KALDNES

MOVING BED BIOFILM REACTOR INDUSTRIAL INSTALLATION LIST

FACILITY & LOCATION		ORGANIC LOAD TO PLANT	REACTOR VOLUME m³	REACTOR VOLUME ft³	INDUSTRY & OBJECTIVE
Opoland Chips Norway	1992	1,400 kg COD/day	3,086 lb COD/day	210	7,415	Potato Industry Complete organic removal
Sande Paper Mill Norway	1993	60,000 kg COD/day	132,276 lb COD/day	1,700	60,027	Paper board mill Complete organic removal
Stora Papyrus Grycksbo Sweden	1993	700 kg BOD₇/day	1,543 lb BOD₇/day	135	4,767	Fine paper mill Complete organic removal
Meieriet Sor Norway	1993	210 kg BOD₇/day	463 lb BOD₇/day	65	2,295	Dairy Complete organic removal
Gjovik Potetindustri Norway	1994	1,000 kg COD/day	2,205 lb COD/day	134	1,201	Potato industry Complete organic removal
Lom & Skjak Norway	1994	250 kg BOD₇/day	551 lb BOD7/day	100	3,531	Dairy Complete organic removal
Laga Mill - Natrix Sweden	1994	4,000 kg COD/day	8,818 lb COD/day	200	7,062	Recycle paper mill Post treatment from Trickling Filter
Hilleshog - Natrix Sweden	1994	--	--	2	70	Seeds Degradation of Specific substance
Wargons Bruk – Natrix Sweden	1994	49,000 kg COD/day	108,025 lb COD/day	2,000	70,620	Sulphite pulp, Groundwood Complete organic removal
Shoalhaven Mill Australia	1995	5,000 kg/COD/day	11,023 lb COD/day	576	20,339	Paper mill Pre-treatment prior to AS

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FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME		INDUSTRY & OBJECTIVE
				m^3	ft^3	
SCA Hygiene Paper Germany	1995	8,900 kg COD/day	19,621 lb COD/day	380	13,418	Paper mill Pre-treatment prior to AS
Borgeskogen Norway	1995	200 kg COD/day	441 lb COD/day	40	1,412	Tank washing facility Complete organic removal
Rockhammers Bruk – Natrix Sweden	1995	20,000 kg COD/day	44,092 lb COD/day	150	5,300	Bleached CTMP mill Pre-treatment prior to AS
Utansjo Bruk – Natrix Sweden	1995	70,000 kg COD/day	154,322 lb COD/day	550	19,420	Sulphite pulp mill Pre-treatment prior to AS
Intercell SA – Natrix Poland	1995	22,000 kg COD/day	48,501 lb COD/day	550	19,420	Kraft, NSSC, recycle fiber, kraft paper Pre-treatment prior to AS
Alier Spain	1996	15,000 kg BOD_7/day	33,069 lb BOD_7/day	600	21,186	Paper Mill COD removal Pre-treatment prior to AS
Ostlandsmeieriet Norway	1996	50 kg BOD_7/day	110 lb BOD_7/day	17	600	Dairy Complete organic removal
IVEX Chagrin Falls, Ohio USA	1996	460 kg BOD_5/day	1,014 lb BOD_5/day	166	5,861	Recycle Paper Mill Complete organic removal
Backhammers Bruk – Natrix Sweden	1996	7,000 kg COD/day	15,432 lb COD/day	1,500	52,965	Unbleached Kraft & Paper Complete organic removal
Edet Bruk – Natrix Sweden	1996	8,000 kg COD/day	17,636 lb COD/day	1,700	60,027	Tissue from recycled fiber mill Complete organic removal

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FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME		INDUSTRY & OBJECTIVE
				m^3	ft^3	
Nyboholms Bruk – Natrix Sweden	1996	2,000 kg COD/day	4,409 lb COD/day	500	17,655	Tissue from recycled fiber mill Complete organic removal
Prior Elverum, Norway	1997	840 kg BOD_7/day	1,852 lb BOD_7/day	150	5,297	Chicken processing Complete organic removal
Campri/Sacessav Italy	1997	228 kg BOD_5/day	503 lb BOD_5/day	120	4,237	Distillery Complete organic removal
Werra Papier Germany	1997	565 kg BOD/day	1,246 lb BOD/day	130	4,590	Paper Mill Pre-treatment prior to AS
Asahimatu Food Japan	1997	70 kg BOD_5/day	154 lb BOD_5/day	42	1,483	Soya Been Steaming Complete organic removal
Hoogoven Kwidzyn, Poland	1997	600 m^3/hr	4 MGD	--	--	Nitrification of ground water
Oltedal Norway	1997	325 kg COD/day	716 lb COD/day	129	4,555	Textile Industry
Katrinefors Bruk – Natrix Sweden	1997	8,000 kg COD/day	17,637 lb COD/day	1,500	52,965	Tissue from recycled fiber Complete organic removal
Astra AB – Natrix Sweden	1997	5,000 kg COD/day	11,023 lb COD/day	1,830	64,617	Pharmaceuticals Complete organic removal
Tasman P&P – Natrix New Zealand	1997	15,000 kg COD/day	33,069 lb COD/day	1,500	52,965	TMP mill Pre-treatment prior to aerated lagoon

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MOVING BED BIOFILM REACTOR INDUSTRIAL INSTALLATION LIST

FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME m^3	ft^3	INDUSTRY & OBJECTIVE
Mcwhorter – Natrix Sweden	1997	100 kg COD/day	220 lb COD/day	90	3,178	Chemicals Pre-treatment prior to municipal plant
Munkedals AB – Natrix Sweden	1997	--	--	200	7,062	Printing Paper Post-treatment to trickling filter
SCA Graphic Ostrand - Natrix Sweden	1997	10,000 kg COD/day	22,046 lb COD/day	2,375	83,861	TCF kraft pulp mill Anaerobic/aerobic condensate treatment
ARLA – Natrix Sweden	1997	--	--	100	3,531	Cheese Denitrification
Laga Mill – Natrix Sweden	1997	4,000 kg COD/day	8,818 lb COD/day	200	7,062	Board from recycled fiber Extension of existing treatment plant
Morinaga Co. Ltd Japan	1998	158 kg BOD/day	348 lb BOD/day	60	2,119	BOD-removal Polishing anaerobic
Nippon Light Metal Japan	1998	40 kg COD/day	88 lb COD/day	117	4,131	Metal Finishing COD removal afterAS process
Exxon Chemical Baton Rouge, LA USA	1998	17,700 kg BOD$_5$/day	39,021 lb BOD$_5$/day	1,670	58,968	Chemical Facility, Pre-treatment prior to AS
Valley Pride Pack Norwalk, WI USA	1998	1,337 kg BOD$_5$/day	2,948 lb BOD$_5$/day	266	9,392	Slaughter House, Pre-treatment prior to AS
Phillips Petroleum Borger, TX USA	1998	5,454 kg BOD$_5$/day	12,024 lb BOD$_5$/day	1,927	68,042	Refinery / chemicals, Pre-treatment prior to AS

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FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME		INDUSTRY & OBJECTIVE
				m^3	ft^3	
Norsk Hydro Rafnes, Norway	1998	230 kg BOD_7/day	507 lb BOD_7/day	85	3,001	Complete organic removal
Maggi Kempthal Switzerland	1996	150 kg TN/day	331 lb TN/day	280	9,887	Nitrification
Maggi Kempthal Switzerland	1998	1,000 kg BOD_5/day	2,205 lb BOD_5/day	600	21,186	BOD-removal + pre denit
Thurpapier, Weinfelden Switzerland	1998	4,760 kg COD/day	10,494 lb COD/day	420	14,830	Paper Mill Pre-treatment prior to AS
Fiskeby Board Sweden	1998	10,000 kg COD/day	22,046 lb COD/day	255	9,004	Recycle Paper Mill, Pre-treatment prior to AS
Casco Products AB Sweden	1998	58 kg TN/day	128 lb TN/day	60	2,119	Chemical Industry Post-Denitrification
Karlshamn Sweden	1998	700 kg COD/day	1,543 lb COD/day	400	14,124	Vegetable Oils COD removal
Nordsjo – Natrix Sweden	1998	--	--	12	424	Chemicals Pre-treatment prior to municipal plant
Haindl Paper – Natrix Germany	1998	20,000 kg COD/day	44,092 lb COD/day	1,800	63,558	Newsprint from recycled fiber Pre-treatment prior to AS
Boras Energi – Natrix Sweden	1998	4,000 kg COD/day	8,818 lb COD/day	1,500	52,965	Energy from biofuel Detoxification of condensate

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MOVING BED BIOFILM REACTOR INDUSTRIAL INSTALLATION LIST

FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME		INDUSTRY & OBJECTIVE
				m^3	ft^3	
Enso Pankakoski – Natrix Finland	1998	5,000 kg COD/day	11,023 lb COD/day	540	19,067	Board from recycled fiber Complete organic removal
AssiDoman Packaging – Natrix France	1998	1,500 kg COD/day	3,307 lb COD/day	350	12,359	Board, liner, fluting from recycled fiber Post-treatment after anaerobic fluidized bed
Schonfelder Paper Germany	1999	2,600 kg COD/day	5,732 lb COD/day	270	9,534	Paper Mill Complete organic removal
Piper Impact I New Albany, MS USA	1999	185 kg BOD$_5$/day	408 lb BOD$_5$/day	153	5,402	Metal Finishing Complete organic removal
Piper Impact II New Albany, MS USA	1999	123 kg BOD$_5$/day	271 lb BOD$_5$/day	284	10,028	Metal Finishing Complete organic removal
Bartow Ethanol Bartow, FL USA	1999	3,320 kg BOD$_5$/day	7,317 lb BOD$_5$/day	855	30,175	Alcohol Production Complete organic removal
Dresden Paper Heidenau, Germany	1999	1,400 kg BOD$_5$/day	3,086 lb BOD$_5$/day	240	8,474	Paper Mill Pre-treatment prior to AS
DENSO Co. Ltd. Kouda Factory, Japan	1999	135 kg TN/day	298 lb TN/day	850	30,113	BOD & Nitrogen removal
Sountry Co. Japan	1999	720 kg BOD/day	1,587 lb BOD/day	750	26,482	BOD & Nitrogen removal
Unipac Optotoelectronics Corp. Taiwan	1999	150 kg TOC/day	331 lb TOC/day	400	14,124	Electronics Complete organic removal

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FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME		INDUSTRY & OBJECTIVE
				m^3	ft^3	
Sapporo Breweries Ltd. Japan	1999	2,400 kg BOD/day	5,291 lb BOD/day	1,800	63,558	Beverage Complete organic removal
Koishiya Foods Japan	1999	200 kg BOD/day	441 lb BOD/day	200	7,062	Food Treatment BOD removal
AS Synnove Finden Meieri Norway	1999	800 kg BOD$_7$/day	1,764 lb BOD$_7$/day	300	10,593	Dairy BOD removal
Hadwaco Ltd., Oy Notaresco, Italy	1999	140 kg COD/day	308 lb COD/day	50	1,765	Leachate Treatment COD removal
Morita Co. Ltd Japan	1999	450 kg BOD/day	992 lb BOD/day	184	6,497	Sake Manufacturing BOD removal
Nisseki Mitsubishi – Beverage Japan	1999	--	--	318	11,228	Beverage producing wastewater AS system modification – BOD removal
Steinbeis Temming – Natrix Germany	1999	4,000 kg COD/day	8,984 lb COD/day	420	14,830	Paper from recycled fiber Pre-treatment prior to AS
UPB – Natrix Malaysia	1999	9,000 kg COD/day	19,841 lb COD/day	900	31,779	Paper from recycled fiber Pre-treatment prior to AS
Cargill Foods – Natrix Austrailia	1999	--	--	100	3,531	Oil seed extraction Complete organic removal
Nordic Synthesis – Natrix Sweden	1999	1,000 kg Nitrogen/day	2,205 lb Nitrogen/day	2,400	84,744	Chemicals Denitrification

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FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME		INDUSTRY & OBJECTIVE
				m^3	ft^3	
Astra AB – Natrix Sweden	1999	--	--	--	--	Pharmaceuticals Extension of existing plant
AE Staley Louden, TN USA	2000	10,909 kg BOD_5/day	24,050 lb BOD_5/day	675	23,834	Beverage Pre-treatment prior to AS
Coca-Cola Apizaco, Mexico	2000	4,197 kg BOD_5/day	9,253 lb BOD_5/day	251	8,880	Beverage Pre-treatment prior to AS
Frito-Lay Beloit, WI USA	2000	3,950 kg BOD_5/day	8,700 lb BOD_5/day	286	10,092	Potato Industry Pre-treatment prior to AS
Coop Kobe Japan	2000	1,440 kg BOD/day	3,175 lb BOD/day	570	20,127	Food Processing wastewater BOD removal
YKK Corporation Japan	2000	360 kg BOD/day	794 lb BOD/day	600	21,186	Dyeing wastewater treatment COD removal
GKN Indugasa Spain	2000	140 kg COD/day	309 lb COD/day	24	847	Manufacturer of Transmissions BOD Removal
Quanta Display Inc. Taiwan	2000	360 kg TOC/day	794 lb TOC/day	120	4,237	LCD wastewater Treatment
Yotsuba Milk Products Japan	2000	240 kg BOD/day	529 lb BOD/day	390	13,771	Dairy wastewater Treatment
Oslo Lufthavn AS Norway	2000	1,000 kg COD/day	2,204 lb COD/day	150	5,295	Glycol removal

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FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME m^3	ft^3	INDUSTRY & OBJECTIVE
Jakusevac Croatia	2000	68 kg BOD/d	150 lb BOD/d	25	882	Slaughter House BOD removal
COOP Marrone Italy	2000	180 kg COD/d	395 lb COD/d	70	2,470	Food Industry Roughing reactor
Greenfield – Natrix France	2000	19,000 kg COD/day	41,887 lb COD/day	1,000	35,310	Pulp from recycled fiber Pretreatment prior to AS
AO Kondopoga – Natrix Russia	2000	120,000 kg COD/day	264,552 lb COD/day	5,600	197,736	Sulfite pulp, newsprint Pretreatment prior to AS
Nordic Synthesis – Natrix Sweden	2000	4,000 kg COD/day	8,818 lb COD/day	2,400	84,744	Chemicals Aerobic Treatment
FMC Biopolymer – Natrix Norway	2000	--	--	--	--	Hybrid reactor
ABAC University – Natrix Thailand	2000	--	--	--	--	Sewage Pretreatment prior to AS
Irving Pulp & Paper – Natrix Canada	2000	10,000 kg BOD/day	22,046 lb BOD/day	1,800	63,558	Bleached Kraft Pulp Sole treatment, thermophilic
Titan Pharmaceutical – Natrix China	2000	--	--	--	--	Pharmaceutical Sole Biological Treatment
Raleigh Dairy – Natrix Argentina	2000	--	--	--	--	Dairy Treatment Sole Biological Treatment

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FACILITY & LOCATION	ORGANIC LOAD TO PLANT			REACTOR VOLUME m³	ft³	INDUSTRY & OBJECTIVE
Cosmo Oil Co. Ltd. Japan	2001	740 kg COD/day	1,631 lb COD/day	336	11,864	Refinery wastewater COD removal
Leprino Foods New Mexico, USA	2001	10,600 Kg BOD/d	23,350 lb BOD/d	738	26,071	Mozzarella Cheese Processing
Merck & Co. Puerto Rico, USA	2001	8,165 Kg COD/d	18,000 lb COD/d	1,326	46,800	Pharmaceutical Wastewater
Pharmacia Georgia, USA	2001	1,000 kg BOD/d	2,200 lb BOD/d	560	19,786	Pharmaceutical Wastewater
Ritek-Oled Taiwan	2001	1,500 kg BOD/d	3,305 lb BOD/d	426	15,040	Electronics, BOD removal
Pu Li Hospital Taiwan	2001	52.5 kg BOD/d	115 lb BOD/d	20	706	Hospital, BOD removal
Sun Ten Pharmaceutical Taiwan	2001	60 kg COD/d	132 lb COD/d	10	353	Pharmaceutical, BOD removal
SembCorp U.T. Singapore	2001	7,600 kg BOD/d	16,755 lb BOD/d	1,200	42,370	Chemical Industry, COD removal
Ripasa S.A. Brazil	2001	10,800 kg COD/d	23,810 lb COD/d	1,662	3,664	Paper Mill COD removal
Stora Enso Hylte – Natrix Sweden	2001	82,000 kg COD/day	180,777 lb COD/day	5,000	176,550	Newsprint, TMP, Recycle Paper Pretreatment prior to AS

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FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME m^3	ft^3	INDUSTRY & OBJECTIVE
Asia Kraft – Natrix Thailand	2001	20,000 kg BOD/d	44,100 lb BOD/d			Recycled Paper Mill Pretreatment prior to AS
Sodra Cell, Varo Mill – Natrix Sweden	2001	100,000 kg COD/d	220,500 lb COD/d			Bleached kraft mill Pretreatment prior to AS
Schauman – Natrix Finland	2001	14,000 kg COD/d	30,900 lb COD/d			Plywood Mill Sole biological treatment
Papierfabrik Hainsberg – Natrix Germany	2001					
Leprino Foods California, USA	2002	17,690 kg BOD/d	39,000 lb BOD/d	2,356	83,200	Mozzarella Cheese Processing
Svensk Brikettenergi – Natrix Sweden	2002	2,000 kg BOD/d	4,410 lb BOD/d			Biofuel Dryer Plant Sole Biological Treatment
Filipstads Kommun - Natrix Sweden	2002	10,000 kg BOD/d	22,050 lb BOD/d			Potato chips, Pretreatment prior to AS
Stora Enso Langerbrugge - Natrix Belgium	2002	72,000 kg BOD/d	158,730 lb BOD/d			Recycled paper mill Pretreatment prior to AS
Chudovo - Natrix Russia	2002	1,000 kg BOD/d	2,205 lb BOD/d			Plywood mill Sole biological treatment
Haapavesi - Natrix Finland	2002	8,800 kg BOD/d	19,400 lb BOD/d			Dairy products Pretreatment prior to AS

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FACILITY & LOCATION		ORGANIC LOAD TO PLANT		REACTOR VOLUME m³	ft³	INDUSTRY & OBJECTIVE
AO Kondopoga - Natrix Russia	2002	84,000 kg BOD/d	185,190 lb BOD/d			Newsprint mill Pretreatment prior to AS
Fraser Papers - Natrix Maine USA	2002	16,330 Kg COD/d	36,000 lb COD/d	2,100	74,150	Pulp & Paper
Solvay Paperboard - Natrix New York USA	2003	3,000 Kg BOD/d	6,615	1,300 (one reactor)	45,900	Post-Treatment Pulp & Paper
Nisseki Mitsubishi Oil Co. Japan	2003	1,650 kg BOD/day	3,637 lb BOD/day	1,100	2,425	Power Plant wastewater BOD & N-removal
Klabin Kimberly S.A. Brazil	2003	5,568 kg COD/day	12,275 lb COD/day	614	21,680	Paper Mill COD removal
Glanbia Cheese North Ireland	2003	4,670 kg BOD/day	10,295 lb BOD/day	220	7,768	Dairy Pretreatment ahead of AS
Norwegian Wind Norwegian Construction	2003	405 kg BOD/day	890 lb BOD/day	78	2,754	Maritime wastewater BOD removal
Norwegian Sun Norwegian Construction	2003	488 kg BOD/day	1,075 lb BOD/day	101	3,566	Maritime wastewater BOD removal
Norwegian Sky Norwegian Construction	2003	488 kg BOD/day	1,075 lb BOD/day	101	3,566	Maritime wastewater BOD removal
OSL Gardermoen Norway	2004	4,500 kg COD/day	9,920 lb COD/day	1,380	48,730	De-icing waste from airport BOD removal

BIOLOGICAL NITROGEN REMOVAL FROM MUNICIPAL LANDFILL LEACHATE IN A PILOT SCALE SUSPENDED CARRIER BIOFILM PROCESS

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Abstract—Leachate from a municipal landfill receiving both household and industrial waste was treated in pilot scale suspended carrier biofilm reactors for removal of nitrogen and organic matter. Nitrification was studied in two parallel aerated reactors, filled to 60% (v/v) with two different models of carrier media with specific surface areas of 210 and 390 m^2/m^3 media, respectively. Denitrification was carried out in a mechanically stirred, anoxic reactor filled to 40% with the carrier media having a surface area of 210 m^2/m^3. Nitrified leachate from one of the aerobic reactors was fed to the anoxic reactor together with external carbon source, initially acetic acid and later methanol, to promote denitrification. The leachate temperature varied between 10 and 26°C during the study. Nitrification proceeded well in this temperature range. The highest volumetric nitrification rate, 24 g N/m^3 reactor h (16°C), was obtained for the media with the largest surface area. The maximum denitrification rate with methanol as carbon source was approximately 55 g N/m^3 h. As the process had reached optimal operation, inorganic nitrogen was almost completely removed and the removal of total nitrogen was approximately 90%. The chemical oxygen demand (COD) removal was around 20% during the study. © 1998 Elsevier Science Ltd. All rights reserved

Key words—nitrification, denitrification, biofilm, suspended carrier, landfill leachate

INTRODUCTION

At many municipal landfills, household waste is deposited together with different kinds of waste of industrial origin, sometimes even hazardous waste. This means that the composition of leachates from municipal landfills might be rather varying depending on differences in the waste deposited. Furthermore, the composition of the leachate from one and the same landfill generally changes considerably with the age of the landfill. In young landfills, containing large amounts of readily biodegradable organic matter, a rapid anaerobic fermentation of this matter takes place, resulting in volatile fatty acids (VFA) as the main fermentation products. This early phase of a landfill's lifetime is called the acidogenic phase, and as much as 95% of the organic content in the leachate can be made up of VFA during this phase (Harmsen, 1983).

As a landfill matures it enters the methanogenic phase. Methanogenic microorganisms develop in the waste, and the VFA are converted to biogas, *i.e.* methane and carbon dioxide. The content of VFA and other readily biodegradable organic compounds in the leachate decreases dramatically and

the organic matter in the leachate becomes dominated by refractory compounds, such as fulvic acid-like substances (Harmsen, 1983). In both the acidogenic and the methanogenic phases, most landfill leachates contain high concentrations of nitrogen, mainly in the form of ammonium. The ammonium constitutes an environmental problem as ammonium is toxic in high concentrations (Forgie, 1988) and has fertilizing effects that may lead to eutrophication of the receiving waters.

In order to minimize the negative effects on the environment, landfill leachate must be treated to remove nitrogen and organic compounds before being discharged. At present, a common solution is to treat the leachate together with municipal sewage in the municipal sewage treatment plant. However, this solution has been increasingly questioned due to the concern over the effects of leachates on the biological sewage treatment process and the quality of the sludge generated in this process. It has been indicated that inhibitory compounds in the leachates may disturb the biological sewage treatment process, and the accumulation of hazardous compounds, such as bioaccumulable organics and heavy metals, from the leachates in the sludge formed in the treatment process may make the use of this sludge as fertilizer in agriculture impossible (Lagerkvist, 1986; Linde, 1995).

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As a consequence of the ambition to disconnect landfill leachate from the municipal sewage treatment, the interest in processes for separate treatment of leachate has grown. Methods that have been studied and adopted for use in leachate treatment include suspended growth systems such as aerated lagoons, conventional activated sludge processes and sequencing batch reactors (SBR) (Robinson and Grantham, 1988; Hosomi et al., 1989). As these methods have been found to suffer from different limitations, such as sensitivity to low temperatures and loss of activity due to poor separability of the biomass (EPA, 1975; Knox, 1985), there has been a continued search for other alternatives. A number of investigators have reported on tests on leachate with attached growth systems such as trickling filters (Knox, 1985), submerged aerobic biological filters (Pedersen and Jansen, 1992) and rotating biological contactors (Spengel and Dzombak, 1991). These systems have the advantage of not suffering from loss of active biomass due to poor separability and nitrification in attached growth systems has been shown to be significantly less negatively affected by low temperatures than in suspended growth systems (Knox, 1985).

The suspended carrier biofilm process (SCBP) or moving bed biofilm process is a new type of attached growth system that has gained a lot of interest in recent years (Ødegaard et al., 1994). This process is based on the use of plastic carrier elements which are kept in suspension and movement in the process while the active microorganisms grow as a biofilm on the surfaces of the carrier media. The media is contained in the system by a screen at the outlet of the reactor. The SCBP has become extensively used for nitrification and denitrification of municipal sewage and recently laboratory tests concerning nitrification of landfill leachate in a SCBP have been conducted (Welander et al., 1997). These tests showed nitrification of leachate in this kind of process to proceed at a relatively high rate at temperatures as low as 5°C, indicating this kind of process to be well suited for treatment of landfill leachate under realistic conditions.

The aim of the present investigation was to study biological nitrogen removal from landfill leachate, in a SCBP, on-site at a landfill under realistic conditions (variations in temperature and leachate composition, etc.), with the goal of obtaining data that could be used for design of a full-scale process.

MATERIAL AND METHODS

Landfill leachate

Leachate from the Hyllstofta landfill, Sweden, was used in the experiments. The Hyllstofta landfill was started in 1975 and was still in use during the time of this study (1995). The landfill is a traditionally mixed landfill receiving both household wastes and industrial wastes. The industrial waste is dominated by plastic material from a large plastic industry located near the landfill. The total amounts of the different kinds of waste deposited at the landfill at the time of study were 330 000 tons of household waste, 430 000 tons of industrial waste, 190 000 tons of sludge and 100 000 tons of ashes. The amount of leachate produced has typically been between 60 000 and 70 000 m³ annually. The leachate is collected in a lagoon with a volume of 12 000 m³ before being discharged to the municipal sewage treatment plant. Leachate from this lagoon was used in the experiments.

The leachate used in the study contained 400–800 mg/l ammonium nitrogen, 800–2000 mg/l COD and a biochemical oxygen demand (BOD_7) of 30–140 mg/l.

Pilot plant trial

The pilot study was carried out in a two-stage suspended carrier biofilm process. The first reactor (reactor 1) was operated aerobically for nitrification and removal of organic matter, while the second reactor (reactor 2) was operated in an anoxic mode for denitrification with addition of external carbon source.

Reactor 1 consisted of a 5 m³ plastic tank filled to 60% with carrier media (Natrix 6/6C, ANOX AB, Lund). The reactor was aerated by means of a fan and fine bubble membrane diffusers on the bottom of the tank. The pH in the reactor was automatically controlled to 7.0 (day 1–15) and 7.8 (from day 16). Initially, before a significant nitrification was established, the pH adjustment was made with sulfuric acid, while sodium hydroxide was used after a high degree of nitrification was obtained. The reactor was inoculated with 5 m³ nitrifying activated sludge from the municipal sewage treatment plant in Helsingborg, Sweden, and leachate was fed to the reactor at a flow rate corresponding to a hydraulic retention time (HRT) of 4 days. The HRT was then decreased in stages during the trial to determine the maximum loading capacity of the process.

Reactor 2 was a plastic tank with a volume of 900 l filled to 40% with the same kind of media as reactor 1. The carrier media in reactor 2 was kept in movement by means of a mechanical stirrer (30 rpm). Acetic acid was used as external carbon source from day 1 to 54, after which methanol was used. The dosage of carbon source was adjusted manually during the test with the aim to dose approximately 4 g COD per g NO_3^-–N entering the anoxic reactor. Phosphorus in the form of phosphoric acid was dosed to reactor 2 throughout the study in an amount of approximately 10 g phosphorus per m³ leachate. Reactor 2 was not inoculated except for the inoculum entering the reactor with the effluent from reactor 1 (Fig. 1).

On day 19, a third reactor (reactor 3) was started up in order to compare the performance of a new carrier model with the one used in reactor 1 and 2. Reactor 3 had a volume of 900 l and it was filled to 60% with carrier media (Natrix 12/12C, ANOX AB, Lund). It was operated aerobically for nitrification in parallel with the two-stage system, and without any subsequent denitrification. The reactor was aerated by the same fan as used for reactor 1, but a course bubble aeration system was used. The reactor was inoculated with 900 l of the same activated sludge as used for reactor 1 and the start-up and operating procedures were also the same as for reactor 1 except for beginning at an HRT of 3 days.

Carrier media

Natrix model 6/6C is a plastic carrier made of polyethylene mixed with lime and small amounts of other compounds. Each carrier element is built up of walls (altogether 12) arranged radially to form a cylindrical body with a length and diameter of 32 and 30 mm, respectively. The carrier has an inner surface area of 210 m²/m³ carrier. Only the inner surface is available for biofilm growth as the outer surface is continuously cleaned from
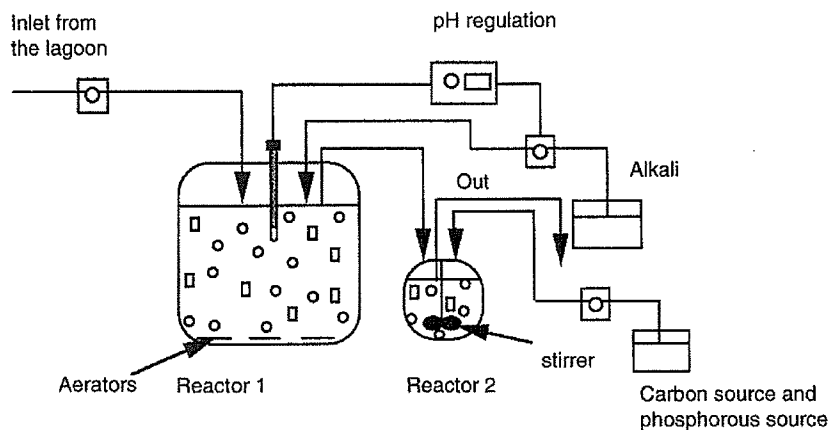
Fig. 1. The experimental setup for reactor 1 and 2.

biofilm by the frequent collisions between the carrier elements. Natrix model 12/12C is similar to model 6/6C except for the number of walls being doubled, resulting in a protected surface area of 390 m^2/m^3 carrier.

Analyses

NH_4^+-N, NO_3^--N + NO_2^--N, NO_2^--N and total nitrogen were measured according to Swedish Standard (SS 02 81 34, SS 02 81 33, SS 02 81 32 and SS 02 81 31, respectively) and COD was analyzed according to Swedish Standard (SS 02 81 42). Total suspended solids (TSS) was analyzed as described in Standard Methods (APHA *et al.*, 1985). BOD_7 was analysed according to Swedish standard SS 02 8193.

RESULTS

The temperature of the leachate varied between a minimum of approximately 10°C and a maximum of close to 26°C during the investigation period (Fig. 2).

Nitrification

During start-up, nitrification increased continuously, resulting in a successively decreasing ammonium concentration in the reactor (Fig. 3). During the first month, nitrite was the main pro-

duct of nitrification (Fig. 4), but as the ammonium concentration reached a low level, the nitrite concentration dropped dramatically and a complete nitrification to nitrate was obtained (Fig. 4).

The HRT in reactor 1 was kept at 4 days during the first 67 days of the experiment, after which it was decreased to 3 days (Fig. 5). As nitrification was still complete at this HRT, the HRT was further decreased to 2.5 days on day 103. This resulted in a slight increase in the ammonium concentration to a few mg/l, indicating that the maximum capacity of the reactor was reached. The nitrification rate achieved during this period was around 12 g N/m^3 reactor h at a temperature of 17°C.

The biofilm on the carrier media was very thin, barely visible to the naked eye, throughout the test. However, microscopical investigations revealed a compact biofilm containing a large fraction of dense bacterial clusters typical of the growth of nitrifiers. The TSS concentration in both the untreated and the nitrified leachate was approximately 150 mg/l.
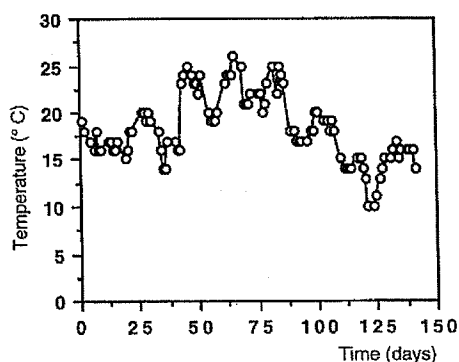
Fig. 2. The temperature as a function of time in reactor 1.
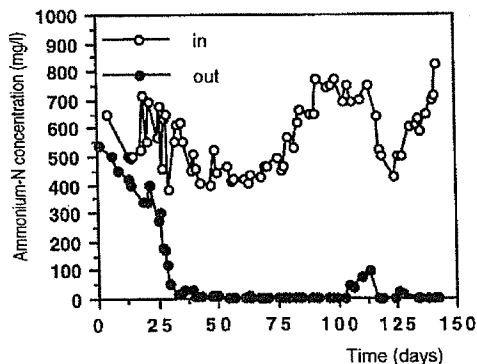
Fig. 3. The NH_4^+-N concentration in the inlet to and the effluent from reactor 1.
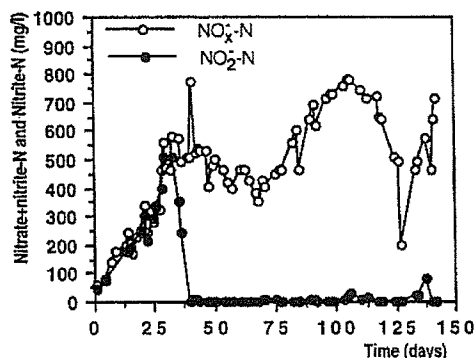
Fig. 4. The NO_x^--N and the NO_2^--N concentrations in the effluent from reactor 1.
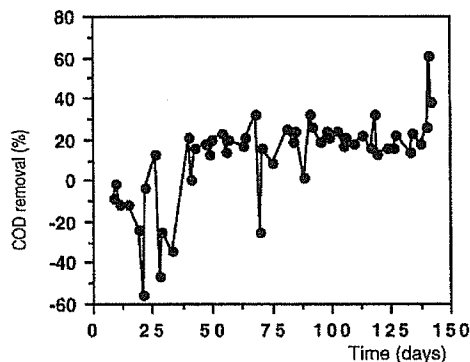

Fig. 6. COD removal in reactor 1 vs. time.

Throughout the experiment, the air flow through reactor 1 was regularly adjusted in relation to the loading and the temperature to maintain a dissolved oxygen (DO) level in the reactor between 80 and 90% of the saturation level. On day 112, the HRT was increased to 4.3 days in order to be able to decrease the air flow through the reactor without causing a lack of oxygen. During the final period of the experiment, more air was needed in reactor 3 to be able to test the loading capacity of this reactor and the capacity of the fan was not high enough to supply both reactors with maximum air flows.

During start-up, the COD-level in the leachate increased in reactor 1 (due to nitrite production), after which a COD-removal around 20% was obtained (Fig. 6).

Reactor 3 was started up at an HRT of 3 days (Fig. 5). After approximately two months of operation, a complete nitrification was achieved (Figs 7 and 8). The nitrite production during start-up was considerably lower in this reactor than in reactor 1 (Fig. 8). On day 68, the HRT was decreased to 2.3 days. Initially, a complete nitrification was maintained at this HRT, but as the ammonium concentration in the influent leachate increased, significant levels of ammonium were detected in the reactor (Fig. 7) and a large portion of the ammonium being nitrified ended up as nitrite (Fig. 8). This incomplete nitrification was clearly related to problems keeping the oxygen concentration at a sufficient level during this period; DO levels around 2 mg/l were measured in the reactor. After an increase in aeration on day 100, the DO level increased to around 5 mg/l and both ammonium and nitrite decreased to insignificant levels (Figs 7 and 8). On day 107, the HRT was further decreased to 1.8 days, maintaining a complete nitrification, although significant nitrite levels were measured on a couple of occasions (Figs 7 and 8). On day 118, the HRT was decreased to 1.2 days and the aeration was further increased. At this loading, the ammonium concentration in the reactor increased to a significant level, indicating the maximum capacity of the reactor being reached (Fig. 7). The nitrification rate achieved during this period was around 24 g N/m^3 reactor h at a temperature of 16°C.

The biofilm on the carrier media in reactor 3 had the same character as described above for the biofilm on the media in reactor 1 while the nitrified leachate contained a relatively large number of free living small bacteria and a few flagellates.
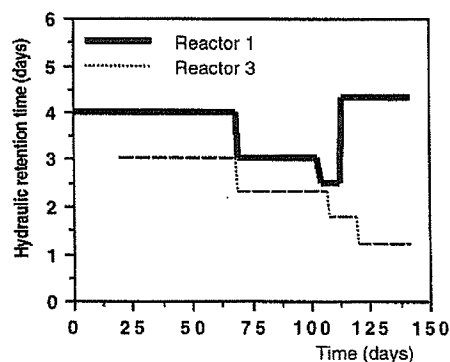

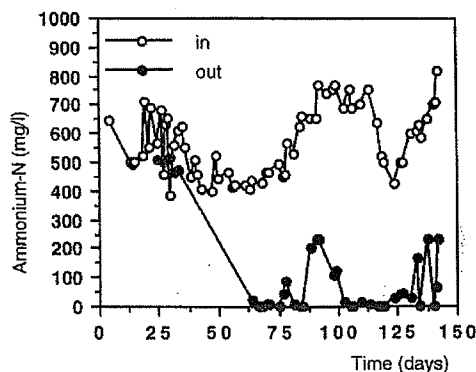Fig. 5. The hydraulic retention time in reactors 1 and 3 vs. time.

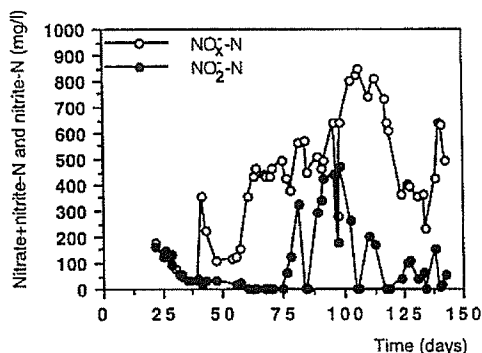Fig. 7. The NH_4^+-N concentration in the inlet to and the effluent from reactor 3.

Fig. 8. The NO_x^--N and the NO_2^--N concentrations in the effluent from reactor 3.

Denitrification

As all effluent from reactor 1 was led to reactor 2, the HRT in reactor 2 followed the HRT in reactor 1 throughout the test, being 18% of the later.

Denitrification started quickly as nitrification increased in reactor 1 and nitrate entered reactor 2. After approximately two weeks, a complete denitrification was achieved in reactor 2 at an HRT of 0.72 days (Fig. 9). However, after almost two weeks of complete denitrification, problems balancing the dosage of acetic acid with the consumption led to an overdosage, resulting in a sharp drop in the pH and a severe disturbance of denitrification (Fig. 9). After a period of more careful dosage of carbon source, denitrification recovered. However, to be sure of avoiding another pH-drop, acetic acid was changed for methanol on day 54.

Acclimation of the denitrification process to methanol proceeded rather slowly, the nitrate level decreasing successively during a period of approximately 40 days until a nearly complete denitrification was achieved around day 100. On two occasions during this period, day 82 and day 89, failure of the methanol dosage led to rapid increases in the nitrate level (Fig. 9). On day 103 the HRT in reactor 2 was decreased from 0.54 to 0.45 days,
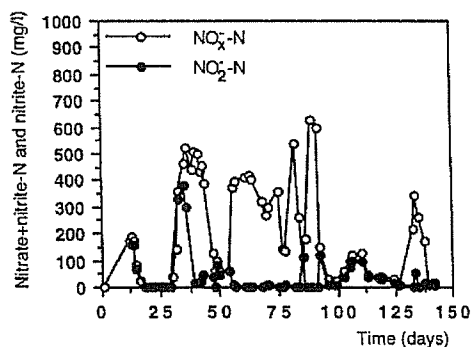
Fig. 9. The NO_x^--N and the NO_2^--N concentrations in the effluent from reactor 2.

resulting in an increase in the nitrite level up to around 130 mg N/l (Fig. 9). As the HRT was prolonged to 0.77 days on day 112, the nitrate concentration decreased to a low level again. A temporary peak in the concentration occurred around day 134 due to another failure of the methanol dosage. The maximum denitrification rate achieved during the test was around 55 g/m^3 reactor h at a temperature around 17°C. The removal of total nitrogen in the system was approximately 90% when denitrification was close to complete.

During the investigation, a rather thick (0.5–1 mm) biofilm developed on the surface of the carrier media in reactor 2. Microscopical investigations showed a dense biofilm, dominated by rod-formed bacteria. No appendaged bacteria of the *Hyphomicrobium* type, often found in systems denitrifying with methanol as carbon source, could be observed. The liquid phase contained large flocs which settled well and a relatively large number of free living bacteria and low concentrations of flagellates. The TSS concentration was between 300 and 400 mg/l.

DISCUSSION

The results of the present study show suspended carrier biofilm technology to be an interesting alternative for biological removal of nitrogen from landfill leachate. Although denitrification suffered from practical problems with the dosing of external carbon source during a considerable part of the test, the high capacity of the process was clearly demonstrated when the process had acclimated to methanol as carbon source and the dosing worked properly. Under these conditions, a nearly complete removal of inorganic nitrogen and a total nitrogen removal of around 90% were achieved.

The volumetric nitrification and denitrification rates achieved in this study, 24 g N/m^3 h (16°C, reactor 3) and 55 g N/m^3 h (17°C), respectively, were at least as high as the highest rates reported for other types of processes operated on leachate (Table 1).

In the present study, the comparison between the two models of media showed a clear proportionality between the nitrification rate and the surface area of the media. Obviously, the carrier model with the largest surface should be the model of choice for full-scale nitrification of leachate. However, increasing the surface area of the media to an even higher level than 390 m^2/m^3 would hardly further increase the nitrification rate as this would require more narrow flow passages through the carrier elements, decreasing mass transfer. Optimal design of carrier elements for the SCBP always requires a proper balance between media surface area and an openness for free flow to enhance mass transfer, as liquid film diffusion has been shown to be an important parameter in the SCBP (Hem, 1991).

Table 1. The volumetric nitrification and denitrification rates reached by different biological processes operated on leachate and SCBP operated on municipal wastewater respectively leachate

Process	Nitrification rate (g N/m^3 h)	Denitrification rate (g N/m^3 h)	Temperature (°C)	Ref.
Trickling filter	3		16	Knox, 1985
Activated sludge	21		13	Knox, 1985
Submerged aerobic biological filter	24		20	Pedersen and Jansen, 1992
Rotating biological contactor (RBC)	17		20	Spengel and Dzombak, 1991
SCBP, municipal wastewater	12.5–16.7			Rusten et al., 1995a
SCBP, laboratory scale	11–40		20	Welander et al., 1997
Trickling filter		6.3	18–21	Werner and Kayser, 1991
Activated sludge		2.5	18–21	Werner and Kayser, 1991
Fluidized bed		22.9	18–21	Werner and Kayser, 1991
SCBP municipal wastewater		28		Rusten et al., 1995b

The biofilm thickness developed in the anoxic reactor in the tests rules out the use of the media with the largest surface in this application, as the media would most certainly be clogged with biomass. Obviously, different media designs are optimal for different applications of the SCBP. The choice of media has to be based on knowledge of the application and the water to be treated, and factors such as expected biofilm thickness, etc., have to be considered.

The comparison of the reaction rates obtained in the present study with the rates reported for other processes shows that nitrification and denitrification of landfill leachate can be accommodated in a compact process when using suspended carrier technology. Other advantages of the SCBP, as compared to suspended growth systems, such as the activated sludge process, are the simplicity of operation, the low risk of losing the biomass due to poor separability and the weak temperature dependence (Welander et al., 1997). Furthermore, attached growth systems are generally considered less sensitive to toxicity and variations in environmental conditions, a characteristic of value when treating landfill leachate with such a high nitrogen content that there may be risks for substrate as well as product inhibition of nitrification, especially if nitrite accumulates (Anthonisen et al., 1976). The main drawback of the SCBP, and other attached growth systems, is the need for operation at a high DO level to maintain high nitrification rates (Ødegaard et al., 1994). The oxygen dependence of the SCBP has been shown to be approximately first order and not half order as for other biofilm processes (Harremoës, 1982; Hem et al., 1994).

When compared to most other types of attached growth systems, the main advantage of the SCBP is the very good mixing of the reactor contents, resulting in an efficient mass transfer and elimination of the risks of liquid short-circuiting and clogging of the media with biomass or other solids.

The low COD removal achieved in the tests, around 20%, is in agreement with the high COD to BOD ratio of the untreated leachate. The poor biodegradability of the organic content is typical for leachates from landfills in the methanogenic phase,

for which most biodegradable organic matter is converted to biogas already in the landfill. The increase in COD over reactor 1 during start-up is explained by the high amount of nitrite formed during this period. As ammonium is not oxidized in the COD analysis, but nitrite is, conversion of ammonium to nitrite in the process results in an increase in COD.

Nitrite accumulation in the aerobic systems occurred mainly during start up and during periods of low DO concentrations (reactor 3) or overloading of the process. Therefore, in a properly designed full scale process, nitrite should not be expected to accumulate except for during start-up. It should also be noted that nitrite accumulation in the aerobic stage is not necessarily a problem, as long as the concentration does not reach a level that is toxic to nitrification. When the aerobic stage is followed by an anoxic stage, nitrite as well as nitrate is removed through denitrification. As a matter of fact, partial nitrification to nitrite followed by denitrification of the latter is an economically favorable process, as less oxygen is needed in the aerobic stage and less carbon source in the anoxic stage than when nitrification goes all the way to nitrate. In view of this, it is more important to avoid nitrite accumulation in the anoxic stage due to partial denitrification (nitrate/nitrite respiration), which could result in discharge of toxic nitrite into the receiving waters. In the pilot plant study, nitrite accumulation in the anoxic stage occurred as this reactor was overloaded. Therefore, when designing a full-scale process, it is important to use a satisfactory safety margin to minimize the risk of nitrite discharge.

The need for process control would be rather limited in a full scale SCBP for leachate treatment. Primarily, the pH in the aerobic stage and the dosing of external carbon source to the anoxic stage would have to be controlled. Even though nitrification is known to have a high optimum pH level, it might be advantageous to start the process at a lower pH, not higher than 7, in order to avoid ammonia toxicity. As nitrification increases, the pH can then be increased to an optimal level, around 7.5, which would also decrease the risk of nitric

acid toxicity if nitrite accumulates. This strategy was partly used in the pilot plant trials to be sure of avoiding toxicity problems.

The dosage of external carbon source to the anoxic stage should most certainly be the most tricky parameter to control in a full scale process, especially when the nitrogen content in the leachate shows great variations, as it did during the present investigation. Too low a dose in relation to the nitrate to be denitrified immediately results in decreased nitrogen removal, while an overdosage results in unused carbon source remaining in the treated leachate with increased COD and BOD discharge as a consequence. The latter problem can be avoided by introducing another aerobic stage for removal of any excess carbon source after the anoxic reactor. In practice, this would probably be necessary to avoid BOD discharge. Nevertheless, an efficient dosage strategy is required, as an overdosage followed by aerobic posttreatment is undesirable from an economic point of view. Possibly, carbon source dosage based on on-line nitrate measurements could be an interesting alternative.

CONCLUSIONS

Suspended carrier biofilm technology is an efficient alternative for biological treatment of landfill leachate. A high degree of nitrogen removal can be achieved at high volumetric nitrification and denitrification rates.

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1,4-DIOXANE BIO-DEGRADATION PILOT STUDY AT THE LOWRY LANDFILL SUPERFUND SITE

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ABSTRACT

Groundwater from the Lowry Landfill Superfund Site just outside of Denver, CO, has been treated using a UV catalyzed hydrogen peroxide advanced oxidation process, which has had difficulty removing a target compound (1,4-dioxane) from some Site groundwater streams to low levels due to UV absorbance/interference issues. Bench-scale studies were completed to investigate biological degradation of the target compound. Seven bench-scale conventional activated sludge systems were operated over six months at a variety of feed mixtures, loading rates, and sludge retention times. The bench scale studies showed that 1,4-dioxane is amenable to biological treatment, with removals to below the detection limit at several operating conditions.

Subsequent the bench-scale studies, a Pilot Study was conducted on-site to demonstrate the sustainability of 1,4-dioxane degradation, the consistency of the level of treatment achievable, and to investigate the effects of different operating parameters on the performance of the system. The Pilot Study was operated as a fixed film, moving bed bio-reactor (MBBR) system utilizing Kaldnes® media to support the biological growth. The Pilot Study demonstrated consistent 1,4-dioxane degradation occurred in the presence of tetrahydrafuran (THF) at various operating conditions. Additionally, the study showed that 1,4-dioxane removal efficiency was not significantly affected by temperatures in the 15-25 °C range, at the mass loadings tested.

KEYWORDS:

1,4-dioxane, tetrahydrafuran, landfill leachate, activated sludge, Kaldnes, MBBR

INTRODUCTION

The Lowry Landfill Superfund Site is located approximately 20 miles southeast of downtown Denver, Colorado, in unincorporated Arapahoe County. From the mid-1960s until 1980, the Site was operated as an industrial liquid waste and municipal solid waste landfill. Liquid wastes disposed of at the Site included substances such as industrial liquids and sludges containing chlorinated solvents, heavy metals, and petroleum hydrocarbons.

As part of the overall remedy for the Site, groundwater is extracted and treated in an on-site water treatment plant (WTP) prior to discharge to the local municipalities wastewater treament facility. The WTP has two primary sources of groundwater, the North Boundary Barrier Wall (NBBW) and the North Toe Extraction System (NTES). The NBBW is a extraction trench used to maintain hydraulic control of groundwater migrating away from the Site and has relatively few organics. The NTES is a trench at the downgradient limit of the landfill and has a high concentration of organics, including 1,4-dioxane and tetrahydrofuran (THF).

The existing groundwater treatment plant consists of (in order of treatment):

- Equalization

- Chemical Softening

- pH Adjustment

- Bag Filtration

- UV/Oxidation

- Activated Carbon

The current treatment is sufficient for most of the permitted parameters, however poor UV transmittance and possible UV interference by inorganics (caused by NTES waters) prevents UV/Oxidation from removing 1,4-dioxane to below the permit limits.

Biological pretreatment was piloted to evaluate whether 1,4-dioxane permit levels could be achieved. Previous bench scale studies indicated that greater than 95% 1,4-dioxane removal efficiencies were possible when feeding a 30% NTES / 70% NBBW Blended water. However, biological reactors fed 100% NTES groundwater showed signs of inhibition, removing just 70% of the 1,4-dioxane. A Pilot Study was performed to further define under what conditions biological treatment was possible, with the four primary objectives of:

1. Evaluating the sustainability and robustness of 1,4-dioxane degradation;

2. Assessing the effect of temperature on the degradation of 1,4-dioxane;

3. Assessing the effect of organic loading on the stability of 1,4-dioxane degradation; and

4. Providing criteria for sizing and design of a full-scale biological treatment system.

Research has shown that both dioxane and tetrahydrofuran (THF) are degradable in fixed film biological systems (Sock, 1993); however, the removal rate was very dependant upon reaction temperature. Sock showed that removal efficiency for dioxane was optimal near 35 ºC with activity decreasing at 20 ºC. Sock's biomass was developed using a mixture of THF and 1,4-dioxane, but was able to sustain growth with 1,4-dioxane as a sole substrate. Researchers at North Carolina State University (Zenker et al., 2000) developed a culture that could only degrade 1,4-dioxane cometabolically in the presence of THF. Furthermore, Zenker showed that THF acted as a competitive inhibitor of 1,4-dioxane degradation, with the enzymes responsible for the ether degradation preferentially degrading THF, when present.

MATERIALS AND METHODS

Design and Operation

Two 350 gallon HDPE tanks (PolyProcessing) were used as biological reactors. The reactors were filled to 50% volume with plastic media (specific gravity of 0.98) (Kaldnes®) in suspension to support thebiomass. Once loaded with biomass, the media approached neutral buoyancy and was more easily mixed throughout the reactor. This configuration is a moving bed bio-reactor (MBBR). The media was retained in the reactor by coarse wedgewire screens over the exit piping. The system was initially operated as a sequencing batch reactor (SBR) with the following cycles (times were the base design and were operator adjustable during the study):

1. Fill (120 minutes): Feed was added to the reactors. The fill period consisted of two steps, anoxic fill with just mechnical agitation and an aerated fill. The length of the anoxic fill and aerated fill cycles were determined by the operator based on operational data. Because the feed was added throughout a large portion of the reaction time, the reactor behaved like a continuously stirred tank reactor (CSTR).

2. React (60 minutes): This was the time that the system was allowed to react without feed. The length of the react period depended on the length of the fill period and the kinetics of the biomass. During this time the reactor was mixed and aerated.

3. Settle (30 minutes): This period was a clarification step where mixing and aeration are stopped and the biomass was allowed to settle to the bottom of the reactor and the Kaldnes® media was allowed to float to the top of the reactor.

4. Decant (30 minutes): The clarified effluent was decanted by pump from the reactor to allow volume for additional treatment. The decant time was dependent on the volume of water to be decanted and the flow rate at which it was removed.

5. Idle (0 minutes): The time between the end of the decant cycle and the start of another fill period. The idle time depends on cycling of other reactors in the system and is there to insure that one reactor is filling at all times. No idle time was used during the study.

A piping and instrumentation drawing of the pilot system is provided as Figure 1. Aeration was by fine bubble diffusion (EDI, 2 cfm/each) with the oxygen being provided by facility instrument air. Flow to both reactors was monitored by an inline rotameter and controlled by a needle valve. The cycles noted above were controlled by PLC to allow for the cycling of equipment to meet operational requirements. Both the agitator and the diffusion air were controlled by a PLC operated, normally closed, solenoid valve on the airline from the air receiver.

Temperatures in the reactors were controlled by immersed temperature elements. Both reactors were provided with low-watt-density resistance-type immersible heaters (Chromolox, 12 kW) to allow operation at elevated temperatures (25°C). Reactor 2 was equipped with immersed cooling coils connected to a small external chiller (Aqualogic 22,000 BTU/hr) to allow for simulation of the normal process water temperature (15°C).

Effluent from the tank was pumped from a small section of the tank separated from the media by a wedge-wire screen (Industrial Screen Products). The screen had 5-mm openings to prohibit the media (10-mm diameter) from passing with the effluent. Decant drainage and effluent sampling were performed from this media-isolated portion of the tank.

Figure 1: Pilot Reactor P&ID
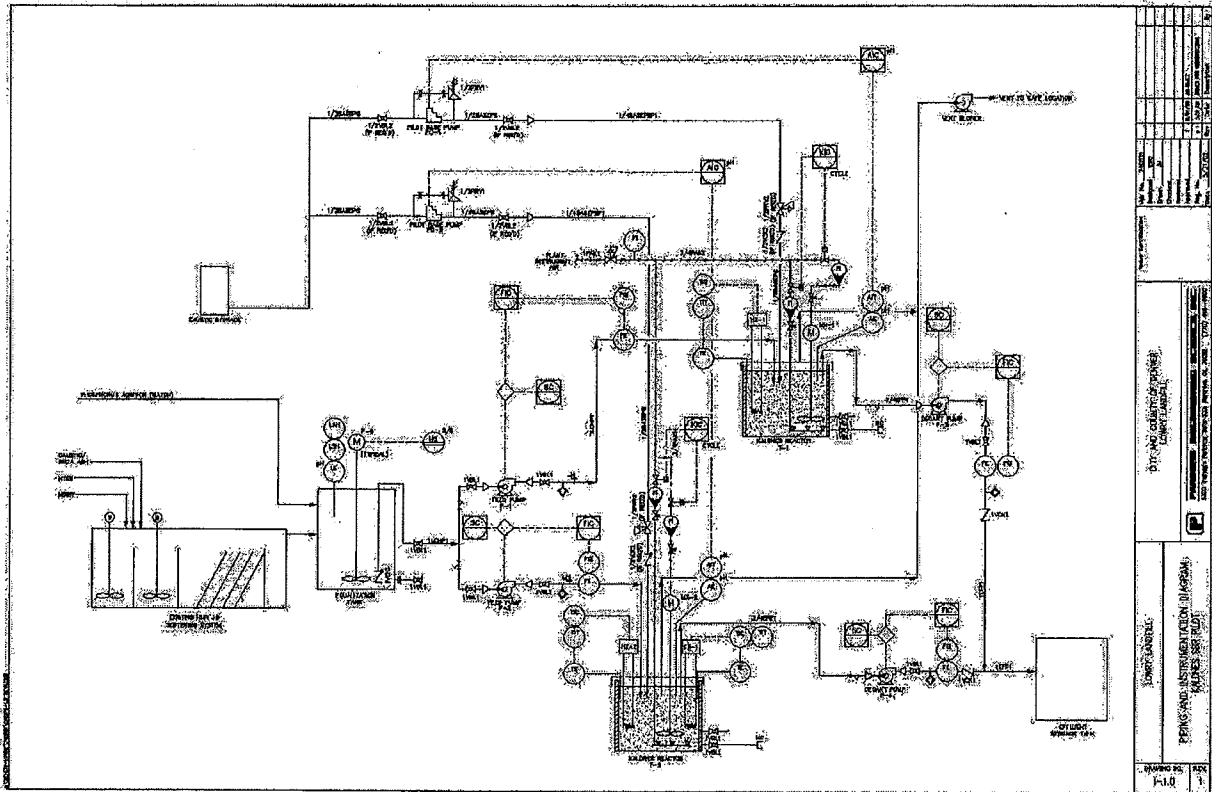
Feed was added with a variable-speed rubber-impeller centrifugal pump (ITT Jabsco) capable of delivering 0.3 to 1.5 gpm. The flow rate was monitored with an inline magnetic flow meter (Rosemount). The system PLC used the flow measurement to control pump speed. After the settle period, effluent was removed by a self-priming rubber-impeller centrifugal pump (ITT Jabsco) capable of 1 to 5 gpm. The pump was attached to a pipe that was lowered into the top of the tank to the desired effluent volume for the loading being tested in SBR operation. The down pipe was placed midway down the tank straight side of the reactor for the continuous flow tests (described below).

Reactor pH and temperature were continuously monitored throughout the pilot study using on-line temperature compensated pH meters (Rosemount). pH control was provided by the PLC, which turned a chemical metering pump (Pulsfeeder) on and off to supply the reactors with 25-percent NaOH to maintain a pH setpoint of 7.0 during the latter stages of testing.

During the Pilot Study, it was determined that denitrification was not possible due to carry over of dissolved oxygen from the aeration cycle; insufficient BOD was provided to the bacteria to remove the excess oxygen. Therefore, after 5.5 months of operation, the treatment system was converted to continuous flow reactors for the duration of the Pilot Study.

Analysis

Sample analysis was split between several laboratories. COD, ammonia, dissolved oxygen, pH, and solids/settling measurements were primarily performed in the field, with some verification by Severn Trent Laboratories (STL) in Denver, CO. Ammonia and dissolved oxygen were performed by ion selective electrode. COD was performed using high range (0-1500 mg/L) HACH vials. Initially the CODs were performed per the HACH standard method; however, it was discovered that the chlorides, bromides and ammonia contained in the groundwaters caused an overestimation of the COD due to oxidation in the COD vials(Belkin et al., 1992; Kim, 1989). Therefore, the COD method was adjusted to minimize the interference. All other parameters were measured using Standard Methods.

TOC, nitrate, nitrite, TKN, BOD$_5$, TDS, and THF were measured by STL using EPA standard methods. 1,4-dioxane measurements were conducted by Parsons and Zymax (San Luis Obispo, California) using EPA method 8021B.

PROCESS WATER CHARACERISTICS

Characteristics of the influent feed to the Pilot Study reactors (30/70 NTES/NBBW Blend) are summarized in Table 1.

Between November 2, 2002 and August 28, 2003, feed water processed through the pilot system was prepared in a batch mode and stored in the Pilot Equalization Tank on a regular basis. During the batch process, the NTES and NBBW groundwaters were blended in the softening tank of the existing WTP where caustic was added to raise the pH to 10.5. Solids were precipitated in the existing WTP clarifier. After softening, the water's pH was adjusted to 8.0 and pumped to the Pilot Equalization Tank (referred to as the "Blend Feed").

TABLE 1: INFLUENT CHARACTERISTICS

Parameter	Units	Blended Feed
Flow	gpm	0.4-0.75
1,4--dioxane	µg/L	6,400-12,000
THF	µg/L	13,500-25,000
Soluble COD	mg/L	150-230
TOC	mg/L	62-92
TDS	mg/L	4300-5000
Ammonia	mg-N/L	80-160
Nitrite	mg-N/L	0.1-3.0
Nitrate + Nitrite	mg-N/L	10-19

Between November 2, 2002 and August 28, 2003, feed water processed through the pilot system was prepared in a batch mode and stored in the Pilot Equalization Tank on a regular basis. During the batch process, the NTES and NBBW groundwaters were blended in the softening tank of the existing WTP where caustic was added to raise the pH to 10.5. Solids were precipitated in the existing WTP clarifier. After softening, the water's pH was adjusted to 8.0 and pumped to the Pilot Equalization Tank (referred to as the "Blend Feed").

Organic Loading:

The Bench Tests showed that a large heal of COD remained in the effluent after treatment; 50% in the Blend reactors and up to 70% of the COD applied to the NTES reactors. This may be a residual affect of the inorganic inhibition of the COD test. Therefore, for the Pilot Test, Parsons defined the organic loading based strictly on the COD associated with 1,4-dioxane and THF. This was done because 1) the majority of the COD removed in the system can be attributed to 1,4-dioxane and THF removal and 2) 1,4-dioxane and THF could be measured independently from COD. The loading due to 1,4-dioxane and THF COD was calculated using the theoretical CODs for 1,4-dioxane and THF of 1.82 and 2.44 gram per gram COD, respectively.

RESULTS

Startup

The Pilot reactors were seeded prior to beginning operation to insure sufficient biomass for oxidation. 30-gallons of nitrifying activated sludge from the Denver Metropolitan Wastewater Reclamation Facility (Metro) was used as seed in each reactor. The initial fill of the Pilot reactor consisted of:

1. Kaldnes® media to 50-percent of the volume of the reactor
2. Approximately 30 gallons RAS from Metro; and
3. Blended waste to fill the reactor to capacity.

The primary objective of the startup period (September 7, 2002 to October 22, 2002) was to begin biomass attachment to the Kaldnes® media. To increase biomass development, additional COD was added to the reactors in the form of dog food. During the startup period the reactors

were run in a batch mode to maintain good contact between the suspended biomass and the media. The dog food was added periodically and a batch was allowed to run until the COD dropped below 100 mg/L. The process was repeated for a month to allow biomass to grow on the media. Then the reactors were switched to SBR operation and the dog food supplements were discontinued. The average performance of the reactors during Startup and Acclimation are shown in Table 2.

Acclimation

As defined here, the period between startup and consistent treatment is acclimation. Acclimation of the reactors occurred between October 23, 2002 and February 13, 2003. The reactors were operated as SBRs and process water feed batches were made once every two weeks. During acclimation, both reactors experienced periods of poor performance, typically following sloughing of biomass from the media. Figure 2 shows the reactor performance with respect to the changing of the feeds and the media solids sloughing events. Figure 3 shows that the sloughing was likely caused by feed/starve cycles of the biomass due to THF biodegradation in the Pilot Equalization Tank. In order to reduce the degradation of THF in the EQ tank, the feed batch frequency was increased to once weekly on February 13, 2003. A further refinement was implemented in April 2003 where the volume of each batch of feed was reduced, eliminating excessive carryover from the previous batch. This feed procedure was continued while the Blend water was being tested.

Table 2 shows that the dissolved oxygen (DO) during the anoxic cycle averaged greater than 2 mg/L during acclimation. This was likely due to excessive carryover of oxygen from the previous aeration cycle coupled with the low organic carbon concentration in the feed. Due to the oxygen concentration, denitrification was inhibited. Since there are no permit limits for nitrogen and 1,4-dioxane degradation was not being affected, the anoxic cycle was removed from SBR operation in December 2002.

During acclimation the reactors were tested for removal efficiencies of volatile organic compounds (VOCs) and semi-volatile organic compounds (SVOCs). In order to get a full understanding of the processes occurring in the reactors, samples were collected from the reactor feed, each reactor effluent, and from each reactor's off-gas ventilation system. VOC and SVOC samples were collected several times in January 2003. Due to the presence of high concentrations of 1,4-dioxane and THF, elevated detection limits masked the presence of other VOCs in the feed and effluent waters (data not shown).

The off-gas was sampled using 8-hour Summa canisters on December 6, 2002 and a thermal anemometer for flow measurement. Based upon the gas flow rates exiting the reactors, it was determined that less than 0.5% of the 1,4-dioxane was being removed by volatilization. This is to be expected since 1,4-dioxane have low Henry's Law Coefficients.

Table 2: Average Reactor Performance During Startup And Acclimation

| Parameter	Units	Startup 09/07/02 - 10/22/02		Acclimation 10/23/02 - 02/13/03	
		Reactor 1	Reactor 2	Reactor 1	Reactor 2
Influent					
Flow	(mL/min)	Batch Operation		0.65	
SCOD	(mg/L)	N/A	N/A	187	
TCBOD	(mg/L)	N/A	N/A	20	
SCBOD	(mg/L)	N/A	N/A	10	
TOC	(mg/L)	N/A	N/A	81	
TDS	(mg/L)	N/A	N/A	4779	
NH_3	(mg-N/L)	N/A	N/A	127	
NO_2	(mg-N/L)	N/A	N/A	1.01	
$NO_2 + NO_3$	(mg-N/L)	N/A	N/A	16.5	
TSS	(mg/L)	N/A	N/A	95	
pH	(S.U.)	N/A	N/A	7.83	
Phosphorus	(mg-P/L)	N/A	N/A	9.5	
1,4-dioxane	(ug/L)	19281		9828	
THF	(ug/L)	N/A	N/A	18909	
Anoxic Cycle					
pH	(S.U.)	N/A	N/A	6.3	6.4
DO	(mg/L)	N/A	N/A	2.1	2.1
ORP	(mV)	N/A	N/A	207	222
Temperature	(°C)	N/A	N/A	26.1	24.2
NH_3	(mg-N/L)	N/A	N/A	47.8	47.7
NO_2	(mg-N/L)	N/A	N/A	8.0	13
$NO_2 + NO_3$	(mg-N/L)	N/A	N/A	23.7	24.3
Aerobic Cycle					
MLSS	(mg/L)	2241	2020	723	550
MLVSS	(mg/L)	N/A	N/A	347	229
Media Solids	(g/m^2)	N/A	N/A	3.98	3.45
pH	(S.U.)	7.1	7.3	6.4	6.5
DO	(mg/L)	7.0	5.7	5.1	5.1
Temperature	(°C)	24.4	25.1	25.8	24.2
SVI		123	105	31	16
OUR	(mg/L*min)	0.64	0.33	0.20	0.17
Organic Loading	(g D&T COD/g TS*d)	N/A	N/A	0.080	0.080
Effluent					
SCOD	(mg/L)	369	497	124	143
TCBOD	(mg/L)	222.0	313.0	6.8	9.6
SCBOD	(mg/L)	N/A	N/A	3.8	4.4
TOC	(mg/L)	149.0	165.0	52.0	61.0
NH_3	(mg-N/L)	137	127	52	52
NO_2	(mg-N/L)	18.0	5.0	3.3	7.8
$NO_2 + NO_3$	(mg-N/L)	25.0	6.0	29.0	29.0
TSS	(mg/L)	N/A	N/A	52	85
VSS	(mg/L)	N/A	N/A	21	34
pH	(S.U.)	7.1	7.3	6.2	6.2
Phosphorus	(mg-P/L)	35.0	21.0	6.6	7.0
1,4-dioxane	(ug/L)	5725.0	5342.0	2763.0	1914.0
THF	(ug/L)	N/A	N/A	725.0	1223.0

Notes: N/A – Not Analyzed

Figure 2: 1,4-dioxane Treatment Issues
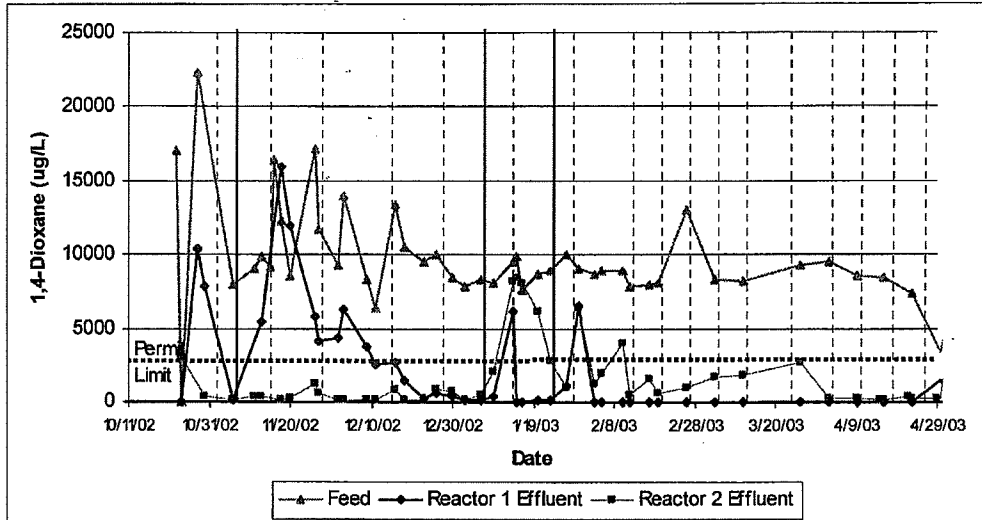
Notes: Solid vertical lines denote sloughing events.
         Dashed vertical lines denote feed changes.

Figure 3: Feed 1,4-dioxane and THF Decline
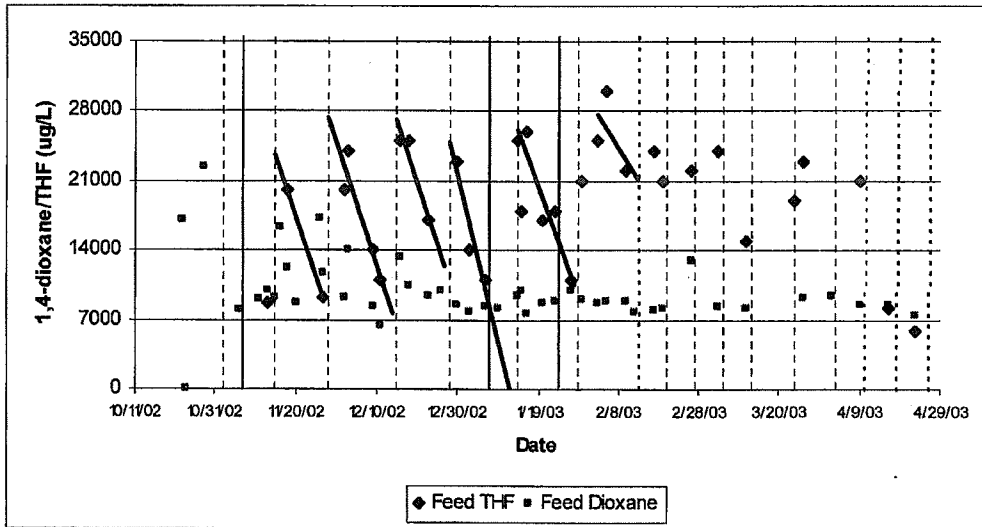
Notes: Solid vertical lines denote sloughing events.
         Dashed vertical lines denote feed changes.
         Solids diagonal lines denote slope of the THF data in that feed cycle.

SBR Operation

After acclimation the reactors were allowed to run for two months in SBR mode to verify that the system could consistently treat the Blend waters. Table 3 shows the average performance characteristics for the two reactors during this period. The extended testing duration verified reactor performance in addition to confirming the previous sloughing events were an artifact of the test approach. As can be seen in Figures 4 and 5, THF and 1,4-dioxane degradation were consistent, with no sloughing events. Reactor 1, which had been operating well before the change in feed procedures, continued to operate at greater than 98-percent removal rates for both 1,4-dioxane and THF. Reactor 2, which had just experienced a sloughing event prior to the change in feed procedure recovered in a little more than a month and returned to greater than 98-percent removal rates.

Figure 6 shows how the mass of attached solids changed after the acclimation period. Both reactors exhibited an increase in attached solids , which eventually leveled off between 2,000 to 2,500 g of solids per reactor. This is equivalent to a mass per media surface area between 6 to 8 g/m², compared to the design value of 7 g/m². However, suspended solids concentrations were consistently less than 400 mg/L, and were continuing to decline as the study progressed. This was primarily because the suspended solids settled very poorly with a large portion remaining dispersed throughout the reactor during the decant cycle. Therefore, a large mass of suspended solids was removed in each decant.

Continous Flow Operation

After two months of operation as SBRs, the reactors were converted to continuous flow bioreactors. This greatly simplified the control and operation of the reactors, as well as allowed the reactors to operate without having the transient concentration spikes associated with batch feeding.

The reactors were converted in a staggered sequence to evaluate any change in reactor performance. Reactor 2 was converted on April 22, 2003 and Reactor 1 was converted on May 14, 2003. Increased 1,4-dioxane effluent concentrations from both reactors were noted after conversion of Reactor 2 that was attributed to low feed concentrations. No change in 1,4-dioxane degradation was attributable to the conversion. However, effluent ammonia concentrations in each reactor declined after changing to continuous operation, and there was a corresponding increase in nitrate during the same period. This is likely because of enhanced nitrification due to the steady application of feed and lower bulk concentrations of contaminants that are typical with continuous feed reactors.

At the same time the reactors were converted from SBR to continuous flow mode, pH control was added to each reactor to evaluate its effect on reactor performance. Pumps dosing 25-percent sodium hydroxide were added to each reactor. Data collected in August 2003, indicated that the reactors were utilizing 1,800 mL/day of the caustic. These data were taken while both reactors were still nitrifying, requiring the use of additional alkalinity.

TABLE 3: AVERAGE REACTOR PERFORMANCE DURING SBR AND INITIAL CONTINUOUS OPERATION

Parameter	Units	SBR		Continuous	
		02/14/03 - 05/14/03	02/14/03 - 04/22/03	05/14/03 - 07/11/03	06/01/03 - 07/11/03
		Reactor 1	Reactor 2	Reactor 1	Reactor 2
Influent					
Flow	(mL/min)	0.68		0.41	
SCOD	(mg/L)	201		N/A	
TCBOD	(mg/L)	15		N/A	
SCBOD	(mg/L)	10		N/A	
TOC	(mg/L)	76		71	
TDS	(mg/L)	4450		4650	
NH_3	(mg-N/L)	115		108	
NO_2	(mg-N/L)	0.22		1.6	
$NO_2 + NO_3$	(mg-N/L)	11.3		8.8	
TSS	(mg/L)	22		16	
pH	(S.U.)	7.6		7.9	
Phosphorus	(mg-P/L)	5.0		N/A	
1,4-dioxane	(ug/L)	9056		8341	
THF	(ug/L)	20410		17760	
Anoxic Cycle					
pH	(S.U.)	N/A	N/A	N/A	N/A
DO	(mg/L)	N/A	N/A	N/A	N/A
ORP	(mV)	N/A	N/A	N/A	N/A
Temperature	(°C)	N/A	N/A	N/A	N/A
NH_3	(mg-N/L)	N/A	N/A	N/A	N/A
NO_2	(mg-N/L)	N/A	N/A	N/A	N/A
$NO_2 + NO_3$	(mg-N/L)	N/A	N/A	N/A	N/A
Aerobic Cycle					
MLSS	(mg/L)	363	102	107	48
MLVSS	(mg/L)	199	36	61	40
Media Solids	(g/m²)	5.3	5.9	7.1	8.2
pH	(S.U.)	6.3	6.4	7.0	7.1
DO	(mg/L)	9.8	12.5	3.6	5.6
Temperature	(°C)	25.5	24.2	25.7	14.8
SVI		20	13	N/A	N/A
OUR	(mg/L*min)	0.17	0.17	0.21	0.22
Organic Loading	(g D&T COD/g TS*d)	0.064	0.069	0.061	0.050
Effluent					
SCOD	(mg/L)	127	146	N/A	N/A
TCBOD	(mg/L)	5.5	8.2	N/A	N/A
SCBOD	(mg/L)	2.8	3.9	N/A	N/A
TOC	(mg/L)	50	58	52	51
NH_3	(mg-N/L)	51	53	32	108
NO_2	(mg-N/L)	0.37	0.27	0.38	0.02
$NO_2 + NO_3$	(mg-N/L)	28	28	67	33
TSS	(mg/L)	30	18	198	234
VSS	(mg/L)	15	9.5	138	165
pH	(S.U.)	6.2	6.0	6.98	7.05
Phosphorus	(mg-P/L)	4.6	4.5	N/A	N/A
1,4-dioxane	(ug/L)	150	1003	57	83
THF	(ug/L)	5	118	46	212

Figure 4: 1,4-dioxane Profiles, Post Acclimation
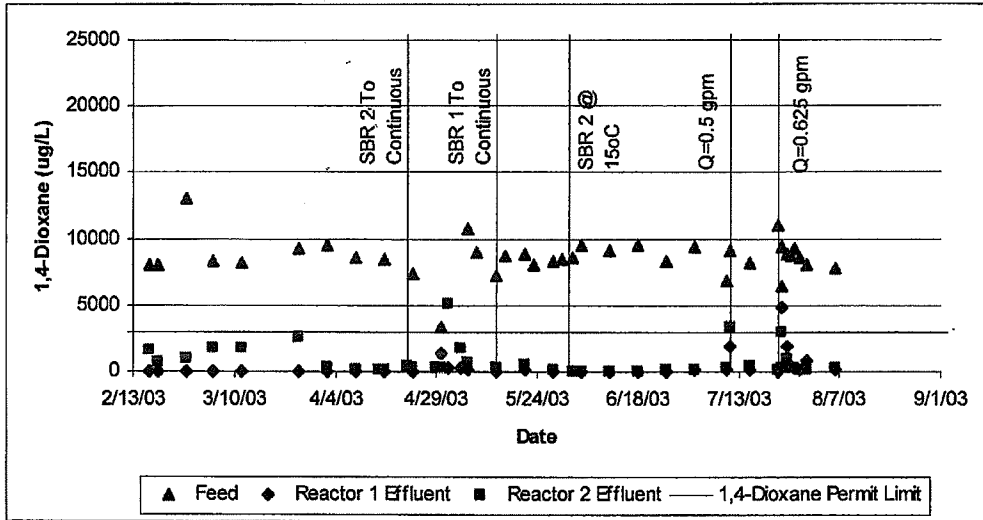
Figure 5: THF Profiles, Post Acclimation
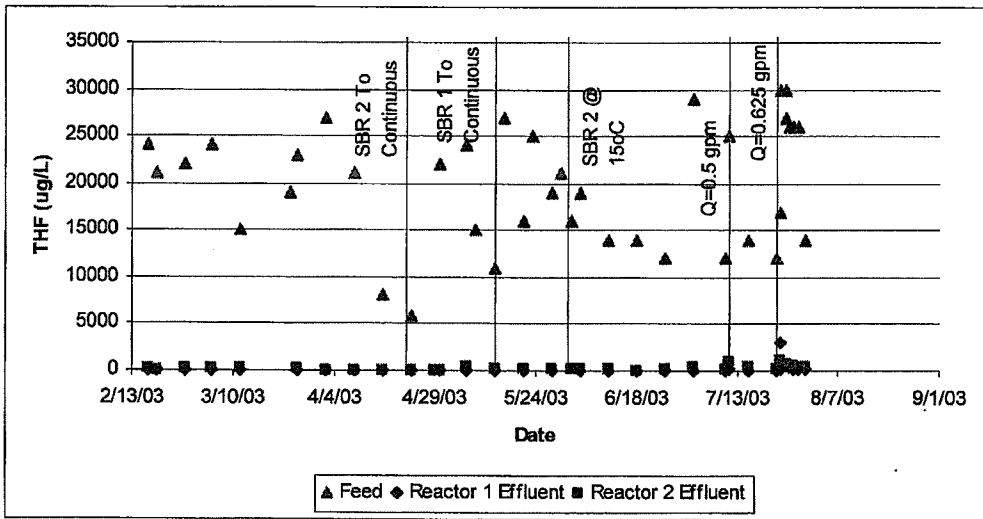
Figure 6: Mass of Attached Solids, Post Acclimation
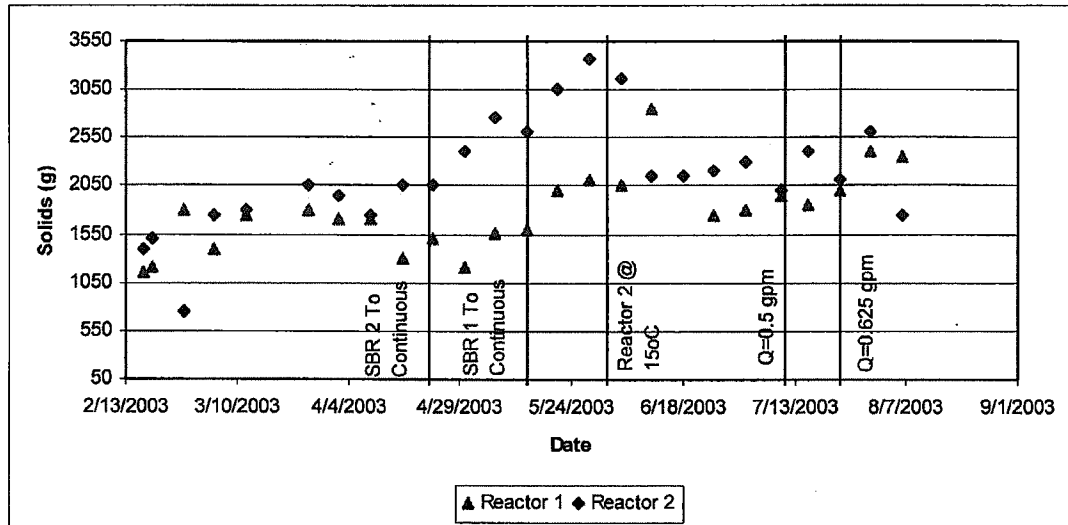
Temperature Variation: With both reactors operating at steady state conditions, the operating temperature of Reactor 2 was reduced from 25 °C to 15 °C. The temperature was reduced by 2 °C per day, with the reactor reaching 15 °C on June 1, 2003. Reactor 2 operated for five weeks at the new condition to show stability and to verify that no adverse affects would be seen from changing the temperature. Reactor 1 acted as a control during this period. At the feed conditions tested, there was no discernable effect on 1,4-dioxane or THF removal by lowering the temperature of Reactor 2.

Lowering the temperature did reduce nitrification efficiencies, resulting in higher effluent ammonia and lower effluent nitrate concentrations (Figures 7, 8, and 9). The increase in effluent ammonia due to decreased nitrification was not considered an adverse affect since 1) there is no nitrogen limit on the facility discharge permit and 2) the increased ammonia concentration appeared to have no inhibitory affects on 1,4-dioxane and THF degradation.

Volumetric Loading Variation: Subsequent to the temperature change discussed above, the organic loading to the reactors were increased to evaluate its effect on reactor performance. Since the composition of the feed was fixed at a constant blend ratio, the organic loading could only be increased by increasing the volumetric flow rate to the system; thereby applying more mass of 1,4-dioxane and THF over a shorter time period. Step changes were made to the flow rate to evaluate recovery time in each reactor.
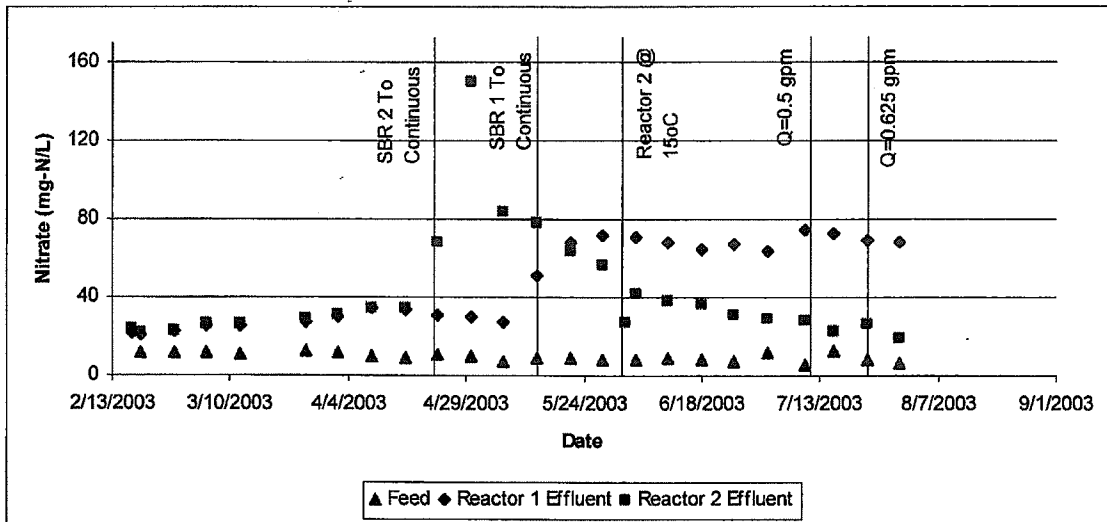
Figure 7: Ammonia Profile, Post Acclimation
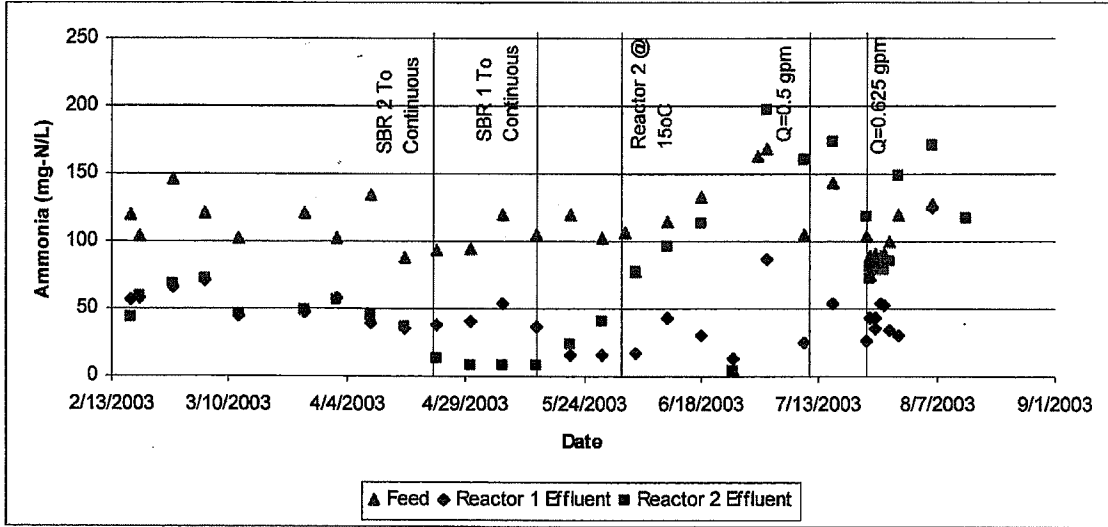
Figure 8: Nitrite Profile, Post Acclimation
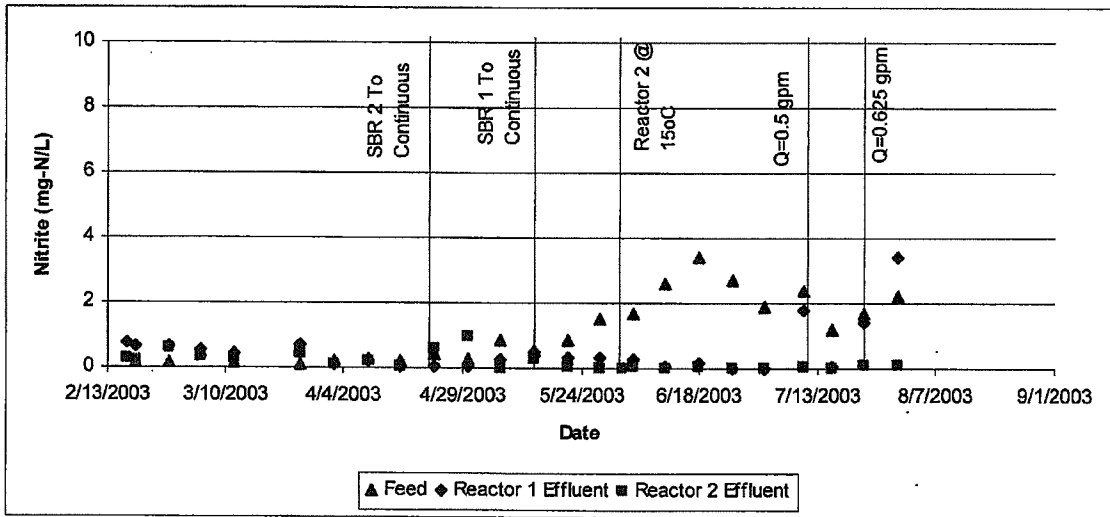
**Figure 9: Nitrate + Nitrite Profile, Post Acclimation**

A 25% increase in organic loading was made on July 11, 2003 (0.4 to 0.5 gpm) Each reactor showed the expected spike in effluent 1,4-dioxane concentrations, with 1,4-dioxane degradation efficiencies dropping from greater than 95% in both reactors to 79% in Reactor 1 and 64% in Reactor 2. A second 25% increase occurred on July 23, 2003 (0.5 to 0.63 gpm), and showed a similar trend in effluent 1,4-dioxane concentrations with increased performance in Reactor 2 (efficiencies only dropping to 89%). Both reactors showed complete recovery to nominal treatment efficiencies in less than five days after both adjustments. THF removal efficiencies declined slightly (2 to 3%) during these same periods.

With increased organic loadings, higher bio-solids loadings were expected on the media. Figure 6 shows that the solids in each reactor did appear to be increasing during the period of increased loading, but there were insufficient samples to make a clear correlation. Figure 6 also shows that the bio-solids in Reactor 2 were consistently higher than Reactor 1 during this period as well, which may explain why Reactor 2 was better able to cope with the second step increase in loading.

Oxygen Uptake Rates

Oxygen uptake rate (OUR) is a measure of bacterial metabolism. As the OUR increases, the faster the biomass is growing and using the food provided. Figure 10 shows that there was little change in the measured OURs for each reactor since acclimation. Note that the OURs in Figure 10 are adjusted with respect to the mass of biomass present in each reactor.

Figure 10: Oxygen Uptake Rate, post Acclimation
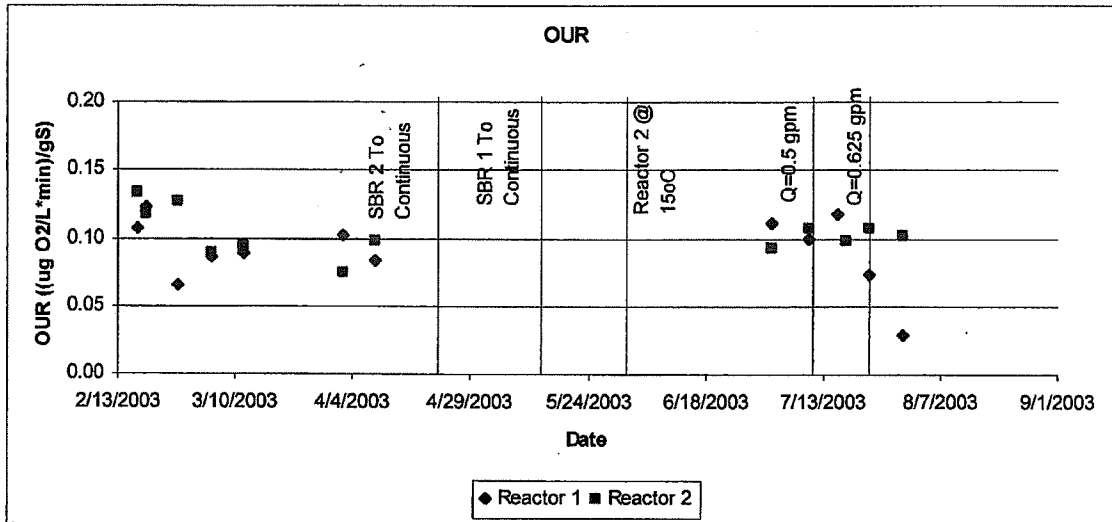
DISCUSSION

SBR Operation

At the start of the steady state period of SBR operation, the reactors had been operating for five months, with varying degrees of success. There was a good base of biomass on the media, and it was proving adequate for 1,4-dioxane degradation at the operating conditions. Once provided with a stable feed of THF, the reactors began to operate as expected from the bench scale tests. It was not surprising for the reactors to operate poorly when under-loaded, as the bench scale tests showed that 1,4-dioxane degradation decreased when the organic loading was too low (data not published). This further supports the theory that the degradation of 1,4-dioxane is linked to THF as a form of cometabolism. The THF provides the primary source of energy for the bacteria and 1,4-dioxane is degraded by the THF enzymes. Therefore, if there is too little THF, then fewer enzymes are produced and little 1,4-dioxane will be degraded.

Figures 11 and 12 show the organic loadings of the two reactors after the acclimation period. The organic loading in the two reactors was inconsistent during SBR operation, primarily because of a combination of variability of the feed concentrations and the increasing biomass on the media. As the media solids increased, the organic loading decreased, such that the average organic loadings during the SBR operation were 0.064 and 0.069 grams of dioxane and THF as COD/grams of Total Solids * day (g D&T COD/g TS*d), for Reactors 1 and 2 respectively.

Conversion to Continuous Flow

The SBRs were converted to continuous flow reactors because:

1. Nitrification was not achieved;

2. Continuous reactors are less complex then SBRs;

3. Continuous flow reactors typically result in a more consistent, low bulk concentration compared to SBRs, possibly reducing inhibition; and

4. Continuous flow reactors eliminate the concentration transients that are inherent in batch feeding.

When changing the reactors from SBR to continuous flow mode, an effort was made to maintain the organic loading level to minimize the number of changed variables so that any changes in operation could be attributed to the change is flow regime rather than changes in organic loading.

However, the media solids continued to increase in each reactor, causing the organic loading to drop during this period. The average organic loadings during the initial continuous flow operation were 0.061 and 0.050 g D&T COD/g TS*d, for Reactors 1 and 2 respectively.

Throughout continuous flow operation the removal of 1,4-dioxane and THF consistently exceeded 95-percent. When ncreases in volumetric loading were implemented, corresponding increases in organic loading were expected. However, because of a combination of decreased THF feed concentration over time and increased media solids, the organic loading only increased slightly during the adjustments.
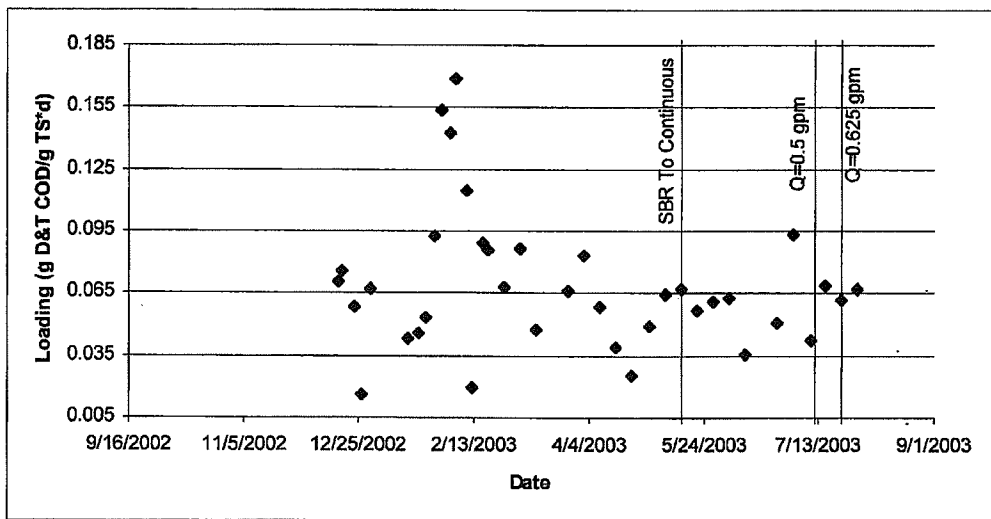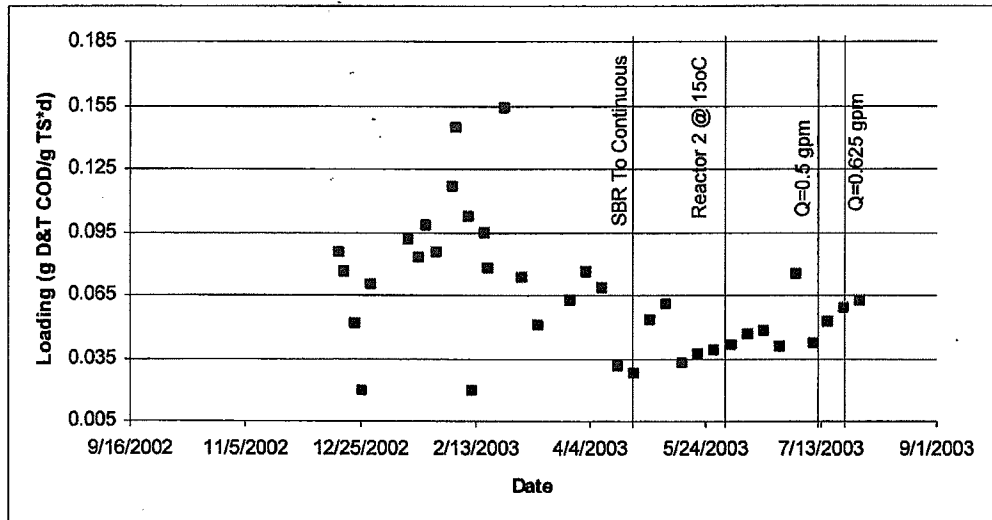
Figure 11: Reactor 1 Organic Loading

Figure 12: Reactor 2 Organic Loading

1,4-Dioxane Degradation By Cometabolism

Research at North Carolina State University (NC State) (Zenker et al., 2000) concluded that THF acts both as a catalyst and competitive inhibitor for 1,4-dioxane degradation. The NC State research showed that their 1,4-dioxane degrading biomass used THF for cell growth and energy, but exhibited no removal of 1,4-dioxane when THF was not present. However, the NC State researchers also discovered that when THF was present at high concentrations it was degraded preferentially, with little 1,4-dioxane degradation until the THF concentration fell to low levels. This is in direct opposition to the biomass grown by Sock (1993); which, while initially fed with THF and 1,4-dioxane, was able to degrade 1,4-dioxane as a sole substrate when the THF was discontinued.

The spikes of effluent 1,4-dioxane that accompanied the two step increases in volumetric loading (See Figures 4 and 5), coupled with the much lower increases in effluent THF concentrations indicates that the biomass in the pilot reactors more closely resembles that researched by Zenker, et.al. When the loading was increased, an additional mass loading of THF and 1,4-dioxane would have been applied to the system, causing an increase in the bulk concentration of both chemicals. The data suggests that the biomass absorbed the increase in THF very quickly, but not the 1,4-dioxane. However, when the biomass had adapted to the increased mass of THF, the 1,4-dioxane was quickly removed.

STUDY OBJECTIVE RESULTS

As previously stated in this report, the pilot study had four primary objectives: ·

1. Evaluate the sustainability and robustness of 1,4-dioxane degradation;

2. Assess the effect of temperature on the degradation of 1,4-dioxane;

3. Assess the effect of organic loading on the stability of 1,4-dioxane degradation; and

4. Provide criteria for sizing and design of a full-scale biological system.

Each of these objectives was met during the Pilot Study and are discussed below.

Sustainability and Robustness

After start-up and acclimation, the reactors were run continuously for seven months. During that time, the only excursions above the effluent 1,4-dioxane limit could be attributed directly to changes in the feed composition or flow. This long period of consistent treatment shows that the biomass is stable. The quickness of response by each reactor to the step increases in volumetric loading, less than five days for each change, shows the robustness of the biological system.

The implementation of the Pilot Study required that the feed be made in batches, with the variability that came with the batch-making procedure. Additional variability was introduced by the unintended, and uncontrollable, biodegradation occurring in the Pilot Equalization Tank. These two variables caused the reactors to experience a higher feed variability than would be anticipated in a full scale system because the full scale system will receive a continuous feed directly from the groundwater extraction trench without the need for an EQ tank. Site experience has indicated that groundwater quality changes tend to be more subtle, as compared to what was observed in the storage tank. The gradual changes in process water feed concentrations should be absorbed by a full scale system without comprising effluent quality.

The one exception to this would be if the THF concentrations in the NTES groundwater were to decrease below that which can sustain the biomass or 1,4-dioxane cometabolism. This is a possibility as more NTES water is pumped and the contaminants are attenuated. Different approachs can be used in a full scale system to diminish the impact of lowered THF concentrations. One such approach would be to add supplemental THF through a chemical addition system. A more efficient way to deal with this possible change in water quality is to design the full scale system with the ability to vary the reaction volume and the percentage of media fill. By lowering the fill percentage of reaction volume, you essentially increase the organic loading per mass of biomass, such that even at lower concentrations of feed THF the biomass is not starved.

Temperature Effects

After changing to continuous feed operation, Reactor 2 was decreased to an operating temperature of 15 °C. This is the typical temperature of the process waters in the WTP. There was little to no affect on 1,4-dioxane degradation with the decrease in temperature. However, Reactor 2 quickly lost its ability to nitrify, allowing ammonia to pass through the system untreated. Since there is no nitrogen limits on the existing WTP, and the ammonia concentration is below that which will inhibit biological growth, the loss of nitrification is acceptable.

Organic Loading

Organic loadings during the Pilot Study ranged from 0.05 to 0.10 g D&T COD/g TS*d. The changes in organic loading were followed by quick adaptation by the reactors. At no time did the organic loading exceed the capacity of the reactors. Based on the bench results, lower (0.04) and higher loadings (0.15) are possible.

CONCLUSIONS

The following conclusions were made based on the results of the Pilot Study:

1. 1,4-dioxane can be degraded to greater than 95-percent efficiency consistently in the process waters at organic loadings of 0.04 to 0.075 g D&T COD/g TS*d.

2. Biodegradation of 1,4-dioxane at 15 °C is possible with little impact on removal efficiencies.

3. 1,4-dioxane is not removed by volatilization.

4. Reactor operation at 15 °C inhibits nitrification.

5. Alkalinity is required to fully nitrify at 25 °C.

6. Too little feed THF can result in reduced 1,4-dioxane degradation efficiency.

7. Instantaneous increases in loadings of 25-percent will result in 2-3 days decreased 1,4-dioxane degradation efficiency, with possible effluent concentrations above the permit limit.

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Credits

Authors

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