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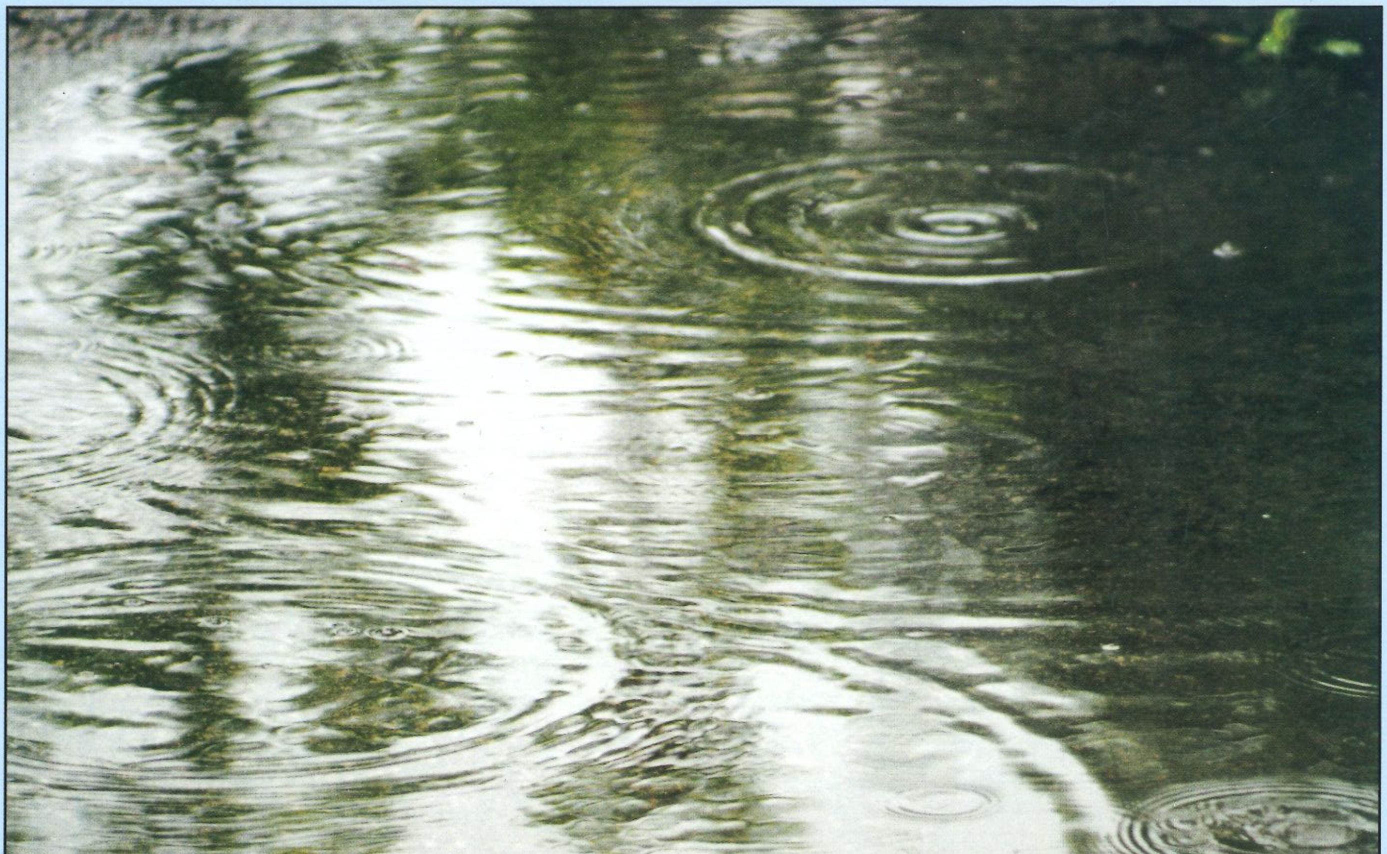

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Integration of membrane technology in communal wastewater treatment: operation and cost analysis

Outdated sewage treatment plants face expansion to meet modern wastewater legislation. An attractive alternative is integration of membrane filtration technology to eliminate post-sedimentation. At the communal wastewater treatment plant in Weiz, Austria, a hollow fibre membrane pilot plant holding 3.7 m^3 activated sludge with 13.9 m^2 submersed membrane area (0.1 µm pore size) was tested over six months at 21 l/m^2h effective throughput. Minimal excess sludge withdrawal, sludge age >30 days and stable biomass at 11 kg/m^3 DS were maintained to achieve 95% and 99% degradation of COD and NH_4-N respectively in the permeate, which also surpassed EU bathing water standards respecting microbiological contamination. Operation was optimal between 3-20 kg/m^3 DS and 10-15 m^3/h membrane aeration, but biofouling became problematic outside these ranges. Extended membrane relaxation effectively reduced trans-membrane pressure and postponed chemical cleaning procedures. Scale-up calculations took k_La, RTD, aeration capacity, membrane area, basin volume and costs into account in comparison to conventional expansion. The price/m^3 wastewater makes both processes competitive, but long-term advantages, including universal applicability, guaranteed effluent purity, volume recovery, reduced excess sludge, modular expansion and presumably longer membrane lifetimes, make membrane technology clearly profitable and adaptable to future socio-economic and ecological needs while efficiently replenishing precious water resources.

INTRODUCTION

The classic method of communal wastewater treatment, which is limited to the removal of solids by sedimentation and dissolved material by microorganisms, is increasingly unable to meet the strict new stipulations of water and health laws, making costly revisions necessary. In many systems, elimination mechanisms are lacking and in others the borders of purification ability have been reached. The primary performance-limiting factor is sedimentation for retention of activated sludge and other solids. Its substitution by microfiltration opens possibilities to improve performance in biological wastewater treatment plants and simultaneously reduce costs.

Typical biological wastewater treatment processes incorporate primary sedimentation, (denitrification), nitrification and secondary sedimentation steps. These are accompanied by many problems such as space limitation, washout danger by rainfall, excess sludge disposal and poor degradation of organic matter and nitrogen. As wastewater amounts increase due to increased canalisation, the old plants are not able to meet the demands of the future and face plant expansion. Expanding capacity under the current activated sludge plant technology is equivalent to doubling or even tripling the size of the holding tanks [1], which naturally is connected with high reconstruction costs [2]. However, even large plants only achieve slight improvement in performance.

An attractive alternative to plant expansion is the innovative application of microfiltration systems employing hollow fibre membrane modules. These modules can be submerged directly in the nitrification basin following the primary sedimentation and denitrification steps, whereby the secondary sedimentation step to separate the wastewater from the sludge becomes obsolete [3]. Such systems have been successfully implemented in industrial wastewater treatment applications [4]. Now the aim is to prove whether the technology is suitable and profitable for application to communal wastewater treatment [5]. Consequently, the need to test membrane filtration in larger communities and update existing biological treatment to meet progressively stricter legal requirements spurred this research.

The site of research was the wastewater treatment plant (Figure 1) in the community of Weiz, Austria (25,000 population equivalents (PE)), which has a population of 10,000 and three primary industries: a slaughter house, schnapps distillery and sheep dairy.

The plant in Weiz was designed in the early 1970s for carbon degradation only. Nowadays the additionally required reduction of nitrogen is impossible for the existing plant. Thus, it faces either volumetric expansion of the conventional method or the specific adaptation of a new technology, i.e. membrane technology, to achieve compliance to the strict rules on effluent emissions. The latter was tested as a pilot membrane plant (PP) run in parallel to the large plant (LP) to provide a direct comparison of both processes.

During seven months of research from March through October 1998 the aims were to

- adapt the membrane system to the specific wastewater composition,
- optimise operation and performance through various experiments,
- determine scale-up parameters and a rough plan for installation in the existing large plant

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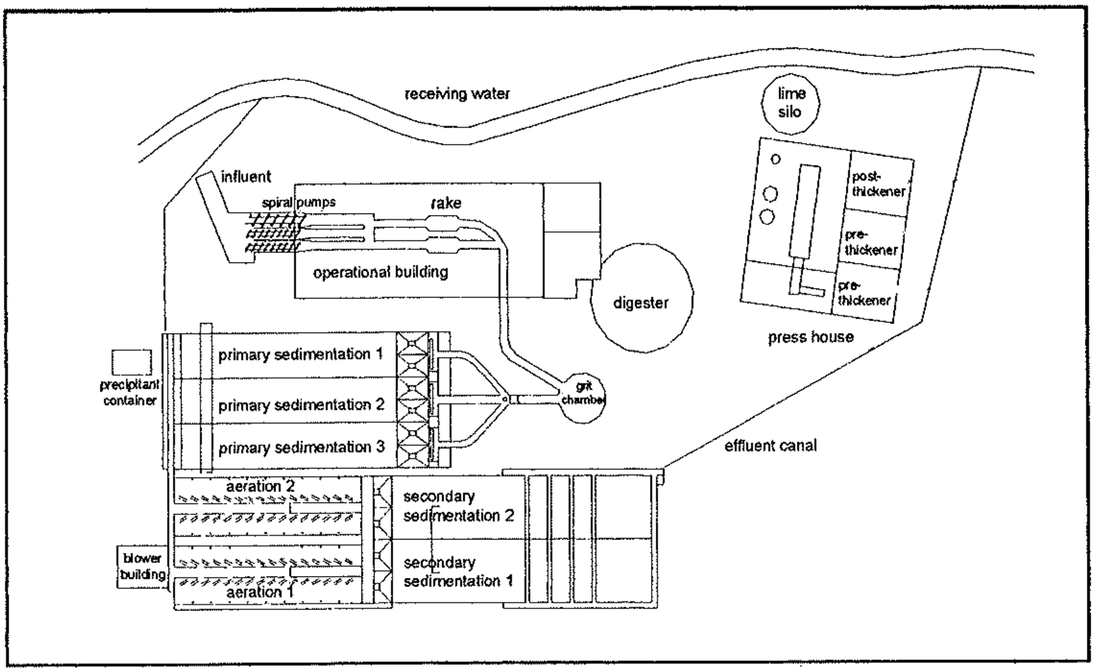

Figure 1. Schematic diagram of the wastewater treatment plant in Weiz, Austria

- and compare costs of conventional expansion vs. membrane technology in wastewater treatment.

These goals were accomplished through experimental emphasis on effluent analysis, sludge attributes, exhaust gas emission, operation ranges, membrane behaviour, energy input and scale-up calculations [6, 7, 8].

MATERIALS AND METHODS

The Pilot Plant

The pilot plant (Figure 2) combined biological wastewater treatment via denitrification and nitrification processes with microfiltration through hollow fibre membranes to separate the purified water from the sludge. It consisted of three primary tanks with a total capacity of 4 m^3, which was only partially utilised.

Systematically, after primary sedimentation in the large plant, wastewater was pumped through a sieve (2 mm pore size) before entering the well-mixed anoxic denitrification tank (0.36 m^3) from which it freely flowed into the nitrification tank (2.5 m^3) for intermittent aeration and was pumped

1) partially back to the denitrification tank,
2) partially as anti-foam spray for the nitrification tank and
3) partially to the continuously aerated filtration tank (0.87 m^3) for microfiltration as permeate.

The filtration tank was equipped with an overflow to the nitrification tank to maintain a constant liquid level to cover the upright submerged membrane module.

The microfiltration could be classified as semi-crossflow filtration since it was driven by application of a low-pressure inner vacuum to create a trans-membrane pressure, together with intensive aeration along the outer membrane surfaces to provide turbulence around the pores. Additionally, this aeration served the double purpose of mixing the tank and supplying oxygen for biological nitrification. Filtration was run intermittently with a periodic permeate backwash cycle

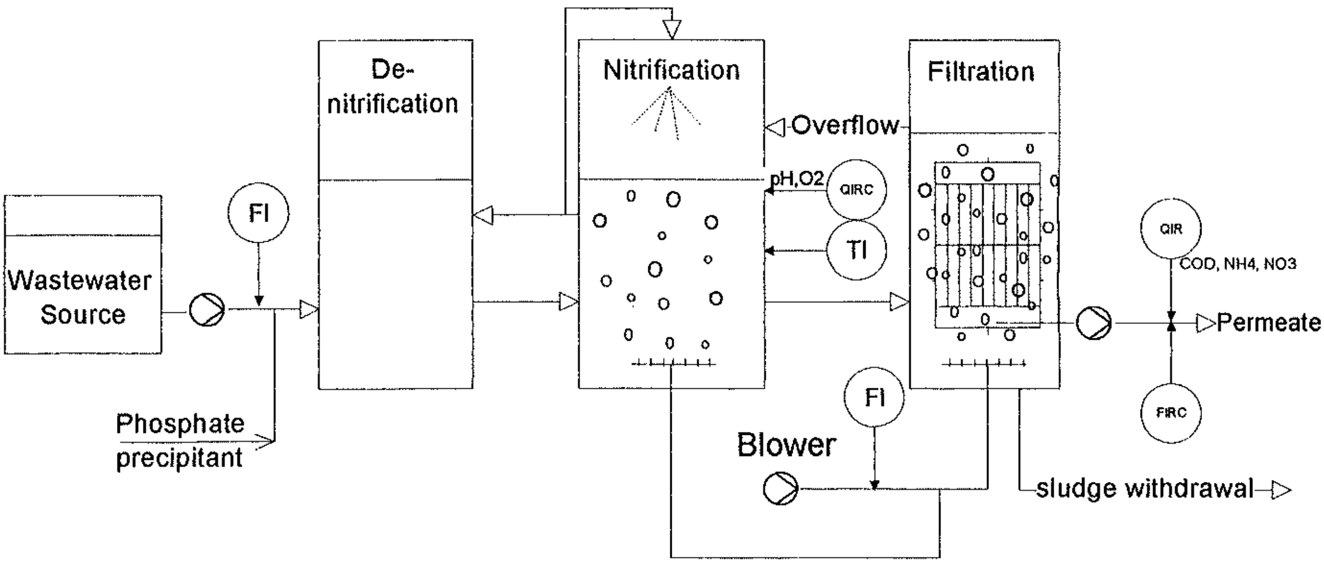

Figure 2. Process flow diagram of the pilot plant (PP)

(30 s every 5 min). The membrane module was specified to be neutral and hydrophilic with a 0.1 μm pore size and 13.9 m^2 total membrane surface area.

After set up, the pilot plant was filled with sludge from the large plant and run at a load corresponding to about 30 PE.

Water analysis

Water samples were taken once daily from the inlets to the primary sedimentation and aeration basins (inlet to pilot plant) and from the permeate and effluent of the large plant. They were analysed photometrically with a WTW Photolab S6 photometer using standard cuvette tests for Chemical Oxygen Demand (COD, WTW C1/25 COD 160), Ammonium (NH_4-N, WTW A5/25), Nitrate (NO_3-N, WTW N1/25), Nitrite (NO_2-N, WTW N4/25) and total-Phosphate (PO_4-P, WTW P5/25 or Merck Spectroquant® Phosphorus Cell test (PMB) 1.14543.0001 P). Heated reactions were carried out with a WTW Thermoreaktor CR 2010. Additionally, the permeate was analysed for total germ count (Merck Standard agar at 30°C for 24 hours), coliforms and faecal coliforms (on Endo agar at 30°C and 37°C, respectively, for 24 hours), *Streptococcus faecalis* (on SF agar (Azide agar) at 37°C, for 48 hours) and *Salmonella* contamination (on m-FC agar at 45°C, for 24 hours).

Sludge attributes

Sludge from the pilot plant was analysed for dry solids (DS) using a Sartorius Moisture Analyser MA 30 to determine its dry weight at 105°C, viscosity using a Haake Viscotester VT-01 rotating drum viscometer and settled sludge volume (SSV) according to the DIN 38 414 Part 10. Occasionally observations of the sludge were made under the microscope at 100x magnification.

According to the Schlumosed Method [9], a comparison between sludge from the large and pilot plants was made with respect to the transparency of its supernatant during undisturbed sedimentation.

Exhaust gas

The exhaust above the pilot plant sludge tanks was analysed by the TÜV (Technische Überwachungs Verein) for greenhouse gas content. Nitrogen oxides were measured with an Ansyco NO/NO_x-Chemilumineszenz-Analysator (0-500 ppm). N_2O was measured using gas chromatography and an ECD detector.

Pilot plant measurements

Tank temperature, membrane aeration, pH (Pro Minet, Duldest Umformer 4-20 mA) and flow (inlet, circulation and phosphate precipitation solution dosage) were documented daily.

In the context of the permeate concentrations, the pilot plant was allowed to reach steady state over two months by adjusting various process parameters. These included permeate flow, trans-membrane pressure, O_2-concentration (Endress & Hauser COS 3, E&H COM 220), nitrification tank blower frequency and throughput, which were recorded online along with parameters from the large plant: conductivity (WTW TetraCon 700-7, WTW LF 171), temperature (with conductivity electrode), pH (WTW Sensolyt 690, WTW pH 170) and O_2-concentration (WTW Trixomatic 690, WTW OXI 170).

Membrane behaviour

Process disturbances were simulated to test their effects on the microfiltration capacity or trans-membrane pressure by varying permeate flux, sludge concentration and membrane aeration. Membrane relaxation as an alternative to the backwash cycle was also examined.

Energy input

The k_La, specific oxygen transport coefficient, for both the large and pilot plants were measured using the Dynamic Method [10] and applied to simulate the energy demand of a membrane system in the large plant. The equation $k_La*t = \ln(O_\infty - O_L)$ was used for calculations, where O_∞ is the oxygen concentration at the final steady state and O_L the oxygen concentration in the liquid phase at time t during the re-aeration period.

Scale-up calculations

The degree of mixing based on residence time distribution tests using a salt tracer in the large plant was determined and evaluated with the Tanks in a Series Model [11].

Additionally, the existing contribution of excess sludge to the pressed sludge quantity was calculated to determine the effects on the nitrogen balance and quantify savings in disposal costs with a membrane system in the large plant.

Finally, three scale-up parameters were calculated for a membrane system in the large plant (expanded to 30,000 PE): aeration capacity, membrane area and basin volume.

From these scale-up considerations, a model was designed for the implementation of membrane technology in the large plant.

Cost comparison of reconstruction

The costs associated with conventional expansion vs. innovative implementation of membrane technology were evaluated using the Standard Price Tables of the State Government of Styria and the standard values for conventional wastewater treatment systems (ATV A131) [12].

RESULTS AND DISCUSSION

Effluent analysis and sludge attributes

A significant divergence in ability to degrade the wastewater existed between both plants (Figure 3).

The deficiency in the large plant confirmed the need for revision of the existing system. Its partial degradation of COD and NH4-N was contrasted by 95% and 99% degradation respectively in the pilot plant. Only with respect to total N-degradation (NH_4-N + NO_3-N + NO_2-N) did the pilot plant come short of the mandatory regulations [13]. The blame lay in a defective oxygen measurement transformer which prevented proper control of aeration in the nitrification tank and inflated the nitrate concentrations. With appropriate aeration regulation, nitrate concentrations should be easily reduced.

Average sludge parameters and quantification indexes for both plants are listed in Table 1.

The average dry solids concentration in the pilot plant over six months of operation was about 9 kg/m^3 (Table 1).

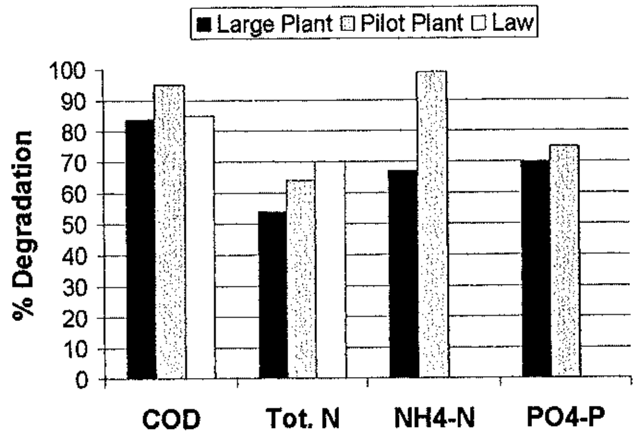

Figure 3. Comparison of degradation in both plants

Table 1. Sludge Parameters for both plants

Sludge Parameter	LP	PP	Unit
Dry solids, DS	2.6	8.6	kg/m³
Viscosity	-	20	mPas
Sludge volume index, SVI	140	131	l/kg DS
Sludge age	2.8	>30	d
Volumetric loading, VL_{COD}	1.7	0.45	kg/m³d
Sludge loading, $SL_{COD,DS}$	0.66	0.05	kg/kgd
Excess sludge	0.66	0.10*	kg_{DS}/kg_{COD_elim}

*Sludge was purged during the sludge concentration experiment, cleaning and sampling; otherwise excess sludge was zero.

Starting at about 3 kg/m³, the sludge rose to an equilibrium concentration of 11 kg/m³ after 2.5 months. The maximum of 14 kg/m³ DS was representative of the significantly higher attainable sludge concentrations as opposed to 4 kg/m³ DS in conventional systems.

Similarly, sludge age rose because sludge loss was controllable and reduced. Complete confinement within the system (with the exception of sludge which was purged in connection with the concentration experiments, cleaning and sampling) enabled slow-growing nitrifying bacteria to propagate, and improved degradation since the older, more concentrated sludge could specialise in and adapt better to the degradation of complex substances.

These effects of altered living conditions upon biomass population were pronounced at the microscopic level. Whereas sludge flocks from the large plant were generally small, compact, containing large lively microorganisms; sludge flocks from the pilot plant were large, diffuse, containing smaller microorganisms and numerous free-swimming and thread-forming bacteria (Figure 4).

Although such characteristics are threatening to conventional sedimentation processes which favour speedy flocculation and low free-swimming or thread-forming bacteria count, they were irrelevant in the pilot plant because the biomass was permanently retained within the membrane system and did not affect the visual purity of the effluent. But even more important was the excellent degradation achieved by the pilot sludge, which was attributed to increased surface contact between the microorganisms and substrate-containing wastewater. Indeed, biochemically efficient sludge contradicts the customary requirement of good secondary sedimentation properties [14].

Consequently the sedimentation properties were tested to investigate the differences in sludge properties at the macroscopic level and confirm the poor separation characteristics of the pilot sludge. Figure 5 shows the percent transparency in the supernatant of settling sludge over time for sensors at three different depths.

Clearly, the supernatant from the large plant reached 90% transparency at all three sensors after 16 minutes, whereas that from the pilot plant hovered at about 10% transparency following the same length of sedimentation time. Hence, after six months of separation from the large

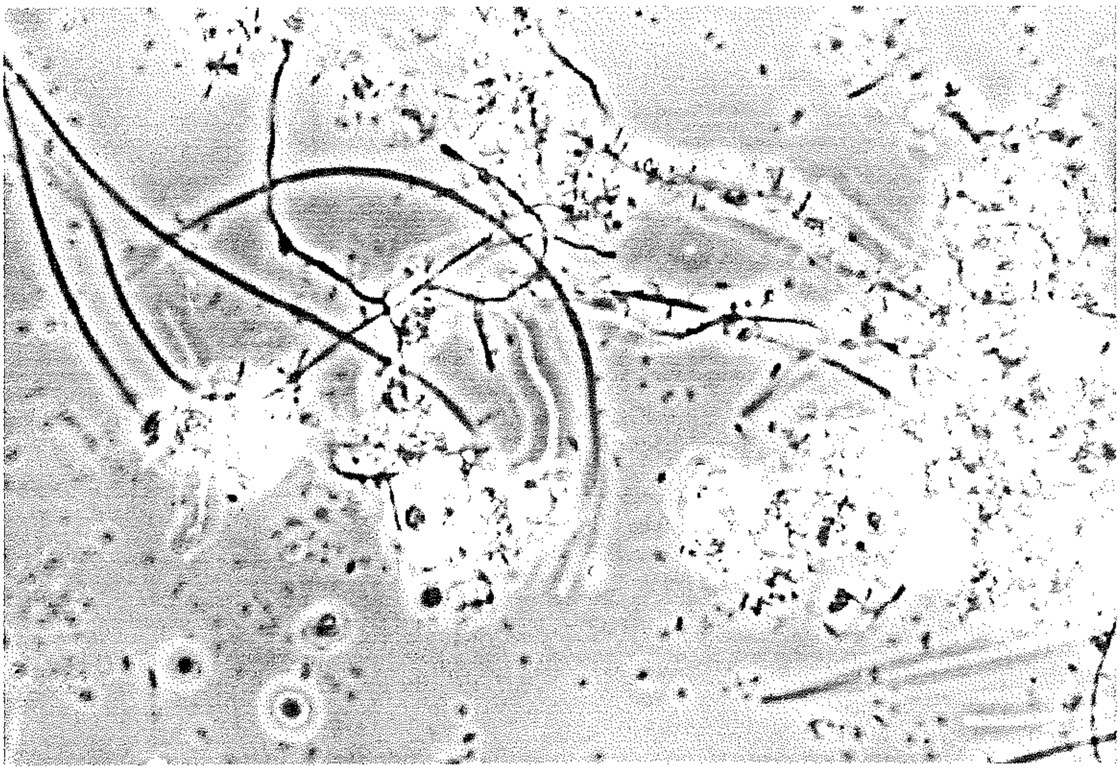

Figure 4. Pilot plant sludge flocks with thread-forming and free-swimming bacteria (100x magnification)

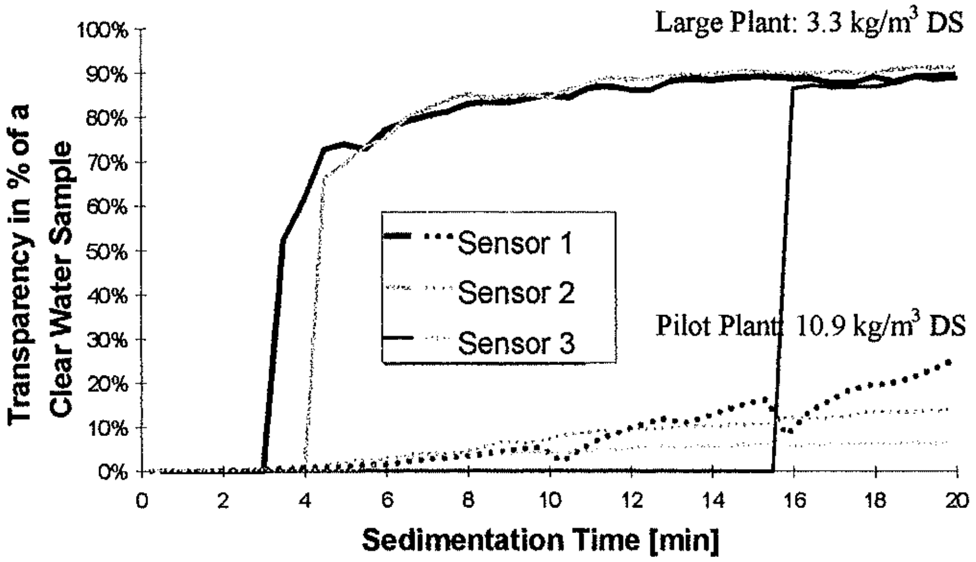

Figure 5. Sludge separation: Large Plant vs. Pilot Plant [9]

plant, the original biomass population evolved and adapted in a strongly divergent manner to cause these profound changes in its physical properties. The continual turbulence and exposure to alternately aerobic and anoxic zones most likely contributed to the formation of free-swimming and thread-forming bacteria that hindered efficient flocculation and sedimentation.

Regarding microbiological contamination of the permeate, the average of four samples (each in tandem) over three months was 1900 cfu/100 ml total germ count, of which none were coliform, faecal coliform, nor *Salmonella*, and only seven were *Streptococcus faecalis*. Therefore, on the basis of these parameters, the permeate passed the microbiological aspects of the European bathing water standards [15] (guide values in cfu/100 ml: 500 total coliform, 100 faecal coliform, 100 *Streptococcus faecalis*) and could be classified as suitable for bathing purposes. Because of the questionable sterility of the sampling and membrane filtration methods, external germs may have been included in the results.

Phosphate precipitation in the large plant was achieved via dosage of Iron Sulphate and yielded an average ß-value of 1.6 (ß-value = mole metal ion delivered/mole eliminated PO_4-P). In the pilot plant, minimal amounts of phosphate precipitation solution VTA 24 (Al^{3+}) were required to meet an effluent concentration of 1 mg/l PO_4-P, which corresponded to a ß-value of 0.37. These results would indicate higher phosphorous removal in the pilot system caused by complex binding of the phosphate, but further investigation is needed.

Exhaust gas

The emission of harmful exhaust gases (NO_x, N_2O) due to incomplete denitrification processes was measured in the pilot plant. Incomplete denitrification is caused by inhibition of reductases by oxygen [16] and becomes critical after transitions from anaerobic to aerobic conditions [17] or low pH.

The results showed insignificant emission concentrations of NO_x at 1 ppm and N_2O under the detection limit (<2 mg/m³ or <1 ppm). Therefore, the conditions in the pilot plant were favourable to biological denitrification and non-harmful to the environment.

Pilot plant operation and membrane behaviour

Pilot plant operation transpired smoothly over the entire research period, with the exception of a few process disturbances (power failure, safety shutoff events, oxygen regulation) and membrane blockage, which was the culmination of rigorous experimental testing at the system's limits.

The operational averages are listed in Table 2.

Table 2. Pilot Plant Operation Parameters

Parameter	Min.	Mean	Max.	Unit
Total volume		3.7		m³
Temperature	16	21*	26	°C
PH	6.6	7.4	8.9	-
Total throughput (net)	3.5	7	8.7	m³/d
	19.9	21	26.2	l/m²h
Permeate flux (nominal)	14.4	26.4	30.9	l/m²h
Backwash flux (nominal)	42.1	45.5	48.4	l/m²h
Membrane pressure_permeate	-0.120	-0.336	-0.519	bar
Membrane pressure_wash	0.420	0.562	0.646	bar
Membrane aeration	0.36	0.88	1.15	m³/m²h

*a simple cooling system was necessary during the months of July and August to prevent sludge temperature from rising above 26°C.

Sludge temperature and pH remained within the optimal limits for nitrification and denitrification reactions. The membrane aeration alone proved to be sufficient for nitrification. Thus the nitrification tank increasingly became the site of denitrification as the intermittent aeration cycle was gradually reduced. For the majority of the research project, the maximum allowed permeate flow was tested with backwashing at a factor of 1.6 that of the permeate.

In the context of various membrane experiments, the operation ranges and limits of the membrane system were explored (Table 3).

Table 3. Ranges for membrane parameters

Test	Min.	Max.	Unit
Sludge concentration	0	20	kg/m³
Membrane aeration	10	15	m³/h
Short membrane relaxation	> 72		s
Extended membrane relaxation	> 12		h

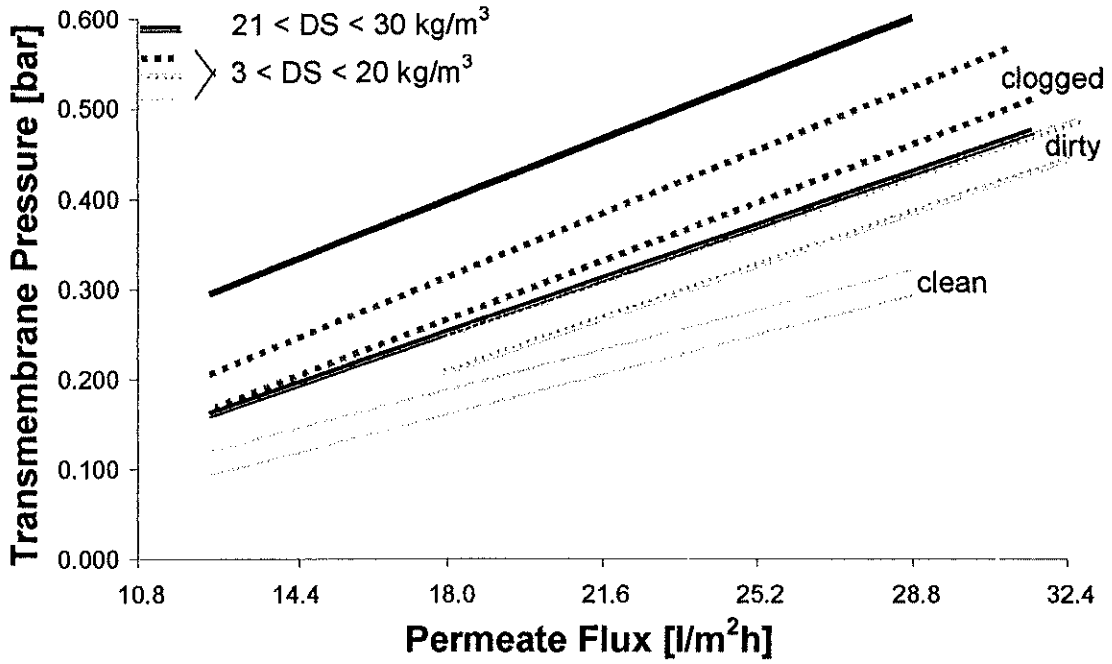

Figure 6. Trans-membrane pressure as a function of permeate flux, sludge concentration and membrane condition

With increasing permeate flow, the membrane pressure rose linearly at 1.25 mbar/l/h. Sludge concentrations under 20 kg/m^3 had little effect on trans-membrane pressure; but above 21 kg/m^3, the pressure rose exponentially until the safety shutoff took action at -0.6 bar and 30 kg/m^3 sludge concentration. Likewise, membrane aeration between 10-15 m^3/h had minimal effect on trans-membrane pressure, but aeration at 5 m^3/h caused severe membrane blockage and system shutdown after 11 hours. Aeration above 15 m^3/h was avoided as the oxygen surplus disturbed denitrification and consequently caused the pH to drop.

As a result of these tests, biofouling became problematic: gradually backwashing no longer took effect, the membrane pressure rose to a critical level (>-0.5 bar) and the membranes became blocked. The behaviour of trans-membrane pressure as a function of permeate flow, sludge concentration and membrane condition is presented in Figure 6.

In the end thorough chemical and mechanical cleaning was necessary to remove the thick sludge cakes covering the outsides of the membranes. This action effectively reduced the trans-membrane pressure, however alternatives were tested which would be less troublesome and more cost effective. Both extended relaxation and controlled flow fluctuations acted to reduce and maintain trans-membrane pressure respectively.

Dynamic flow was used to match fluctuations in influent loads but also had the advantage of relieving the membranes during the night when the permeate was withdrawn at half the rate, yet backwashing was carried on as usual (a rate of 1.6 times the peak permeate flux). A long-term equilibrium set in and trans-membrane pressure remained stable.

The benefits of extended relaxation in contrast were accidentally discovered after process shutdowns when the filtration was halted but membrane aeration was maintained. Extended relaxation is defined as >12 hours of membrane aeration without backwashing or permeate withdrawal. Drops in pressure of up to 100 mbar were recorded. The novelty of extended relaxation is the possibility of its application in a multi-module membrane system. In essence, recurrent, extended relaxation of consecutive modules in a membrane system could become part of a routine maintenance schedule, which could replace the need for chemical cleaning, or at least extend the cleaning interval.

Short membrane relaxation (72 s every 5 min) was not found to be a suitable alternative to the backwashing cycle since the trans-membrane pressure rose and productivity would have been lost at longer relaxation times.

Economic evaluation of energy input, scale-up and reconstruction

To consider energy input and scale-up of the pilot plant for application in the large plant, k_La, residence time distribution and other parameters were determined. These were used to compare reconstruction with conventional expansion vs. membrane technology.

The primary energy investment in both plants centred around aeration and the transfer of oxygen to the microorganisms. Typically, most of the air was lost to the atmosphere, making aeration inefficient. In confirmation of this phenomenon, only 3% of the total oxygen input could be attributed to biological utilisation. Likewise k_La experiments yielded k_La < 10 [h^{-1}] in both plants. The poor oxygen transfer rates were attributed in part to the open systems but primarily to the absence of mechanical mixers to disperse the bubbles and prevent coalescence. Since the use of mechanical mixers could damage the membranes, the higher costs due to inefficient aeration are most likely unavoidable.

Because all mixing in the large plant transpired because of gravity flow and the above-mentioned aeration, the determination of hydraulic mean residence time (MRT) was necessary for considerations pertaining to scale-up with membrane technology. Individual MRT of 5.5 h and 4.2 h were measured in the primary sedimentation and aeration basins respectively. Mixing in the aeration basins was nearly ideal (the equivalent to 1.2 stirred tanks according to the Tanks in a Series Model [11]), but the primary and secondary sedimentation basins demonstrated regions of dead volume and channelling. The experimental measurements were compiled according to a structured model for hydrodynamic and oxygen transfer [4, 18, 19] coupled with kinetics according to the activated sludge model number 1 (ASM 1) [20].

Two other factors for scale-up were the determination of the nitrogen balance and quota of pressed sludge originating from excess sludge withdrawal. Calculation of the nitrogen balance was difficult because the influence of air stripping, organically bound nitrogen, and emission of gaseous intermediates during nitrification and denitrification was unknown. However, the amount of ammonium produced during sludge digestion and returned to the plant in the press water was calculated to be 17 kg/d NH_4-N. This amount would be reduced in a membrane system in the large plant, based on the following result of the excess sludge determination.

As previously mentioned, the withdrawal of excess sludge is reduced in membrane systems. Thus the existing quantity of excess sludge was determined to evaluate the economic advantages of its absence. Without excess sludge, a 53% reduction of pressed sludge would be achieved, implying significant advantages to disposal costs and reducing press water amounts.

Together with the previous considerations, three scale-up parameters were determined: blower capacity, membrane area and basin volume.

Because of geometric differences between the two plants, blower capacity could not be directly scaled up from the pilot plant. Instead, standard values for conventional wastewater treatment systems were assumed which yielded a theoretical oxygen consumption of 118 kg O_2/h in the large plant. At a 2.5 m basin depth, this corresponds to 4,732 m^3/h air, which was multiplied by a factor 2 to account for air lost to membrane aeration. Thus the total quantity of air required would be 9,464 m^3/h, or 158 m^3/min. At the moment, 3 blowers are installed in the large plant, which deliver a total of 104 m^3/min air: 1 at 36.8 kW delivers 53 m^3/min, and 2 at 22 kW deliver 25.6 m^3/min. Thus an additional blower with 36.8 kW would need to be installed. According to the membrane manufacturer, it is necessary to aerate the membranes at 1 m^3/m^2h, or in this case with 12,500 m^3/h air (see Table 4). However, this value cannot be taken as a set guideline before aeration parameters are fully researched.

In contrast, membrane area is only dependent upon influent flow and can be easily scaled by membrane flux. Although the producer suggested a 20 l/m^2h flux, flux as high as 30 l/m^2h was tested in the pilot plant over 2 months without adverse effects on the membranes. Flux above 30 l/m^2h may even be temporarily applied by raising the trans-membrane pressure to cover daily peaks or rain events. For the scale-up calculation, the data in Table 4 were used to yield a total required membrane area of 12,500 m^2, or 272 membrane modules which could be organised into 34 blocks of 8 modules each (see Figure 7).

Table 4. Scale-up Data

	Value	Unit
Dry weather peak influent	250	m^3/h
Dry weather flux	20	l/m^2h
Rainy weather peak influent	437.5	m^3/h
Rainy weather flux	up to 35	l/m^2h
Membrane area	12,500	m^2
Modules	272	Qty.

Regarding basin volume, a direct scale-up from the pilot to the large plant was impossible since the proportions were not similar. Instead, a control calculation of required aeration basin volume was made using standard values [12] and assuming 0.1 kg DS/kg BOD_5 excess sludge, 40-day sludge age and 15 kg/m^3 dry solids sludge concentration. The result yielded 360 m^3, which is well under the existing 624 m^3 aeration basin volume. Assuming adequate denitrification at 38% of the aeration basin volume, only 137 m^3 would be necessary. This denitrification volume could be accommodated in one of the two existing secondary sedimentation basins (406 m^3).

Figure 7 shows a diagram of these process revisions for a model membrane system in the large plant.

Compared with Figure 1, the basic process would not be altered much by installation of membrane technology. Contrary to conventional expansion, no supplementary basin volume would be necessary.

Revision of the existing system would transpire as follows: Since the primary sedimentation process is over scaled, one of the basins could be used as a stormwater-balancing tank. Next, the flow through the aeration basins would be reversed, requiring a trough to be erected from the primary sedimentation basins to the current effluent (end of the secondary sedimentation basins). Here as mentioned above, one (or both if necessary) of the secondary sedimentation basins would be converted into a denitrification tank, and the other could be used as an additional reserve for rain events. The present aeration basins would be used as nitrification tanks with submersed membrane modules, which serve as substitutes for secondary sedimentation. Between the existing secondary sedimentation basins and aeration basins,

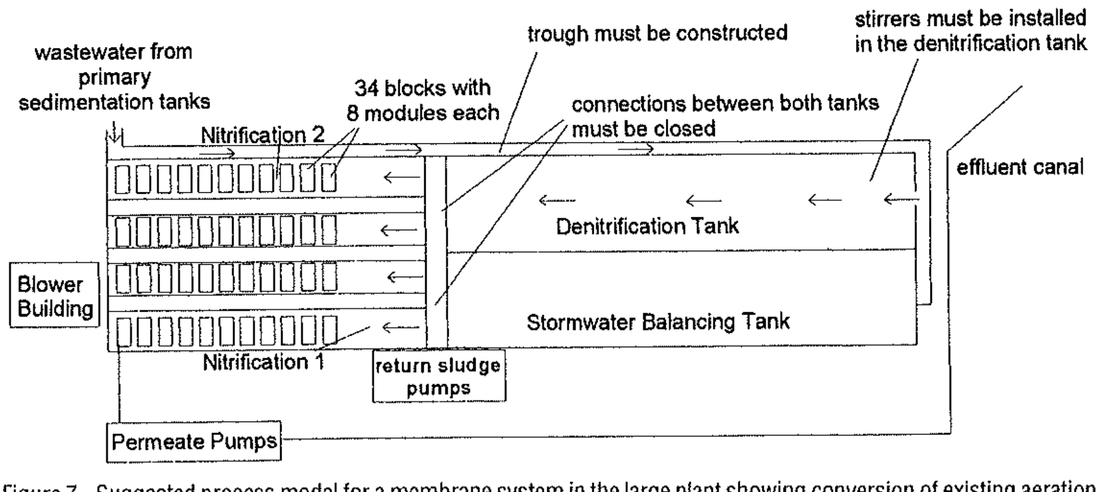

Figure 7. Suggested process model for a membrane system in the large plant showing conversion of existing aeration and secondary sedimentation basins into nitrification and denitrification tanks respectively.

a separating wall with a passage would have to be built to minimise the exchange of oxygen between the nitrification and denitrification tanks. The flow of return sludge would also have to be reversed such that sludge from the nitrification tanks could be withdrawn (at best from several points in the nitrification tank to ensure even distribution) for return into the denitrification tank. The blowers would be equipped with a frequency transformer to regulate a constant concentration of oxygen in the nitrification tank. Additionally, a permeate pump would be installed, and finally the denitrification basin would be equipped with stirrers.

The costs of these revisions provide a direct comparison to the alternative of conventional expansion. As calculated according to the ATV A131 [12] the additional volume expansions listed in Table 5. Here the requirements for membrane technology are also summarised.

Table 5. Reconstruction Requirements: Conventional Expansion vs. Membrane Technology

	Existing Capacity	Additionally Required Capacity		Unit
		Conventional Expansion	Membrane Technology	
Pre-clarifiers	864	0	0	m^3
Nitrification	624	2,683	0	m^3
Denitrification	0	2,227	0*	m^3
Post-clarifiers	812	5,267	0**	m^3
Total volume	2,300	10,177	0	m^3
Aerator capacity	80	3	38	kW
Membrane area	0	0	12,500	m^2

*1 secondary sedimentation basin would be converted to denitrification by reversing the flow
**membranes replace secondary sedimentation

Plainly conventional technology would result in a five-fold expansion of the present total volume. In terms of expenditures, Table 6 compares both alternatives to yield a final price of wastewater per cubic meter.

The values listed for conventional expansion were calculated according to the Standard Price Tables of the State Government of Styria for erection of a wastewater treatment plant larger than 10,000 PE, which list the base cost at 305 Euro/PE (subdivided into 60% construction (183 Euro/PE) and 40% machinery (122 Euro/PE)). In the case of the large plant, construction costs were multiplied by a factor of 0.6 to account for the existing basin volume and utility buildings; and machinery costs were multiplied by a factor of 0.5 to account for continued use of existing equipment.

Table 6. Reconstruction Expenditures: Conventional Expansion vs. Membrane Technology

	Costs (kEuro)	
	Conventional Expansion	Membrane Technology
Construction costs	3,296.4	360
Machinery costs	1,831.4	1,090
Membrane costs*	0	1,859.6
Yearly operating costs**	732.5/yr	584.3/yr
Yearly total costs***	1,216.5/yr	1,287.3/yr
Cost/m³ wastewater	**0.56 Euro**	**0.59 Euro**

*6,800 Euro per module
**14.5 Euro/PE general operation + 9.9 Euro/PE for sludge disposal
***yearly total costs taking construction service (depreciated over a 25 year life at 7% interest), machinery service (dep. 15 years at 7% interest), membrane service (dep. 6 years at 7% interest), general operating costs, sludge disposal and additional costs for membrane technology (electricity and chemicals) into account.

Regarding the expenditures listed for membrane technology, minimal reconstruction costs were assumed at €360,000 and machinery costs were estimated to be €1,090,000. In comparison to conventional expansion, machinery costs could be saved, general operating costs would remain the same and sludge disposal costs would be reduced up to 50% because of the reduction of excess sludge.

Additional costs for membrane technology include costs for electricity and chemicals. These were included in the yearly total costs in Table 6 at 127,900 and 34,000 Euro/year respectively. Chemical costs could be saved if extended relaxation were applied in a maintenance schedule as mentioned previously.

Since the current guaranteed lifetime of the membranes is six years, a membrane system could become costly. However, longer service times are expected in the future which would lower the membrane costs and make this process profitable. In any case, the bottom prices, 0.56 and 0.59 Euro/m^3 wastewater for conventional expansion and membrane technology respectively, indicate that both alternatives are competitive. We believe the advantages demonstrated during this research project outweigh the slight apparent economic advantage of conventional expansion and provide a sure step into the future demands of wastewater treatment.

CONCLUSIONS

During this pilot project for integrated membrane filtration of communal activated sludge, the biomass population adapted to the specific conditions and wastewater components to maintain a stable equilibrium, which improved effluent quality and degradation of complex organic substances even during overloads and toxic surges. COD and ammonium were effectively reduced without the emission of toxic gases. Although the permeate surpassed the EU bathing water guidelines regarding the absence of faecal germs and parasites; heavy metals, viruses, hormones and other soluble toxic substances likely passed through the pores of the membrane.

The extraction of excess sludge was unnecessary. However, the 'infinite' sludge age and thickness could create ideal conditions for increased genetic transfer rates and elevated resistance to antibiotics [21]. With certainty, the sludge would not be suitable for emergency clarification in sedimentation basins because of its high thread-forming and free-swimming bacteria content hence poor separation ability.

Temperature variations had little effect on sludge concentration or its purification properties. Due to increased biomass concentration, more heat per volume was generated during biological degradation, which would be advantageous to nitrification reactions in the winter months but might require cooling systems in the summer depending on the specific application.

The pilot plant functioned optimally at sludge concentrations under 20 kg/m^3 and with 12 m^3/h membrane aeration. Operation outside the limits led to increased trans-membrane pressure, biofouling, membrane blockage and

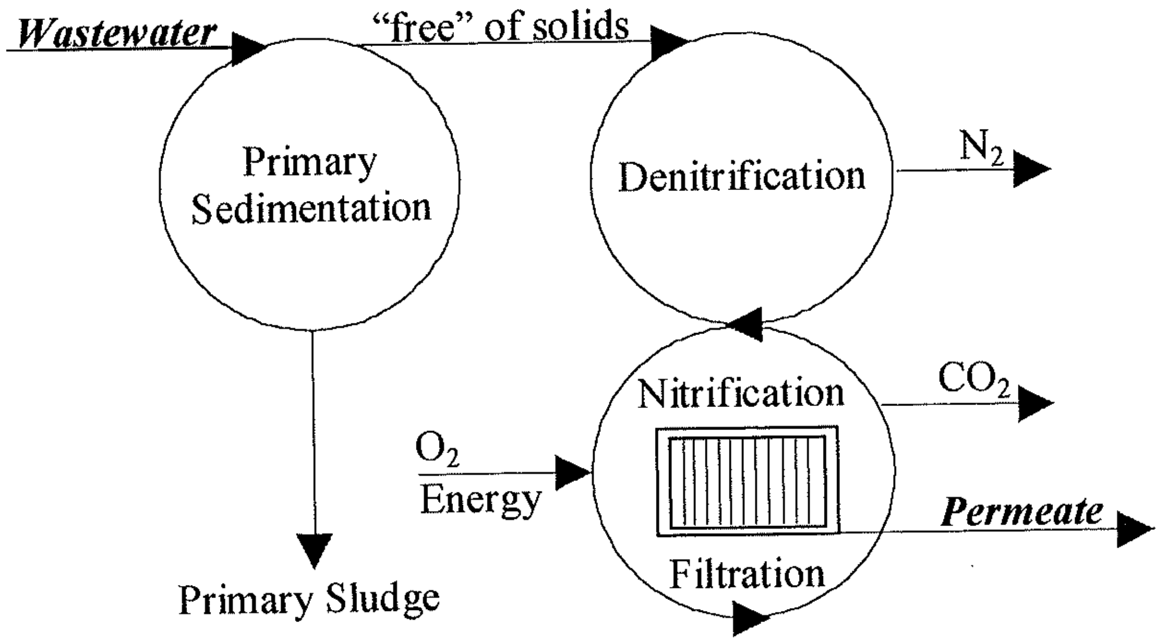

Figure 8. Biological wastewater treatment with integrated membrane filtration

intensive cleaning procedures. Chemical cleaning was avoided by application of extended relaxation, which effectively reduced membrane pressure.

Implementation of membrane technology to modernise the large plant would reduce excess sludge production, save sludge disposal costs by up to 50%, recover two basins for stormwater balancing, allow sufficient capacity for future expansion of the Weiz community and guarantee long-term effluent purity. In contrast, conventional plant expansion would require a five-fold volume increase and significantly higher initial investment costs. Long-term suitability, adaptability and cleansing performance would remain uncertain especially since more restrictive water purification standards are required in the future.

Outlook

We recommend the use of membrane technology in small-scale applications especially in arid regions, but we also hope to demonstrate that large-scale applications are equally successful and profitable to growing communities like Weiz, Austria.

In the future, as the membrane industry develops, longer standing times and definite price reduction will probably make the process clearly profitable. Presently, an important economic advantage exists in the fact that membrane systems can be gradually and flexibly up-scaled and adapted on a modular basis to meet the specific needs of a particular community. In contrast, plant expansion has to be scaled for future maximum loads from the start, providing little leeway for deviation and requiring enormous initial investment.

Altogether, the essential innovation is the elimination of plant expansion as the sole method to meet strict wastewater regulations. Attaining compliant effluent purity is no longer questionable and tank volume is won back for possible use as a buffer during rain events or for future coverage of a community's growing wastewater treatment needs. Thus, future widespread integration of membrane technology in wastewater treatment could prove to be significantly more ecologically friendly and cost-effective as it enables efficient transformation of polluted water into clear purified permeate (Figure 8).

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