysis of variance indicated that sludge age and C/N ratio had significant effect (95% level) on the diffusion coefficients of oxygen in the biological material. Sludge age and C/N ratio in the growth media also significantly affected the glucose unit reaction rate. The oxygen diffusion coefficient decreased with longer sludge age and lower C/N ratio growth media as indicated in Figure 3.7. The trend for glucose was not consistent. The glucose diffusivity in biological material averaged 30% of that in water, and the glucose unit reaction rate was decreased with longer sludge age and higher C/N ratios as shown in Figure 3.8 and Figure 3.9 respectively. The reactant diffusion coefficient has been found significant if mass transport through the biological aggregate potentially rate limiting. The models have been established by other researchers with referring indicated that overall reaction rates could be sensitive to reactant diffusivity and concluded this was particularly true of fixed films where thick microbial aggregates were likely. The ability of a given reactant to penetrate into a microbial film was found to be function of film thickness, microbial activity, reactant bulk liquid concentration and reactant stoichiometry. Mass transfer limitations are possible when the liquid film or biological film so

thick that the reactant can not penetrate through it; the microbial activity so high that the reactant is consumed in the outer layer or reactant concentration in the bulk liquid so low that insufficient driving gradient is available.

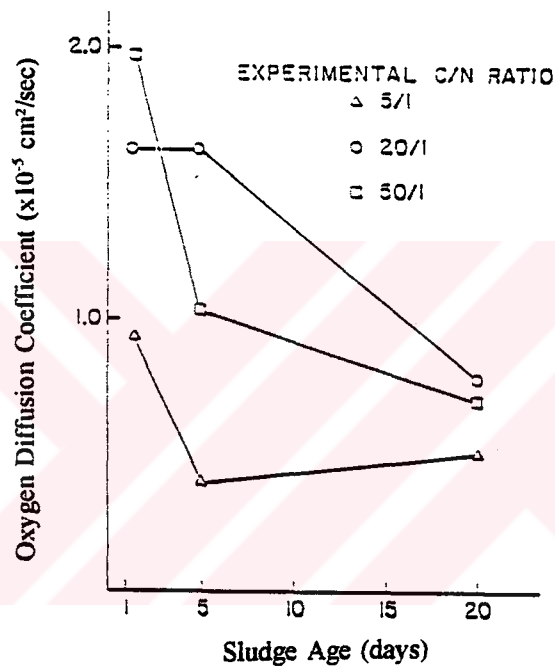


Figure 3.7 Oxygen Diffusion Coefficient as a function of Sludge Age and Carbon to Nitrogen Ratio (Matson and Charaklis, 1976)

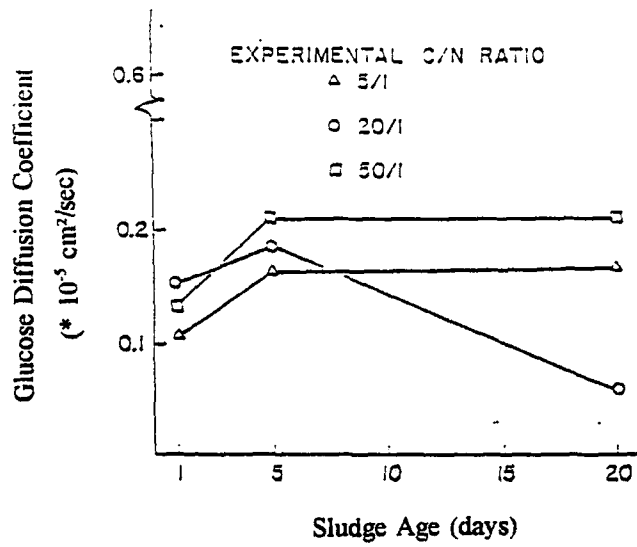


Figure 3.8 Glucose Diffusion Coefficient as a function of Sludge Age and Carbon to Nitrogen Ratio (Matson and Charaklis, 1976)

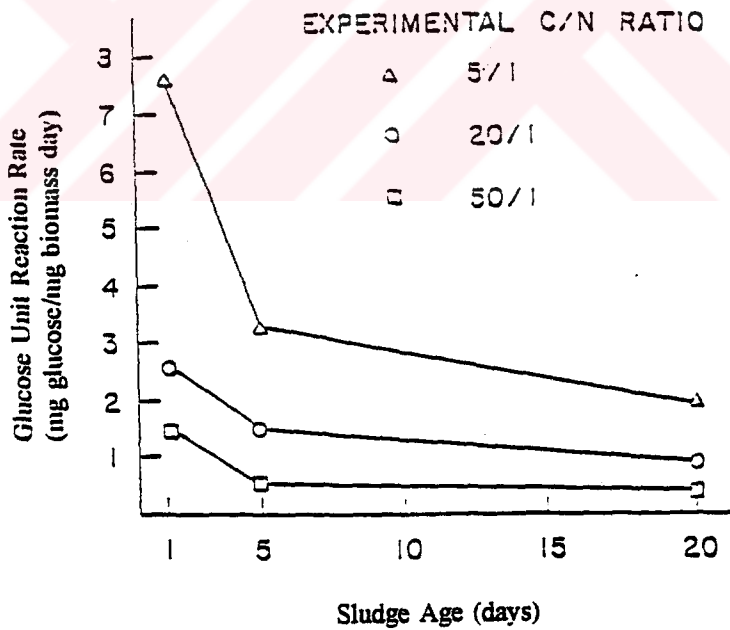


Figure 3.9 Glucose unit reaction rate as a function of sludge age and carbon to nitrogen ratio (Matson and Charaklis, 1976)

In the modeling study, steady state diffusion and reaction of a given substrate through a spherical floc matrix containing a homogenous concentration of micro-organisms are described and expressed in the following equations :

$$\frac{\partial C_s}{\partial t} + V \nabla C_s = D_{sf} \nabla^2 C_s - r_s \quad (3.18)$$

steady state no convection diffusion reaction

where

$$r_s = D_{sf} \nabla^2 C_s$$

In spherical coordinates ;

$$r_s = D_{sf} \left[\frac{\partial^2 C_s}{\partial R^2} + \frac{2}{R} \frac{\partial C_s}{\partial R} \right] \quad (3.19)$$

where

D_{sf} : the diffusivity of a given substrate through the floc matrix

C_s : concentration of substrate

R : radial distance measured from the floc center

r_s : rate of substrate reaction

According to Matson and Characklis (Matson and Characklis,1976) analytical solutions can be obtained if the reaction order considered either zero or first between substrate and the micro-

organisms. The actual reaction order inside the floc matrix is a function of the substrate, organisms species, their physiological state, etc.

For the zero order reaction order case, the reaction rate r_s , can be expressed as follows:

$$r_s = \rho_f k_0 \quad (3.20)$$

where

r_s : substrate reaction rate

ρ_f : floc density

k_0 : zero order reaction rate constant

Integrating of equation (3.20) in equation (3.19) yielded,

$$\frac{\partial^2 C_s}{\partial R^2} + \frac{2 \partial C_s}{R \partial R} = \frac{\rho_f \cdot k_0}{D_{sf}} \quad (3.21)$$

The boundary condition of the left hand-side of the equation (3.21)

1. $C_s = \text{finite for all } R$
2. $C_s = C_f \quad \text{at} \quad R = R_f$

The integration of equation (3.21) yields,

$$C_f - C_s = \frac{\rho_f \cdot k_0}{6 D_{sf}} (R_f^2 - R^2) \quad (3.22)$$

solution of above equation (3.22) for at the point $C_s = 0$ yields,

$$R = \left[R_f^2 - \frac{6 D_{sf} C_f}{\rho_f k_0} \right]^{1/2} \quad (3.23)$$

In this equation the distance, substrate penetrates into the floc particle is defined as $(R_f - R)$. The penetration of oxygen or any substrates into a floc particle calculated from given reaction rate constant (k_0), the diffusivity (D_{sf}) and the size and density (ρ_f) of the floc particle. The radial distance from the center of spherical floc oxygen penetrates is set to equal zero, then new equation is expressed as follows;

$$R = \left[R_f^2 - \frac{6 D_{sf} C_f}{\rho_f k_0} \right]^{1/2}$$

$$R = 0 \quad C_f = C_s$$

$$R_f = \left[\frac{6 C_s D_{sf}}{\rho_f k_0} \right]^{1/2}$$

$$C_s = R_f^2 \frac{\rho_f k_0}{6 D_{sf}} \quad (3.24)$$

where

R_f : radius of floc

As a result of the study, an illustrative graphic has been prepared by Matson and Characklis (Matson and Characklis, 1976) that shows oxygen diffusion limitations as a function of the various floc sizes

with using of zero order reaction order case for oxygen transfer, is given in Figure 3.10.

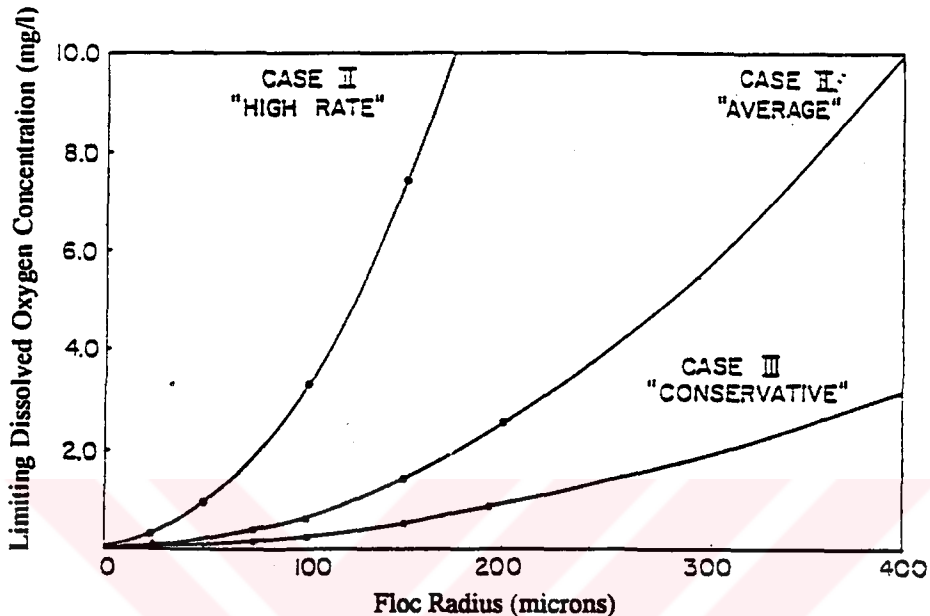


Figure 3.10 Oxygen Diffusion Limitations as a Function of Floc Size (Matson and Characklis, 1976)

Oxygen and glucose diffusion coefficients in biological floc were measured and found to be lower than their values in pure water in most cases. This indicated that potential oxygen and substrate limitations in biofilms were exist. A simple zero order diffusion-reaction model constructed by Matson and Characklis (Matson and Characklis, 1976) showed that the potential for oxygen diffusion limitations in "high" rate treatment process and "average" rate treatment process containing large floc particles also possible in activated sludge. In Figure 3.10, Case I shows the parameters D_{sf} , ρ_f and k_0 that may have represents those found in the "average" aeration basin. Case II represents an estimate of the "high rate process". Case III represents the most "conservative" estimates of the process parameters.

For Case I, the oxygen diffusivity was taken as an average of the actual field measurements, and the oxygen may become limiting at larger floc sizes in conventional activated sludge systems, which are usually designed to maintain a minimum of 2.0 mg/l dissolved oxygen in the basin .

For Case II, the rate constant was double that of the other cases; the density was highest and diffusivity the lowest. This case was the illustration of the most severe oxygen diffusion limitations. Even at small floc sizes (r less than 100 microns) the minimum dissolved oxygen concentration required to keep the floc aerobic was greater than 3.0 mg/l .

For Case III, the most conservative estimates of parameters were made and oxygen diffusion limitations will probably not result in this situation irrespective of floc size.

It is stated by Matson and Characklis (Matson and Characklis, 1976) that, many assumptions have been made to arrive at these conclusions, adding that which may or not be valid, and would affect the results, assumed that the electron donor substrate was not diffusion limited in the floc particle. In completely stirred tank reactors, the substrate concentration throughout the basin is low, and may, indeed, only diffuse into the outer shell of floc. In this case the substrate will be diffusion limited. The stoichiometry of the reaction as well as the respective oxygen and substrate concentration will influence the true diffusion regime in the floc particles. On the other hand the non-spherical nature of the floc particles was probably the most serious limitations to the model. Any shape irregularity will increase the surface area to volume ratio and thereby lower the potential for diffusion limitations was suggested. Lastly the zero order reaction assumption created a sharp division between an aerobic zone and anaerobic zone in a floc particle. A first order rate assumption forces an exponential decrease in oxygen concentration approaching but not reaching zero at

the floc center. Both zero order and first order reaction zones probably were exist in a floc particle. A Monod type hyperbolic relationship may be most appropriate.

3.4 Heide and Kruize Model, 1979

A model for the oxygen transfer in activated sludge also presented by Kossen (Kossen, 1978), and experimentally investigated by Heide and Kruize (Heide and Kruize, 1979). In this study the determination of the so-called critical oxygen emphasized, which was the oxygen concentration in the sludge suspension, at which the core of the floc is just not yet becoming anaerobic. The activated sludge flocs were assumed spherical and also homogenous in composition. No distinction was made between microflocs and macroflocs. In the modelling study, the simple oxygen transfer model considered is given in Figure 3.11.

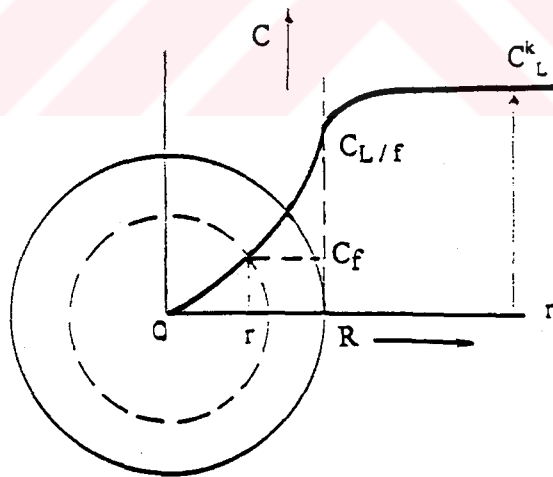


Figure 3.11

Oxygen transfer at the critical oxygen concentration C_L^k in activated sludge (liquid phase).
(Heide and Kruize, 1979)

The oxygen consumption rate is described by zero order reaction in oxygen if no limitation of mass transfer occurs. At a distance, r in the sludge floc it is defined as follows :

$$r_f = - \frac{dC_f}{dr} \quad (3.25)$$

where

r_f : oxygen consumption rate in the floc

C_f : oxygen concentration at the radial distance, r in the sludge floc.

In the mass balance situation the oxygen consumption in the inner sphere with the volume is equal to the oxygen supply through the surface area of the inner sphere. Oxygen supply takes place by diffusion, so that ;

$$r_f v_i = a_i \frac{dC_f}{dr} D_f \quad (3.26)$$

where

v_i : volume of inner sphere

a_i : surface area of inner sphere

D_f : molecular diffusion coefficient of oxygen in the floc

For volume and surface area of inner sphere, below equations is written as follows ;

$$v_i = \frac{4}{3} \pi r^3 \quad (3.27)$$

$$a_i = 4 \pi r^2 \quad (3.28)$$

substituting of equations (3.27) and (3.28) into equation (3.26) which gives,

$$r_f \frac{4}{3} \pi r^3 = 4 \pi r^2 \frac{dC_f}{dr} D_f \quad (3.28)$$

with rearranging of equation (3.28) results ;

$$r_f \frac{1}{3} r = D_f \frac{dC_f}{dr} \quad (3.29)$$

by integration of equation (3.29),

$$r_f \frac{1}{3} \int_0^r r dr = D_f \int_0^C \frac{dC_f}{dr}$$

solution for above equation at the boundary conditions ;

1. C_f finite for all r
2. $C_f = C_{L/f}$ at $r = R$

$$C_f = \frac{r_f}{6D_f} r^2 \quad (3.30)$$

than equation (3.30) now becomes;

$$C_{L/f} = \frac{r_f}{6 D_f} R^2 \quad (3.31)$$

where

$C_{L/f}$: oxygen concentration at the surface of the floc-liquid interface

R : radial distance in the floc.

The oxygen concentration at the surface of the floc-liquid interface, $C_{L/f}$ is in equilibrium with the critical oxygen concentration in the fluid denoted by C_L^k . The oxygen supply from the fluid to the surface of the sludge floc is expressed by ;

$$r_f = K_L a_f (C_L^k - C_{L/f}) \quad (3.32)$$

where

K_L : mass transfer coefficient

a_f : specific surface area of the sludge floc

Specific surface area of the sludge floc in spherical form is expressed as follows;

$$a_f = A_f / V_f \quad (3.33)$$

where

V_f : volume of sludge floc

A_f : surface area of sludge floc

The surface area and volume of the sludge in equation (3.33) can be written by the form;

$$A_f = 4\pi R^2 \quad (3.34)$$

$$V_f = 4/3 \pi R^3 \quad (3.35)$$

from the definition of the specific surface, in equation (3.33),

$$a_f = 3/R \quad (3.36)$$

Assuming the difference in velocity of sludge floc and that of the surrounding fluid is very small, thus, the mass transfer coefficient, K_L can be derived from the Sherwood number. For the so called rigid spherical model the Sherwood number is given by;

$$Sh = 2 R K_L / D_L = 2 \quad (3.37)$$

or,

$$K_L = D_L / R \quad (3.38)$$

where

D_L : diffusion coefficient of oxygen in the fluid

by combining of the equation (3.31) and (3.38) and after substituting of the equation (3.36),

$$r_f = (D_L / R) (3 / R) (C_L^k - C_L / f)$$

$$r_f = 3 D_L / R^2 (C_L^k - C_L / f)$$

$$C_L^k = (r_f R^2 / 3 D_L) + C_L / f \quad (3.39)$$

by substituting equation (3.31) into the equation (3.38) it can be arrived to;

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] r_f R^2 \quad (3.40)$$

where

C_L^k : the critical oxygen concentration in the sludge suspension at which just not yet occurs the anaerobic state ($M L^{-3}$)

D_L : diffusion coefficient of oxygen in the sludge suspension ($L^2 T^{-1}$)

D_f : diffusion coefficient of oxygen in the sludge floc ($L^2 T^{-1}$)

R : radius of the sludge floc (L)

r_f : oxygen consumption rate in the sludge flocs ($ML^{-3}T^{-1}$)

According to Heide and Kruize (1979), the oxygen consumption rate in the sludge floc, r_f can be derived from the oxygen consumption rate in the sludge suspension. They introduced that the oxygen consumption rate in sludge flocs could be determined in the respiration measurements used in the experimental study, and expressed as follows;

$$r_f = r_L \frac{\text{total volume}}{\text{wet sludge volume}} \quad (3.41)$$

where

r_L : oxygen consumption rate in the sludge suspension ($ML^{-3}T^{-1}$)

It was concluded that, the wet sludge volume depends on the porosity of the sludge flocs. The looser the structure of the flocs, the larger the volume of the sludge will be. As an indication for the wet sludge volume was the sludge volume index, (SVI). This index is represented the volume of the sludge settled per gram of dry matter. In addition the constant volume was found after 40 minutes settling time, but without any explanation. They combined the SVI and the dry sludge content and obtained the wet sludge volume, together with equation (3.41) this led them to arrive;

$$r_f = \frac{r_L}{SVI * d_s} \quad (3.42)$$

where

SVI = sludge volume index ($L^3 M^{-1}$)

d s = dry solids content of the sludge ($M L^{-3}$)

Combining of the equations (3.40) and (3.41) it can be arrived to ;

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] \frac{r_L}{SVI \cdot d s} R^2 \quad (3.43)$$

From the above equation (3.43), for a given activated sludge mixture, assuming $SVI \cdot ds = \text{constant}$ and at a given temperature D_f and D_L constant, the critical oxygen concentration dependent exclusively on the oxygen consumption rate and the diameter of the sludge can be calculated if C_L^k , r_L and dry solids content are measured and the diffusion coefficients estimated. For calculation of the theoretical critical oxygen concentration, the rounded values of $D_L = 2 \cdot 10^{-9} \text{ m}^2/\text{s}$ and $D_f = 1 \cdot 10^{-9} \text{ m}^2/\text{s}$ has been used with referring to Matson's study (Matson and Characklis, 1976) that have been proposed for the different shapes for various types of biomass .

Further it was assumed that for well settleable sludge $SVI = 75 \text{ ml/g}$ and $d.s = 4 \text{ g/l}$, introducing also a mean sludge diameter it is arrived to below equation,

$$C_L^k = 3 \cdot 10^8 r_L d^2 \quad (3.44)$$

According to Heide and Kruize (Heide and Kruize, 1979) the oxygen consumption rate in the sludge suspension, r_L is between $2\text{-}20 \text{ g O}_2/\text{kg dry solid hr}$ or at 4 g ds/m^3 between globally $2\text{-}20 \cdot 10^{-6} \text{ kg O}_2/\text{m}^3 \text{ sec}$ depends on the average and/or momentous sludge load. In the range of oxygen consumption rates assumed by the investigators critical oxygen concentrations for floc diameters between 0.1 and 2 mm have been calculated and the results are shown in Figure 3.12

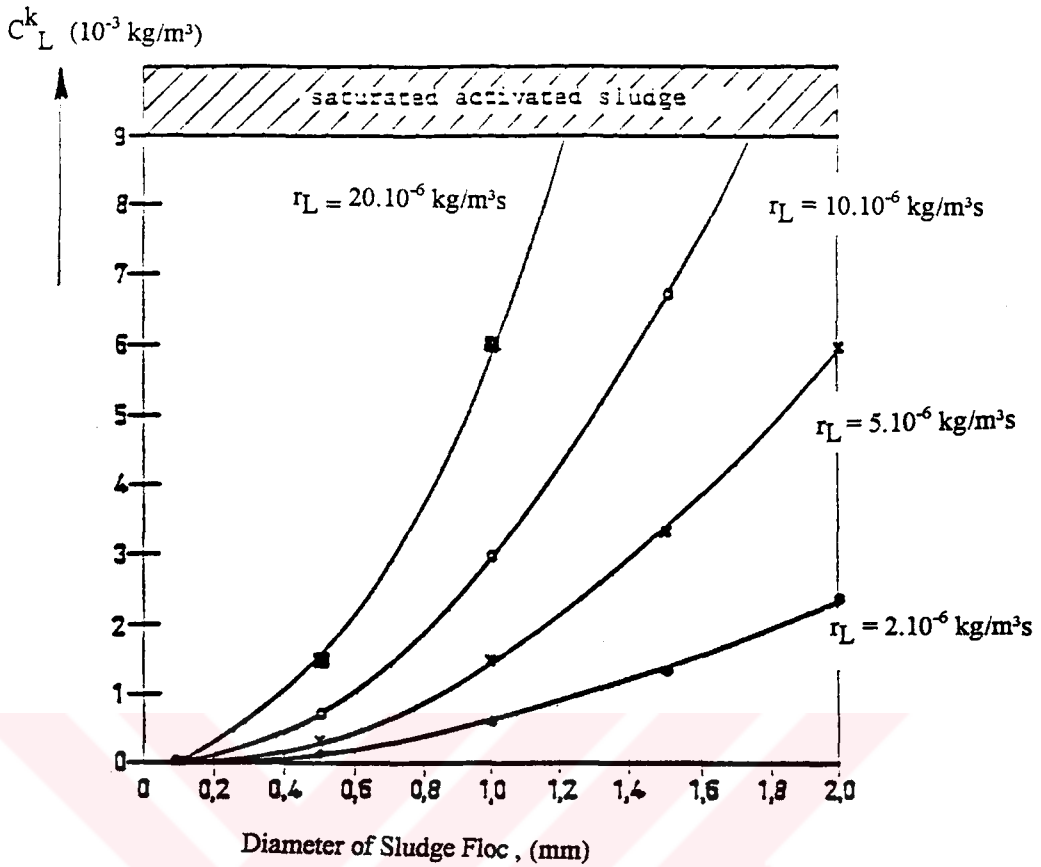


Figure 3.12 Critical oxygen concentration C_L^k as a function of the floc diameter, (d) and the oxygen consumption rate of the sludge suspension, r_L (SVI $d_s = 0.35$) (Heide and Kruize, 1979)

3.5 Smith and Coackley Model, 1984

In order to obtain a more complete understanding of diffusivity coefficients in activated sludge, it was found necessary to make further studies into diffusivity and the pore structure of activated sludge by Smith and Coackley (Smith and Coackley, 1984). A diffusion cell used in their study is shown in Figure 3.13. One side of the cell was continually fed with pure oxygen to maintain oxygen saturation. The other side contained dissolved oxygen meter and a magnetic follower with a magnetic stirrer below. The center compartment separated from

the two outer compartments by oxygen permeable membranes. The diffusion cell calibrated by filling with de-ionized water. A cross member fitted between the two membranes to stop them distorting .

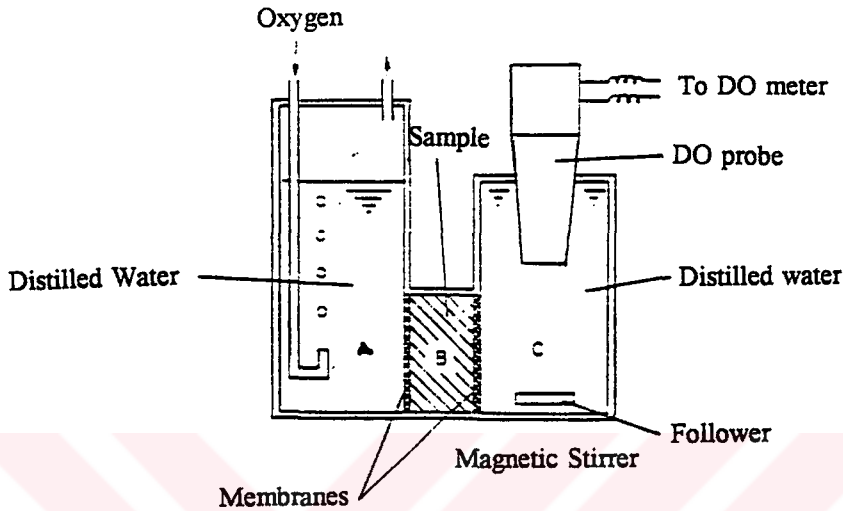


Figure 3.13 Diagram of diffusion cell
(Smith and Coackley, 1984).

The diffusion flux, F is calculated from the below equation,

$$F = \frac{DO_{t1} - DO_{t2}}{t_1 - t_2} \cdot \frac{1}{SA} \quad (3.45)$$

where

DO_{t1} : dissolved oxygen in compartment C at $t = t_1$

DO_{t2} : dissolved oxygen in compartment C at $t = t_2$

V : volume of compartment C

SA : surface area of membrane between compartments A and B.

The definition of diffusivity or diffusion coefficient derived from the concept that, in an isotropic substance, the rate of transfer of

diffusing substances through a unit area of a section was proportional to the concentration gradient measured normal to that section,

$$F = -D \frac{\partial c}{\partial n} \quad (3.46)$$

where

F : diffusion flux, mass of oxygen/unit area of section unit time.

C : concentration of diffusing substances, dissolved oxygen concentration

D : diffusivity or diffusion coefficient

n : distance measured normal to the area of section.

Diffusion coefficients often vary with the concentration of diffusion substance but for dilute solutions the dependence is slight and can be assumed to be constant for practical purposes. Hence D can be calculated from equations (3.45) and (3.46).

$$D = \frac{F d}{C_{\text{sat}} - C_{\text{av}}} \quad (3.47)$$

where

C_{sat} : pure dissolved oxygen saturated concentration in water

C_{av} : average dissolved oxygen concentration in compartment C

d : distance across cell

As one of the result of this study, the diffusion cell gave a value of oxygen diffusivity in pure water of $2.25 \times 10^{-9} \text{ m}^2/\text{s}$ with $SD = \pm$

$0.20 \cdot 10^{-9} \text{ m}^2/\text{s}$. This was compared with the published value of Sheerwood of $2.1 \cdot 10^{-9} \text{ m}^2/\text{s}$ at 20°C (Smith and Coockley, 1984).

The activated sludge thickened by gravity up to about 3 % solids but centrifugation was found necessary to increase the solids concentrations any further. Four different sludge samples were used. The change of the diffusivity with concentration is shown in Figure 3.14. It was interesting that, over a solid range of about 3.0 - 9.0 % solids, the values of diffusivity were very similar. The average diffusivity value for the four sludges was found as $7.42 \cdot 10^{-10} \text{ m}^2/\text{s}$.

In order to establish typical values of the diffusivity in microscopic flocs, it was found necessary by Smith and Coackley (Smith and Coackley, 1984) to relate diffusivity measurements on relatively large values of thickened sludge to the microscopic floc found in aeration tanks, and difficult to measure diffusivity in a small floc size of 10 - 100 μm diameter, especially when the size and shape of a given floc in an aeration tank were probably not static. The relevance of the diffusivity measurements to flocs established by examining of the similarity in physical structure of thickened sludge and the flocs. In particular, diffusivity was closely related to the pore structure of a heterogeneous solid and a comparison of the pore structure in flocs and thickened sludge would assist in relating the diffusivity measurements to activated sludge flocs. Similarities in the physical structure were density, specific surface area, mean pore radius and tortuosity.

As internal geometries of porous solids are poorly understood, an empirical factor often introduced to quantify the nature of the porous solid. The mean pore radius, R_p and its definition by Satterfield (Satterfield, 1970) were adopted for activated sludge and expressed by Smith (Smith, 1984) as follows;

$$R_p = \frac{2 \epsilon}{S_g \rho_f 10^6} \quad (3.48)$$

where

R_p : the mean pore radius (m)

S_g : surface area per unit mass of porous media (m^2/g)

ρ_f : bulk density of floc (g/cm^3)

ϵ : porosity of floc

$$\text{and } \epsilon = 1 - \frac{X_f}{\rho_s} \quad (3.49)$$

where

X_f : solid concentration of floc (g/cm^3)

ρ_s : density of dried sludge (g/cm^3)

The factor of 2 in equation (3.49) was derived from assuming pores are cylindrical or silt shaped and do not intersect. Broeckhoff and Linsen (Broeckhoff and Linsen, 1970) considered a factor of 2.5 to be appropriate, which is in between considering the pores to be cylindrical and spherical. Purely spherical pores that give a factor of 3. Hence equation (3.49) for purely spherical pore,

$$R_p = \frac{3 \epsilon}{S_g \rho_f 10^6} \quad (3.50)$$

Surface area of porous solids measured by dye adsorption techniques. A method for applying these techniques to activated sludge discussed by Smith and Coackley (Smith and Coackley, 1983). The method used in these experiments was for measuring the surface area of

activated sludge and thickened sludge from the same sample. The results are shown in Figure 3.15 of the dye adsorption per unit mass of mixed liquor and centrifuged sludge was similar, and indicated that the specific surface area of the sludge has not been altered as the sludge thickened. Thus the geometry of the pores in activated sludge has not been changed in thickening the sludge. (Smith and Coackley, 1984)

The solids concentration of an activated sludge floc, X_f is, calculated from, ρ_f and ρ_s with using of below equation ,

$$X_f = \frac{\rho_f - 1}{\rho_s - 1} \rho_s \quad (3.51)$$

It has been stated by the investigators that, the actual value of R_p was only approximate due to the assumptions in calculating the specific surface area and density of flocs. The relevance of these values of R_p was discussed together with tortuosity. For diffusion in a heterogeneous solid, the diffusion rate would be affected by the tortuous path that the diffusing molecules have to follow due to the irregular pattern of the solid. The geometry of the pore structure makes it impossible to determine accurately the effective length of the diffusion path. Interconnections within the pore structure, the tortuous character of the pores, variations in the cross-section along the pores all contribute to the complex geometry. In heterogeneous solids it is usual to characterize mass transport phenomena in terms of effective diffusion coefficients so diffusion calculations are based on pores plus solids. The difference between the effective diffusivity in the sludge and the diffusivity in pure water is related to the tortuosity and the available liquid in the sludge,

$$\text{Relative diffusivity ratio} = \frac{\text{effective diffusivity in sludge } \varepsilon}{\text{diffusivity in pure water } \tau} = \frac{\varepsilon}{\tau} \quad (3.52)$$

where

ϵ is the porosity i.e. a measure of the total available liquid

τ is the tortuosity factor, which will be unity if there is no tortuosity and increase as the diffusion path becomes more tortuous.

Theoretical models of tortuous pores have been used to estimate the value of tortuosity factor (τ) by Smith and Coackley (Smith and Coackley, 1984) with referring to Study of Satterfield (Satterfield, 1970). Tortuosity in activated sludge calculated using equations (3.50) and (3.52) and the diffusivity results is shown in Table 3.2. The change in tortuosity factor (τ) with solid concentrations is shown in Figure 3.16.

As a results of this study below conclusions have been made by Smith and Coackley (Smith and Coackley, 1984) which are as follows ;

The diffusivity of activated sludge is found to be $7.4 \cdot 10^{-10} \text{ m}^2/\text{s}$. Pore morphology and diffusivity are closely related in heterogeneous solids, as a comparison of the pore structure of thickened sludge and activated sludge flocs is required in order to relate the diffusivity values from the macro to micro scale.

In activated sludge, the mean pore radius was found to be 108-130 Å. The tortuosity between 3-9 % solids concentration was 2.73. The calculated specific surface area and mean pore radius was similar in the mixed liquor and thickened activated sludge taken from the same sample. The tortuosity of activated sludge was relatively constant in the solid range 3-9 % solids.

From the general discussion of density, porosity, specific surface area, mean pore radius and tortuosity of activated sludge, the pore structure hence the diffusivity, are similar in thickened sludge and activated sludge flocs. Thus the value of diffusivity of $7.4 \cdot 10^{-10} \text{ m}^2/\text{s}$ which was

obtained for the bulk thickened activated sludge can be adopted for the diffusivity in activated sludge flocs concluded by the investigators.

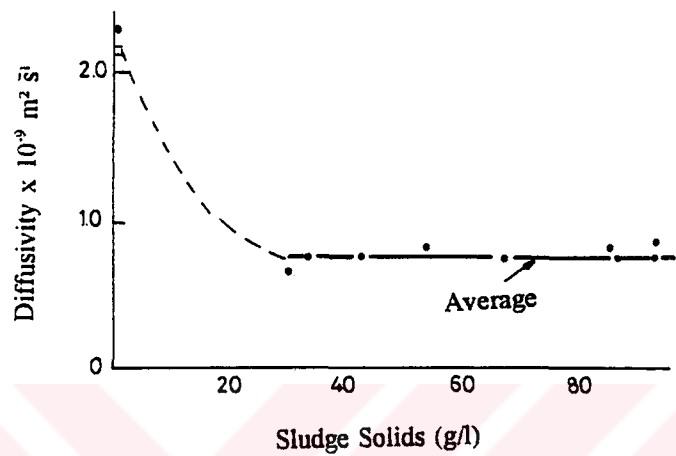


Figure 3.14 Effect of concentration of sludge solids on diffusivity (Smith and Coackley, 1984)

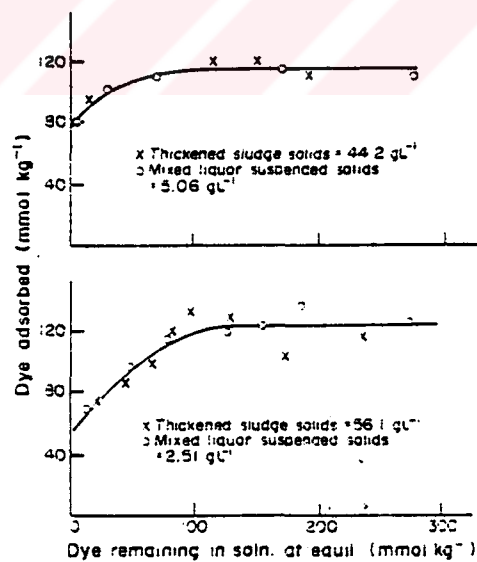


Figure 3.15 Comparison of adsorption of the dye Lissamine Scarlet 4R onto mixed liquor and thickened sludge (Smith and Coackley, 1984)

Table 3.2 Diffusivity and Tortuosity Results
(Smith and Coackley, 1984)

Sample	Solids Concentration (g/l)	Effective Diffusivity 10^{-10} (m ² /s)	Tortuosity
A	84.0	7.93 8.40 8.42 7.96 7.74 -	2.44
Mean		8.10	2.44
B	45.7 65.1 32.0 52.8 85.4	6.68 6.97 7.01 7.77 6.91	3.04 2.90 2.93 2.60 2.93
		6.93 8.18 8.14 7.67	2.42 2.48
C	42.0	7.64	2.66
D	29.4	6.62 7.36 -	2.94
Mean		7.42	2.73
SD		± 0.56	± 0.23

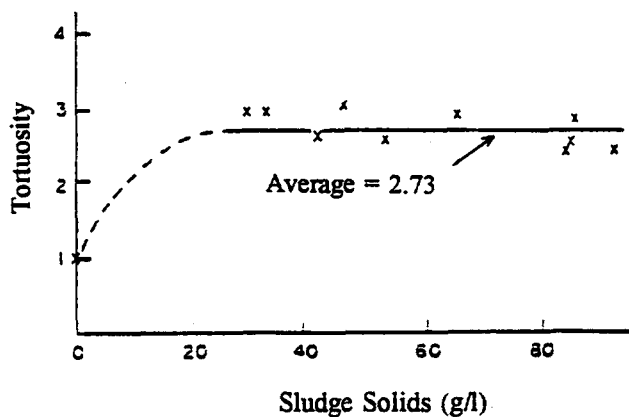


Figure 3.16 Relationship between tortuosity and sludge solids
(Smith and Coackley, 1984)

CHAPTER 4. A NEW OXYGEN MODEL FOR MINERALIZATION AND NITRIFICATION IN AEROBIC SUSPENDED CULTURE SYSTEMS

The model in this study is partly based on Heide and Kruize's Model reviewed in Chapter 3.4. A new approach with some modifications and additions is described. It can be concluded that the proposed model would be adequate for explanation of the meaningful oxygen transfer mechanisms in mineralization and nitrification which takes place in aerobic suspended culture system of biological wastewater treatment.

4.1 Geometric Representation

In the light above explanation, the proposed model for mineralization and nitrification considers the simple diffusion model for flocs is represented by Figure 4.1:

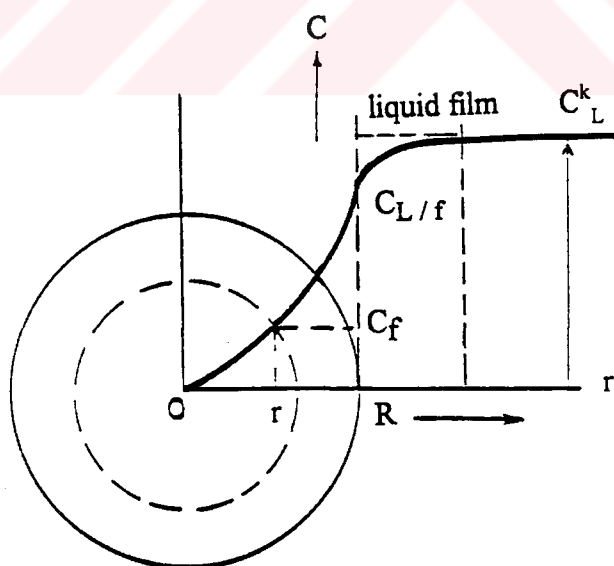


Figure 4.1 Simple Diffusion Model for Single Floc in Mineralization and Nitrification Used in the Modelling Study

Assumptions have been made for this model are as follows:

- 1) Flocs are in spherical form
- 2) Bulk concentration is diffusing first into the liquid film than into the floc
- 3) Concentration profile in the liquid film is linear
- 4) Concentration profile inside the floc is represented by equation given below

$$D_f \frac{dC_f}{dr} a_i = r_f v_i$$

where

D_f : molecular diffusion coefficient of oxygen in the floc

C_f : oxygen concentration at the radial distance, r in the sludge floc

r_f : oxygen consumption rate in the flocs

r : radial distance in the sludge floc

a_i : surface area of inner sphere

v_i : volume of inner sphere

- 5) steady-state conditions
- 6) endogeneous respiration of cells not considered
- 7) BOD means carbonaceous BOD_s

4.2 Reaction Rate for Mineralization and Nitrification

Considered equation as a starting point of the model is as follows;

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] r_f R^2 \quad (4.1)$$

where

C_L^k : the critical oxygen concentration in the sludge suspension at which just not yet occurs the anaerobic state ($M L^{-3}$)

D_L : diffusion coefficient of oxygen in the sludge suspension ($L^2 T^{-1}$)

D_f : diffusion coefficient of oxygen in the sludge floc ($L^2 T^{-1}$)

R : radius of the sludge floc (L)

r_f : oxygen consumption rate in the sludge flocs ($M L^{-3} T^{-1}$)

The oxygen consumption rate of the flocs in aerobic biological process for zero-order reaction case can be defined;

$$r_f = r_o = k_o \quad (4.2)$$

where

k_o : reaction rate constant of oxygen in the floc.

For such a case, the definition of the meaningful oxygen rate constant with covering the operational parameters in suspended culture system which can take place and effects the transfer mechanisms in reaction is necessary.

One of the most important wastewater characteristic which can be considered in the environmental engineering point of view in aerobic biological treatment process is chemical oxygen demand COD, and the other is biochemical oxygen demand BOD. Therefore, COD removal rate constant in terms of oxygen for zero order reaction can be expressed by:

$$k_o = \frac{(S_o - S_e)_{\text{COD}}}{\theta} \quad (4.3)$$

where

S_o : influent COD concentration (M L⁻³)

S_e : effluent COD concentration (M L⁻³)

θ : hydraulic retention time (T)

The reaction rate constant k_o , in oxygen transfer for zero order reaction case is one of the meaningful rate constant for COD removal in aerobic biological treatment process. Normally, it is directly related to the measurable chemical oxygen demand parameter, which is easy and very fast to determine its value when it is compared with the others. However, it is known that sometimes very difficult to determine its exact value in the effluent because of the higher value may be encountered by the reason of the internal cell degradation.

With the same approach the biological oxygen demand BOD removal rate in terms of oxygen for zero order reaction can be expressed as follows:

$$r_f = r_o = k_{\text{BOD}} \quad (4.4)$$

$$k_o = k_{\text{BOD}} \quad (4.5)$$

where

k_{BOD} : BOD removal rate constant for zero order reaction

the new equation for oxygen rate constant accordingly becomes,

$$k_o = \frac{v (S_o - S_e)_{\text{BOD}}}{\theta} \quad (4.6)$$

where

S_o : influent BOD concentration (M L^{-3})

S_e : effluent BOD concentration (M L^{-3})

θ : hydraulic retention time (T)

v : stoichiometric oxygen consumption ratio for BOD (MM^{-1})

4.3 F/M Ratio and Removal Efficiency

Taking into account that the flocs which have been formed from the biological microorganism mass are playing key role in the transfer mechanisms, such a result of this fact, the oxygen reaction rate constant has a direct relationship with the microorganisms is defined as follows;

$$k_o = k_{\text{BOD}} = k_f X_{\text{MLSS}} \quad (4.7)$$

$$k_f = \frac{v (S_o - S_e)_{\text{BOD}}}{\theta X_{\text{MLSS}}} \quad (4.8)$$

$$k_o = \frac{v (S_o - S_e)_{\text{BOD}}}{\theta X_{\text{MLSS}}} X_{\text{MLSS}} \quad (4.9)$$

Considering the biological oxygen demand BOD load as a single substrate in the biological removal process where the mass of substrate loading per mass of microorganism F/M , in the aeration tank is known by;

$$F/M = \frac{Q (S_o)_{BOD}}{V X_{MLSS}} = \frac{(S_o)_{BOD}}{\theta X_{MLSS}} \quad (4.10)$$

where

F/M : food to microorganisms ratio	(T^{-1})
Q : inflow rate to aeration tank	$(L^3 T^{-1})$
S_o : influent BOD concentration	(ML^{-3})
V : Volume of tank	(L^3)

In the environmental engineering, it is well known that for any substrate, substrate removal efficiency in any reaction can be defined by;

$$\xi = \frac{S_o - S_e}{S_o} \quad (4.11)$$

where

ξ : removal efficiency of substrate (%)

4.4 Diffusion and Reaction

Reconsidering of the equation (4.9), this can be rearranged for oxygen rate constant in the form of the below equation for BOD.

$$k_o = \frac{v ((S_o - S_e) / S_o)_{BOD}}{(\theta X_{MLSS} / S_o)} X_{MLSS} \quad (4.12)$$

$$k_o = \frac{v(\xi)_{\text{BOD}}}{(1/F/M)_{\text{BOD}}} X_{\text{MLSS}} \quad (4.13)$$

$$k_o = [v(\xi)_{\text{BOD}} (F/M)_{\text{BOD}}] X_{\text{MLSS}} \quad (4.14)$$

4.5 Oxygen Model for Mineralization

With substituting of the above equation (4.14) into the equation (4.1) results ;

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] R^2 [v(\xi)_{\text{BOD}} (F/M)_{\text{BOD}}] X_{\text{MLSS}} \quad (4.15)$$

In the above equation (4.15), the radius of floc R, can be written for the diameter of spherical floc form;

$$R = \frac{d}{2} \quad (4.16)$$

where

d : diameter of floc (L)

Finally, substituting of equation (4.16) into the equation (4.15) it is arrived to the general model equation for mineralization as follows :

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] \frac{d^2}{4} [v(\xi)_{\text{BOD}} (F/M)_{\text{BOD}}] X_{\text{MLSS}} \quad (4.17)$$

It is reminded that, the proposed oxygen model for mineralization is valid with the below limitations which are as follows ;

- The BOD removal efficiency in the system defined with the term ξ_{BOD} , valid if the effluent BOD concentration S_e , is equal or higher than 19.7 mg/l to prevent from nitrification (Kubota and Fujie, 1977; Gönenç and Harremoes, 1990).
- The model assumes that nutrient concentration exists in the system is enough for the biological growth of microorganism .
- The model considers that, the zero order reaction takes place for oxygen (Matson and Characklis, 1976., Heide and Kruize, 1979., Gönenç and Harremoes, 1990) so far there is no substrate limitation in the bulk liquid and flocs.

4.6 Oxygen Model for Mineralization and Nitrification

As it was shown by Kubota and Miyaji when the effluent BOD concentration falls less than a certain value, the nitrification of ammonia nitrogen proceeds abruptly (Kubota and Fujie, 1977). Under such conditions the oxygen transfer due to nitrification must also be taken into account in the oxygen transfer mechanisms.

If completely mixed flow of the liquid in the aeration tank is assumed, the biological ammonia nitrogen removal rate for zero order reaction can be expressed as follows;

$$r_N = k_N \quad (4.18)$$

When the nitrification occurs in the system the equal amount of the oxygen must be transferred into the biological flocs in zero order reaction can be expressed by ;

$$r_f = r_0 = r_N = k_N$$

$$r_0 = k_0$$

$$r_f = k_0 = k_N$$

$$k_0 = (k_{0 \text{ BOD}} + k_{0 \text{ NH}_4^+ \text{-N}}) \quad (4.19)$$

where

$$k_{0 \text{ BOD}} = \frac{v (S_0 - S_e)_{\text{BOD}}}{\theta}$$

$$k_{0 \text{ NH}_4^+ \text{-N}} = \frac{v_N (N_0 - N_e)_{\text{NH}_4^+ \text{-N}}}{\theta}$$

On the other hand, with considering of the equations (4.7) and (4.8), the new equation for nitrification can be defined as follows:

$$r_f = k_f \cdot X_{\text{MLSS}}$$

$$k_f = (k_{f \text{ BOD}} + k_{f \text{ NH}_4^+ \text{-N}}) \quad (4.20)$$

where

$$k_{f \text{ BOD}} = \frac{v (S_0 - S_e)_{\text{BOD}}}{\theta X_{\text{MLSS}}},$$

$$k_f \text{ NH}_4\text{-N} = \frac{v_N (N_o - N_e) \text{ NH}_4\text{-N}}{\theta X_{\text{MLSS}}}$$

Substituting of the equation (4.20) into the equation (4.19) it is arrived to:

$$r_f = k_o = \left[\frac{v (S_o - S_e)_{\text{BOD}}}{\theta X_{\text{MLSS}}} + \frac{v_N (N_o - N_e) \text{ NH}_4\text{-N}}{\theta X_{\text{MLSS}}} \right] X_{\text{MLSS}} \quad (4.21)$$

According to the equation (4.12), above equation (4.21) can be written for oxygen rate constant in the form of below equations which can be applied for BOD and ammonia nitrogen in nitrification as follows :

$$k_{f \text{ BOD}} = \frac{v (S_o - S_e / S_o)_{\text{BOD}}}{(\theta X_{\text{MLSS}} / S_o)} = \frac{v (\xi)_{\text{BOD}}}{1 / (F/M)_{\text{BOD}}} \quad (4.22)$$

$$k_{f \text{ NH}_4\text{-N}} = \frac{v_N (N_o - N_e / N_o) \text{ NH}_4\text{-N}}{(\theta X_{\text{MLSS}} / N_o)} = \frac{v_N (\xi)_{\text{NH}_4\text{-N}}}{(1 / F/M)_{\text{NH}_4\text{-N}}} \quad (4.23)$$

with substituting of equations (4.22) and (4.23) into the equation (4.21), it is arrived to:

$$r_f = [v (\xi)_{\text{BOD}} (F/M)_{\text{BOD}} + v_N (\xi)_{\text{NH}_4\text{-N}} (F/M)_{\text{NH}_4\text{-N}}] X_{\text{MLSS}} \quad (4.24)$$

Finally, substituting of the previously defined term by equation (4.24) into the basic equation (4.15), it gives the model equation for mineralization and nitrification as follows,

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] \frac{d^2}{4} \left[v(\xi)_{\text{BOD}} (F/M)_{\text{BOD}} + v_N (\xi)_{\text{NH}_4^+ - \text{N}} (F/M)_{\text{NH}_4^+ - \text{N}} \right] X_{\text{MLSS}} \quad (4.25)$$

Infact, the relation between the effluent BOD concentration S_e and that of ammonia nitrogen concentration N_e , when the nitrification was proceeding is given by ,

$$N_e = \frac{k_n S_e}{\beta_n - S_e} \quad (4.26)$$

where constants k_n and β_n were determined empirically as 1.47 and 19.7 mg/l respectively by Kubota and Miyaji (Kubota and Fujie, 1977).

Under the light of this information from the literature review, the consideration is given to nitrogen removal efficiency that was defined previously in the equation (4.24),

$$\xi_{\text{NH}_4^+ - \text{N}} = \frac{N_o - N_e}{N_o}$$

or,

$$\xi_{\text{NH}_4^+ - \text{N}} = 1 - \frac{N_e}{N_o} \quad (4.27)$$

Substituting of the effluent nitrogen concentration value for nitrification given by equation (4.26) into the equation (4.27) results in the equation (4.25),

$$C_L^k = [K] \left[v (\xi)_{\text{BOD}} (F/M)_{\text{BOD}} + v_N \left(1 - (k_n S_e / \beta_n - S_e / N_0) (F/M)_{\text{NH}_4^+ - \text{N}} \right) \right] X_{\text{MLSS}} \quad (4.28)$$

where

$$K = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] \frac{d^2}{4}$$

According to the equation (4.28), if the effluent BOD concentration S_e value is bigger than 19.7 mg/l, N_e value should be set as N_0 in equation (4.27), that is the nitrification does not proceed and therefore the right hand side of the equation can be dropped and the equation (4.28) becomes as it was given for mineralization only as follows;

$$C_L^k = [K] \left[v (\xi)_{\text{BOD}} (F/M)_{\text{BOD}} \right] X_{\text{MLSS}}$$

If nitrification proceeds, the given equation (4.28) is valid since the effluent BOD concentration less than 19.7 mg/l. In this situation, the effluent ammonia nitrogen concentration N_e value can be calculated from the equation (4.26) for equation (4.28).

If one can considers that nitrification is taking place in zero order reaction than the equation (4.25) can be written in the form now;

$$C_L^k = [K] \left[\frac{v(\xi)_{\text{BOD}} (F/M)_{\text{BOD}}}{v_N(\xi)_{\text{NH}_4^+ - \text{N}} (F/M)_{\text{NH}_4^+ - \text{N}}} + 1 \right] X_{\text{MLSS}} \quad (4.29)$$

or, equation (4.29) can be simplified to:

$$C_L^k = [K] \left[\frac{v(\xi)_{\text{BOD}} S_o}{v_N(\xi)_{\text{NH}_4^+ - \text{N}} N_o} + 1 \right] X_{\text{MLSS}} \quad (4.30)$$

It is reminded that, this form of the equation (4.30) for mineralization with nitrification is valid with the below limitations, which are as follows;

$$\begin{aligned} S_o &> S_e < 19.7 \text{ mg/l} \\ S_o &> N_o \\ N_o &> N_e > 0 \end{aligned}$$

It is also possible to make last simplification on the equation (4.30) to arrive the model equation for mineralization and nitrification as follows:

$$C_L^k = [K] \left[\left(\frac{v(S_o - S_e)}{v_N(N_o - N_e)} \right) + 1 \right] X_{\text{MLSS}}$$

Substituting of the term K as it was defined previously, it can be arrived to the basic equation for mineralization and nitrification as follows:

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] \frac{d^2}{4} \left[\left(\frac{v(S_o - S_e)}{v_N(N_o - N_e)} \right) + 1 \right] X_{\text{MLSS}} \quad (4.31)$$

where

C_L^k : critical oxygen concentration in the sludge suspension at which just not yet occurs the anaerobic state (ML^{-3})

D_L : diffusion coefficient of oxygen in the sludge suspension liquid (L^2T^{-1})

D_f : diffusion coefficient of oxygen in the sludge floc (L^2T^{-1})

d : diameter of floc (L)

v : stoichiometric oxygen consumption ratio for BOD (MM^{-1})

ξ_{BOD} : % removal efficiency of BOD

F/M_{BOD} : food to microorganisms ratio for BOD in nitrification (T^{-1})

v_N : stoichiometric oxygen consumption ratio for nitrogen (MM^{-1})

$\xi_{NH_4^+-N}$: % removal efficiency of nitrogen

$F/M_{NH_4^+-N}$: food to microorganisms ratio for nitrogen in nitrification (T^{-1})

X_{MLSS} : mixed liquor suspended solids concentration in nitrification (ML^{-3})

It is reminded that the proposed oxygen model for mineralization and nitrification is valid with the limitations which are as follows;

- The nitrogen removal efficiency in the system defined with the term $\xi_{NH_4^+-N}$ is valid if the effluent BOD concentration is less than 19.7 mg/l for nitrification can be proceeds.

- The model assumes that there is a enough nutrient concentration exist in the system for the biological growth of the flocs and not related to nitrification.
- The model considers that, zero order reaction takes place for oxygen. There is no substrate limitation in the bulk liquid and flocs except stated value of the effluent BOD concentration in the nitrification.



CHAPTER 5. GRAPHICAL REPRESENTATION FOR MINERALIZATION AND NITRIFICATION

5.1 Parameter Evaluation for Model

The diffusion of oxygen in the floc is important in the performance of activated sludge especially at low levels of turbulence when the floc size is large.

Pasveer (1954) in his mathematical analysis of Fickian diffusion he assumed a value of diffusivity equal to oxygen in pure water $2.1 \cdot 10^{-9} \text{ m}^2/\text{sec}$.

Wuhrman (1960) has used the value of the diffusivity of oxygen in the floc of $0.5 \cdot 10^{-9} \text{ m}^2/\text{sec}$ at 15°C .

Mueller et al. (1967) reported a diffusivity value of $0.5 \cdot 10^{-9} \text{ m}^2/\text{sec}$ for limiting oxygen concentration calculations for nominal floc diameter. He concluded that a value of $0.5 \cdot 10^{-9} \text{ m}^2/\text{sec}$ seems reasonable to use for smaller dominal diameters but for the nominal diameter of 115 microns, a value of $1.2 \cdot 10^{-9} \text{ m}^3/\text{sec}$ is recommended as a better value to use.

Matson and Characklis (1976) reported an average diffusivity for oxygen through live and deactivated sludge has a value of $1.27 \cdot 10^{-9} \text{ m}^2/\text{sec}$.

Heide and Kruize (1979) have used the rounded value of diffusion coefficient of oxygen in the sludge suspension $D_L = 2.0 \cdot 10^{-9} \text{ m}^2/\text{sec}$ and in the sludge floc $D_f = 1.0 \cdot 10^{-9} \text{ m}^2/\text{sec}$ for calculation of the

theoretical critical oxygen concentration.

Smith and Gackley (1984) have investigated that the diffusion coefficient of oxygen in pure water of $2.25 \cdot 10^{-9} \text{ m}^2/\text{sec}$ which was found comparable with the published value of Sherwood of $2.1 \cdot 10^{-9} \text{ m}^2/\text{sec}$ at 20°C . From the general discussion of density, porosity, specific surface area, mean pore radius and tortuosity of activated sludge, the diffusivity was found similar in thickened sludge and activated sludge flocs, hence the value of oxygen diffusion coefficient of $0.74 \cdot 10^{-9} \text{ m}^2/\text{sec}$ for bulk thickened activated sludge could be adopted for the diffusivity in activated sludge flocs.

Mueller et al. (1978) reported that the stoichiometric consumption ratio for BOD and nitrogen has a value of $0.80 \text{ gO}_2/\text{gBOD}$ and $4.57 \text{ gO}_2/\text{gN}$ respectively.

Gönenç (1982) reported that the stoichiometric consumption ratio for BOD $0.80 \text{ gO}_2/\text{gBOD}$ and for nitrogen $4.2 \text{ gO}_2/\text{g N}$.

Kbota and Fujie (1977) used in their calculation the stoichiometric consumption ratio for BOD and nitrogen $0.84 \text{ gO}_2/\text{gBOD}$ and $4.3 \text{ gO}_2/\text{g N}$ respectively.

A list of basic parameters which have been evaluated is given in Table 5.1.

5.2 Design Conditions Of Activated Sludge Systems

In practice, the classical design parameters of aerobic suspended culture - activated sludge systems for mineralisation and nitrification are mainly organic loading rate (F/M), mixed liquor suspended solid concentration (X_{MLSS}), removal efficiency (ξ) and sludge age (θ_c).

Table 5.1 List of Basic Parameters Evaluated

Parameter	Symbol	Unit	Magnitude	Reference
Stoichiometric consumption ratio for BOD removal	ν	kg O ₂ /kg BOD	0.84 0.80 0.5-1.0 0.80	Kubota and Fujie (1977) Mueller et al. (1978) Harremoes (1982) Gönenç (1982)
Stoichiometric consumption ratio for Nitrogen	ν_N	kg O ₂ /kg N	4.30 4.57 4.20 4.20	Kubota and Fujie (1977) Mueller et al. (1978) Harremoes (1982) Gönenç (1982)
Diffusion coefficient of oxygen in the sludge suspension (liquid)	D_L	m ² /sec	$2.1 \cdot 10^{-9}$ $2.0 \cdot 10^{-9}$ $2.25 \cdot 10^{-9}$	Pasveer (1954) Heide and Kruize (1979) Smith and Coackley (1984)
Diffusion coefficient of oxygen in the floc	D_f	m ² /sec	$0.5 \cdot 10^{-9}$ $0.5 - 1.2 \cdot 10^{-9}$ $1.27 \cdot 10^{-9}$ $1.0 \cdot 10^{-9}$ $0.74 \cdot 10^{-9}$	Wuhrman (1960) Mueller et al. (1967) Matson and Characklis (1976) Heide and Kruize (1979) Smith and Coackley (1984)
For mineralization For nitrification Diffusion coefficient of oxygen in the sludge suspension Diffusion coefficient of oxygen in the floc	ν ν_N D_L D_f	kg O ₂ /kg BOD kg O ₂ /kg N m ² /sec m ² /sec	0.84 4.20 $2.1 \cdot 10^{-9}$ $0.74 \cdot 10^{-9}$	The values are used in the model

An activated sludge system is designed in three types due to the organic loading rates, within the range of mixed liquor suspended solid concentration and sludge age as follows:

Heavy loaded plant: $F/M = 0.5 - 2.0 \text{ day}^{-1}$
 $X_{MLSS} = 4000 - 10000 \text{ mg/l}$
 $\theta_c = 2 - 5 \text{ days}$

Conventionally loaded plant: $F/M = 0.2 - 0.5 \text{ day}^{-1}$
 $X_{MLSS} = 1500 - 3000 \text{ mg/l}$
 $\theta_c = 3 - 10 \text{ days}$

Lightly loaded plant: $F/M = 0.05 - 0.2 \text{ day}^{-1}$
 $X_{MLSS} = 3000 - 6000 \text{ mg/l}$
 $\theta_c = 15 - 30 \text{ days}$

The performance of an activated sludge treatment plant is influenced by the variations in the wastewater influent flows and quality, as well as by the fluctuations in the process variables. Traditional design models employed to estimate process performance are generally based on the steady-state assumption. Time varying loadings frequently result in short-term process upsets, instability, and adverse effects on effluent quality. To control the activated sludge process, quality variability should be minimized and the effluent quality maintained below the effluent standard values. Because both wastewater flowrate and pollutant concentrations change with time, determination of process response to influent variations is of considerable importance.

Today, one of the method used largely in practice for determination of the process behaviour of activated sludge plants is oxygen consumption. The difference in oxygen consumption in the differently loaded types of activated sludge plants is due primarily to the following factors:

- a. mineralisation
- b. nitrification
- c. mixing

Mineralisation means the conversion of organics into new cells and inorganic matter. This process depends only on the total sludge quantity and the age of the sludge. As sludge age is inversely proportional to sludge loading, greater mineralisation will take place in lightly loaded plants than heavily loaded ones.

The oxygen consumption for decomposition of the carbon compounds in the wastewater depends on the quantity and concentration of the effluent, so that this will mean varying oxygen consumption and therefore a need for control and regulation. The same applies to the supply of nitrogen compounds in nitrifying plants.

However, the oxygen consumption for mineralisation is the dominating factor in the case of lightly loaded plants, while it plays a modest part in heavily loaded ones, it will be seen that the need for aeration control is greater in the case of heavily loaded plants than of lightly loaded plants.

With regard to mixing in the aeration tank, a certain supply of energy is needed in order to keep the activated sludge in suspension in one hand, on the other hand, it promotes to break down the flocs to be in smaller diameters. The mixing is normally a side effect of the blowing in of air, but in the case of lightly loaded plants the energy required for mixing may be primary factor, so that the oxygen transfer takes place as part of the necessary mixing. This means that oxygen control and regulation can be made impossible in lightly loaded plants, as controlled blowing in of oxygen of the activated sludge.

Today, in practice, the field oxygen transfer rate is calculated according to the equation (2.11) given in Chapter 2, Literature Review.

As it can be seen from the formula, it covers all possible correction factors and interferences, except design value of oxygen concentration (C_L).

A value of 2.0 mg DO/l for design or operation which is used by the design engineers has no scientific indication.

The determination of the diameter of the activated sludge floc related to the power input for oxygen consumption rate have been studied by Heide and Kruize (Heide and Kruize, 1979). In their study, the determined diameters with a microscope has indicated the microflocs. They have also used the Kolomogorov's theory to calculate the diameter of the sludge floc under operation circumstances, but found not applicable, if the power supply is decreased, or in other words the requirement of isotropic turbulence is not met. For highly loaded activated sludge systems, the average power supply is about 10 Watt/m³; for heavily loaded systems, the oxygen consumption rate is higher, but the power supply is also higher for the transfer of oxygen to the activated sludge system by means of aeration. The average floc diameters as a function of the power inputs given by the investigator are as follows:

<u>Power Input</u> <u>Watt/m³</u>	<u>Average Floc</u> <u>Diameter (mm)</u>
0.2	1.0
1.3	0.85
2.5	0.8
4.3	0.7
10.0	0.6
20.0	0.5

With regard to mineralisation and nitrification in aerobic suspended culture systems, the calculated critical oxygen concentrations for different classical activated sludge design parameters for varying floc

diameters, may be found in aeration tank are shown by graphics on the following

5.3 Graphical Representation for Mineralization

Calculated critical oxygen concentrations for various expected floc diameters that may found in the aeration tank for different mixed liquor suspended solid concentrations $X_{MLSS} = 1000, 2500, 4500$ mg/l, which depend on the design and operation of the aeration tank at constant organic loading rate $F/M = 0.60$ day⁻¹ with the 90%BOD removal efficiency requirement in mineralization are given in Table 5.2 and shown on Figure 5.1.

Calculated critical oxygen concentrations for various floc diameters that may found in the aeration tank for different organic loading rates that can be changed due to inflow rate and/or influent BOD concentration $F/M = 0.1, 0.2, 0.2, 0.4, 0.6, 1.0$ day⁻¹ which depend on the operation of aeration tank at low mixed liquor suspended solid concentration $X_{MLSS} = 2.000$ mg/l with the 90% BOD removal efficiency requirement in mineralization are given in Table 5.3 and shown on Figure 5.2

Calculated critical oxygen concentrations for various expected floc diameters that may found in the aeration tank for different organic loading rates that can be changed due to inflow rate and/or influent BOD concentration $F/M = 0.1, 0.2, 0.4, 0.6, 1.0$ day⁻¹ which depend on the operation of aeration tank at constant concentration of mixed liquor suspended solid $X_{MLSS} = 3.000$ mg/l with the 90%BOD removal efficiency requirement in mineralization are given in Table 5.4 and shown on Figure 5.3.

Calculated critical oxygen concentrations for various expected floc diameters that may found in the aeration tank for different organic loading rates that can be changed due to inflow rate and/or influent

BOD concentration, $F/M = 0.1, 0.2, 0.4, 0.6, 1.0 \text{ day}^{-1}$ which depend on the operation of the aeration tank at constant mixed liquor suspended solid concentration $X_{MLSS} = 5000 \text{ mg/l}$ with the 90% BOD removal efficiency requirement in mineralization are given in Table 5.5 and shown on Figure 5.4

Calculated critical oxygen concentrations for various expected floc diameters that may found in the aeration tank for the 50,70,98 % BOD removal efficiency that can be changed due to rapid increase or decrease in the influent BOD concentration which depend on the operation of the aeration tank at constant mixed liquor suspended solid concentration $X_{MLSS} = 3500 \text{ mg/l}$ and $F/M = 0.45 \text{ day}^{-1}$ in mineralization are given in Table 5.6 and shown on Figure 5.5

Calculated critical oxygen concentrations for various organic loading rate F/M ratios that may vary time by time in the aeration tank which mostly dependent inflow rate and influent BOD concentration at constant mixed liquor suspended solid concentration $X_{MLSS} = 2675 \text{ mg/l}$ with the 90% BOD removal efficiency requirement for varying organic loading conditions in mineralization are given in Table 5.7 and shown on Figure 5.6.

5.4 Graphical Representation for Transition from Mineralization to Nitrification

In the frame of the model for mineralization and nitrification with using of the model equation (4.26) which reflects the transition concentration of BOD and nitrogen from mineralization to nitrification, the calculated effluent ammonia nitrogen concentrations for effluent BOD concentrations are given in Table 5.8 and shown on Figure 5.7.

5.5 Graphical Representation for Mineralization and Nitrification

Calculated critical oxygen concentrations for various expected floc diameters that may be found in the aeration tank in mineralization and nitrification for different effluent nitrogen concentration that can be changed due to effluent BOD particularly in nitrification, which depends on design and operation of the aeration tank at constant mixed liquor suspended solid concentration $X_{MLSS}=3500\text{mg/l}$; $(F/M)_{BOD}=0.48\text{day}^{-1}$, influent BOD and ammonia nitrogen concentrations 350 mg/l and 35 mg/l respectively given in Table 5.9 and shown on Figure 5.8.



Table 5.2 Calculated critical Oxygen concentrations for various mixed liquor suspended solid concentrations when organic loading rate $F/M = 0.6 \text{ day}^{-1}$ and 90% BOD removal efficiency

Floc Diameter d (mm)	$X_{MLSS} \text{ (mg/l)}$		
	1.000	2.500	4.500
	$C_L^k \text{ (kg } O_2/\text{m}^3)$		
0.5	$0.13 \cdot 10^{-3}$	$0.33 \cdot 10^{-3}$	$0.59 \cdot 10^{-3}$
1.0	$0.50 \cdot 10^{-3}$	$1.25 \cdot 10^{-3}$	$2.25 \cdot 10^{-3}$
1.5	$1.14 \cdot 10^{-3}$	$2.85 \cdot 10^{-3}$	$5.13 \cdot 10^{-3}$
2.0	$2.02 \cdot 10^{-3}$	$5.05 \cdot 10^{-3}$	$9.00 \cdot 10^{-3}$

Assumed oxygen saturation concentration of the activated sludge is maximum 9.0 mg/l.

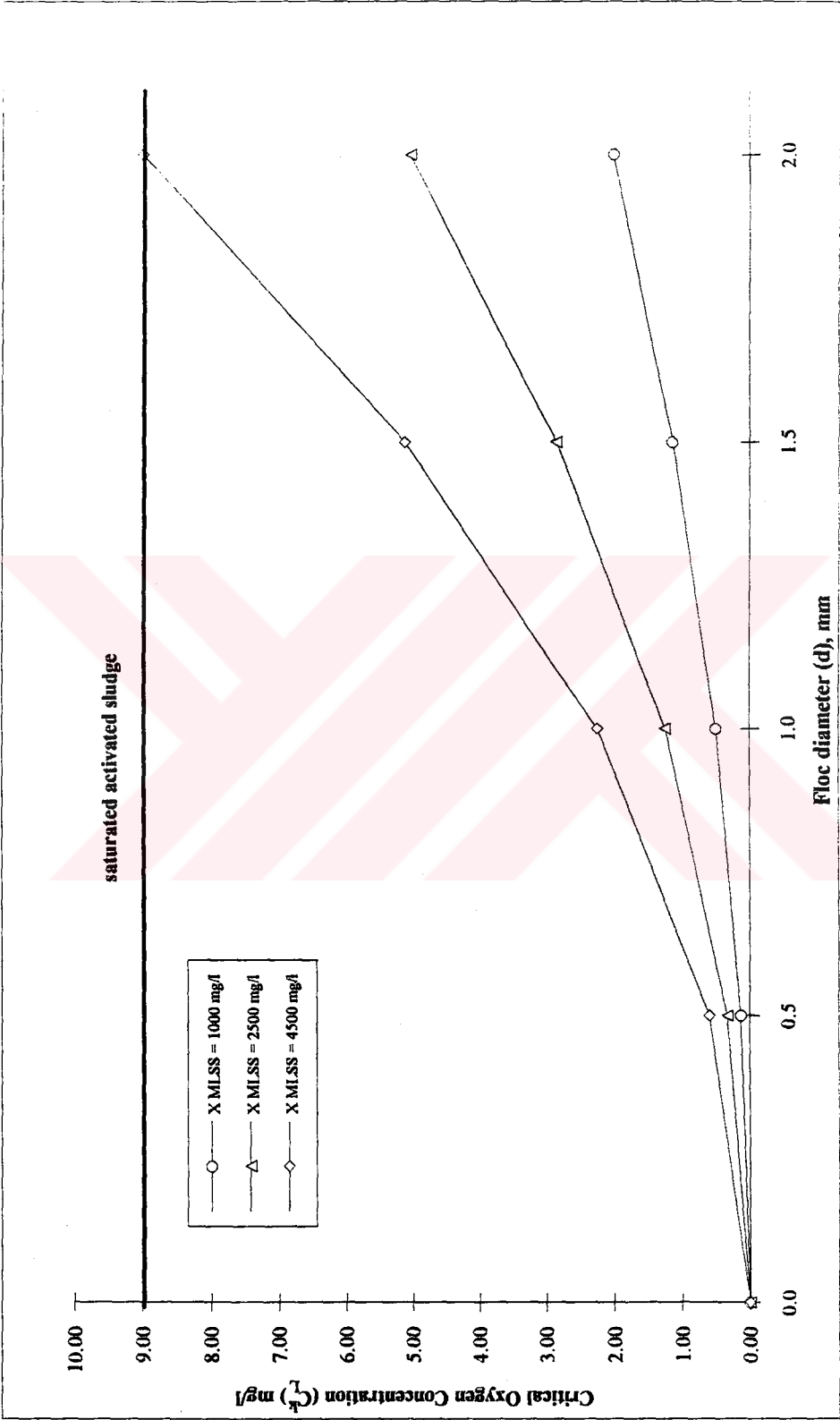


Figure 5.1 Calculated critical oxygen concentrations for various floc diameters in mineralization, $F/M=0.60 \text{ day}^{-1}$, $\xi_{BOD} = 90\%$.

Table 5.3 Calculated critical oxygen concentrations for various organic loading rates when mixed liquor suspended solid concentration $X_{MLSS} = 2000 \text{ mg/l}$ and 90% BOD removal efficiency

Floc Diameter d (mm)	F/M (day^{-1})				
	0.10	0.20	0.40	0.60	1.00
	$C_L^k (\text{kg O}_2/\text{m}^3)$				
0.5	$0.42 \cdot 10^{-4}$	$0.84 \cdot 10^{-4}$	$0.17 \cdot 10^{-3}$	$0.25 \cdot 10^{-3}$	$0.42 \cdot 10^{-3}$
1.0	$0.17 \cdot 10^{-3}$	$0.34 \cdot 10^{-3}$	$0.68 \cdot 10^{-3}$	$1.02 \cdot 10^{-3}$	$1.68 \cdot 10^{-3}$
1.5	$0.38 \cdot 10^{-3}$	$0.76 \cdot 10^{-3}$	$1.52 \cdot 10^{-3}$	$2.28 \cdot 10^{-3}$	$3.78 \cdot 10^{-3}$
2.0	$0.67 \cdot 10^{-3}$	$1.34 \cdot 10^{-3}$	$2.68 \cdot 10^{-3}$	$4.02 \cdot 10^{-3}$	$6.73 \cdot 10^{-3}$

Assumed oxygen saturation concentration of the activated sludge is maximum 9.0 mg/l.

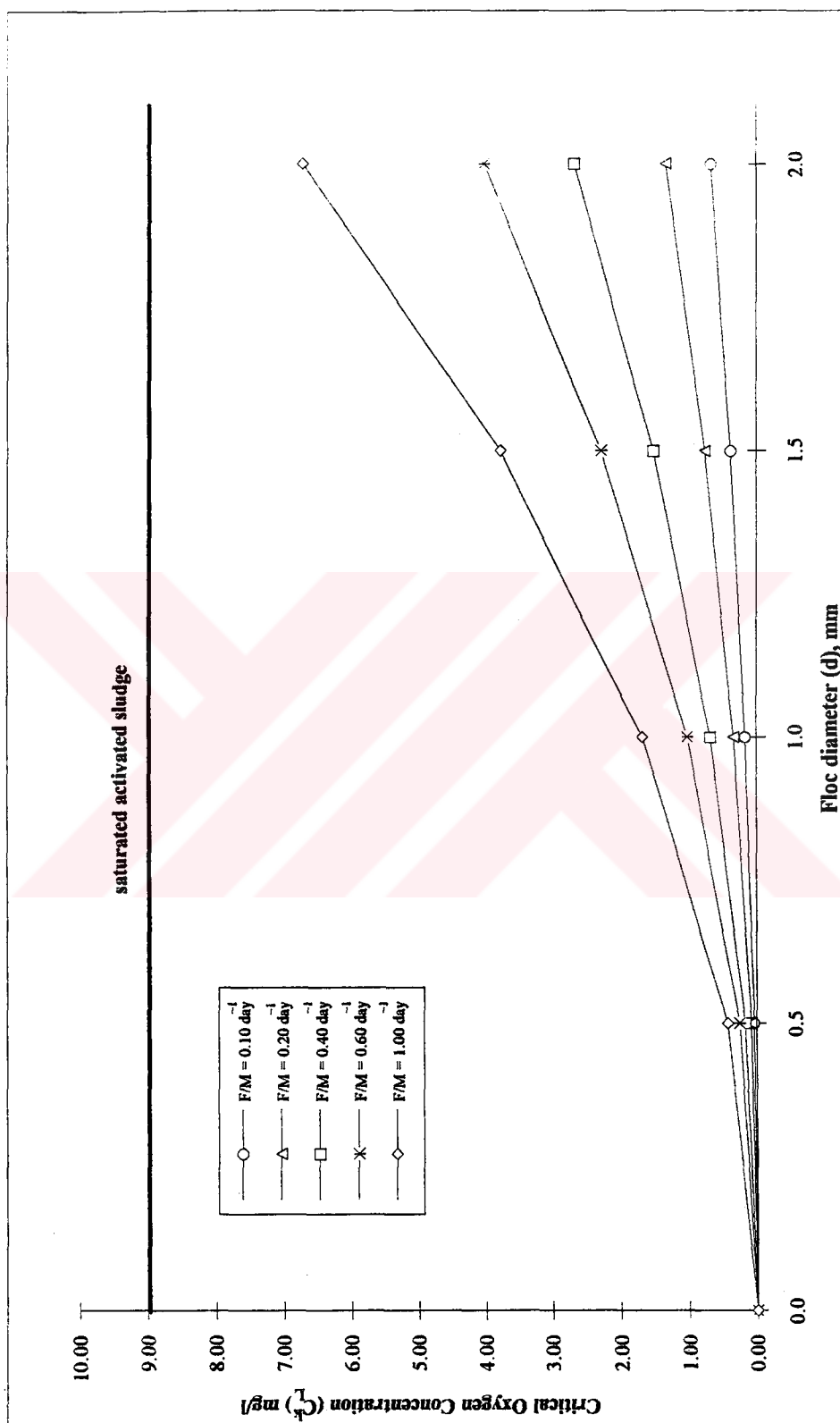


Figure 5.2 Calculated critical oxygen concentrations for various floc diameters in mineralization, $X_{MLSS} = 2000$ mg/l, $\xi_{BOD} = 90\%$

Table 5.4 Calculated critical oxygen concentrations for various organic loading rates when mixed liquor suspended solid concentration $X_{MLSS} = 3000 \text{ mg/l}$ and 90% BOD removal efficiency

Floc Diameter d (mm)	F/M (day^{-1})				
	0.10	0.20	0.40	0.60	1.00
	$C_L^k (\text{kg O}_2/\text{m}^3)$				
0.5	$0.63 \cdot 10^{-4}$	$0.13 \cdot 10^{-3}$	$0.25 \cdot 10^{-3}$	$0.38 \cdot 10^{-3}$	$0.63 \cdot 10^{-3}$
1.0	$0.25 \cdot 10^{-3}$	$0.50 \cdot 10^{-3}$	$1.01 \cdot 10^{-3}$	$1.51 \cdot 10^{-3}$	$2.50 \cdot 10^{-3}$
1.5	$0.57 \cdot 10^{-3}$	$1.14 \cdot 10^{-3}$	$2.27 \cdot 10^{-3}$	$3.41 \cdot 10^{-3}$	$5.70 \cdot 10^{-3}$
2.0	$1.00 \cdot 10^{-3}$	$2.02 \cdot 10^{-3}$	$4.04 \cdot 10^{-3}$	$6.06 \cdot 10^{-3}$	> saat.conc.

Assumed oxygen saturation concentration of the activated sludge is maximum 9.0 mg/l.

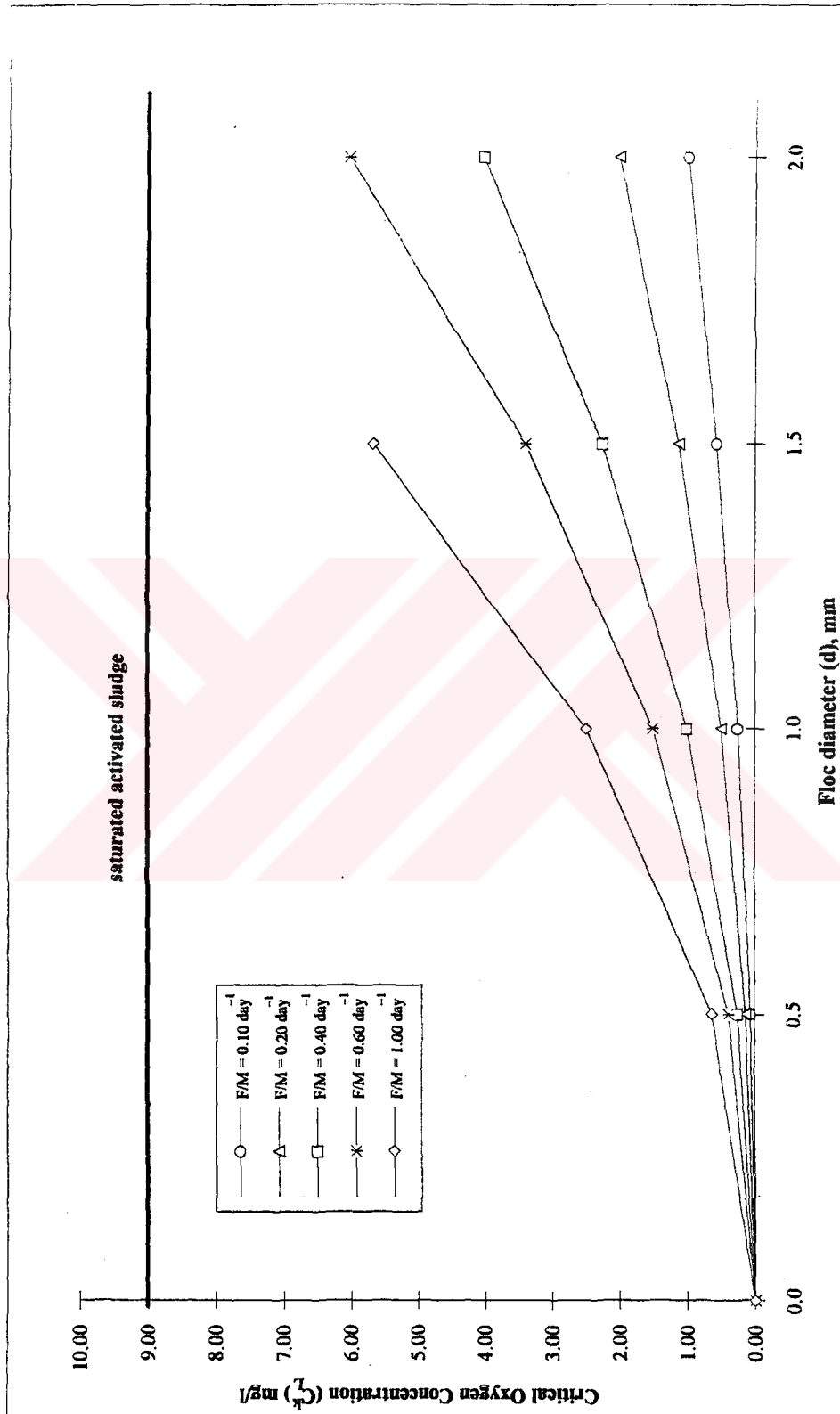


Figure 5.3 Calculated critical oxygen concentrations for various floc diameters in mineralization, $X_{MLSS} = 3000$ mg/l, $\xi_{BOD} = 90\%$.

Table 5.5 Calculated critical oxygen concentrations for various organic loading rates when mixed liquor suspended solid concentration $X_{MLSS} = 5000 \text{ mg/l}$ and 90% BOD removal efficiency

Floc Diameter d (mm)	F/M (day ⁻¹)					
	0.10	0.20	0.40	0.60	1.00	
	$C_L^k \text{ (kg O}_2\text{/m}^3\text{)}$					
0.5	$0.11 \cdot 10^{-3}$	$0.21 \cdot 10^{-3}$	$0.42 \cdot 10^{-3}$	$0.63 \cdot 10^{-3}$	$1.05 \cdot 10^{-3}$	
1.0	$0.42 \cdot 10^{-3}$	$0.84 \cdot 10^{-3}$	$1.68 \cdot 10^{-3}$	$2.52 \cdot 10^{-3}$	$4.21 \cdot 10^{-3}$	
1.5	$0.96 \cdot 10^{-3}$	$1.89 \cdot 10^{-3}$	$3.78 \cdot 10^{-3}$	$5.68 \cdot 10^{-3}$	> sat. conc.	
2.0	$1.68 \cdot 10^{-3}$	$3.36 \cdot 10^{-3}$	$6.73 \cdot 10^{-3}$	> sat. conc	> saat. conc.	

Assumed oxygen saturation concentration of the activated sludge is maximum 9.0 mg/l.

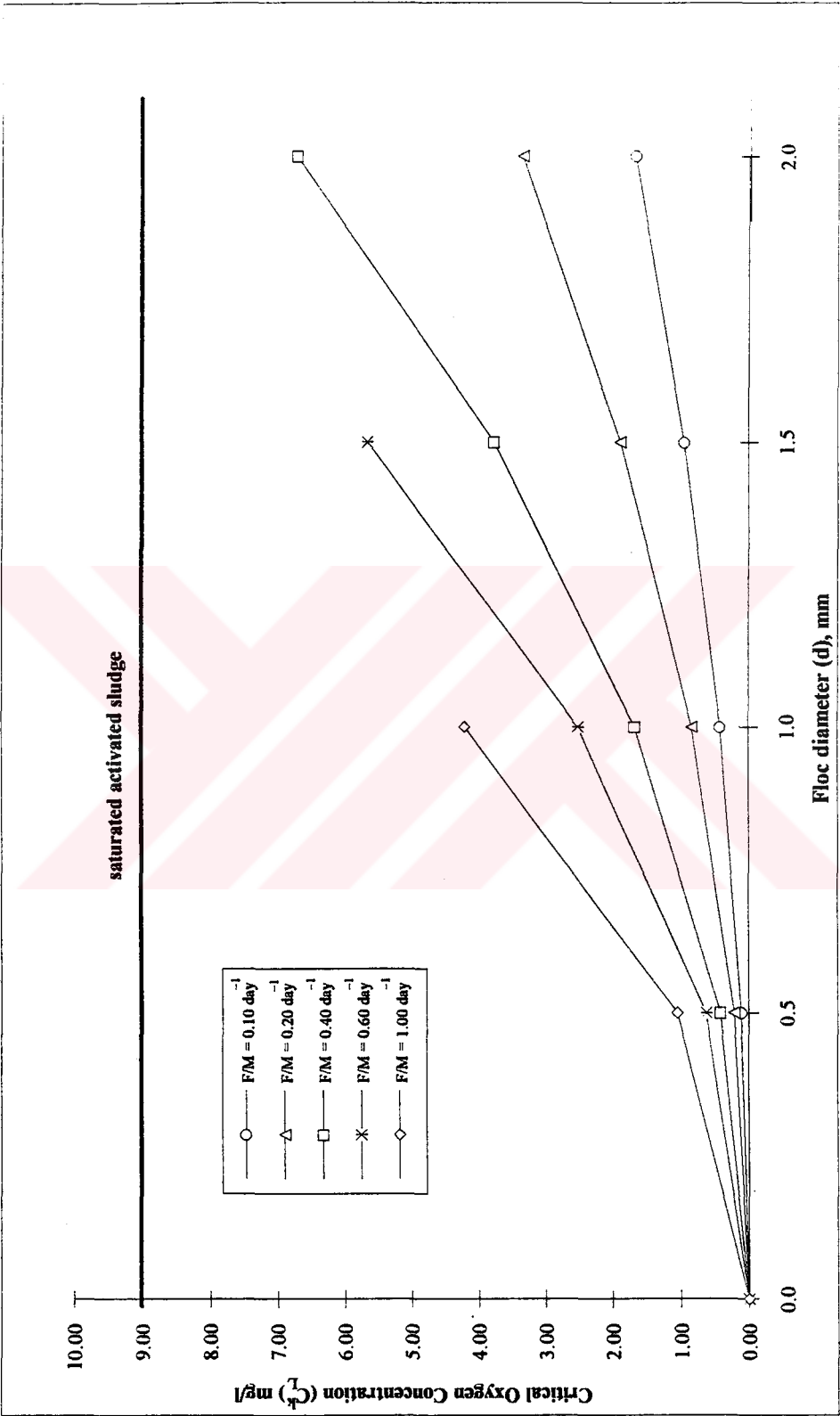


Figure 5.4 Calculated critical oxygen concentrations for various floc diameters in mineralization, $X_{MLSS} = 5000$ mg/l, $\xi_{BOD} = 90\%$.

Table 5.6 Calculated critical oxygen concentrations for BOD removal efficiency when mixed liquor suspended solid concentration $X_{MLSS} = 3500 \text{ mg/l}$ and $F/M = 0.45 \text{ day}^{-1}$

Floc Diameter d (mm)	ξ BOD		
	50%	70%	98%
	$C_L^k (\text{kg } O_2/\text{m}^3)$		
0.5	$0.20 \cdot 10^{-3}$	$0.29 \cdot 10^{-3}$	$0.40 \cdot 10^{-3}$
1.0	$0.82 \cdot 10^{-3}$	$1.14 \cdot 10^{-3}$	$1.60 \cdot 10^{-3}$
1.5	$1.84 \cdot 10^{-3}$	$2.58 \cdot 10^{-3}$	$3.60 \cdot 10^{-3}$
2.0	$3.27 \cdot 10^{-3}$	$4.58 \cdot 10^{-3}$	$6.41 \cdot 10^{-3}$

Assumed oxygen saturation concentration of the activated sludge is maximum 9.0 mg/l .

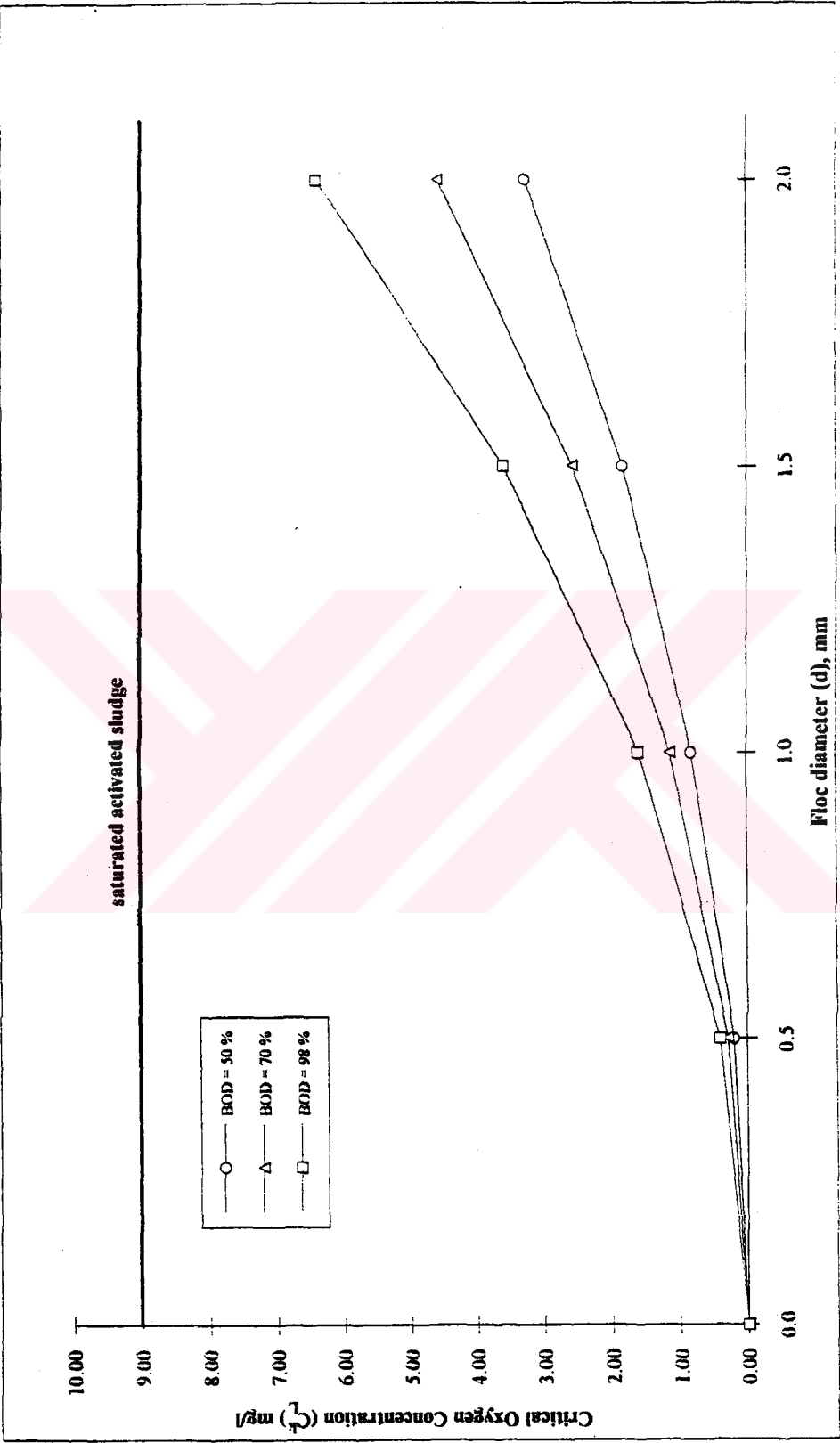


Figure 5.5 Calculated critical oxygen concentrations for various floc diameters in mineralization, $X_{MLSS} = 3500$ mg/l, $F/M = 0.45$ day⁻¹

Table 5.7 Calculated critical oxygen concentrations for organic loading rates $F/M = 0.2, 0.4, 0.6$ and 1.0 day^{-1} , when mixed liquor suspended solid concentration $X_{M,SS} = 2675 \text{ mg/l}$ and 90% BOD removal efficiency

Floc Diameter d (mm)	F/M day^{-1}			
	0.20	0.40	0.60	1.00
	$C_L^k \text{ (kgO}_2\text{/m}^3\text{)}$			
0.5	$0.11 \cdot 10^{-3}$	$0.22 \cdot 10^{-3}$	$0.33 \cdot 10^{-3}$	$0.55 \cdot 10^{-3}$
1.0	$0.45 \cdot 10^{-3}$	$0.90 \cdot 10^{-3}$	$1.35 \cdot 10^{-3}$	$2.25 \cdot 10^{-3}$
2.0	$1.80 \cdot 10^{-3}$	$3.60 \cdot 10^{-3}$	$5.40 \cdot 10^{-3}$	$9.00 \cdot 10^{-3}$

Assumed saturation concentration of the activated sludge is maximum $9.0 \text{ mg O}_2\text{/l}$

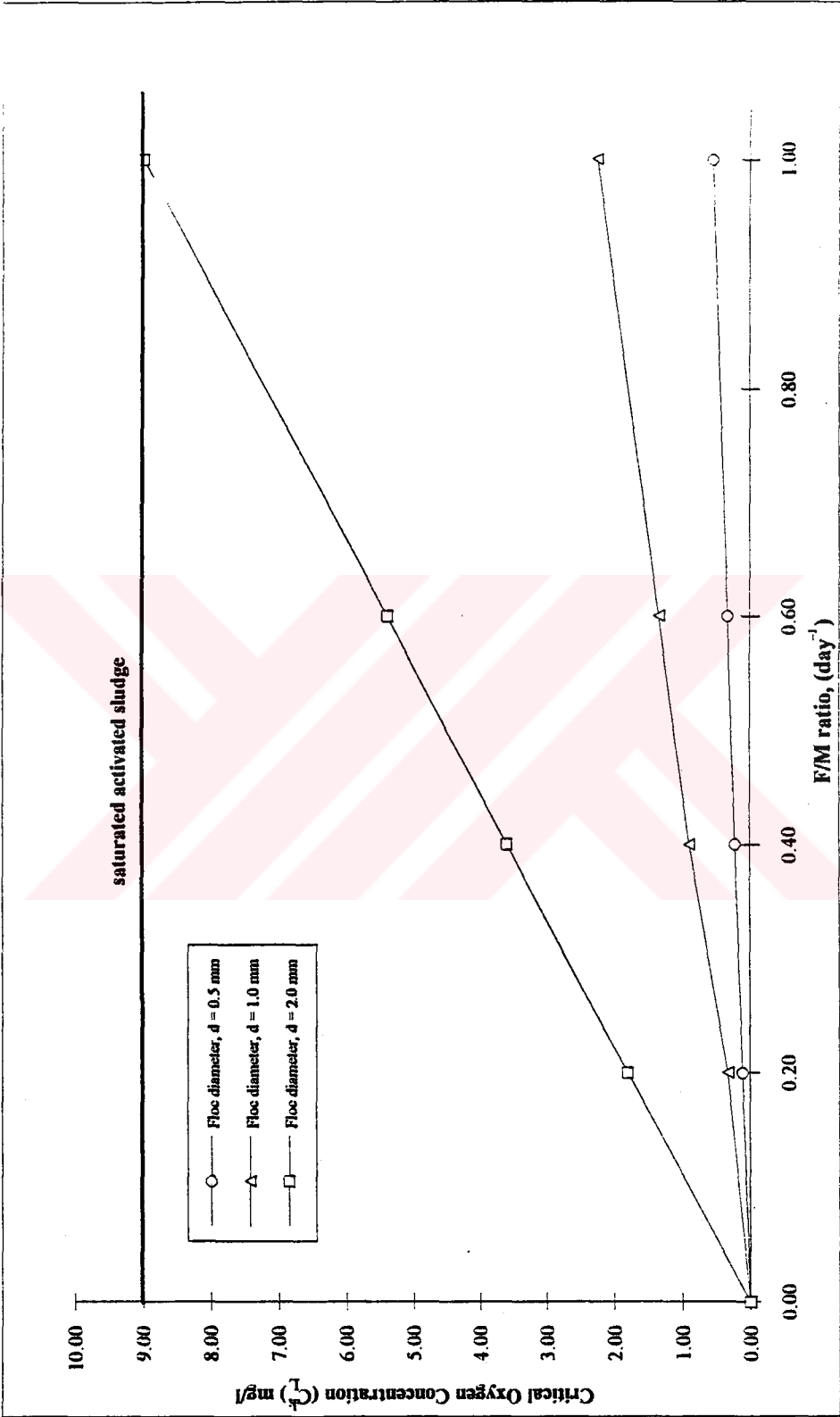


Figure 5.6 Calculated critical oxygen concentrations for various F/M ratios in mineralization, $X_{\text{MLSS}} = 2675 \text{ mg/l}$, $\xi_{\text{BOD}} = 90\%$

Table 5.8 Calculated effluent ammonia nitrogen concentrations from the model equation (4.26) for various effluent BOD concentration in nitrification process

S _e (mg BOD/l)	N _e (mg NH ₄ ⁺ -N/l)	Remarks
19.7	0	Transition from Mineralization to Nitrification
18.7	27.5	Mineralization + Nitrification
17.7	13.0	Mineralization + Nitrification
16.7	8.2	Mineralization + Nitrification
15.7	5.8	Mineralization + Nitrification
13.7	3.4	Mineralization + Nitrification
10.7	1.7	Mineralization + Nitrification
5.7	0.6	Mineralization + Nitrification
0	0	No biological growth, no oxygen transfer proceeds into flocs

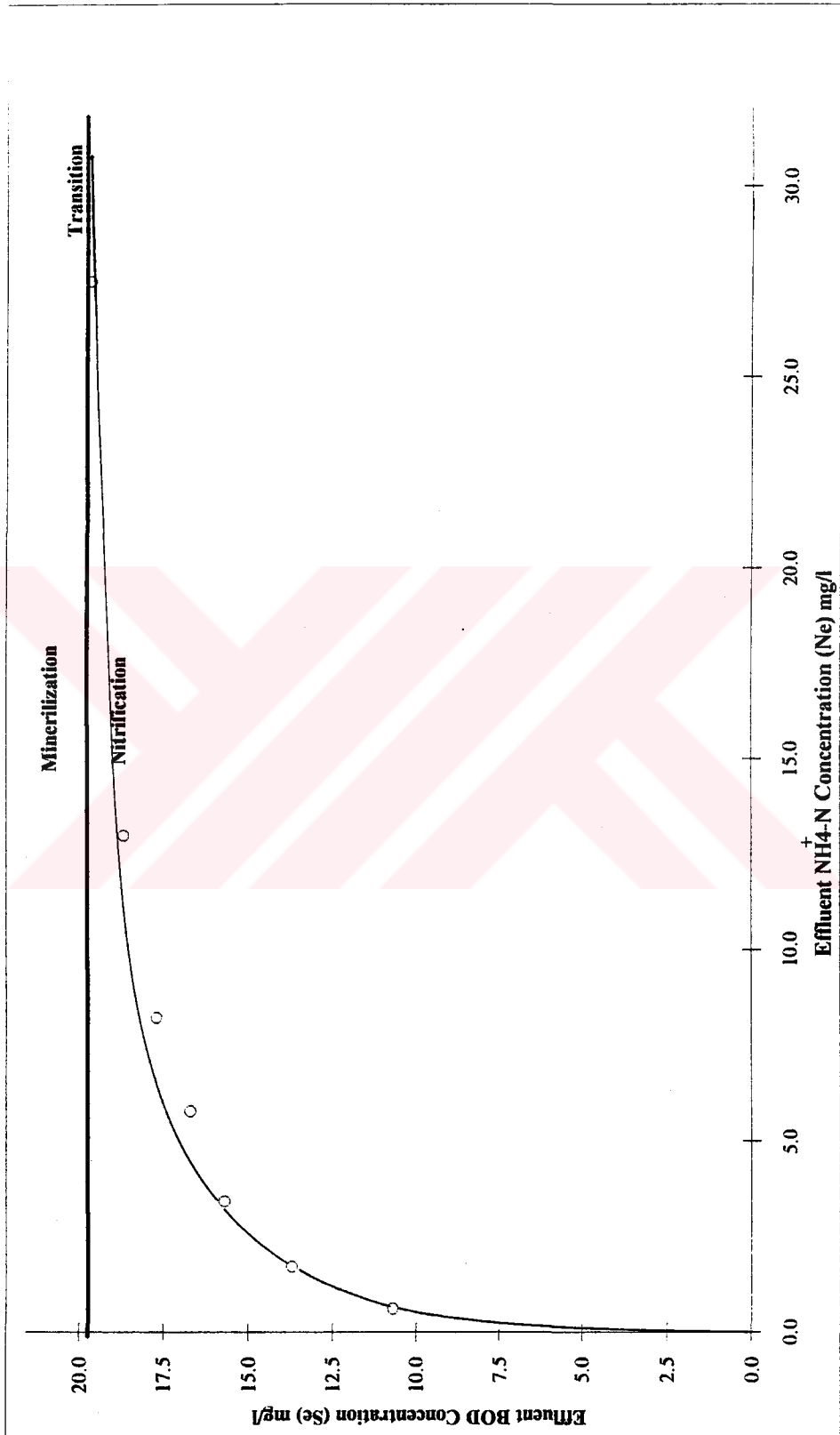


Figure 5.7 The calculated effluent Ammonia - nitrogen concentrations v.s. different effluent BOD concentrations in minerilization and Nitrification

Table 5.9 Calculated Critical Oxygen Concentrations for Various Floc Diameters when $X_{MLSS} = 3500 \text{ mg/l}$ and $(F/M)_{BOD} = 0.48 \text{ day}^{-1}$ in Mineralization and Nitrification

Floc Diameter d (mm)	MINERALIZATION		NITRIFICATION	
	$S_o = 350 \text{ mgBOD/l}$ $S_o = 19.7 \text{ mgBOD/l}$ $N_o = 35 \text{ m NH}_4^+ \text{ -N/l}$	$S_o = 18.7 \text{ mg BOD/l}$ $N_o = 27.5 \text{ mg NH}_4^+ \text{ -N/l}$	$S_o = 15 \text{ mg BOD/l}$ $N_o = 4.7 \text{ mg NH}_4^+ \text{ -N/l}$	$S_o = 10 \text{ mg BOD/l}$ $N_o = 1.5 \text{ mg NH}_4^+ \text{ -N/l}$
	$C^*_{L} \text{ (kgO}_2\text{/m}^3\text{)}$			
0.1	$1.5 \cdot 10^{-5}$	$3.8 \cdot 10^{-4}$	$1.3 \cdot 10^{-4}$	$1.2 \cdot 10^{-4}$
0.2	$5.9 \cdot 10^{-5}$	$1.5 \cdot 10^{-3}$	$5.0 \cdot 10^{-4}$	$4.7 \cdot 10^{-4}$
0.3	$1.3 \cdot 10^{-4}$	$3.4 \cdot 10^{-3}$	$1.1 \cdot 10^{-3}$	$1.1 \cdot 10^{-3}$
0.4	$2.4 \cdot 10^{-4}$	$6.1 \cdot 10^{-3}$	$2.0 \cdot 10^{-3}$	$1.8 \cdot 10^{-3}$
0.5	$3.7 \cdot 10^{-4}$	> sat. conc.	$3.1 \cdot 10^{-3}$	$2.9 \cdot 10^{-3}$
0.6	$5.3 \cdot 10^{-4}$	> sat. conc.	$4.5 \cdot 10^{-3}$	$4.2 \cdot 10^{-3}$
0.7	$7.3 \cdot 10^{-4}$	> sat. conc.	$6.1 \cdot 10^{-3}$	$5.8 \cdot 10^{-3}$
0.8	$9.5 \cdot 10^{-4}$	> sat. conc.	$8.0 \cdot 10^{-3}$	$7.5 \cdot 10^{-3}$
0.9	$1.2 \cdot 10^{-3}$	> sat. conc.	> sat. conc.	> sat. conc.
1.0	$1.5 \cdot 10^{-3}$	> sat. conc.	> sat. conc.	> sat. conc.
1.5	$3.3 \cdot 10^{-3}$	> sat. conc.	> sat. conc.	> sat. conc.
2.0	$5.9 \cdot 10^{-3}$	> sat. conc.	> sat. conc.	> sat. conc.

Assumed saturation concentration of the activated sludge is maximum $9.0 \text{ mg O}_2\text{/l}$

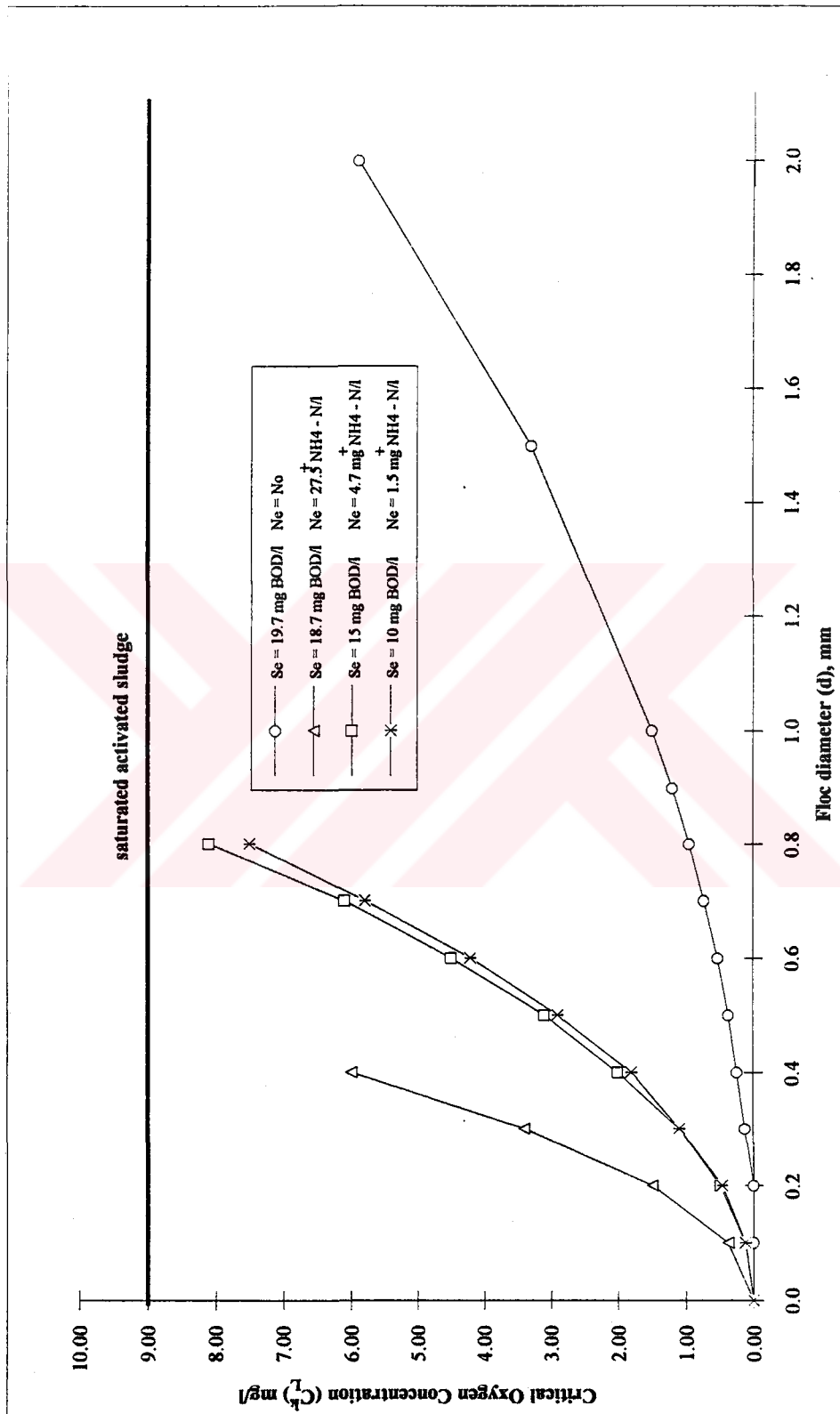


Figure 5.8 Comparison of calculated critical oxygen concentration for various floc diameters in mineralization and nitrification
 $X_{NH_4-N} = 3500$ mg/l, $S_o = 350$ mg BOD, $N_o = 35$ mg NH_4-N /l

RESULTS AND DISCUSSION

In this study, a conceptual model for the calculation of the oxygen required by the aerobic suspended culture systems used for mineralization and nitrification processes is developed. This model can be defined by the following equation:

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] \frac{d^2}{4} \left[\left(\frac{v (S_o - S_e)}{v_N (N_o - N_e)} \right) + 1 \right] X_{MLSS}$$

or,

$$C_L^k = \left[\frac{1}{3 D_L} + \frac{1}{6 D_f} \right] \frac{d^2}{4} \left[\left(v(\xi)(F/M) \right)_{BOD} + \left(v(\xi)(F/M) \right)_{NH_4^+ - N} \right] X_{MLSS}$$

According to the model, the critical dissolved oxygen concentration which does not allow anaerobic conditions in the suspended flocs is related to:

- the molecular diffusion coefficient of oxygen in the liquid film on the outermost of the flocs
- the molecular diffusion coefficient of oxygen in the floc
- the dimensions of the floc (diameter)

- the stoichiometric coefficients of the oxidation reactions of BOD and $\text{NH}_4\text{-N}$ proceeded in the flocs
- the amount of BOD removed by mineralization and amount of $\text{NH}_4\text{-N}$ removed by nitrification in the system.
- the BOD and $\text{NH}_4\text{-N}$ loading rates per unit mass of microorganisms in the system
- the concentration of the suspended floc in the system

Given these relations, critical oxygen concentration variation that can be expressible functionally in the system is;

- inversely proportional to the diffusion coefficient in the liquid film and floc
- directly proportional to the diameter of the floc
- directly proportional to the stoichiometric coefficients and loading rates of BOD and $\text{NH}_4\text{-N}$ or,
- directly proportional to the stoichiometric coefficients and removal rates of BOD and $\text{NH}_4\text{-N}$
- directly proportional with the floc concentration in the system

This functional relations are shown graphically in the last chapter of the thesis. By using these graphics, the required oxygen concentration for suspended culture systems under the different conditions can easily be defined. Also the amount of oxygen that has to be supplied to the system can be computed in the design phase or, the

system can be operated according to the requirements. By use of these graphics, some practical recommendations are as follows:

- To reduce the amount of oxygen for a known loading rate of BOD and removal efficiency of BOD, one should reduce the floc diameter.
- To reduce the amount of oxygen for a known floc diameter, concentration and removal efficiency of BOD, one should decrease the BOD loading.
- To other reduce the amount of oxygen for a known floc diameter, concentration and BOD loading, one should decrease the amount of BOD removed..
- To reduce the amount of oxygen in the system for all above, one should reduce the diameter of floc as much as possible.

Some important results from the developed model are:

- In practice, the traditional design model of the aerobic suspended culture systems for mineralization and nitrification are based on the steady-state assumptions. So far, the design aeration tank is always under influence of the variations in the influent flow and concentrations, as well as by the fluctuations in the process variables, as a consequences, time varying loadings frequently result in short term process upsets, instability and adverse effects on the designed effluent quality. Apart from that, all possible variations should be considered in the design phase by the design engineers.

- In the design phase of the activated sludge to control the process, quality variability should also be considered and minimized, and the effluent quality be maintained below the effluent standard values, because both wastewater flowrate and pollutant concentrations change with by time, therefore, determination of process response to influent variations is very important.
- Keeping in mind, one of the method used largely in practice for determination of the process behaviour of the activated sludge plants is oxygen consumption, therefore, the difference in oxygen consumptions which have a direct relation with the critical oxygen concentration in heavily, coventionaly and lightly loaded types of the activated sludge plant is due primarily to minerilization, nitrification and mixing.
- The oxygen consumption for minerilization is dominating factor in the case of lightly loaded plants while it is compared in heavily loaded ones at constant design parameters such as mixed liquor suspended solid concentration, removal efficiency for selected floc diameter. Therefore, the need for aeration control in terms of critical oxygen concentration control in the aeration tank is greater in the case of heavily loaded plants than of lightly loaded plants.
- For mixing in the aeration tank, a certain supply of energy is required to keep the activated sludge flocs in suspension and breakdown the flocs to be in smaller diameters. In the case of lightly loaded plants, the energy required for mixing of activated sludge is the primary factor, so that the oxygen transfer takes place as part of the necessary mixing and as a result of this , the BOD concentration in the effluent

decreases or removal efficiency of minerilization increases. This means that oxygen control can be made impossible in lightly loaded plants.

- By using of the developed model, it is possible to verify that the meaningful nitrification starts only when the BOD concentration is less than 19.7 mg/l in the aeration tank as indicated by a number of investigator. This transition concentration of the BOD is critical from an oxygen supply and oxygen control point of view in minerilization. Consequently, ability of the mechanical aerators or constant speed blowers to supply oxygen is a potential risk for operation of large wastewater treatment plants.
- If the concentration in the effluent decreases or the removal efficiency of minerilization increases, the efficiency of nitrification will increase. If nitrification to proceed with minerilization, the floc diameter should be reduced, because it is more important than minerilization in control of the process, decreasing of amount of oxygen by reducing the floc diameter which is most important than over solely minerilization.
- In practice, in the designing of conventional activated sludge system, such a value of organic loading $F/M = 0.40 \text{ day}^{-1}$, removal efficiency $\xi = 90 \%$, sludge age $\theta_c = 3$ days and mixed liquor suspended solid concentration $X_{MLSS} = 2675 \text{ mg/l}$ are chosen. For example, if, a value of organic loading such as $F/M = 0.60 \text{ day}^{-1}$ comes into the system during the operation, according to the result of the established model in this study, the removal efficiency will decrease to 60 %. However, due to Figure 5.6 given in Chapter 5, if floc diameter in the system can be arranged to about 0.8 mm, it will be possible to cope with the first design removal efficiency of the system at this loading rate.

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