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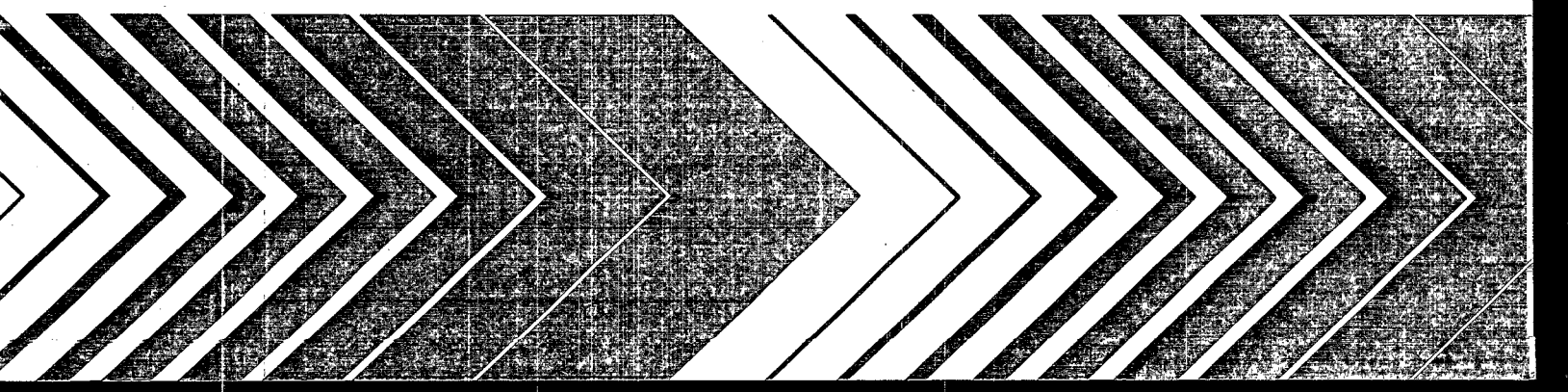
Technology Assessment of Anaerobic Systems for Municipal Wastewater Treatment:

1. Anaerobic Fluidized Bed; 2. ANFLOW



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TECHNOLOGY ASSESSMENT of ANAEROBIC SYSTEMS FOR MUNICIPAL WASTEWATER TREATMENT: 1. ANAEROBIC FLUIDIZED BED 2. ANFLOW

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and asthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

The study summarized in this report assesses the potential of two recently developed approaches for the anaerobic treatment of municipal wastewater.

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ABSTRACT

This report discusses two developing technologies for the treatment of municipal wastewaters. These technologies are anaerobic fluidized bed systems and an anaerobic fixed-film bioreactor (ANFLOW). In both systems wastewater is treated at ambient temperature. The objective of this report is primarily to provide guidance to those individuals involved with reviewing new processes as part of the Innovative and Alternative Technology program.

Fluidized bed systems have previously been utilized for wastewater treatment. However, anaerobic fluidized bed treatment of municipal wastewater has only been evaluated in the laboratory. Available data show that primary effluent can be successfully treated to provide an effluent of acceptable secondary quality (30/30 mg/l of BOD and SS). To accomplish this, the anaerobic fluidized bed systems are operated to provide extremely high solids retention times. This report discusses: available laboratory data on system performance; fluidized bed expansion and voidage-velocity relationships; the influence of bacterial growth on changes in fluidization characteristics; power requirements for fluidization; potential cost and energy savings compared to activated sludge secondary treatment plants; and provides estimates of anaerobic fluidized bed treatment costs. Because of the limited data available, the technology is still unproven and further information in several areas discussed in this report is needed. Anaerobic fluidized bed processes are considered eligible for funding as innovative technology on a case by case basis where all relevant factors affecting process performance have been carefully considered in the design.

ANFLOW is an acronym for an anaerobic treatment process evaluated at Oak Ridge National Laboratory and reported on at several conferences. The process treats raw wastewater introduced into the bottom of a bed containing Raschig ring packing. The upflow velocities are quite low (4 - 12 m/hr or 100 - 300 gpd/sq ft), and detention times of several hours are required. Data obtained to date show that the process is not capable of achieving an effluent of acceptable secondary quality. Previous favorable analyses of process potential and economics have been based on questionable process expectations and assumptions. This report analyzes and reevaluates the data and process economics previously reported.

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TECHNOLOGY DESCRIPTION

INTRODUCTION

This report discusses two developing technologies for the treatment of municipal wastewaters. The processes are an anaerobic fluidized bed and an anaerobic fixed-film bioreactor(ANFLOW). In both systems, wastewater at ambient temperature is treated anaerobically to promote suspended solids capture and conversion of organic materials to bacteria, carbon dioxide and methane.

The anaerobic treatment of wastewater offers several potential advantage- when compared to aerobic treatment systems. The power requirements for activated sludge aeration typically represent from 40 to 65 percent of the energy demand in plants treating municipal wastewater; power requirements of 0.119 to 0.172 kwh/cu m (450 to 650 kwh/MG) are not uncommon. When aerobic sludge digestion is incorporated into the plant design, the power requirements for aeration can easily double. Furthermore, a 3785 cu m (1 MGD) activated sludge system following primary clarification will normally contribute an additional 272 to 363 kg (600 to 800 lb) of secondary sludge which requires stabilization and dewatering. Anaerobic treatment processes can produce a high degree of waste stabilization, require no oxygen and produce significantly less biological sludge than aerobic processes. In addition, a potentially usable fuel, CH_4, is also produced. However, the perceived disadvantages of these systems for the treatment of average strength municipal wastewaters have limited their consideration as viable treatment technologies.

PROCESS DESCRIPTION

Anaerobic Fluidized Beds

A schematic diagram of an anaerobic fluidized bed is shown in Figure 1. Wastewater is pumped up through a distribution system which supports the media and provides for uniform flow through the bed. The bed consists of a solid support medium such as sand, carbon, anthracite or synthetic particles to provide a surface area for biological film attachment and growth. The combination of particle size, shape, density, bacterial film thickness and properties, and wastewater viscosity determine both the fluid velocity needed for bed fluidization and the bed expansion characteristics following the onset of fluidization. By keeping the particles in a fluidized state, blockages of the bed are avoided. A more in-depth discussion of fluidization principles is presented in Section III.

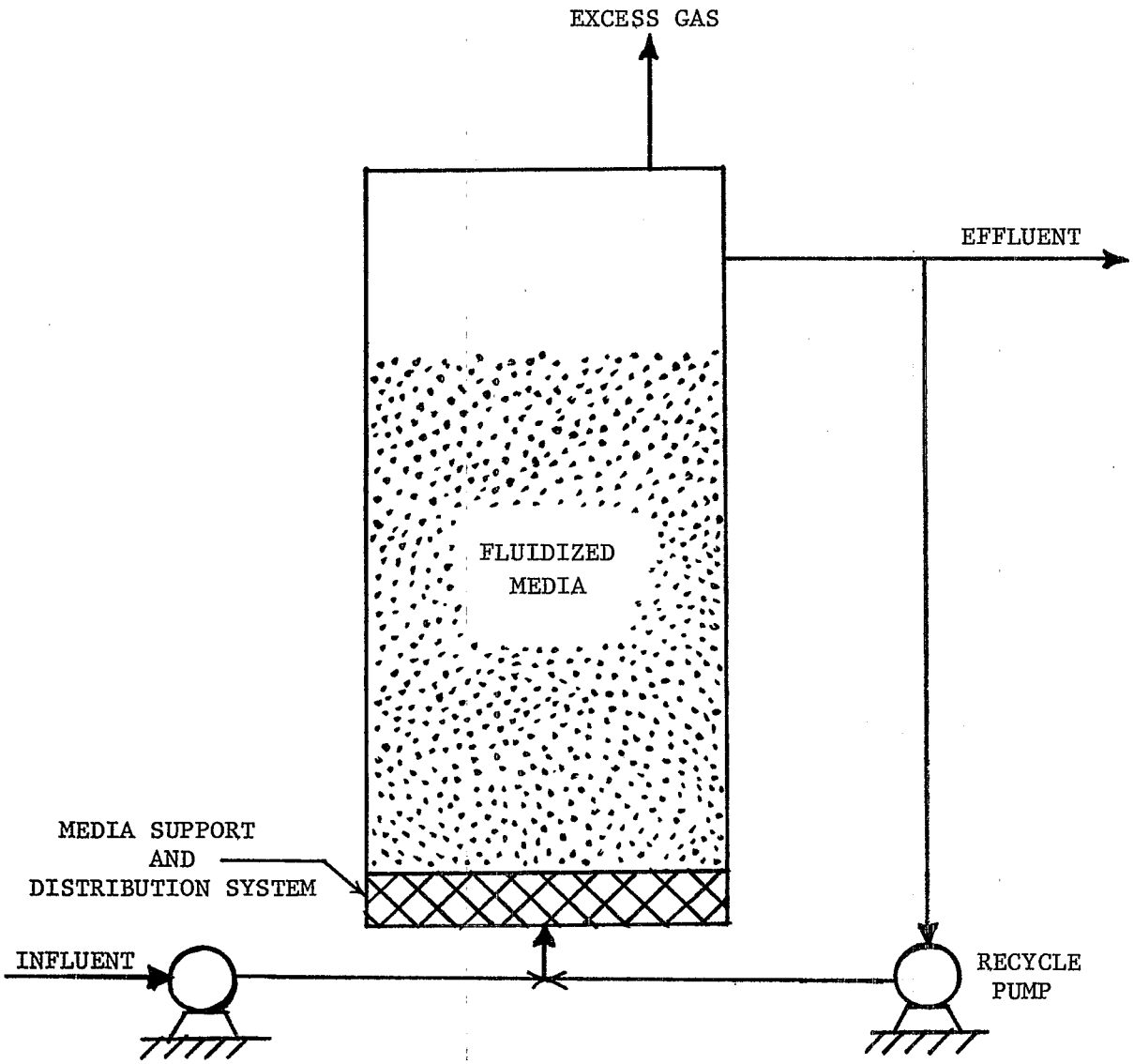

FIGURE 1. Schematic Diagram of a Fluidized Bed System

ANFLOW

ANFLOW is an acronym for an anaerobic upflow fixed-film system developed by Oak Ridge National Laboratory (ORNL) under contract with the U.S. Department of Energy (DOE). A schematic diagram is shown in Figure 2. The system investigated contained a 3.05 m (10 ft) depth of 2.5 cm (1 in) unglazed ceramic Raschig ring packing. Upflow velocities are roughly an order of magnitude less than commonly observed in fluidized bed operation and result in reactor detention times of several hours.

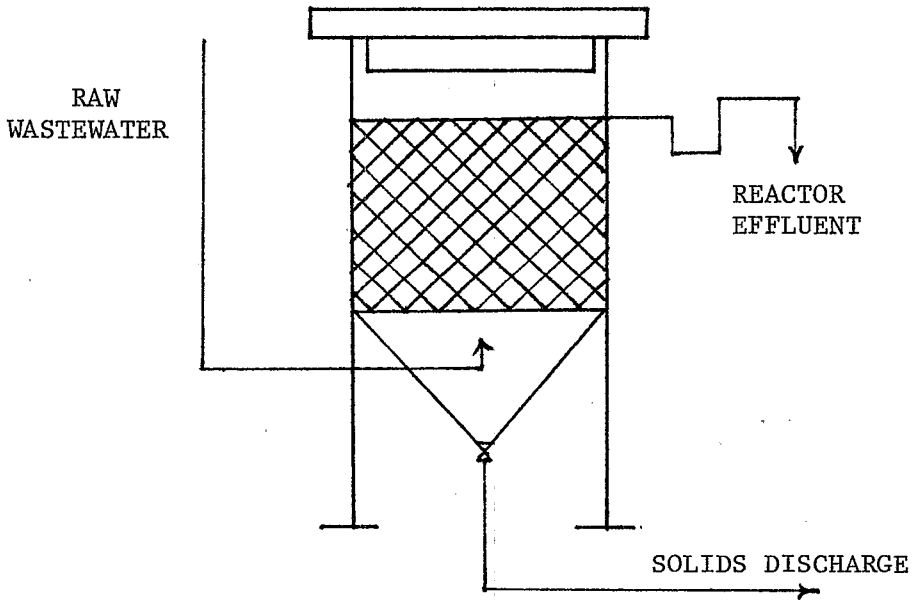

FIGURE 2. Schematic Diagram of an ANFLOW Reactor

SECTION 2

DEVELOPMENT STATUS

ANAEROBIC FLUIDIZED BEDS

Treatment of domestic wastewater in anaerobic fluidized beds has been investigated in the laboratory (1,2); these studies are discussed in Section III. No pilot scale investigations have been reported in the literature. No full scale plants treating municipal sewage are in operation.

Fluidized bed use in wastewater treatment is not a new concept. Aerobic fluidized bed systems have been evaluated for carbonaceous and nitrogenous oxidation, and commercial systems, such as the Oxitron System marketed by Dorr-Oliver, are available. Fluidized bed technology has been demonstrated for wastewater denitrification at pilot plant scale (3) and a full-scale expanded bed system for wastewater denitrification is currently awaiting startup in Pensacola, Florida. Anaerobic fluidized bed systems have been investigated at pilot plant scale for treatment of liquors from thermal sludge conditioning (4) and for a corn starch waste (5). A 3.66 m (12 ft) diameter full-scale anaerobic system for treatment of high strength waste has recently gone into operation in Birmingham, Alabama (6).

ANFLOW

ORNL has conducted a 2-year pilot plant investigation with the ANFLOW system. Results from this study have been presented (7,8,9), and are discussed in Section III. ORNL plans to construct a pilot plant designed for a nominal flowrate of 189 cu m/day (50,000 gpd) in the Fall of 1980 on a site provided by the City of Knoxville, Tennessee. Funding will be provided by DOE. A workshop describing the system and soliciting industrial participation in the process was held in May, 1980 (10).

SECTION 3

TECHNOLOGY EVALUATION

BACKGROUND AND GENERAL PROCESS THEORY

Anaerobic processes for the treatment of high strength industrial wastewaters have been in use for some time. Cillie et al., (11) reviewed the results of a number of studies on anaerobic treatment of industrial wastes and discussed performance data from full-scale plants. Mueller and Mancini (12) summarized a number of instances where the anaerobic filter has been investigated for industrial waste treatment of high strength wastes. The upflow anaerobic sludge blanket process has also been recommended where waste strengths greater than 1000 mg/1 COD are present (13).

Anaerobic sludge digestion has been widely used in municipal wastewater treatment plants. Anaerobic processes are also under investigation for the treatment of concentrated sidestreams such as liquors arising from heat treatment of waste sludge (4,14).

The use of anaerobic processes for direct treatment of influent wastewater in facilities other than septic or Imhoff tanks or anaerobic lagoons is also not a new idea. Over twenty years ago, Coulter et al., (15) reported on laboratory studies in which raw domestic wastewater was passed through an upflow anaerobic sludge contact chamber and then through a rock filter. BOD removals at room temperature averaged 82 percent and effluent solids varied from 2 - 20 mg/1. The study was continued in a one-year investigation in a 6.2 cu m (1650 gpd) pilot plant treating raw wastewater (16). The wastewater first flowed upward through a conical sludge contact tank with an 18-hour detention time. Suspended solids removals in the sludge contact tank of 74 and 88 percent and COD removals of 54 and 77 percent were obtained in winter and summer, respectively. Below upflow velocities of 4.57 m/day (15 ft/day or 112 gpd/sq ft) the sludge zone exhibited a definite line of demarcation. An additional 44 percent suspended solids removal was obtained in the rock column but COD removals only averaged 10 percent, with gas production in warm weather causing increased solids carryover. Hydrogen sulfide generation presented an odor problem. Overall, the combined units removed 91 percent of the suspended solids and 66 percent of the COD.

When Fall and Kraus (17) evaluated a full-scale anaerobic upflow contact tank receiving domestic waste over a 20-month operating period, the suspended solids removal averaged 77 percent but the average BOD removal was only 34 percent. Large amounts of silt and clay in the influent contributed to the solids removal efficiency which was stated to remain unchanged when

the retention period was reduced to 13.4 hours (hydraulic loading of 8.67 m/day or 213 gpd/sq ft). BOD removal was worse in the summer due to acid fermentation of the sludge and escape of organic acids in the effluent.

Pretorius (18) used an upflow sludge contact chamber followed by a stone and sand filter and found that up to 90 percent of the COD in raw sewage (excluding effluent solids) could be removed in 24 hours at a temperature of 20 ^{o}C. Thirty-five to forty percent of the solids captured in the contact chamber were reported to be hydrolyzed, with the remainder requiring periodic removal. More gas was produced in the biofilter than in the digester.

Young and McCarty (19) evaluated biological treatment at 25 ^{o}C in anaerobic rock filled filters receiving synthetic wastes. Effluent quality was inversely proportional to hydraulic detention time. The absence of requirements for solids separation and return, heating above 25 ^{o}C, and a minimum solids disposal problem suggested that the filter had a number of economic advantages for sufficiently concentrated wastes. They further indicated that although satisfactory treatment of low strength wastes may be possible, the anaerobic filter appears to operate best at waste strengths above about 1000 mg/l of ultimate BOD.

The underlying biochemical principles which are operative in either ANFLOW or fluidized bed reactors are the same as in any anaerobic waste treatment process. BOD is removed by the entrapment of suspended material, by the formation of bacterial cells which do not escape in the effluent and by the production of CH_4 gas. A generalized scheme for methane formation is shown in Figure 3. Contrary to previous beliefs, propionate and butyrate are not substrates for methanogenic bacteria but are converted to H_2, CO_2 and acetate by a hydrogenogenic microflora (20). As shown by McCarty (21), organic waste concentrations of 5000 mg/l or above are required before methane production is sufficient to raise the waste temperature significantly by combusting the gas produced. The minimum solids retention time (SRT) for 18.3 ^{o}C (65 ^{o}F) operation was reported to be 11 days. Mueller and Mancini (12) reported that a sludge age of 100 days or greater will yield optimal COD removals in anaerobic filters. Hence unheated anaerobic biological systems treating dilute wastes require long SRT's to maintain suitable conditions and acceptable kinetic rates for methane formation to occur in reactors of reasonable size.

ANAEROBIC FLUIDIZED BEDS

Process Theory

As noted by Miller (22), there is some potential for confusion in the terminology associated with expanded and fluidized beds. In distinguishing between the two conditions, Cooper and Wheeldon (23) indicated that in an expanded bed the particles remain in stationary contact while in a fluidized bed the particles are in free motion.

As fluid is passed upward through a bed of particles there is a linear relationship between the log of the pressure drop vs. the log of the fluid velocity. As the velocity approaches the minimum fluidizing velocity, some

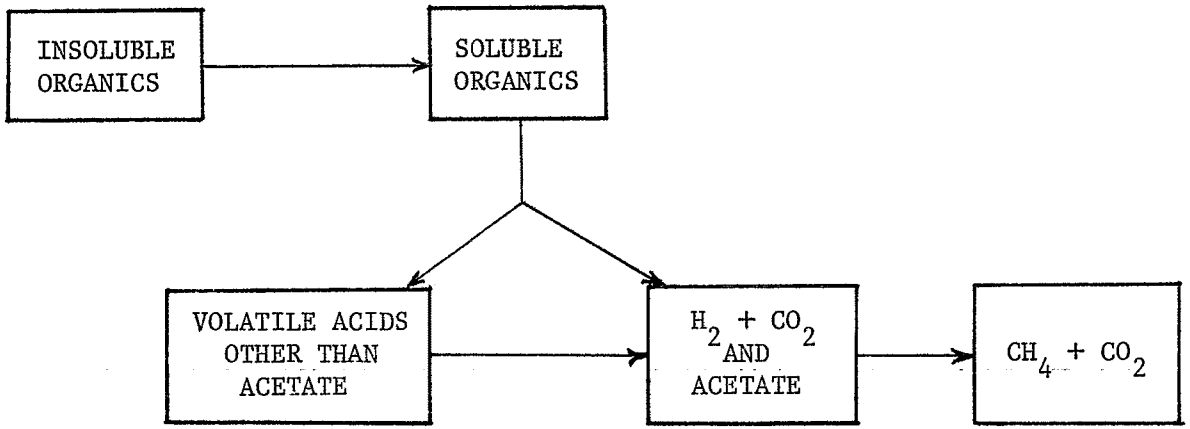

FIGURE 3. Scheme for the Formation of Methane from Organic Matter

bed expansion (up to 5 % according to Cooper and Wheeldon) will normally occur before the pressure drop has reached the buoyant particle weight per unit area of bed. This effect will be most marked when the bed is initially highly consolidated (24). A small amount of localized movement can occur but if the bed is viewed as a whole the particles are still in stationary contact. If flow were further increased a point is reached at which the pressure drop is just equal to the buoyant particle weight and this is the point of minimum fluidization. Further flow increases result in expansion of the fluidized bed while the pressure drop remains essentially constant.

Local variations in the permeability of a randomly packed bed tend to cause a gradual rather than sudden transition to a fluidized bed. Furthermore in beds containing variable sized particles, some differential fluidization may be observed. The behavior of both ideal and real beds has been described by Richardson (24). An in-depth review of various expansion models and procedures to characterize bed expansion has been given by Cleasby and Baumann (25). The minimum fluidization velocity can be estimated from the following relationships (25):

$$V_{mf} = 0.00381 \quad \frac{d^{1.82} \left[\gamma(\gamma_s - \gamma) \right]^{0.94}}{\mu^{0.88}} \qquad (1)$$

where

V_{mf} = minimum fluidization velocity in gpm/sq ft
d = particle diameter in mm
γ, γ_s = fluid and particle specific weights in lb/cu ft
μ = viscosity in centipoise

Between 2°C and 40°C, μ can be adequately represented by the following:

$$\mu = 1.778227 - 0.05671\ T + 0.001067\ T^2 - 0.00000885\ T^3 \qquad (2)$$

where

T = temperature, °C

If the Reynolds number at minimum fluidization (Re_{mf}) is greater than 10, the following multiplication correction factor, k_{mf}, should be applied to the velocity.

$$k_{mf} = 1.775\ Re_{mf}^{-0.272} \qquad (3)$$

The head loss through a fluidized bed is given by

$$\Delta P = \frac{L\ (\gamma_s - \gamma)\ \ (1. - \varepsilon)}{\gamma} \qquad (4)$$

where

ΔP = head loss in ft
L = bed height in ft
ε = porosity of the expanded bed

In addition to the pressure drop within the bed, the designer must also account for the head loss in the fluid distributor at the base of the bed.

Figure 4 shows the variation in minimum fluidization velocity for spherical particles of different diameters and specific gravities. Decreasing water viscosity with increasing water temperature results in an increase in the fluidization velocity for particles of given characteristics.

A number of workers have shown (24) that the voidage-fluidizing velocity relationship for particulate bed expansion in a fluidized bed can be represented for many systems by the following expression:

$$\frac{V}{V_i} = \varepsilon^n \qquad (5)$$

where

V is the fluidization velocity

V_i is approximately equal to the free-falling sedimentation velocity of the particle in the liquid

n is an exponent which varies between roughly 2.4 to 4.5 for spherical particles and which is normally higher for nonspherical particles. For spherical particles it can be correlated to the Reynolds number computed from the free-falling particle velocity.

This equation has no theoretical basis but is widely used because it is simple in form, readily applied and reasonably accurate. A number of other relationships have also been developed (26). Cleasby and Baumann (25) have presented considerable information showing expansion height vs. flow rate for different sizes of clean silica and garnet sands.

Process Capabilities

Jewell (1,2) reported on an upflow expanded bed reactor (Jewell's terminology) with the support media consisting of a mixture of PVC particles and ion exchange resin with diameters less than 1 mm. This laboratory study utilized a 1-liter reactor with 5.1 cm (2 in) I.D. After fifty days startup operation which included seeding with anaerobic sludge, experiments with primary effluent as feed were conducted for a period of 200 days. The primary effluent was a weak domestic waste with an average influent COD of 186 mg/l. Primary effluent was blended with recycle, with the recycle pumping rate maintained constant at about 100 ml/min. Except for some shock loading studies, the temperature was maintained at 20°C. Effluent quality was monitored by unfiltered COD and SS measurements.

During the 200 day study, the hydraulic retention time (HRT) was varied from 24 hours down to a low of 0.08 hours near the end of the investigation. For the first 95 days the HRT was 4 hours or greater and after approximately a 20-day period of operation at HRT's of 2 - 0.5 hours the HRT was again returned to 8 hours for several weeks. During the last ten days of the study the HRT was varied from 0.25 down to 0.08 hours.

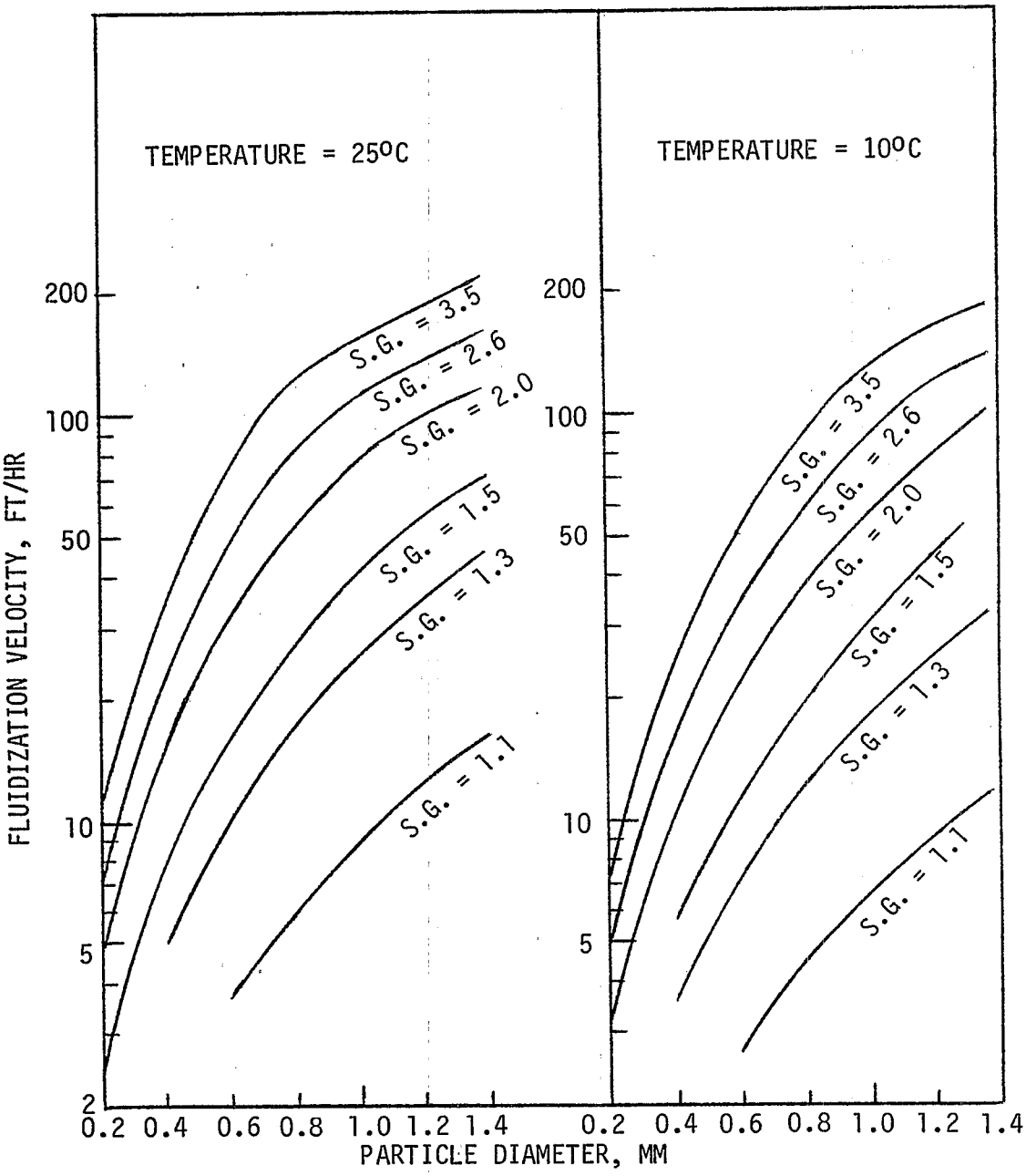

FIGURE 4. Fluidization Velocity vs. Particle Diameter at Different Specific Gravities

The effective biomass concentration in the reactor was reported to vary from 20 to 30 kg VSS/cu m. The cell yield was estimated at about 0.15 g VSS/g COD destroyed. If a reactor removed 150 mg/1 of COD, had a 4 hour HRT, effluent VSS of 6 mg/1 and a net yield of 0.15, it would take 252 days to accumulate a biomass concentration of 25 kg VSS/cu m. Alternatively, if a reactor were operating at equilibrium with a 4 hour HRT, reactor VSS of 25 kg/cu m, effluent VSS of 6 mg/1, and no deliberate sludge wasting, the SRT would be 694 days. These calculations are intended to illustrate the long SRT's which are characteristic of the system investigated by Jewell (1,2).

The data indicated that the anaerobic expanded bed system could treat primary effluent and consistently produce a secondary effluent of excellent quality (COD of $\sim$ 30 mg/1 and SS of $\sim$ 4 mg/1) when operating at an 8 hour HRT and at 20°C. Good effluent quality was also obtained during operation at a 4 hour HRT. The data suggest that long term operation at HRT's of 1-2 hours may also be possible. Although the system produced acceptable effluent quality during the brief periods of operation at 1-2 hour HRT's, it is not known what would happen over a long time period if operation were continued under these conditions. In view of the long SRT's associated with equilibrium operation under a given set of conditions, the successful operation for a few days at the high loadings does not insure that the same effluent quality would be achieved at the new equilibrium conditions which ultimately develop.

Switzenbaum and Jewell (27) also evaluated the expanded bed concept in a laboratory study with a feed of glucose and nutrient salts. This small scale study used 5.1 cm (2 in) I.D. columns with a fluidization media of aluminum oxide particles that were approximately 0.5 mm in size. The bed was expanded from an initial volume of 400 ml to an operating volume of 500 ml. Three reactors were operated at 10, 20 and 30°C, respectively with steady state feed concentrations ranging from 200 to 600 mg/1 of COD. Solids concentrations in the reactor were reported to range between 15,000 to 38,000 mg/1 TVS. At feed concentrations of 200 and 400 mg/1, the COD removals resulting from a combination of cell synthesis and CH_4 production were as shown in Table 1.

TABLE 1. SOLUBLE COD REMOVALS REPORTED BY SWITZENBAUM AND JEWELL (27)

	SOLUBLE COD REMOVAL %					
HRT	Feed, 10°C, mg/1		Feed, 20°C, mg/1		Feed, 30°C, mg/1	
hours	200	400	200	400	200	400
6	73	83	83	88	79	82
4	70	81	74	86	72	83
2	55	65	72	81	66	77
1	50	54	57	67	61	70

These data show that fluidized bed systems are operable over the range of wastewater temperatures which are encountered throughout most of the United

States. On the average, about 80 percent of the COD removal resulted from CH_4 formation. Whether results from municipal wastewater treatment will be comparable to those obtained from glucose at the lower temperatures has not yet been ascertained.

Design and Energy Considerations

Selecting the most appropriate media for a given application in an anaerobic fluidized bed system will depend upon the HRT required and the overall characteristics of the bacteria film/inert media particles. One of the most interesting observations in the study by Switzenbaum and Jewell (27) was the extremely thin bacterial film thicknesses encountered. Film thickness was estimated by viewing the particles under a light microscope with a calibrated ocular. The thicknesses ranged from a minimum of 7.1 to a maximum of 14.4 microns. It was also reported that the unattached entrapped biomass comprised between 4 to 6 percent of the total biomass present.

Changes in particle characteristics resulting from bacterial growth will impact fluidization characteristics of the bed. The degree of impact can be estimated by calculating changes in bed characteristics which would be observed from an assumed spherical bacterial growth of different thicknesses developing around spherical support media. The results of such calculations for one set of assumed parameters is shown in Table 2. A bacterial specific gravity of 1.50 (dry weight basis) with a film concentration of 0.15 gm/cu cm represents a bacterial film with an apparent specific gravity of 1.05 (0.15 + (1. - 0.15/1.5)). The changes resulting from the 0.015 mm assumed bacteria thickness in Table 2 indicate that the thin dense films reported by Switzenbaum and Jewell (27) should have very little impact on the fluidization characteristics of the bed as a whole. For example, the 15 micron film modeled in Table 2 would decrease the fluidization velocity of a 0.5 mm particle by only 0.52 m/hr (1.7 ft/hr) i.e., from 10.03 m/hr (32.9 ft/hr) to 9.51 m/hr (31.2 ft/hr). On the other hand, the 0.65 mm activated carbon particles in the denitrification columns operated by Jeris et al. (28) were reported to reach particle sizes of 3 to 4 mm diameter. The influence that various film thicknesses would have on the support particles and the resulting changes in the bed characteristics in the absence of some positive mechanism to limit the particle size can be discerned from Figure 5. This Figure is also based on the same model for spherical particles that was used in Table 2, although some of the parameter estimates are different in this example. These results illustrate that in designing fluid bed systems it is important to know the nature and thickness of the bacterial growth to be expected. This will influence the optimal media size and density, tha amount of bed expansion observed, the need to control media-bacteria particle size, and the importance of diffusional considerations within the films in controlling the biofilm kinetics (29).

As an operational expedient, the systems studied by Jewell (1,2) and Switzenbaum and Jewell (27) used a very high recycle rate to maintain bed expansion. The recycle rate in Jewell's system was maintained at 100 ml/min which corresponds to an upward velocity of 70.4 m/day (1730 gpd/sq ft); for operation at a 4-hour HRT the recycle:influent pumping ratio was 24:1. Switzenbaum and Jewell used even higher recycle flows (211 m/day or 5200 gpd/

TABLE 2. Fluidization Characteristics for Spherical Particles

Temperature = 20.0 Degrees C

Media Specific Gravity..... 2.650
Bacteria Specific Gravity..... 1.500
Bacteria Film Thickness, mm..... 0.015
Bacteria Film Concentration, gm/cu cm..... 0.150
Initial Porosity of Unfluidized Bed..... 0.400
Final Porosity of Unfluidized Bed..... 0.400

INITIAL VALUES:									
Particle Diameter, mm	0.20	0.30	0.40	0.50	0.60	0.70	0.80	0.90	1.00
Particle Volume, cu mm	0.0042	0.0141	0.0335	0.0654	0.1131	0.1796	0.2681	0.3817	0.5236
Particle Mass, mg	0.0111	0.0374	0.0888	0.1734	0.2997	0.4759	0.7104	1.011	1.388
Particle Number Per Liter	1.43×10^8	4.24×10^7	1.79×10^7	9.17×10^6	5.31×10^6	3.34×10^6	2.24×10^6	1.57×10^6	1.15×10^6
Fluidization Velocity, gpm/sq ft	0.773	1.617	2.729	4.097	5.709	7.557	9.637	11.94	14.46
Fluidization Velocity, ft/hr	6.20	12.97	21.89	32.86	45.79	60.62	77.30	95.78	116.0
Reynolds Number	0.1049	0.329	0.741	1.389	2.323	3.588	5.229	7.289	9.811
Fluidization Head Loss, ft/ft	0.990	0.990	0.990	0.990	0.990	0.990	0.990	0.990	0.990
FINAL VALUES:									
Particle Diameter, mm	0.230	0.33	0.43	0.53	0.63	0.73	0.83	0.93	1.03
Particle Volume, cu mm	0.00637	0.0188	0.0416	0.0779	0.1309	.2037	0.2994	0.4212	0.5721
Particle Specific Gravity	2.102	2.252	2.338	2.393	2.432	2.461	2.483	2.500	2.514
Particle Mass, mg	0.0134	0.0424	0.0973	0.1866	0.3184	0.5012	0.7433	1.053	1.438
Unexpanded Particle Number Per Liter	9.42×10^7	3.19×10^7	1.44×10^7	7.70×10^6	4.58×10^6	2.95×10^6	2.00×10^6	1.42×10^6	1.05×10^6
Bacteria in Unexpanded Bed, mg/l	3.08×10^4	2.24×10^4	1.76×10^4	1.44×10^4	1.23×10^4	1.06×10^4	9.41×10^3	8.43×10^3	7.64×10^3
Ratio of Unexpanded Bed Volumes	1.521	1.331	1.242	1.191	1.158	1.134	1.117	1.103	1.093
Fluidization Velocity, gpm/sq ft	0.6821	1.484	2.556	3.886	5.461	7.275	9.32	11.59	14.08
Fluidization Velocity, ft/hr	5.472	11.90	20.51	31.17	43.81	58.35	74.75	92.96	112.9
Reynolds Number	0.1064	0.3321	0.7456	1.397	2.334	3.602	5.246	7.31	9.836
Head Loss Per ft of Media Initially Present, ft	1.006	0.999	0.997	0.9957	0.9947	0.994	0.993	0.9931	0.9928
Bacteria Per Liter of Media Initially Present, kg	0.0469	0.0297	0.218	0.0172	0.0141	0.0121	0.0105	0.0093	0.0083

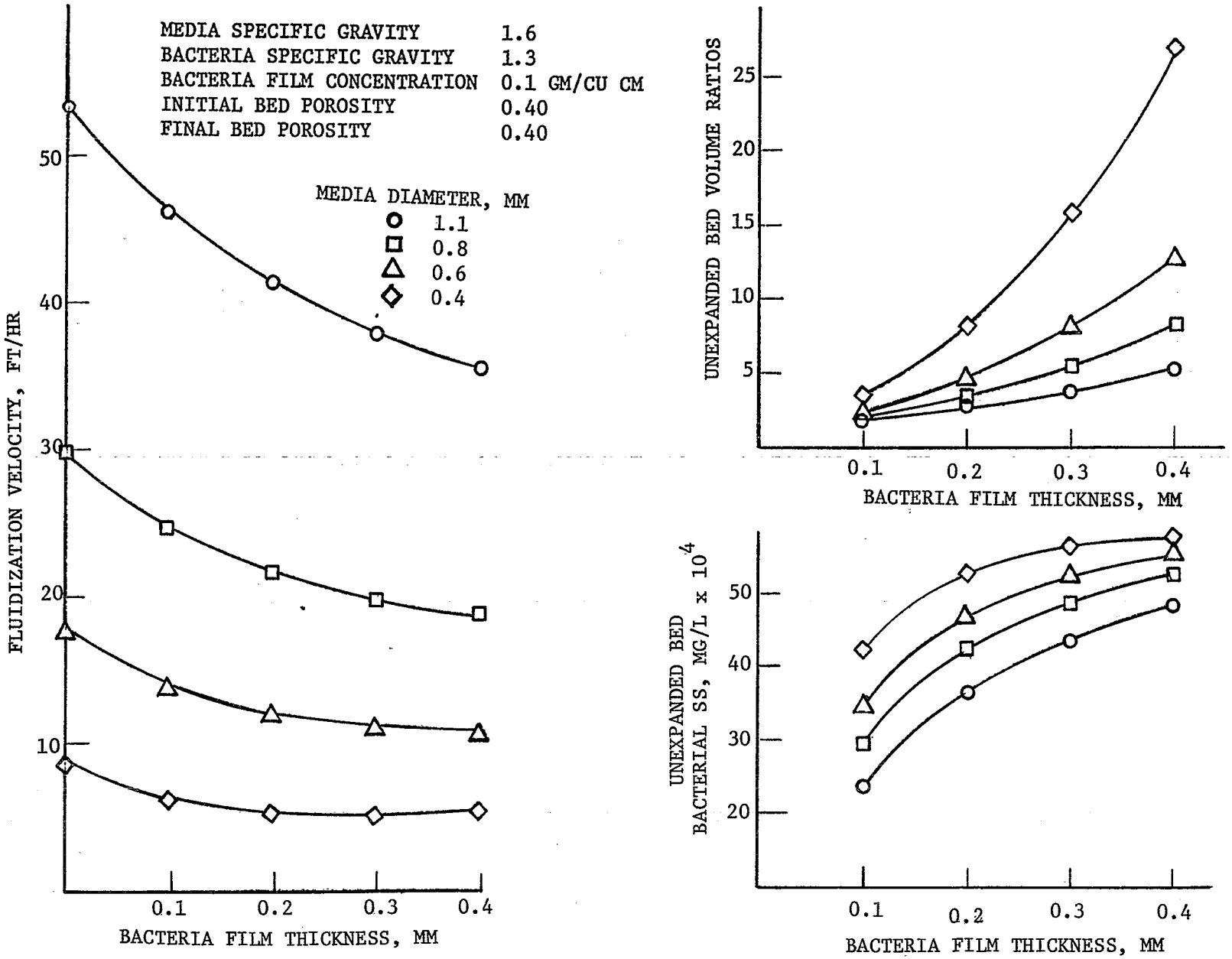

FIGURE 5. Influence of Bacteria Film Thickness on Fluidization Properties and Bed Characteristics

sq ft) for studies with the aluminum oxide media. Although the fluidized bed system will save energy compared to activated sludge because no oxygen is added to the reactor, it is clear that the energy savings could be negated through excessive pumping requirements.

For any specific set of design parameters, Figure 4 and Equation 4 can be used to estimate the pumping energy requirements for fluidization. As an example, consider a reactor containing 3.05 m (10 ft) of silica sand of specific gravity 2.65 at ε_{mf} of 0.40. The head loss through the bed is 3.02 m (9.9 ft). If the design called for a 2-hour HRT and no recycle pumping was contemplated, the particle sizes would need to be exceedingly small (< 0.2 mm) as shown by the curves in Figure 3. Assuming a wire to water pumping efficiency of 65 %, the energy expended to overcome the headloss through the bed (excluding the losses in the distribution system) would be

$$\frac{(3.02\ \text{m})\ (9.806\ \text{newton/kg})\ (1000\ \text{kg/cu m})}{(3600\ \text{sec/hr})\ (1000\ \text{watt/kw})\ (0.65_{eff})} = 0.0126\ \text{kwh/cu m}\ (47.8\ \text{kwh/MG})$$

If the proposed design called for using sand particles of approximately 1 mm size, the minimum fluidization velocity would increase to 30.5 m/hr (100 ft/hr) and providing a 2-hour HRT in the 3.05 m (10 ft) bed would require that the recycle:influent pumping ratio rise to greater than 19:1 to achieve more than minimum bed expansion. In this case, the pumping requirements at an overall efficiency of 65 percent would rise to 0.252 kwh/cu m (955 kwh/MG) of wastewater treated, excluding the additional losses in the distributor system. The distributor losses will vary with the type of distribution system and flow rates chosen, and will probably add an additional 0.3 to 1.2 m (1 to 4 ft) of head loss to the system.

It can be seen that the energy requirements for fluidized beds will be determined by the HRT required, the size and specific gravity of the media selected, and the extent to which the bacterial film characteristics alter the particles behavior. For the thin films observed by Switzenbaum and Jewell, silica sand particles of around 0.3 to 0.4 mm size should produce acceptable fluidization characteristics and bacterial concentrations (Table 2) and result in a headloss of 8.5 to 13.7 m (28 to 45 ft) for a 2-hour HRT. Pumping 3785 cu m/day (1 mgd) with a headloss of 15.2 m (50 ft) requires 242 kwh/day at an overall efficiency of 65 percent, so the headloss for such a system would be reasonable.

In actual practice it is not clear what flow control strategy would be optimal for plant operation. One approach is to incorporate a flow equalization basin ahead of the reactor to insure that it receives a relatively constant hydraulic loading and a more uniform organic loading. This approach will minimize the amount of recycle pumping required. Alternatively, it may be more desirable to pass the incoming flow directly through the system and vary the recycle ratio as required for adequate bed expansion. If the ratio of maximum flow to minimum flow and maximum organic concentration to minimum concentration synchronously varied by 3:1, the organic mass loading would vary by 9:1 during the day. The optimal combination of flow equalization, reactor size and recycle rate can only be calculated when considered in conjunction with the expansion characteristics of the media selected and the biological

kinetic response of the attached growth.

Biological sludge production in anaerobic systems is substantially less than in aerobic systems. McCarty (21) lists the following growth constants and endogenous respiration rates:

Waste	Growth Constant	Endogenous Respiration Rate $days^{-1}$
Fatty Acid	0.054	0.038
Carbohydrate	0.240	0.033
Protein	0.076	0.014

The combination of a low cell yield coupled with extremely long SRT's in the reactor will lead to a very low net sludge production. This net solids production may be low enough to obviate the need for final clarifiers and still meet secondary effluent standards. The data presented by Jewell suggest that the net sludge production is low enough so that the excess solids can be discharged in the effluent. If gas bubble formation and subsequent attachment to the particles tends to float media from the reactor, some stripping and final settling may be required. The settling velocity of a 0.4 mm particle of 1.1 specific gravity is approximately 21.3 m/hr (70 ft/hr) so settling these particles can be accomplished in clarifiers/settling tanks with high overflow rates. Larger or more dense particles will, of course, settle faster.

The expanded bed system investigated by Jewell was operated on primary effluent. Hence there would still be primary sludge to be processed and disposed of. A comparison of sludge quantities and thickening characteristics between a primary plant and a secondary activated sludge plant illustrates several potential advantages which may be realized by substituting an anaerobic reactor for an activated sludge system. Table 3 lists sludge quantities and volumes for municipal wastewater treatment plants containing primary clarifiers followed by activated sludge systems which receive incoming BOD and suspended solids concentrations of 200 mg/l each for Case No. 1 or concentrations of 250 mg/l of each for Case No. 2. In both cases the primary clarifier is assumed to provide 60 percent suspended solids removal and 35 percent BOD removal. The solids processing scenarios will depend upon the size of the plant and the options for solids disposal. If the net solids production in an anaerobic fluidized bed system with a high SRT was 0.075 g VSS/g BOD removed, then substitution of an anaerobic system for the activated sludge process in the two cases in Table 3 would result in effluent suspended solids of 12 or 15 mg/l. If the soluble effluent BOD was approximately 10 to 15 mg/l with an anaerobic system, it would be possible to achieve secondary effluent quality with no provision for solids capture from the anaerobic reactor.

Primary sludge can be gravity thickened to around 9 percent solids and is also easy to dewater with relatively low chemical conditioning dosages and high solids yields (30). For a small plant, lime conditioning (or stabilization if required) followed by vacuum filtration and landfilling may be the least cost sludge disposal option. Where anaerobic digestion

TABLE 3. SLUDGE QUANTITIES AND VOLUMES REQUIRING PROCESSING PER MILLION GALLONS TREATED IN A TYPICAL ACTIVATED SLUDGE PLANT WITH PRIMARY CLARIFICATION

	Case No. 1	Case No. 2
Influent BOD_5 and Suspended Solids, mg/1 of each	200	250
Primary Sludge, lb	1001	1251
Secondary Sludge, lb	636	837
Unthickened Primary Sludge Volume at 4% Solids, gal	3000	3750
Thickened Primary Sludge Volume at 9% Solids, gal	1334	1667
Unthickened Secondary Sludge Volume at 1% Solids, gal	7626	10036
Thickened Secondary Sludge Volume at 3% Solids, gal	2542	3345
Thickened Combined Sludge Volume at 5.5%, gal	3569	4552
Primary Sludge Volatile Solids, lb	651	813
Activated Sludge Volatile Solids, lb	477	628

Design Assumptions:

Primary Clarifier Solids Removal 60%
Primary Clarifier Sludge 65% Volatile Solids
Primary Clarifier BOD Removal 35%
Cell Yield 0.75 lb. VSS/lb BOD_5 Removed
Cell Decay 0.07 days $^{-1}$
Soluble Effluent BOD_5 3 mg/l
SRT 5 days
Effluent Suspended Solids 15 mg/l
Effluent Solids are 75% Volatile

is to be used for sludge stabilization, Table 3 illustrates that the volume occupied by thickened primary sludge alone is substantially less than achieved by gravity thickening of primary and activated sludges. The small sludge volumes require smaller anaerobic digesters. Of course, in any real design situation the advantages and cost of flotation thickening of secondary sludge should also be considered.

The oxygen required for the activated sludge systems in Table 3 operated at a 5-day SRT should be nearly the same as the influent BOD_5's to the aeration basin, i.e., 130 and 163 mg/l for the two cases shown. If air is input with 8 psig adiabatic compression, 70 percent efficiency of compressor and motor, and an aeration device with a 7 percent oxygen transfer efficiency, the power requirements are 0.124 and 0.156 kwh/cu m (471 and 590 kwh/MG). A mechanical aerator with an oxygen transfer efficiency of 1.1 kg O_2/kwh (1.8 lb O_2/Hp·hr) would require 449 kwh or 563 kwh for the higher oxygen demand. When these values are compared to a fluidized bed reactor it provides a rough measure of the pumping energy which can be expended in the anaerobic system and still be competitive with activated sludge on the basis of energy criteria.

If the activated sludge plants summarized in Table 3 employed anaerobic digestion for the stabilization of the primary and secondary sludges, the primary sludge volatile solids would comprise about 57 percent of the total volatile solids loading to the digester. In this hypothetical example roughly 50 percent of the influent degradable carbon to the activated sludge system would be oxidized and the remaining carbon removed would be transformed into biological solids; some would escape in the effluent. It is this remaining transformed organic matter that is available for CH_4 production in the anaerobic sludge digestion process. In contrast, if an anaerobic fluidized bed were substituted for an activated sludge system the BOD removal in a system operated with no sludge wasting or effluent solids capture would result from CH_4 formation entirely. For the two wastewaters characterized in Table 3, there would be 130 or 163 mg/l of BOD_5 available for CH_4 formation if an anaerobic fluidized bed were used in place of activated sludge. The amount of CH_4 formed will be determined by the efficiency of waste utilization and the net biological sludge production.

An interesting aspect of CH_4 formation with an anaerobic reactor is shown by the solubility data in Table 4 (31).

TABLE 4. SOLUBILITY OF CARBON DIOXIDE AND METHANE GASES

Temperature °C	Solubility*, mg/l		Solubility as C, mg/l	
	CH_4	CO_2	CH_4 -C	CO_2-C
10	29.6	2318	22.2	632
15	26.0	1970	19.5	537
20	23.2	1688	17.4	460
25	20.9	1449	15.7	395
30	19.0	1257	14.3	343

*When the pressure of the gas plus that of the water vapor is 760 mm Hg

In contrast to an anaerobic sludge digester where the high sludge feed concentrations make the amount of CH_4 exiting in solution negligible in comparison to the amount which is recovered in the overlying gas phase, the amount of CH_4 which leaves the reactor in a dissolved phase from an anaerobic fluidized bed reactor can represent a substantial part of the CH_4 formed.

Methane production from the anaerobic decomposition of any organic compound can be accurately predicted by a number of techniques. Symons (32) developed the following equation:

$$C_nH_aO_b + (n - a/4 - b/2)H_2O \longrightarrow (n/2 - a/8 + b/4)CO_2 + (n/2 + a/8 - b/4)CH_4$$

Equal proportions of methane and CO_2 result from the decomposition of carbohydrates and also from acetic acid. Proteins, fats and long chain acids will yield gas compositions higher in CH_4 than CO_2. Typical municipal wastewaters have total organic carbon concentrations (TOC's) in the primary effluent of 80 to 180 mg/1. If 85 percent of this TOC were converted to CO_2 and CH_4 in an anaerobic system in the ratio of 40:60, the carbon in the methane produced would range from 41 to 92 mg C/1. A comparison of these values with the methane solubility data in Table 4 shows that in all cases the quantity of methane produced which exits as dissolved methane gas must be considered in any design situation where recovery of the methane from the gaseous space overlying the reactor will be practiced. These data show that the amount of CH_4 which remains dissolved in the liquid phase can be a significant fraction of the total CH_4 production. Of course, the partial pressure of the methane in the gaseous phase will affect the equilibrium solubility concentration. Whether the dissolved CH_4 concentration will tend toward the equilibrium concentration dictated by the overlying partial pressure, remain near the saturation concentrations shown in Table 4, or be somewhat supersaturated will be influenced by the reactor design, the hydraulic residence time, and the degree of gas transfer across the gas-liquid interface. In contrast to an anaerobic sludge digester with an overlying atmosphere of 25 to 35 percent CO_2, the CO_2 overlying an anaerobic fluidized bed reactor will be much less. Assuming influent TOC of 80 to 180 mg/1, the CO_2 production would be 27 to 61 mg/1 as C. When these values are compared to the solubility limits in Table 4 it is clear that the equilibrium partial pressure of CO_2 will be quite small. The actual values will depend upon wastewater pH and mass transfer across the gas-liquid interface, but should be less than 10 percent of the off-gas volume. Also the N_2 concentration in the overlying gas volume could be 5 to 15 percent of the total gas volume because of evolution of the nitrogen gas initially dissolved in the wastewater.

Another consideration in estimating methane production is the sulfate concentration of the wastewater. The sulfate concentration in natural waters varies widely as shown by the values in Table 5 (33). In anaerobic systems the sulfate can serve as a terminal electron acceptor in biologically mediated reactions. This can be represented by the following half reaction (34):

$$\frac{1}{8} SO_4^{-2} + \frac{19}{16} H^+ + e^- \longrightarrow \frac{1}{16} H_2S + \frac{1}{16} HS^- + \frac{1}{2} H_2O$$

TABLE 5. SULFATE CONCENTRATIONS AT SELECTED LOCATIONS IN THE UNITED STATES (33)

Location	SO_4^{-2} Range, mg/l	Sampling Period
Connecticut River at Thompsonville, CT	11 - 16	1/66 - 9/66
Hudson River at Poughkeepsi, NY	20 - 28	10/65 - 9/66
Neuse River at Goldsboro, NC	5 - 11	10/65 - 9/66
Sacramento River at Freeport, CA	6 - 15	4/66 - 9/66
Colorado River near Grand Canyon, AZ	123 - 246	10/65 - 9/66
Ohio River near Huntington, WV	100 - 207	10/65 - 12/65

According to Bryant (35), it is well known that methanogenesis in natural ecosystems does not occur when sulfate is present. Conversion of acetate to CO_2 with sulfate reduction to sulfide is thermodynamically more favorable than acetate conversion to CO_2 and CH_4. With wastewaters containing influent COD's of 200 to 250 mg/l and SO_4^{-2} concentrations of 200 mg/l (133 mg/l as O_2), the majority of the organic material could be oxidized through sulfate reduction with a corresponding decrease in methane formation. Hydrogen sulfide gas is extremely soluble in water (3850 mg/l at 20°C), whereas most heavy metals form insoluble sulfides. The partioning of the H_2S gas between the liquid and overlying gas phase will depend on the distribution of sulfur species and the degree to which the equilibrium conditions predicted by Henry's law are approached.

Process Economics

Bell et al. (36) made a preliminary design for an anaerobic fluidized bed system treating heat treatment liquor. The reactors were arranged into four modules with each module consisting of three reactors in series. Reactor volume was 4106 cu m (145,000 cu ft) with 464 sq m (5000 sq ft) of surface area. System components included individual recycle pumps for each reactor and controlled gas release. The installed cost, first quarter 1980, was estimated to be $3,436,000. This corresponds to $837. per cu m ($23.70 per cu ft) of reactor volume.

Anaerobic fluidized bed treatment was considered for treatment of vertical tube reactor (VTR) effluent in the facility plan for Montrose, Colorado (37). In the proposed design the VTR was used in place of a primary clarifier. The design flow was 10,900 cu m/day (2.88 mgd). Four anaerobic fluidized bed units, each with dimensions of 7.0 m x 7.0 m x 7.6 m deep with 0.61 m additional freeboard (23 ft x 23 ft x 25 ft + 2 ft freeboard) were planned to provide a 2-hour HRT at design flow. Reinforced concrete covered tanks with common wall construction were envisioned. The construction cost estimate for the anaerobic fluidized bed and recycle pumps was $946,000 (ENR = 3350). Ignoring the 0.61 m (2 ft) of freeboard, this amounts to $631. per cu m ($17.88 per cu ft) of reactor volume or $86.66 per cu m ($328,000 per mgd) of design flow capacity.

An anaerobic fluidized bed system is also under consideration for the town of Hanover, NH (38). JI Associates' initial estimate of the construction cost is approximately $706. per cu m ($20. per cu ft) of reactor volume (March, 1981). This includes pumps and all controls thought to be necessary for the facility. Thus the estimated construction cost for the 8700 cu m/day (2.3 mgd) design would be $934,000 or slightly more than $105. per cu m ($400,000 per mgd) of design flow capacity to provide an average 3.6 hour HRT.

Pumping cost per 3785 cu m (1 MG) of flow per 0.305 m (1 ft) head loss is 24.2¢ at 5¢ per kwh and 65% overall wire to water efficiency. Hence a system with a head loss of 3.66 m (12 ft) per pass through the bed and an overall recycle:influent ratio of 2:1 would represent a power cost of 0.23¢ per cu m ($8.71 per MG) treated, or an annual cost of 84.0¢ per cu m/day ($3180 per mgd) of design capacity.

Summary

The results reported by Jewell (1,2) and Switzenbaum and Jewell (27) have demonstrated that better than secondary effluent quality can be obtained from a laboratory anaerobic expanded bed reactor treating primary effluent at 20°C. The process was also shown to provide good COD removal with a glucose feed when the temperature was 10°C and the HRT was four hours or greater. Since wastewater temperatures in much of the United States fall to 8 to 12°C during wintertime operation, the response at lower temperatures is quite important. Previous studies by O'Rourke (39) with homogenized primary sludge established that methane fermentation was drastically reduced at 15°C and that efficient digestion could not be accomplished even at a 60-day retention time. The lipid fraction of the waste was not utilized. However there was a measurable reduction in the total COD due to the methane fermentation of formic and acetic acids resulting from cellulose and protein degradation. Whether or not anaerobic treatment of municipal wastewaters at low temperature is economically attractive has yet to be demonstrated.

Because of the limited data available, the long time required for such systems to come to equilibrium, and the scale of the studies reported, there are a number of questions related to anaerobic fluidized bed technology which remain to be answered before the design approach can be optimized. These include:reaction kinetics as a function of temperature and reactor response under dynamic loading; optimal reactor depth, media density and

size; need for equalization basins and an overall flow control strategy for adequate bed expansion; net solids production; solids levels attainable in the reactor; biological film properties; effect of biological growth on media expansion characteristics; solids control strategies in the reactor if any; need for final clarifiers; influence of wastewater sulfate concentration on the desirability and performance of the process; and long term process stability and reliability at pilot scale. The process is presently considered to be eligible for funding as innovative technology on a case by case basis where all relevant factors affecting process performance have been carefully considered.

ANFLOW

Process Theory

In contrast to fluidized bed systems where extensive surface area is provided for biological attachment (0.5 mm particles provide about 92.9 sq m (1000 sq ft) of surface area per 0.028 cu m (1 cu ft) of bed), the 2.5 cm (1 in) Raschig ring packing investigated in the ANFLOW system provides about 5.39 sq m (58 sq ft) of surface area per 0.028 cu m (1 cu ft) (40). Upflow velocities investigated in the ANFLOW system were normally in the range of 4.07 - 12.2 m/day (100 - 300 gpd/sq ft) or 0.17 - 0.51 m/hr (0.56 - 1.67 ft/hr). Hence the fluid velocities and the requirements for biological attachment and growth are substantially different in ANFLOW and fluidized bed systems. The ANFLOW system relies on sludge settling and solids entrapment to provide a substantial portion of the BOD removal. Another difference between the two systems is that ANFLOW treats raw rather than settled wastewater.

Process Capabilities

Oak Ridge National Laboratory undertook a two-year pilot plant investigation of an ANFLOW system. The cylindrical reactor had a diameter of 1.52 m (5 ft) and an overall height of 5.58 m (18.3 ft). It contained 3.05 m (10 ft) of 2.5 cm (1 in) unglazed ceramic Raschig ring packing. The bottom of the cylindrical column was a 45 degree cone with a flanged outlet at the bottom. A schematic diagram of the process was previously shown in Figure 2. The column was seeded with a mixture of rumen fluid and anaerobically digested sludge. Feed for the ANFLOW unit was taken from the headworks of the Oak Ridge East Sewage Treatment Plant immediately downstream of a comminutor. The unsettled wastewater was fed directly to the ANFLOW column at a constant flow rate which was periodically varied as desired. An overflow weir and a collection trough in the top of the column were used to collect reactor effluent. The total off-gas volume was measured by a wet test meter.

Results from the pilot plant operation have been presented in papers at three conferences (7,8,9). The data presentations consist primarily of average monthly values provided in histogram form for selected parameters or other condensations of process data. None of the papers presents day-to-day parameter values so that the degree of variation can be quantified.

A summary of ANFLOW reactor performance reported by Genung et al. (7) is shown in Table 6. It can be seen that the ANFLOW reactor does not produce an effluent of acceptable quality for discharge as secondary effluent. The ANFLOW reactor produced BOD and TSS removals which averaged 53 percent and 69 percent respectively, for months 3 through 21. For months 3 through 15, TSS removals averaged 76 percent. Genung et al. (7) drained the bioreactor (at some point near the end of the study) and washed with wastewater fed at 30.3 cu m/day (8000 gpd or 407 gpd/sq ft) for 24 hours to test the feasibility of periodically removing solids. TSS removal rates were then reevaluated at 3.78, 18.9 and 26.5 cu m/day (1000, 5000 and 7000 gpd); in all cases the TSS removals were about 75 percent.

Based on the pilot plant data, an ANFLOW reactor treating municipal wastewater would be expected to produce effluent qualities as shown in Table 7. Since the ANFLOW reactor does not produce an effluent of acceptable secondary quality, further treatment will be required.

TABLE 7. EXPECTED EFFLUENT QUALITIES FROM AN ANFLOW REACTOR

BOD and TSS, mg/l Influent	Effluent BOD, mg/l		Effluent TSS, mg/l	
	50 % Removal	60 % Removal	70 % Removal	80 % Removal
200/200	100	80	60	40
250/250	125	100	75	50

Energy Considerations

Since the ANFLOW process is being advocated for energy conservation and methane production, which is stated to represent a significant and recoverable energy source (7), it is appropriate to consider the information on gas generation and recovery potential.

The summary data persented by Koon et al. (8) on average influent and effluent BOD and TOC values lead to the ratios shown in Table 8.

TABLE 8. AVERAGE TOC:BOD RATIOS MEASURED BY KOON ET AL. (8)

Operating Period, days	TOC:BOD Ratio	
	Influent	Effluent
50	0.473	0.381
72	0.725	0.680
62	0.445	0.341
95	0.666	0.873
36	0.766	0.689
33	0.985	0.867

The overall time-weighed TOC:BOD ratios for influent and effluent are 0.651 and 0.648, respectively.

TABLE 6.

Summary of ANFLOW Reactor Performance as Reported by Genung et al (7)

Month	TSS, mg/l		BOD, mg/l		% Removals		Hydraulic Loading	
	Feed	Effluent	Feed	Effluent	BOD	TSS	m/day	gpm,/ft^2
3	63	22	60	32	47	65	2.17	.037
4	100	28	110	43	61	72	2.17	.037
5	210	48	165	76	54	77	2.17	.037
6	110	36	100	50	50	67	4.16	.071
7	156	28	180	95	47	82	4.16	.071
8	165	28	181	92	49	83	4.16	.071
9	192	43	170	63	63	78	4.16	.071
10	110	24	126	49	61	78	6.22	.106
11	102	26	110	51	54	75	8.27	.141
12	67	17	97	48	51	75	8.27	.141
13	120	26	82	43	48	75	8.27	.141
14	120	24	160	62	61	78	3.40	.058
15	107	29	125	73	42	80	12.4	.212
16	140	66	150	69	54	53	12.4	.212
17	250	87	119	50	58	65	12.4	.212
18	160	104	212	92	57	35	12.4	.212
19	134	68	105	60	43	49	14.5	.248
20	213	55	150	60	60	74	13.5	.230
21	132	55	151	82	46	58	12.4	.212

Several figures in the paper by Genung et al. (7) provided information on average monthly parameters for flow, temperature, gas volume and percent CH_4, and influent and effluent BOD. No monthly summary of TOC data was provided. As shown in Table 9, the amount of off-gas recovered from the top of the reactor was quite small. In fact, if the factor of 0.651 is used to estimate the influent TOC (this results in some error for any given month as shown by the above TOC:BOD variation but does not affect the overall conclusion for the aggregate data), it is clear that only 1 to 2 percent of the influent TOC was recovered in the gaseous CH_4 phase when the reactor was operated near its design loading of 18.9 cu m/day (5000 gpd or 255 gpd/sq ft). The unweighed average fraction of TOC recovered as methane in the overlying gas phase for months 10 through 21 was 1.3 percent.

No information was provided on the total amount of methane generated in the ANFLOW reactor. As shown in Table 10, CH_4 solubility (when the partial pressure of CH_4 and H_2O vapor are 760 mm Hg) varied from about 21 to 29 mg/l over the range of effluent temperatures which were encountered.

Since the pressure range of operation of a typical wet test gas meter is 0.76 to 15.2 cm (0.3 to 6 in) of water (41), and a hydraulic head of 5.1 to 10.2 cm (2 to 4 in) of water was adequate to produce flow through the reactor (7), the pressure overlying the liquid was quite close to atmospheric. The concentration of dissolved CH_4 for the case where the CH_4 in solution is in equilibrium with the overlying gaseous methane (i.e., net methane flux is zero) is also shown in Table 10. The concentration of methane exiting the reactor as dissolved gas should have been between these two limits.

The data in Table 10 were used to develop the estimates of CH_4 production shown in Table 11. The MAX and MIN values refer to the upper and lower limits anticipated for the methane exiting in solution. Again, it is noted that the values for percent carbon removal or influent and effluent TOC values are only approximations for any given month because no monthly TOC data were provided and the values were estimated from the BOD data assuming a constant TOC:BOD ratio. As shown in Table 10, most of the methane produced exits the reactor dissolved in the liquid phase. For months 6 through 21 the amount of carbon removed in the ANFLOW reactor that was converted to methane has averaged between 25 percent (MIN) and 45 percent (MAX). For the same period the amount of carbon that entered the reactor and was converted to methane was between 13 percent (MIN) and 23 percent (MAX). The average BOD removal for months 5 through 21 was 53 percent.

The following statements are made by Genung et al. (7). " The methane produced was approximately 33 percent of that which could theoretically have been produced as calculated from measurements of the organic carbon removed from the wastewater by processes in the bioreactor. This efficiency was difficult to estimate, however, since carbon was removed by many mechanisms, some involving solubilization phenomena, for instance, which occurred over undefined periods." The exact meaning of these statements is not clear, but the figure of 33 percent may correspond to the 25 and 45 percent minimum amd maximum concentrations estimated above.

Based upon the above calculations it is clear that any significant

TABLE 9. Methane Recovered as Off-Gas from the ANFLOW Reactor

Month No.	Influent Flow gpd	Influent Flow eu m/day	Gas Characteristics Composition % CH_4	Gas Characteristics Production l/day	Gas Characteristics Carbon g/day	Influent BOD mg/l	Influent Carbon* g/day	Fraction of Influent C Recovered in CH_4 Gas %
1	1050	3.97						
2	1050	3.97						
3	1050	3.97						
4	1050	3.97						
5	1050	3.97	15	17	1.4	165	427	.33
6	2000	7.57	36	16	3.1	100	493	.63
7	2000	7.57	69	40	14.8	180	887	1.7
8	2000	7.57	72	94	36.2	181	892	4.1
9	2000	7.57	66	80	28.3	170	838	3.4
10	3000	11.3	56	111	33.3	126	931	3.6
11	4000	15.1	53	95	27.0	110	1084	2.5
12	4000	15.1	48	57	14.6	97	956	1.5
13	4000	15.1	43	46	10.6	82	808	1.3
14	1650	6.24	40	40	8.6	160	651	1.3
15	6000	22.7	36	37	7.1	125	1848	.38
16	6000	22.7	43	19	4.4	150	2218	.20
17	6000	22.7	56	30	9.0	119	1759	.51
18	6000	22.7	55	27	8.0	212	3134	.26
19	7000	26.5	61	110	35.9	105	1811	2.0
20	6500	24.6	63	86	29.0	150	2402	1.2
21	6000	22.7	82	38	16.7	151	2232	.75

* Assumes TOC = BOD x .651

TABLE 10. Dissolved Methane Concentrations Expected with the ANFLOW Reactor

Month No	Effluent Temperature °C	CH_4 Solubility* mg/l	CH_4 Solubility* mg/l as C	CH_4 Solubility at Equilibrium with Overlying Gas mg/l	CH_4 Solubility at Equilibrium with Overlying Gas mg/l as C	Influent TOC** mg/l	Effluent TOC*** mg/l
1							
2							
3							
4							
5	20	23.2	17.4	3.5	2.6	107	49.2
6	22	22.2	16.7	8.0	6.0	65	32.4
7	24	21.3	16.0	14.7	11.0	117	61.6
8	24.5	21.1	15.8	15.2	11.4	118	59.6
9	24	21.3	16.0	14.1	10.6	111	40.8
10	21.5	22.4	16.8	12.5	9.4	82	31.8
11	17	24.8	18.6	13.1	9.8	72	33.0
12	14	26.7	20.0	12.8	9.6	63	31.1
13	11.5	28.4	21.3	12.2	9.2	53	27.9
14	11	28.8	21.6	11.5	8.6	104	40.2
15	13	27.3	20.5	9.8	7.4	81	47.3
16	16	25.4	19.1	10.9	8.2	98	44.7
17	17	24.8	18.6	13.9	10.4	77	32.4
18	20	23.2	17.4	12.8	9.6	138	59.6
19	23	21.8	16.4	13.3	10.0	68	38.9
20	23	21.8	16.4	13.7	10.3	98	38.9
21	23.5	21.5	16.1	17.6	13.2	98	53.1

* Solubility when CH_4 and water vapor pressure are 760 mm Hg

** Assumes TOC = BOD x .651

*** Assumes TOC = BOD x .648

TABLE 11. Estimated Methane Production in the ANFLOW Reactor

Month No.	TOC* Removed mg/l	CH_4 Recovered in Off-Gas mg C per liter of Influent Flow	Total CH_4 Produced mg C per liter of Influent Flow		Carbon in CH_4 as % of C Removed		Carbon in CH_4 as % of C Entering Reactor	
			MAX	MIN	MAX	MIN	MAX	MIN
5	58	.35	17.8	3.0	31	5.2	17	2.8
6	33	.41	17.1	6.4	52	19	26	10
7	55	2.0	18.0	13.0	33	24	15	11
8	58	4.8	20.6	16.2	36	28	17	14
9	70	3.7	19.7	14.3	28	20	18	13
10	50	2.9	19.7	12.3	39	25	24	15
11	39	1.8	20.4	11.6	52	30	28	16
12	32	.96	21.0	10.6	66	33	33	17
13	25	.70	22.0	9.9	88	40	42	19
14	64	1.4	23.0	10.0	36	16	22	9.6
15	34	.31	20.8	7.7	61	23	26	9.5
16	53	.19	19.3	8.4	36	16	20	8.6
17	45	.40	19.0	10.8	42	24	25	14
18	78	.35	17.8	10.0	23	13	13	7.2
19	29	1.4	17.8	11.4	61	39	26	17
20	59	1.2	17.6	11.5	30	19	18	12
21	45	.74	16.8	13.9	37	31	17	14

* Computed per Table 10

recovery of the methane gas produced in the ANFLOW pilot reactor would have involved recovery from the liquid phase presumably through vacuum degasification. This is in contrast to conventional sanitary engineering design for anaerobic sludge digestion. With sludge digestion, the incoming waste stream is thickened to organic carbon concentrations more than two orders of magnitude greater than those entering an ANFLOW reactor and as a consequence the methane exiting in solution represents less than one percent of that generated.

Total methane gas production which represents 13 to 23 percent of the total TOC which entered a plant as in the ANFLOW pilot reactor, is no more than produced in a conventional plant with anaerobic digestion of primary sludge only.

Design and Economic Considerations

Griffith (42) developed cost information for a conceptual ANFLOW reactor design based on a hydraulic loading of 10.4 m/day (255 gpd/sq ft). The estimated capital costs for various plant sizes and media costs were approximately as shown in Table 12.

TABLE 12. CAPITAL COSTS FOR AN ANFLOW REACTOR REPORTED BY GRIFFITH (42)

Design Flow		Capital Cost, Millions of Dollars		
		Media Cost, $/ cu ft		
cu m/day	MGD	5	10	15
1135	0.3	0.18	0.26	0.33
3785	1.0	0.59	0.85	1.1
18925	5.0	2.8	4.0	5.1

Based on a media cost of $353./cu m ($10./cu ft), the packing comprised 50 percent of the capital cost for a 3785 cu m (1 MGD) ANFLOW reactor. The capital costs are based on an ENR index of 2700. The capital cost given by Griffith can be estimated by:

$$COST = e^{[.973\ \ln(FLOW)]} \times e^{b}$$

where cost is in millions of dollars, flow is in MGD and b = -0.54034, -0.17735 and .06299 for packing costs of 5, 10 and 15 dollars/cu ft, respectively.

The capital cost for a 3785 cu m (1 MGD) ANFLOW treatment system based on packing at $353./cu m ($10./cu ft) and an ENR of 2860 was estimated by Koon et al. (8) at $2,981,250. Based on the information in Table 4 of Genung et al. (7), this estimate includes 50 percent extra for related costs plus a $100,000 base cost for surge and reaeration tanks. Hence, the base cost of the ANFLOW reactor was estimated at $1,887,500. The conflicting information on reactor design criteria makes it impossible to know the basis of the design with certainty, but in all probability the cost refers to a reactor receiving an influent flow rate of 5.87 m/day (0.1 gpm/sq ft or

144 gpd/sq ft). In this case the equivalent cost estimate by Griffith (1.77 MGD and ENR of 2860) would be $1,546,000 assuming the 50 percent cost increase to cover all installed costs.

The ANFLOW system process flow diagram (8) envisioned for a complete wastewater plant includes the following major components:

1. Bar Screen
2. Grit Chamber
3. Comminutor
4. Equalization Basin
5. Grinder Pumps following the Equalization Basin
6. ANFLOW Column
7. Surge Tank (for sludges and backwash water)
8. Reaeration and Chemical Addition Tank for ANFLOW Column Effluent
9. Upflow Sand Filters following Chemical Addition
10. Chlorine Contact Basin

Hence, any analysis of the design and cost effectiveness of an ANFLOW system entails more than an economic analysis for just an ANFLOW reactor.

It is desirable to present and discuss, where appropriate, several observations concerning published information on ANFLOW system design and economics. Specific points worth noting in the paper by Koon et al. (8) are as follows:

1. This paper discusses conceptual designs to treat a raw wastewater with a BOD and TOC of 300 mg/1 and VSS of 275 mg/1 with TSS of 350 mg/1. Plant sizes of 0.05 and 1.0 MGD are discussed.

2. The average hydraulic design loading rate is stated in Table 3 to be 0.1 gpm/sq ft with a peak loading rate of 0.15 gpm/sq ft. In Table 4, the ANFLOW filter is stated to have a detention time of 18 hours and 14,120 sq ft of surface area for a 1 MGD flow. The total volume is given as 141,200 cu ft which corresponds to the 10 ft depth specified on Page 14. However, 0.1 gpm/sq ft and a 10 ft depth correspond to a detention time of only 12.5 hours. A flow of 0.1 gpm and 14,120 sq ft of surface area corresponds to an average flow of 2 MGD for the 1 MGD plant. Furthermore, it is stated on Page 14 that the design was based on a hydraulic loading rate of 0.15 gpm/sq ft which corresponds to a flow of 3 MGD for the 1 MGD plant. In addition to the ANFLOW reactor, the design calls for an equalization basin and aeration system which adds an additional 8 hours of detention time based on the influent flow (i.e., a 1 MGD flow to a 1 MGD plant). Additional facilities include a grit chamber, upflow sand filter, and chlorine contact tank.

3. It was felt that both comminution and subsequent grinding of the influent to produce particle sizes less than 0.5 mm would be required to insure that the anaerobic filter did not prematurely clog.

4. The ANFLOW reactor was anticipated to have a soluble effluent BOD of

25 mg/1 and effluent TSS of 35 mg/1. The design effluent BOD from the entire treatment system (ANFLOW reactor, aeration and filtration) was to be 30 mg/1. A design effluent TSS of 35 mg/1 from the ANFLOW reactor corresponds to 90 percent TSS removal. If the effluent TSS are assumed to be 75 percent volatile, this would represent a solid BOD of 20 mg/1 based on the stated insoluble BOD/VSS ratio of 0.76. Hence, the total effluent design BOD from the ANFLOW reactor would be roughly 45 mg/1 which corresponds to a BOD removal of 85 percent. Table 6 summarizes the information on BOD and solids removal obtained from 21 months operation of the pilot plant reactor; these data were estimated from the figures in Reference 7. As shown in Table 6, at no time during their pilot plant studies did they observe solids removal of 90 percent or BOD removal of 85 percent. Even after the reactor was drained and washed to remove excess solids (Figure 9, Reference 7), the solids removals after restarting were only 75 percent. Hence, the entire economic analysis is based on an ANFLOW reactor which is assumed to perform significantly better than observed during the pilot plant operation.

5. Energy available from the gas generation for the 1 MGD plant was stated to be 225 hp based on an off-gas volume of 35,820 cu ft/day. How these figures were determined is not given. In the subsequent section on system costs (Page 19 of Reference 8) it was stated that a 40 percent conversion efficiency was used to determine the equivalent cost of recovered power. It should be noted that this is a higher efficiency than obtained in the average fossil fuel electric generation plant which is 33 percent (43).

   The information on power generation can be used to estimate the assumed efficiency of TOC conversion to CH_4 and CO_2. The heat of combustion of methane gas at 25°C to $H_2O_{(g)}$ and $CO_{2(g)}$ is 21,502 BTU/lb(44). At STP, 35,820 cu ft of CH_4 will weigh 1595.95 lb and is equivalent to 224.8 hp at a 40 percent conversion efficiency. This presumably is the basis for their anticipated energy recovery and apparently the figure given for the off-gas refers to the methane only. This conclusion is based on the power recovery of 225 hp which is specified in Table 5 (in Reference 8).

   The wastewater characteristics used for the analysis were 300 mg/1 of TOC or 2500 lb of TOC/MG. Of this, 1197 lb of C is apparently presumed to show up in the off-gas as CH_4. Based on roughly 20 mg/1 of CH_4 lost in the effluent (31) an additional 125 lb of C will exit in the water as dissolved CH_4.

   The relationship for power generation in hp is stated to be 0.98 · Q · TOC with Q in mgd and TOC in mg/1. If all incoming C were converted to CH_4 only, the factor used to multiply the product of Q and TOC removed to obtain horsepower is

$$8.34 \cdot \frac{16}{12} \cdot \frac{21502 \cdot 778.1 \cdot 0.4}{550 \cdot 60 \cdot 1440} = 1.566$$

Thus $hp = 1.566 \cdot Q \cdot TOC_{removed}$

Since 0.98/1.566 = 0.626, they have apparently assumed a 62.6 % conversion of TOC to CH_4 which is not unreasonable.

However, 1322 lb of CH_4 as C plus the CO_2 expressed as C represents 1816 lb of C converted to CO_2 and CH_4 per MG treated. If the carbon removal in the ANFLOW reactor is approximated by BOD removal (85%) and ignoring the BOD recycled from the filter backwash, then 85.5% of the removed TOC was apparently assumed to be converted to CO_2 and CH_4. It is clear that this amount of CH_4 generation far exceeds anything observed in their pilot plant studies.

6. The solids yield coefficient for the ANFLOW system was given in Table 3 of Reference 8 as 0.2 g TSS/g BOD removed. If the 255 mg/l of BOD is removed in the ANFLOW reactor, this would correspond to a net solids production of 425 lb/day for a 1 mgd flow. If an additional 20 mg/l of TSS are removed in the filters and assuming an additional 10 mg/l production of TSS across the filters due to BOD removal (since their design assumes BOD removal), the solids returned to the surge tank in the backwash water would be an additional 250 lb for a 1 mgd flow. However, Tables 3 and 4 also indicate that 10,000 gpd are to be drained from the ANFLOW column at 2% solids for a stated solids accumulation in the surge tank of 1650 lb/day. This presumably means that none of the solids in the backwash water are assumed to settle in the surge tank but that they are all recycled to the ANFLOW reactor where they are all removed.

   Of the 1650 lb/day of sludge to be produced, it appears that the design calls for the alum and polymer addition to produce 975 lb of chemical sludge/day for the 1 mgd plant. If one assumes this sludge is all $AlPO_4$ (influent P of $\sim$ 30 mg/l) the corresponding alum dosage is about 1600 lb/day for a 1 mgd flow. Bagged commercial grade aluminum sulfate is currently selling for $146 - $154/ton (45).

   As an alternate calculation, if 255 mg/l of TOC are removed in ANFLOW and 14.5% of this is not converted to CO_2 and CH_4, the C accumulation is 308 lb/day for 1 mgd. If the organic material is 50% C and the total solids accumulation is 70% volatile (this assumption was not made in the paper being reviewed), then the total solids accumulation ignoring backwash would be 880 lb/day.

7. The effluent from the ANFLOW reactor will receive 3.2 hours of aeration followed by alum and polymer addition and upflow filtration. The sand filter is either 6 ft deep (Table 3 in Reference 8) or 5 ft deep (Page 17 in Reference 8). No data are given on anticipated alum or polymer dosages.

8. The design air flow rate of 362 cfm to the 1 mgd aeration system will provide 9020 lb of oxygen at 20°C. Providing 8 mg/l of D.O. for 1 mgd requires 66.7 lb of O_2. At 20°C and a 5% transfer efficiency,

the proposed air flow rate would dissolve 451 lb of O_2 per MG.

The stoichiometric combustion of 1596 lb of CH_4 to CO_2 and H_2O requires 6384 lb of O_2. If an additional 167 lb of CH_4 is stripped in the aeration process, the stoichiometric oxygen requirements rise to 7052 lb. The volume occupied by 167 lb of CH_4 at 20°C is 4023 cu ft or 0.77% of the air volume supplied to the 1 mgd aeration basin. The CH_4 in the contained off-gases from the aeration unit will not be present in explosive concentrations.

The cost analysis was said to include covering of the equalization basin for off-gas containment, but no information was given concerning the treatment/disposal of these gases. However, it is clear that the off-gas volume is compatible with that required for off-gas combustion (based on their assumed CH_4 generation) so this is not a problem. In fact, this may have been the basis for choosing the 362 cfm air flow rate.

9. According to Table 4 (in Reference 8), 15 hp (11.2 kw) will be required to operate the aeration basin in the 1 mgd plant. However, in Table 10 (in Reference 8) where the power requirements for ANFLOW are compared to activated sludge, the complete effluent polishing step (aeration plus filtration) is stated to require only 6.66 kw.

10. The inconsistency in stated power requirements and other information makes it difficult to estimate backwash frequency for the filters. The design calls for the removal of 20 mg/l of TSS in the filters plus an assumed soluble BOD reduction of at least 5 mg/l due to bacterial growth. Table 3 (in Reference 8) lists the maximum filter headloss and gives a value of 0.05 for specific deposit defined as lb SS/sq ft/ft headloss. Filter area is 242 sq ft (for 1 mgd) and backwash requirements are 350 gal/sq ft. Hence, the removal of 167 lb of TSS (ignoring solids production in the filter) will require 2.3 backwashes/day and produce 195,000 gal/day of backwash water for the 1 mgd plant. On the other hand, it was noted in Item 2 above that the design flow for the 1 mgd plant is at least 2 mgd. This indicates that they are estimating a backwash requirement which is 100% of the influent flow. Again it is difficult to determine just what the design actually calls for.

11. It is stated in the section on cost analysis that the comparative cost information is considered accurate to ± 50%.

12. A cost credit was taken for the energy to be recovered from CH_4 generated in the ANFLOW process at the 1 mgd size (Page 19). However, no capital costs are assigned to gas collection, cleaning, storage, or power generation equipment. Since an overall conversion efficiency of 40% was assumed, both electric generation and waste heat recovery must have been contemplated. The cost credit is probably \$44,000/year (225hp · 365 · 24 · 0.07457 · 0.03¢/kwh) although the value is not specifically stated. The paper states that this cost was used to offset unit process power costs for the 1 mgd flow rate case.

13. The 1 inch ceramic Raschig rings used in the pilot plant have an installed cost of $10./cu ft. This results in a capital cost of $3.85 million for a 1 mgd complete ANFLOW plant and a capital cost which is 2.03 times that of the activated sludge system against which it is compared. At 0.05 mgd, the capital costs of the two systems were essentially the same. When 3 inch plastic ring packing was used for the 1 mgd ANFLOW cost estimate, the capital cost estimate decreased to $2.45 million. The 1 inch ceramic rings used in their pilot studies had an approximate surface area of 58 sq ft/cu ft (40) whereas 3 inch Raschig rings would reduce the surface area to approximately 20 sq ft/cu ft. For a given film thickness, the amount of attached biological growth in the system with 3 inch rings would be 1/3 that in the piloted system. It is not clear how this change will enhance the performance of the reactor.

The paper by Genung et al. (7) also gives cost and energy comparisons between ANFLOW and similar activated sludge configurations to those used in Reference 8. A few points worth noting in this paper are:

1. The capital costs for the ANFLOW system are all based on using the 3 inch plastic packing.

2. Capital costs and O&M labor requirements for both ANFLOW and activated sludge are said to be the same for pumping. Yet Koon et al. (7) states that grinding to particle sizes of 0.5 mm or less was felt to be needed for ANFLOW. The flow scheme only calls for grit removal, comminution and flow equalization prior to the grinding/pumping operation.

3. The total anaual cost for ANFLOW at 1 mgd ($297,000 on Page 35 of Reference 7 vs. $300,000 in Reference 8) again apparently takes credit for energy recovery but allocates no capital for this to occur.

4. There is no breakout of O&M costs for ANFLOW so it is not clear how much money, if any, has been allocated for alum and polymer. The sludge production data given in (8) suggest an alum cost alone of $43,800/yr at 1 mgd. Yet the capital cost for ANFLOW of $2.451 million and labor requirements of 6490 man hr/yr represent an annual cost of $261,000 at 1 mgd when using the labor rates and amortization specified in (8). Since the total annual cost is said to be $297,000, alum and polymer costs were probably not considered.

5. Even for the weak wastewater (BOD = 100 mg/1), Table 4 indicates that power recovery is feasible for ANFLOW and will produce at least 954 kwh of power. Based on the results of their pilot plant operation, this is unrealistic.

By this point, it is clear that the economic analysis for a complete ANFLOW system is based on attaining greater BOD and TSS removals and more CH_4 production than ever attained in the pilot plant operation. To make the

process competitive with activated sludge requires using an untested packing which will provide about 1/3 the surface area used in the pilot plant reactor. Neither paper reviewed here provides a flow and materials balance, and it is not clear upon what the ANFLOW design assumptions are actually based. Items such as purchasing and maintaining grinder pumps, power recovery equipment, alum and polymer etc., seem to have been "lost" in the cost accounting procedures used. The vagueness is exacerbated by the manner in which data and design assumptions are presented. The limited number of data parameters makes it impossible to make calculations concerning reactor performance and the performance of the subsequent processes needed. For example, effluent BOD's are not subdivided into particulate and dissolved values. Yet there is the statement (7) that the volatile acids were not efficiently converted to CH_4 in the colder months and tended to be discharged with the effluent, but the acid concentrations are never given. Since acetic acid is the most prevalent volatile acid intermediate formed in the methane fermentation of fats, carbohydrates and proteins, and about 70 percent of the methane produced results from its degradation (46), quantitative data are needed. Similarly, Figure 3 in Reference 8 indicates that for one steady state period the COD reduction within the ANFLOW reactor averaged 99%, which is better than any aerobic system could be expected to perform, but this astounding observation is not explained or developed. Successfully meeting secondary effluent standards will require terminal suspended solids and soluble BOD removal. However, the only results mentioned from the 2-year program are batch filtration studies with 0.25 to 0.50 mm sand, which is probably too small for full-scale operation, and 3-weeks operation with a downflow dual media filter which presented operational difficulties.

SECTION 4

COMPARISON WITH EQUIVALENT TECHNOLOGIES

The use of anaerobic systems in place of aerobic systems for wastewater treatment offers several potential advantages. There are no oxygenation requirements and biological sludge production is much lower. Energy requirements may prove to be lower than with conventional systems and a potentially usable fuel, CH_4, is produced.

To obtain a better overall perspective of the cost of anaerobic systems in relation to conventional activated sludge plants, the cost of primary treatment plants was compared to conventional activated sludge plants for flows of 3785 and 37850 cu m/day (1 and 10 mgd). From this comparison, one could estimate the range of costs for anaerobic reactors that would make the total system cost competitive with the activated sludge system. Version 1.2 of EXEC/OP (47,48) with the single design evaluation feature was used to generate the costs of primary and activated sludge treatment systems. This computer program computes cost and energy requirements for a specified sequence of unit processes and design parameters. A partial listing of the input design parameters is presented in Table 13. Output parameters for cost and energy requirements are summarized in Tables 14 and 15.

As previously shown, the sludge handling sequence for an anaerobic fluidized bed plant would likely be the same as for a primary plant. Hence a comparison of the processes in Tables 14 and 15 gives an indication of the costs which can be associated with a fluidized bed reactor and any needed ancillary equipment (e.g. postaeration) and still result in a treatment sequence competitive with conventional technologies. The two solids handling options considered in Tables 14 and 15 were either gravity thickening, anaerobic digestion, elutriation and vacuum filtration; or gravity thickening, lime stabilization and vacuum filtration. This accounts for the two different costs given for primary plants in Table 14 and for secondary plants in Table 15. All energy requirements for heat, fuel or electric power are expressed in units of equivalent kwh/MG.

The total annual costs for the plants summarized in Tables 14 and 15 are shown in Table 16.

The difference in costs at 1 mgd between the complete activated sludge and primary plants was 324 and 275 $/MG; the cost of the activated sludge tank and final settler was $189/MG. At 10 mgd the difference in costs between the complete plants was 145 and 138 $/MG; the costs of the activated sludge tank and final settlers were 84 and 83 $/MG with the difference reflecting

TABLE 13. Summary of Selected Input Parameters Used for EXEC/OP

Construction Cost Index (3rd quarter 1973 = 1.0)	1.55
Wholesale Price Index (1967 = 1.0)	2.4
Discount Rate, decimal	.07125
Planning Period, years	20.
Direct Hourly Wage, $/hr	7.00
Heat Energy Conversion Efficiency, decimal	.30

Process	Design Parameter	Value
Raw Wastewater	Influent TSS, mg/l	240.
	Influent VSS, mg/l	190.
	Influent Suspended BOD, mg/l	170.
	Influent Dissolved BOD, mg/l	80.
Pumping	Pumping Head, ft	30.
Preliminary Treatment	Grit Removal	yes
	Bar Screens	yes
Primary Sedimentation	TSS removal, %	60.
	Underflow TSS, %	2.
Activated Sludge and Final Settler	Effluent BOD, mg/l	25.
	Effluent TSS, mg/l	25.
	MLVSS, mg/l	2000.
	Oxygen Transfer Efficiency, %	6.
	True Yield Coefficient, lb/lb BOD	.7
	Biomass Decay Coefficient, 1/day	.12
Chlorination	Chlorine Dose, mg/l	8.
Gravity Thickening	Solids Recovery, %	95.
	Hydraulic Loading, gpd/sq ft	600.
	Solids Loading, lb/day/sq ft	25. (PRI)
		8. (MIX)
	Underflow TSS, %	9. (PRI)
		4.5 (MIX)
Anaerobic Digestion	Temperature, ^{o}C	35.
	Retention Time, days	20.
Elutriation	Washwater Ratio	3.
	Underflow TSS, %	9. (PRI)
		5. (MIX)
	Solids Recovery Ratio, %	85.
	Hydraulic Loading, gpd/sq ft	500.
	Solids Loading, lb/day/sq ft	20. (PRI)
		10. (MIX)

TABLE 13. (Continued)

Process	Design Parameters	Value
Lime Stabilization	Lime Dose, lb/ton dry solids	400.
Vacuum Filtration	Primary Elutriated Digested Sludge	
	$FeCl_3$ Dose, lb	40.
	Lime Dose, lb.	0.
	Mixed Elutriated Digested Sludge	
	$FeCl_3$ Dose, lb.	100.
	Lime Dose, lb	0.
	Primary Lime Stabilized Sludge	
	$FeCl_3$ Dose, lb	30.
	Lime Dose, lb	0.
	Mixed Lime Stabilized Sludge	
	$FeCl_3$ Dose, lb	60.
	Lime Dose, lb	0.
	Dewatering Rate, gph/sq ft	
	Primary Elutriated Digested Sludge	16.25
	Mixed Elutriated Digested Sludge	10.
	Primary Lime Stabilized Sludge	17.5
	Mixed Lime Stabilized Sludge	11.25
	$FeCl_3$, $/lb	.064
	Lime, $/ton	45.
Truck Transport and Land Disposal	Hauling Distance, miles	5.
	Depreciation Period for Trucks, years	7.
	Fuel Cost, $/gal	1.
	Land Cost, $/acre	3000.
	Landfilling	yes

TABLE 14. Cost and Energy from EXEC/OP for a 1 MGD Flow

	SECONDARY PLANT			PRIMARY PLANT NO. 1			PRIMARY PLANT NO. 2		
	Total Annual Cost $/MG		Energy KwH/MG	Total Annual Cost $/MG		Energy KwH/MG	Total Annual Cost $/MG		Energy KwH/MG
	Capital	O&M		Capital	O&M		Capital	O&M	
Raw Water Pumping	50.10	23.63	149.5	50.10	23.63	149.5	50.10	23.63	149.5
Preliminary Treatment	17.17	26.30	18.8	17.17	26.30	18.8	17.17	26.30	18.8
Primary Sedimentation	52.30	43.23	25.4	51.69	42.52	25.3	52.34	43.36	25.3
Activated Sludge and Final Settler	103.60	84.96	543.5	--	--	--	--	--	--
Effluent Chlorination	22.48	22.40	18.2	22.48	22.41	18.2	22.54	22.49	18.3
Gravity Thickener	17.41	12.11	7.2	16.90	12.14	8.0	16.84	12.13	7.8
Anaerobic Digestion	--	--	--	--	--	--	78.24	24.04	57.3
Sludge Elutriation	--	--	--	--	--	--	17.20	12.16	8.4
Lime Stabilization	6.16	36.41	7.7	4.28	29.70	5.3	--	--	--
Vacuum Filtration	79.52	69.25	247.7	34.83	44.21	84.2	30.15	30.20	68.3
Truck Transport and Land Disposal	66.01	137.86	91.8	57.49	91.01	61.5	53.22	63.60	44.0
TOTAL	414.75	456.14	1109.8	254.94	291.91	370.8	337.80	257.90	397.7

TABLE 15. Cost and Energy from EXEC/OP for a 10 MGD Flow

	SECONDARY PLANT NO. 1			SECONDARY PLANT NO. 2			PRIMARY PLANT		
	Total Annual Cost $/MG		Energy KwH/MG	Total Annual Cost $/MG		Energy KwH/MG	Total Annual Cost $/MG		Energy KwH/MG
	Capital	O&M		Capital	O&M		Capital	O&M	
Raw Water Pumping	23.48	8.80	141.4	23.48	8.80	141.4	23.48	8.80	141.4
Preliminary Treatment	7.14	7.50	5.2	7.14	7.50	5.2	7.14	7.50	5.2
Primary Sedimentation	18.76	14.29	8.9	18.24	13.82	8.8	18.24	13.85	8.8
Activated Sludge and Final Settler	45.81	37.72	529.5	45.03	37.60	528.2	--	--	--
Effluent Chlorination	7.05	15.45	18.4	6.99	15.19	18.0	7.01	15.27	18.2
Gravity Thickener	3.70	1.63	1.6	3.54	1.57	1.6	2.19	1.22	.9
Anaerobic Digestion	21.15	6.52	155.6	--	--	--	10.67	3.54	44.5
Sludge Elutriation	3.22	1.44	1.4	--	--	--	2.04	1.22	.8
Lime Stabilization	--	--	--	2.89	13.60	3.9	--	--	--
Vacuum Filtration	23.28	23.24	97.0	32.11	34.50	144.9	8.19	12.24	26.5
Truck Transport and Land Disposal	14.87	34.52	21.5	20.76	50.86	33.0	8.69	23.35	15.8
TOTAL	168.46	151.11	980.5	129.56	183.44	885.0	87.65	86.99	262.1

TABLE 16. SUMMARY OF COSTS GENERATED WITH THE EXEC/OP PROGRAM

Plant Size (mgd)	Activated Sludge Plant ($/MG)		Primary Treatment Plant ($/MG)	
	No. 1	No. 2	No. 1	No. 2
1	871	--	547	596
10	320	313	175	--

the variation in recycle streams from sludge processing. In both cases it can be seen that the combined capital and operation cost of an anaerobic fluidized bed and ancillary equipment can exceed that of an activated sludge tank and final settler (by 1.45 to 1.73 for the cases considered) and still produce a treatment system with the same annual cost as the activated sludge plants modeled.

If an anaerobic fluidized bed to provide a 2-hour HRT could be installed at $1060/cu m ($30/cu ft) for a 3785 cu m/day (1 mgd) flow, the capital cost would be $334,000 or a total annual cost (at 7 1/8%) of $87/MG for the capital cost portion of the plant. At 37850 cu m/day (10 mgd) and $706./cu m ($20./cu ft) the annual capital cost would be $58./MG. Pumping costs for an anaerobic fluidized bed hopefully would by no more than $10/MG (see Section III). A rough guide to estimated maintenance material costs and labor requirements can be obtained by considering the reported requirements (49) for gravity filtration structures. Estimated requirements are shown in Table 17.

TABLE 17. OPERATION AND MAINTENANCE SUMMARY FOR GRAVITY FILTRATION STRUCTURES

Total Filter Area		Building Energy	Maintenance Material	Labor	Total Cost*
sq m	sq ft	kwh/yr	$/yr	hr/yr	$/yr
13	140	44,120	800	900	11,120
65	700	151,850	2,510	1,500	22,070
130	1,400	279,070	4,020	2,100	33,390
650	7,000	1,190,160	13,200	4,600	94,900
1300	14,000	2,165,890	21,600	7,000	156,580
2600	28,000	4,123,490	36,700	18,000	340,400

*Calculated using $.03/kwh and $10./hr of labor

When these cost estimates are compared to the cost differences in Tables 14 and 15, it is clear that anaerobic fluidized bed systems offer potential cost advantages when compared to conventional treatment systems.

Fluidized bed systems and ANFLOW systems have certain disadvantages or characteristics which are not typical of aerobic biological systems. These are:

1. The effluent will contain no dissolved oxygen. Post aeration will

be required.

2. Problems associated with hydrogen sulfide production relate to undesirable odors and the corrosive characteristics of this gas. Either anaerobic system may find this to be a potential problem.

3. Since the biological sludge production is less than with aerobic systems, there will be a corresponding reduction in N and P removal. Furthermore, none of the nitrogen will be oxidized.

4. Methane forming bacteria have slow growth rates and are sensitive to toxic materials. If a plant receives toxic materials there may be long periods during reestablishment of the methane bacteria when treatment efficiency suffers.

SECTION 5

ASSESSMENT OF NATIONAL IMPACT

Anaerobic systems used for treatment of municipal wastewaters offer potential for reducing operating energy requirements compared to conventional activated sludge systems. However, owing to the fact that these systems are still in the developmental stage, little data are available upon which to base firm estimates of the energy required to operate an anaerobic reactor, and the energy which may be recoverable as a result of its operation. Similarly, since no full scale systems have been designed or constructed, cost assumptions are difficult to justify. For these reasons, no attempt was made to project the national impact of implementing full scale anaerobic systems for treating municipal wastewater.

There are a number of studies planned or underway to expand the knowledge of anaerobic system process performance. The Department of Energy has recently funded several studies in their program to evaluate submerged media anaerobic reactor (SMAR) concepts. These include:

1. A study to identify and evaluate factors affecting SMAR performance conducted by Dr. Young at Iowa State University.

2. A study to investigate and evaluate the mechanisms of anaerobic filter treatment of dilute wastewater by Drs. Rittmann and Pfeffer at the University of Illinois.

3. A study with Dr. Jeris at Ecolotrol, Inc. to operate an 18.9 cu m (5000 gpd) pilot plant on municipal sewage to evaluate the anaerobic expanded and/or fluidized bed process.

4. Installation of a 189 cu m (50,000 gpd) ANFLOW reactor at Oak Ridge, Tennessee. This is a joint effort between DOE/ORNL, the Norton Company and the City of Oak Ridge.

Projects planned or underway under the sponsorship of EPA include:

1. Installation of pilot scale fluidized bed reactors for anaerobic wastewater treatment at the EPA Test and Evaluation facility in Cincinnati, Ohio.

2. An active I/A effort with the City of Hanover, the State of New Hampshire, and the consulting firms of Hoyle and Tanner and J.I. Associates to design an expanded/fluidized bed process to treat

2 mgd of domestic primary effluent at the City of Hanover plant.

3. A research grant with Dr. Jewell and the City of Hanover to evaluate the performance of retrofitted activated sludge plants redesigned to operate anaerobically.

These studies and hopefully others to follow will refine operating and economic characteristics of alternative approaches to anaerobic wastewater treatment.

SECTION 6

RECOMMENDATIONS

The scale of the proposed ANFLOW project at Oak Ridge, Tennessee should be more than sufficient to determine the operating characteristics and performance of an ANFLOW reactor. Presumably the $3.50/cu ft packing will be installed to test the performance with a more economically competitive media. However, it is doubtful that this will improve the performance. Since the process appears to be possible competitive with conventional systems only at very small scale, the 189 cu m (50,000 gpd) pilot plant should not be significantly smaller than any full scale systems which could follow. Additional supportive experimental studies are judged to be unwarranted at this time.

Effort should be directed to further evaluation of expanded/fluidized beds to gather needed design, performance and economic data. Areas where further information is required are summarized in Section III.

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