

ANAEROBIC-AEROBIC COMBINED PROCESS FOR THE TREATMENT OF SEWAGE WITH NUTRIENT REMOVAL: THE ANANOX[®] PROCESS

G. Garuti*, M. Dohanyos** and A. Tilche*

**ENEA, Dipartimento di Ingegneria Ambientale, CORLAB DIM2, Via Martiri di Monte Sole, 4 - 40129 Bologna, Italy*

***Department of Water Technology and Environmental Engineering, Prague Institute of Chemical Technology, Technická 5, 166 28 Prague 6, Czechoslovakia*

ABSTRACT

Results of a three year experience on a combined anaerobic-anoxic-oxic municipal wastewater treatment process - named ANANOX[®] - are presented.

This process demonstrated to be highly efficient, with 89.6% COD, 89.2% TSS and 81.2% N removal, and a sludge production of only 0.2 kg TSS·kg COD removed⁻¹, a value which is roughly 50% less if compared with traditional nitrification/denitrification processes.

Sulphates play a very significant role in the process because, after being reduced in the anaerobic step, where they give a contribution to the organic matter degradation, they are reoxidized in the anoxic step by nitrates, reducing the organic matter need for denitrification.

Due to the high dependence of efficiency on temperature, the system proposed has advantageous uses for sewage treatment, particularly in warm climates and in tourist and recreational areas where the population increases during the warm season.

KEYWORDS

Anaerobic treatment of sewage; nitrification; denitrification; Anaerobic Baffled Reactor.

INTRODUCTION

Anaerobic treatment of sewage is not a new process. As McCarty (1981 and 1985) has noted, first attempts dated from 1881, and after that date a continuous development led to the Imhoff tank in 1905. After the development of aerobic treatments, anaerobic processes were used in large municipal treatment plants only for sludge treatment, and the anaerobic septic tanks remained just for the treatment of individual households.

The reproposal of anaerobic processes for domestic wastewater treatment is quite recent (Switzenbaum and Jewell, 1980; Lettinga *et al.*, 1981).

The Seminar/Workshop held in Amherst (Switzenbaum ed., 1985) first gave the possibility to compare the results obtained by several research groups around the world. Among the several different processes presented, the

Anaerobic Attached Film Expanded Bed (AAFEB) showed to be the most efficient (Jewell, 1985). The author claimed to obtain very good effluent quality ($50 \text{ mg COD}\cdot\text{l}^{-1}$) at 2 h HRT (Hydraulic Residence Time), due to the very long SRT (Sludge Retention Time) values, around 300 d. Very good results were also presented by Yoda *et al.* (1985), using a fluidized bed. A little lower efficiency was shown with the use of UASB (Upflow Anaerobic Sludge Blanket) reactors (Grin *et al.*, 1985; Nucci *et al.*, 1985; Schellinkhout *et al.*, 1985) and Anaerobic Filters (Genung *et al.*, 1985).

The UASB design was anyway considered to be the most promising technology, due to its simplicity that makes construction, operation and maintenance easier.

In the more recent years, new developments based on the UASB process were presented (de Man *et al.*, 1986; de Man *et al.*, 1988a, b) also at larger scale in the 120 m^3 plant of Sao Paulo (Brazil) (Vieira, 1988), in the 240 m^3 plant of Senigallia (Italy) (Collivignarelli *et al.*, 1990), and in the 64 m^3 plant of CALI (Columbia) (HASKONING, 1989).

A summary table of some published results is reported (Table 1).

TABLE 1 Summary of Some of the Published Results on the Anaerobic Treatment of Sewage: Rounded Data, in Some Case Derived from Graphs

Reference	React.	Vol (m ³)	HRT (h)	Temp. °C	COD _t In mg/l	COD _t Out mg/l	COD _s Out mg/l	COD _t remo-val %	TSS In mg/l	TSS re-moval %
Lettinga <i>et al.</i> (1981)	UASB	6	12	10–20	300–1000	~200	100–200	55–75	—	—
Lettinga <i>et al.</i> (1983)	UASB	0.12	12	18–21	700–1200	—	—	70–78*	—	—
Jewell (1985)	AAFEB	0.001	.5–24	20	~200	~50	—	55–90	—	—
Yoda <i>et al.</i> (1985)	FB	0.009	3.9	15–30	180–400	50–150	40–100	50–75	50–150	70–90
Genung <i>et al.</i> (1985)	AF	190	~24	15–25	—	—	—	50–70	—	78–93
Grin <i>et al.</i> (1985)	UASB	6	8	15–19	100–900	—	100–200	60–75	—	—
Schellinkhout <i>et al.</i> (1985)	UASB	64	3–3.5	~25	200–300	—	—	~75	~200	~60
Nucci <i>et al.</i> (1985)	UASB	3.7	17–23	~30	~550	—	—	30–55	~200	~50
de Man <i>et al.</i> (1986)	UASB	6	12–16	10–18	100–700	—	—	45–60	—	—
Derycke and Verstraete (1986)	PCR	0.12	1.6–6	13–20	~450	wide range	110–250	54–70*	~330	—
de Man <i>et al.</i> (1988a)	UASB	0.12	7–8	12–20	190–1180	—	—	30–75	—	—
Vieira (1988)	UASB	120	4.7–9	21–25	~290	~115	—	~60	~140	~70
Van Rompu and Verstraete (1988)	PU-PPS	0.16	1.2	20	—	—	—	57–67	—	45–70
de Man <i>et al.</i> (1988b)	UASB	20	13–14	10–15	750–1280	—	—	16–48	—	64–78
Orozco (1988)	ABR	7	9–12	13–16	340	—	—	75	—	—
Sanz <i>et al.</i> (1988)	FB	0.54	1.5	10	760	175	—	70	—	89
Collivignarelli <i>et al.</i> (1990)	UASB	240	12–42	14–23	200–325	100–200	50–110	30–55	—	—
HASKONING (1989)	UASB	64	6–8	~25	250–350	120	70	~60	~150	~70
Vincenzi (1989)	ABR	0.35	~15	18–28	750	170	130	~75	—	—

* Calculated as $\text{COD}_t \text{ In} - \text{COD}_s \text{ Out}$

Most of the authors have used some kind of sludge bed reactor. Usually these authors observe that the limitation, particularly for TSS (Total Suspended Solids) removal, is due to the upflow velocity and therefore to the hydraulic retention time and not to the loading rate.

Lettinga and Hulshoff Pol (1991) report that the daily average upflow velocity for raw domestic sewage ranges from $0.33 \text{ m}\cdot\text{h}^{-1}$ at 16°C to $0.67 \text{ m}\cdot\text{h}^{-1}$ for temperatures over 26°C , with acceptable peaks over 2–6 h of 3 times higher values.

Due to the high flow rates and to the risk of channelling for the small mixing effect of the gas production, feed distribution on the reactor bottom is considered a crucial design parameter; one feed inlet point every one or two m^2 is recommended.

Some authors (Derycke *et al.*, 1986; Vincenzi, 1989) have noticed that by increasing influent COD concentrations, the COD removal efficiency increases. This means that separate sewer systems are preferable for the application of anaerobic treatments, also because of the higher average temperature of the sewage, which is another important limitation to the treatment. Many authors in fact show a severe drop in efficiency of the anaerobic treatment, particularly for temperatures decreasing below 13°C . Therefore anaerobic treatment of sewage shows the promise to develop more in warm climates, and particularly in non-developed countries where the recovery of methane can be better valorized.

The main advantages of anaerobic treatments are recognized to be the very low sludge production - when compared to aerobic - and the low energy needs.

All the authors then agree in considering anaerobic processes as pre-treatment of sewage, that must be followed by other treatments in order to reach acceptable water quality levels for the discharge into surface waters.

The effluent of anaerobic treatment contains a consistent amount of ammonia nitrogen and sulphides that must be removed. Also total COD, as can be seen from Table 1, is very often higher than 100 mg·l⁻¹; this depends on the complexity of anaerobic processes and on the necessity to maintain in the bulk liquid a minimum concentration of substrate for each physiologic group of microorganisms which plays a role in the process (McCarty, 1985; Rittmann and Baskin, 1985).

The literature on downstream processes is anyway quite poor. Some authors demonstrated the feasibility of aerobic post-treatment (Yoda et al., 1985), of nitrification (Bovendeur and Klapwijk, 1986) and of chlorination/ozonization (Vieira, 1988).

Among the known literature, only Collivignarelli et al. (1990) presented an anaerobic/aerobic process designed for optimum removal of COD, TSS and nitrogen. In the pilot plant described, the raw sewage, after microscreening, goes to anaerobic treatment in a 240 m³ UASB, then to an anoxic fluidized bed for the denitrification of the recycled effluent from the following nitrification step carried out in a fixed film aerobic reactor.

TABLE 2 Average 1989 Results (E1 after anaerobic; E1/2 after anoxic step)

At ENEA, research activity on the anaerobic treatment of sewage started in 1988.

The interest of ENEA in the anaerobic treatment of sewage came from the analysis of the potential application of this technology in Italy, if completed with adequate post-treatment systems. In many areas in fact, particularly of Central-Southern Italy, Sicily and Sardinia, there are appropriate climatic conditions. On the coasts, many tourist villages must then face problems of a very large increase of population during the warm summer, when anaerobic treatment works better.

Parameters	Influent	E1	E1/2	Effluent
COD total mg·l ⁻¹	762	n.d.	n.d.	64.2
COD soluble mg·l ⁻¹	172	187	75	44.5
TSS mg·l ⁻¹	488	n.d.	1770	73
N-NH ₄ ⁺ mg·l ⁻¹	26.1	24.3	16.5	0
N-NO ₂ ⁻ mg·l ⁻¹	n.d.	0	0	0
N-NO ₃ ⁻ mg·l ⁻¹	n.d.	0	0	12.4
P-PO ₄ ³⁻ mg·l ⁻¹	11.1	17	14	7.8
SO ₄ ²⁻ mg·l ⁻¹	190	85	128	178
TOC mg·l ⁻¹	42.7	53.2	19.1	11.1
TC mg·l ⁻¹	167.5	202.5	145.2	105.5

During 8 months, two different reactors, a 3 chamber Anaerobic Baffled Reactor and a Hybrid Reactor (Sludge Bed + Filter in the top third of the reactor), were studied for COD and TSS removal (Vincenzi, 1989).

The ABR showed better performances, with average removals of 77% COD and 89.5% TSS at quite long HRTs - around 15 h - and temperature ranging from 18 to 28°C.

The ABR (McCarty, 1981; Bachmann *et al.*, 1985) works like a series of UASBs, and shows several advantages in comparison to a one-step reactor, like better capacity of suspended solid retention, lower bed bypass, staged reaction, higher tolerance to shock loadings.

The main results obtained in 1988 were: 1) very low sludge production, ranging from 0.3 to 0.5 l sludge·p.e.⁻¹·d⁻¹; 2) very long SRT in the ABR, around 180d; 3) considerable gas production, around 8.5–10 l CH₄·p.e.⁻¹·d⁻¹; 4) more or less constant effluent COD values also varying with the influent COD concentrations; 5) very high ammonia nitrogen values in

TABLE 3 Average Wastewater Characteristics in 1990

Parameters	Values (mg·l ⁻¹)
COD tot	585.2 ± 321.2
COD sol	170.0 ± 78.1
TOC	137.6 ± 26.5
IC	56.6 ± 21.3
TSS	321 ± 236
VSS	238 ± 172
TKN	88.8 ± 25.1
N-NH ₄ ⁺	52.7 ± 19
N-NO ₃ ⁻	1.25 ± 1.63
P-PO ₄ ³⁻	6.34 ± 2.22
Ptot	10.32 ± 3.45
SO ₄ ⁼	110.7 ± 60.7

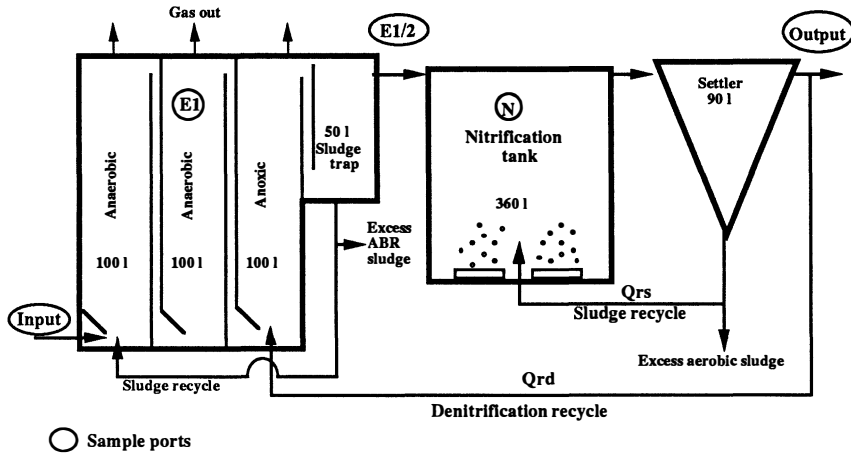


Fig. 1. Schematic of the experimental plant

the effluent, around $50 \text{ mg N-NH}_4^+ \cdot \text{l}^{-1}$.

The plant was modified in 1989 in collaboration with the Dept. of Water Technology and Environmental Engineering of the Prague Institute of Chemical Technology, with the addition of a suspended growth nitrification step and the use of the third chamber of the ABR as an anoxic reactor for the denitrification of the recycled effluent from the final settler.

During the summer campaign, with an average wastewater temperature of 24°C , the HRT was reduced to 5.5 h in the anaerobic phase (9.5 h in the whole ABR).

The most significant data obtained are reported in Table 2.

The very promising results were the basis of the 1990 experience that is described in the present article.

Recently, ENEA has applied for patenting the process, and the name ANANOX® (ANAerobic-ANoxic-OXic) has been deposited.

MATERIALS AND METHODS

The experimental plant (Fig.1), slightly modified from the 1989 campaign, was installed at the municipal wastewater treatment plant of San Giovanni in Persiceto (Bologna) in the north of Italy.

The influent was directly derived from the sewer after microscreening on a 0.75 mm self-cleaning rotating screen (Hydrascreen made by IDRACOS, Italy). The composition of the influent (Table 3) was slightly different from common strong raw waste because of the contribution of a dyeing factory which discharged into the municipal sewer. In this kind of sewer the sewage was separated from the rainfall water.

An Anaerobic Baffled Reactor was used for the anaerobic and the anoxic phase, and a conventional aerated mixing tank for the oxic step.

The ABR reactor consisted of 3 up-flow chambers ($0.21 \times 1.6 \times 0.3 \text{ m}$ LxHxW each one) followed by an internal settler which provided a total liquid volume of 350 l. The third chamber was filled with polyurethane packing material (Filtren TR 30 made by RECTICEL, Belgium) in order to enhance the retention of the biological solids under conditions of high flow rate.

The oxidation tank was a cylinder of 360 l volume with 0.67 m water level; the air was diffused from the bottom by two fine bubble spargers.

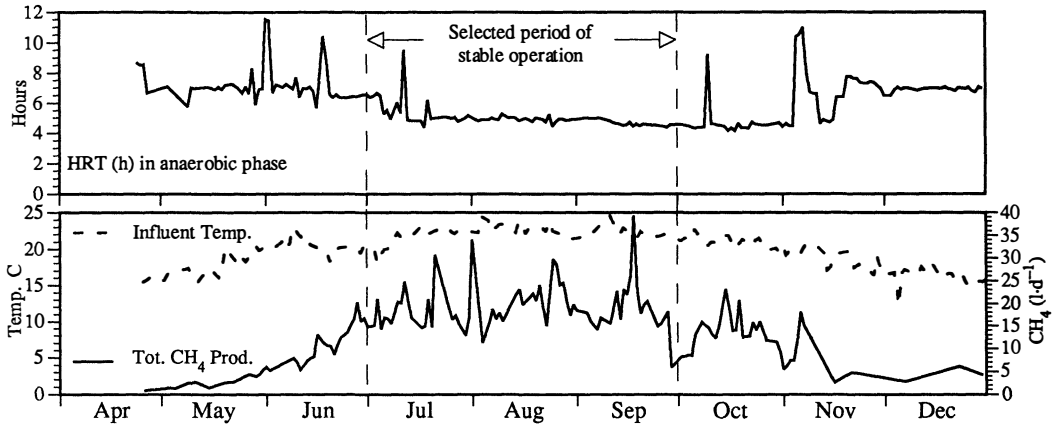


Fig. 2. 1990 data of HRT, influent temperature and methane evolved in biogas

TABLE 4 Average Loading Rates on the Basis of Influent Values

	COD tot kg·m ⁻³ ·d ⁻¹	COD sol kg·m ⁻³ ·d ⁻¹	TSS kg·m ⁻³ ·d ⁻¹	HRT h
①) Anaerobic	2.17	0.8	1.24	4.76
②) ① + Anoxic	1.24	0.46	0.71	8.33
③) ① + ② + Oxidic	0.61	0.22	0.35	16.9
Total plant	0.54	0.2	0.31	19.04

TABLE 5 Average Data from the Selected Period (E1: after anaerobic step; E1/2: after anoxic step)

Parameters	Influent	E1	E1/2	Effluent
Temp. °C	22.1	22.5	22.4	22.1
pH	8.06	7.72	7.82	7.99
COD _{tot} mg·l ⁻¹	427.5	258.1	196.2	44.6
COD _{sol} mg·l ⁻¹	156.5	91.2	45.3	21.3
TOC mg·l ⁻¹	52.9	28.6	19.8	18.0
IC mg·l ⁻¹	134.9	149.3	125.6	94.5
TSS mg·l ⁻¹	250	169	196	27
VSS mg·l ⁻¹	190	117	132	14
N-NH ₄ ⁺ mg·l ⁻¹	47.15	33.90	16.91	1.31
N-NO ₃ ⁻ mg·l ⁻¹	1.56	1.07	2.13	10.39
N-NO ₂ ⁻ mg·l ⁻¹	—	—	—	0.77
SO ₄ ⁼ mg·l ⁻¹	113	52.5	106.5	138.8
PO ₄ ³⁻ mg·l ⁻¹ *	9.77	6.23	5.32	5.85

* Input and output as total P; E1 and E1/2 as P-PO₄³⁻

After gravity sedimentation, in a 90 l sedimentation tank, the clarified effluent was partially recycled to the anoxic zone for denitrification, while the settled sludge was returned to the aerobic tank.

The plant was restarted in April 1990 utilizing the adapted anaerobic seed sludge developed in the previous trial (Vincenzi, 1989) and activated sludge of the full-scale plant in the oxidation tank. After three weeks of technical set-up, sampling and regular loading began.

The loading rates were kept at relatively high levels in order to approach the organic loading rates of the conventional sewage treatment plants (Table 4).

The samples were collected two times a week from five points (Fig. 1). Periodically, sampling of influent was made by a timer-controlled peristaltic pump which pumped 250 ml·h⁻¹ into a container refrigerated at 4°C.

Gas production from each separated chamber was measured with water displacement gas meters (QGM 100 made by APES, Italy); wastewater flow rate was metered using a custom instrument made in our laboratory.

COD and TKN (Total Kjeldahl Nitrogen) analyses have been performed according to Standard Methods (APHA, 1989). Total and Volatile Suspended Solids have been determined by filtration over 1.2 µm with Whatman filters GF/C. Sludge Volume Index was measured according to "Standard Methods".

Nitrogen compounds, phosphates and sulphates were analyzed with a High Performance Ion Chromatograph (DIONEX 4000i).

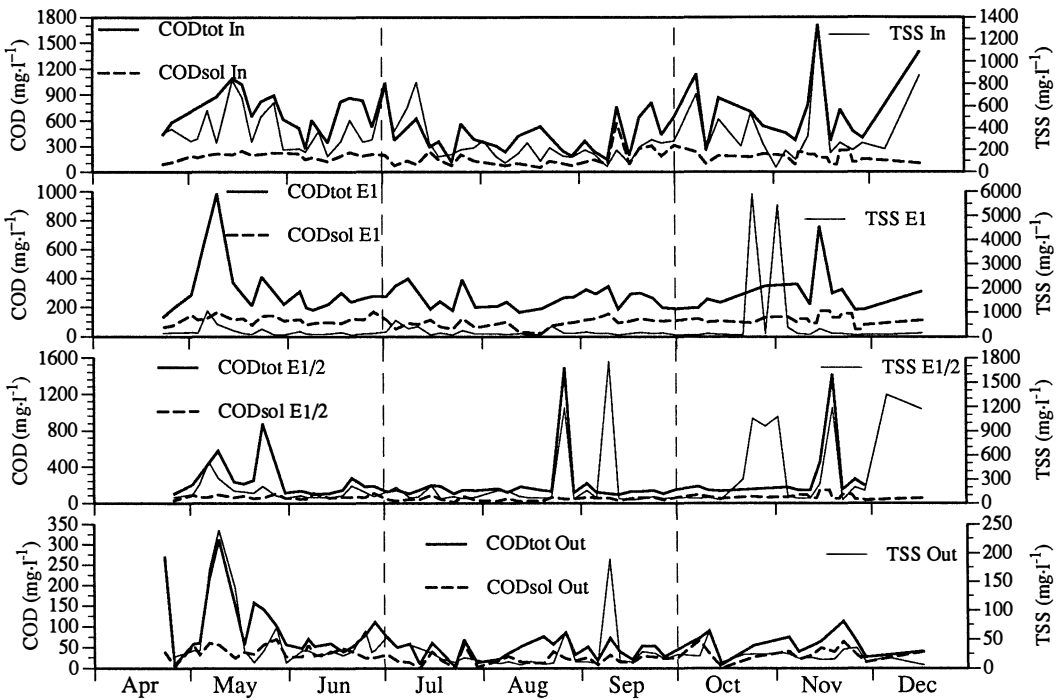


Fig. 3. COD and TSS results of the 1990 campaign (E1: after anaerobic step; E1/2: after anoxic step)

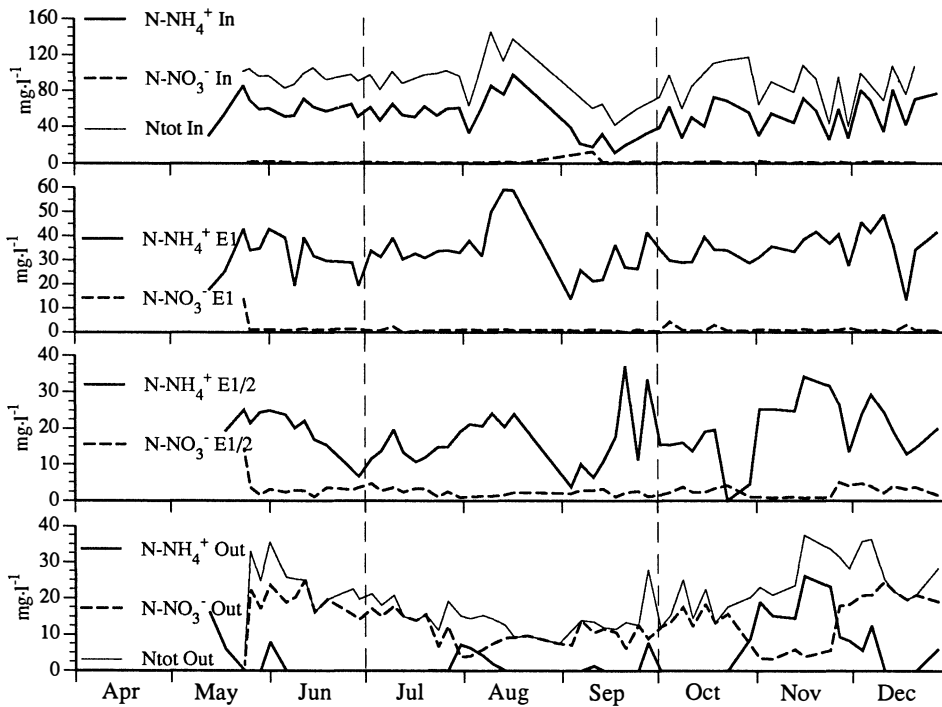


Fig. 4. Nitrogen results of the 1990 campaign (E1: after anaerobic step; E1/2: after anoxic step). Ntot values are partially calculated from correlations.

TABLE 6 Removal Efficiencies of the Various Steps in the Selected Period: negative values represent increments

Parameters	Influent	Removed				Total	Removal eff. (%)
		Anaerobic	Anoxic	Oxic			
COD _{tot} g·h ⁻¹	17.952	7.112	-5.22	14.196	16.088	89.6	
COD _{sol} g·h ⁻¹	6.573	2.743	0.696	2.244	5.683	86.4	
TOC g·h ⁻¹	2.223	1.023	0.276	0.172	1.471	66.2	
IC g·h ⁻¹	5.666	-0.606	-0.587	2.889	1.695	29.9	
N-NH ₄ ⁺ g·h ⁻¹	1.98	0.556	-0.09	1.459	1.925	97.2	
N-NO ₃ ⁻ g·h ⁻¹	0.065	0.021	0.384	-0.768	-0.363	—	
N-inorg. g·h ⁻¹	2.045	0.577	0.333	0.614	1.524	74.5	
SO ₄ ⁼ g·h ⁻¹	4.746	2.539	-0.567	-3.052	-1.08	—	
TSS g·h ⁻¹	10.5	3.42	-9.9	15.84	9.36	89.1	

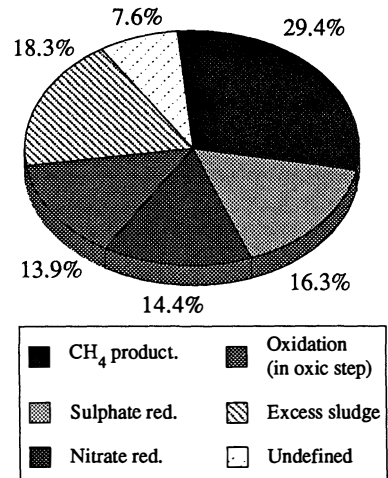


Fig. 5. Distribution of removed COD in the whole process (Tot. = 16.09 g/h)

TC (Total Carbon), IC (Inorganic Carbon) and TOC (Total Organic Carbon) were measured on filtered samples using a DOHRMANN TOC DC 90.

Gas quality was analyzed with a DANI 3865 GC equipped with a thermal conductivity detector.

The plant was equipped with a HP 3852A data acquisition system for the continuous recording of temperatures, gas production, waste water flow and dissolved oxygen.

RESULTS AND DISCUSSION

A general view of the main experimental results, from the whole experimental period, is given in Fig. 2, 3 and 4.

The average input values of total COD were 696 ± 245 mg·l⁻¹, 493 ± 280 mg·l⁻¹ and 637 ± 389 mg·l⁻¹ in spring, summer and autumn respectively and the total COD removal efficiency was 83.4%, 90.1% and 90.7% respectively.

More detailed evaluation has been done from the experimental period from July 1st to October 1st, 1991 during which the treatment plant operated at more stable conditions. The loading parameters B_v (organic loading rate) and HRT from this period are given in Table 4.

Table 5 shows the average values from the selected period of the influent to the plant and the average values of effluents from the individual stages of the process.

In Table 6, results from the mass balances of the most significant process parameters are presented. The mass balances were calculated over the single stages and over the whole plant.

The mass balance of organic matter is based on COD. It should be noticed that the presence of dissolved methane and of reduced forms of sulphur interferes with the organic substances in COD_i determination; anyway, this interference that comes out from the wide variation of the COD/TOC ratio is difficult to quantify. Attempts to measure this interference have shown that COD_{sol} analysis is reliable, because of vacuum filtration that strips out methane and dissolved sulphides.

From the mass balance it is evident that the main part of COD is removed in the anaerobic step. The high negative value of COD_i removed in the anoxic step is caused by wash-out of suspended solids to the oxic step, where they

are removed. The distribution of removed COD in the whole system is shown in Fig. 5. It is evident that only 30% of removed COD was converted to methane.

A significant amount of organic matter is used for reduction of NO_3^- , SO_4^{2-} and for biomass synthesis. Only about 14% of applied organic matter is oxidized in the oxic step.

A relatively small biogas production was observed (Fig. 2). Average daily figures were 11.47 l·d⁻¹ and 16.61 l·d⁻¹ from anaerobic and anoxic steps respectively. Average values of methane concentration were almost constant, 73% and 60% in anaerobic and anoxic steps respectively. The remaining gas was composed of carbon dioxide (3-5%) and nitrogen (24-35%). A significant amount of methane was found in the gas from the anoxic chamber. It would be reasonable to consider that the part of the methane gas initially dissolved in the effluent from the anaerobic chamber was stripped from the liquid phase by nitrogen gas in the anoxic-denitrification chamber. This was taken into account in the calculation of the total methane production. Given a partial pressure of methane in the overlying gas phase, the amount of methane that leaves the reactor dissolved in the liquid phase can be calculated according to Henry's law. The overall methane production was 39.5 l CH₄·m⁻³ of treated sewage,

TABLE 7 Average operating parameters of the oxic unit

HRT	h	8.57
Bv (CODt)	kg·m ⁻³ ·d ⁻¹	0.023
TSS	mg/l	4.08
VSS	g/l	2.67
SVI	ml/g	125
Qx	d	18.3
dissolved O ₂	mg/l	3.6
Q rec-sludge		0.85
Q rec-denit.		1.23

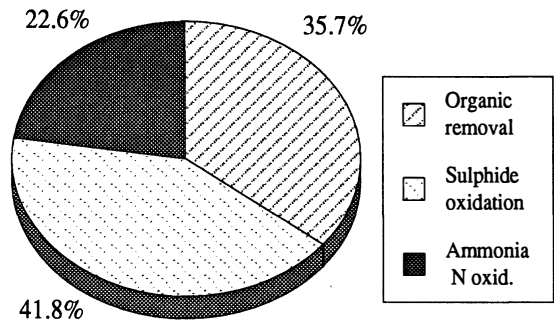


Fig. 6. Distribution of the oxygen utilization in the oxic step (Total = 9.72 g/h)

representing 103.1 l CH₄·kg COD removed⁻¹. Evolved methane represented only 46% of the whole methane produced.

About 28% of the applied N-NH₄⁺ was removed in the anaerobic step. The rest was oxidized to NO₃⁻ in the nitrification tank. The efficiency of nitrification was almost always higher than 92%. It would be reasonable to consider that nitrification was partially inhibited by sulphides. The nitrification rate was found to be 0.05 kg N·m⁻³·d⁻¹ (calculated from the N-NO₃⁻ produced) and the concomitant sulphide oxidation rate was 0.07 kg S·m⁻³·d⁻¹ (calculated from the SO₄²⁻ produced).

Part of the clarified effluent from the final settler (55%) was recirculated into the anoxic chamber for denitrification. The basic variable for denitrification is the carbon source (electron donors). The main advantage of the ANANOX process is that for denitrification there are available electron donors produced in the anaerobic step (VFA, sulphides, CH₄). Denitrification in this arrangement was not limited by the source of available reducing equivalents, but is

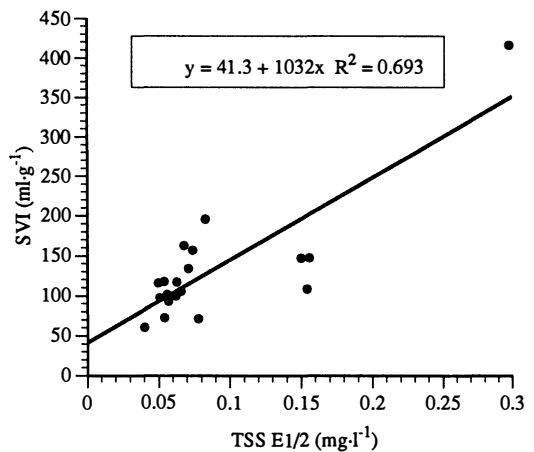


Fig. 7. Relationship between the washout from ABR and SVI of the nitrification sludge

affected only by process parameters (temperature, HRT, inhibitory compounds, etc.).

The denitrification rate was found to be $0.06 \text{ kg N}\cdot\text{m}^{-3}\cdot\text{d}^{-1}$ and efficiency of denitrification 63.5% (calculated from the mass balance over the anoxic unit).

Total Nitrogen removal, computed partly on analytical data and partly on data reconstructed from the correlations calculated in the influent and in the effluent between N-NH_4^+ and Total Nitrogen, was around 82%.

One of the main characteristics of the municipal wastewater used is the significant concentration of sulphates ($113 \text{ mg SO}_4\cdot\text{l}^{-1}$). Sulphate reducing bacteria are responsible for dissimilatory microbial sulphate reduction in the anaerobic reactor. Sulphate acts as electron acceptor in the oxidation of organic matter. Consequently, sulphate reducing bacteria compete with methanogens for electrons available from the substrate, but most of the organic degradation in the anaerobic step was calculated to occur by way of methane production (i.e. 30% of the whole removed COD). For sulphate reduction only 16.3% was used, and about 53% of sulphates were reduced in the anaerobic unit.

As results from the mass balance in the anoxic reactor, a significant amount of sulphates is produced.

It would be reasonable to consider that the sulphur compounds initially reduced in the anaerobic reactor are the electron donors for denitrification in the anoxic step. In the anoxic chamber $0.567 \text{ g SO}_4\cdot\text{h}^{-1}$ were produced that correspond to only 22% of the amount removed in the anaerobic step. The rest of reduced sulphur compounds are oxidized in the oxic step of the process. Effluent sulphate concentration is about 22% higher than influent, due to oxidation of organic sulphur.

Total removal of phosphorus was about 40%. Some of it was removed in the anaerobic unit, probably due to chemical precipitation.

Influent and effluent TSS and VSS concentrations are shown in Table 5. From the mass balance it is evident that about 33% of TSS entered into the reactor are entrapped in the sludge bed of the anaerobic step of ABR.

The mass balance anyway shows that a considerable wash-out of suspended solids from the ABR occurred. The high wash-out of suspended solids was probably the combined result of the high TSS load, of the need of further optimization of the sludge trap and of the quite high surface load applied. During this experimental period the superficial velocity was in fact about $0.66 \text{ m}\cdot\text{h}^{-1}$. Suspended solids washed-out from the ABR were entrapped in the oxic unit where they caused sludge accumulation and significant dilution of activated sludge biomass. Therefore, the only excess sludge during the selected period was taken out from the final settler. Average value of daily excess sludge was $52.8 \text{ g TSS}\cdot\text{d}^{-1}$, while the suspended solids washed out in the effluent were $27.36 \text{ g}\cdot\text{d}^{-1}$; this would represent a total sludge production of $0.207 \text{ kg TSS}\cdot\text{kg COD removed}^{-1}$ ($0.135 \text{ kg VSS}\cdot\text{kg COD removed}^{-1}$) if no suspended solid were washed out in the effluent.

The oxic unit was designed to remove the remaining organic pollutants and to oxidize ammonia nitrogen and reduced sulphur compounds. The main operating parameters of the oxic unit are given in Table 7. From the mass balance it is evident that the main part of oxygen was used for the oxidation of inorganic compounds, while only about 35% of it was used for the oxidation of organic compounds (Fig. 6).

Aerobic post-treatment following anaerobic treatment is by no means without problems from a process point of view. More than the needs of oxygen to reoxidize the reduced N and S compounds, another problem could be the lower biodegradability of organic compounds after anaerobic step and some possible inhibition due to sulphides.

There is no evidence in our experience if sulphides have some influence on SVI.

SVI is therefore affected by suspended solids washed out from the anoxic step, as can be seen from Fig.7.

This shows the importance of a good separation between anaerobic/anoxic and oxic sludges in order to improve the efficiency and stability of the ANANOX process.

CONCLUSIONS

The results of 1990 confirmed the very good performance of the process obtained in the previous campaign and provided much information needed for the optimization of the process and for its management.

The ANANOX process was demonstrated to be highly efficient, with 89.6% COD, 89.2% TSS and 81.2% N removal, and a sludge production of only 0.2 kg TSS·kg COD removed⁻¹. The 40% P removal needs to be further investigated because it is higher than that expected due to sludge growth.

In comparison with a traditional nitrification/denitrification process, the ANANOX produces roughly 50% less sludge, as it comes out from a simulation carried out using the ASCAM model (Di Pinto *et al.*, 1988).

Further investigations need to be carried out on the degree of stabilization of the sludge. One of the future modifications of the process will be the recycle of the thickened excess aerobic sludge to the anaerobic step and the extraction of excess sludge of the whole process only from the sludge trap, positioned at the end of the ABR.

Another advantage of the process is represented by the production of methane ($\sim 10 \text{ l CH}_4 \cdot \text{p.e.}^{-1} \cdot \text{d}^{-1}$), that is anyway very small and can provide only a marginal economic benefit.

The energy needs for oxygen delivery are more or less equal to that of an equivalent aerobic treatment, but the major part of the oxygen needed in the present experience depended on the oxidation of sulphides, because the wastewater contained a fairly high amount of sulphates, probably due to some industrial wastewater discharged into the sewer.

Anyway, sulphates play a very significant role in the process that need to be further studied.

The sulphate reduction in the anaerobic step, although being in competition with methane production, gives a noticeable contribution to the organic matter degradation. In the anoxic step, sulphides are oxidized to sulphates by nitrates, reducing the organic matter need for denitrification. This process can be biologically mediated by a microorganism - *Thiobacillus denitrificans* - that could find here favourable process conditions.

If sulphates are not in excess, like it was in the present experience, they could be completely reoxidized in the anoxic step, reducing the oxygen need in the oxic step.

The ANANOX process proved to be an attractive and feasible alternative to the traditional nitrification/denitrification processes.

This technology could find a particular diffusion in the Mediterranean area, especially for the wastewater treatment of tourist towns or villages that have large seasonal summer growth of the resident population.

Another big area of interest for the application of the ANANOX process is the revamping of overloaded existing plants.

Specific demonstrative projects for these two applications are going to be exploited.

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