EM TASK 9 – CENTRIFUGAL MEMBRANE FILTRATION

Final Report

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EM TASK 9 – CENTRIFUGAL MEMBRANE FILTRATION

EXECUTIVE SUMMARY

The DOE has a critical need to minimize the volume of the liquid mixed-waste streams that occur at the various weapons complex sites so that remediation of these waste streams is more cost-effective. SpinTek Membrane Systems, Inc., of Huntington Beach, California, owns a novel centrifugal membrane filtration technology that makes use of ultrafiltration and centrifugal force to separate suspended and dissolved solids from liquid waste streams, producing a filtered-liquid stream and a low-volume contaminant-concentrate stream. The EERC assisted SpinTek in the continued development of its process by modifying the SpinTek system so as to enhance filtration performance; by comparing the effectiveness of SpinTek's filtration with that of traditional filters, especially when placed prior to a technetium-removal cartridge designed by 3M; and by performing a preliminary cost analysis for the use of SpinTek as a waste-minimization technique.

Statistically designed experimental matrices were used to evaluate variations in turbulence promoter design to determine their effect on filtration performance. The results indicated that the permeate flux rate was greater and the power consumption lower when the turbulence promoter was beveled.

An economic analysis was performed to determine the approximate cost of filtering liquid waste with the SpinTek system and to compare this cost to the cost of other, competing technologies. The information can be used to identify situations in which the SpinTek technology could be most effectively applied. It was found that the cost of using SpinTek technology compares favorably with other ultrafiltration technologies that are currently available and that the choice of using SpinTek rather than another technology would depend on a number of factors, including waste volume and contaminant type and concentration. Because its use is not limited to specific contaminants, the SpinTek process is applicable to a wider variety of liquid waste streams than the other technologies.

SpinTek's filtration is a crosscutting technology with a number of applications for cleanup of DOE weapons complex liquid wastes. One application is the enhancement of downstream unit operations such as adsorption or ion exchange processes where even low levels of suspended solids create operation problems. Tests were performed to evaluate the use of SpinTek filtration as prefiltration prior to a 3M technology that selectively removes dissolved radionuclides from liquid wastes. The 3M technology is limited by the quantity of suspended materials contained in the waste stream, which can cause plugging well before its radionuclide-removal capability is exceeded. Three types of tests were performed: without filtration, with prefiltration using conventional cartridge-style paper filters, and with prefiltration by SpinTek centrifugal membrane filtration. The results showed that prefiltration was required for effective operation of the 3M technology and that the SpinTek prefiltration was much more efficient at removing suspended solids than prefiltration using conventional cartridge-style filters. Because of their minimal cost, conventional paper prefilters may be adequate for filtration of small volumes or material with low

suspended solids content. As waste volumes and the suspended solids loading increase, economics begin to favor the more effective SpinTek prefiltration.

According to current DOE EM needs, there are 90.6 million gallons of tank waste requiring treatment. SpinTek filtration could also be applied to the remediation of contaminated groundwater plumes, the treatment of secondary liquid waste streams from other remediation processes, and the filtration of liquid waste streams generated during decontamination and decommissioning activities.

EM TASK 9 – CENTRIFUGAL MEMBRANE FILTRATION

1.0 BACKGROUND

This project was designed to establish the utility of a novel centrifugal membrane filtration technology for the remediation of liquid mixed-waste streams at U.S. Department of Energy (DOE) facilities in support of the DOE Environmental Management (EM) Program. The Energy & Environmental Research Center (EERC) has teamed with SpinTek Membrane Systems, Inc., a small business and owner of the novel centrifugal membrane filtration technology, to establish the applicability of the technology to DOE site remediation and the commercial viability of the technology for liquid mixed-waste stream remediation.

The technology is a uniquely configured process that makes use of ultrafiltration and centrifugal force to separate suspended and dissolved solids from liquid waste streams, producing a filtered-water stream and a low-volume contaminated-concentrate stream. This technology has the potential for effective and efficient waste volume minimization, the treatment of liquid tank wastes, the remediation of contaminated groundwater plumes, and the treatment of secondary liquid waste streams from other remediation processes, as well as the liquid waste stream generated during decontamination and decommissioning activities.

2.0 OBJECTIVES

The overall project consists of several integrated research phases related to the applicability, continued development, demonstration, and commercialization of the SpinTek centrifugal membrane filtration process. Work performed during this reporting period consisted of Phase 2 evaluation of the SpinTek centrifugal membrane filtration technology and Phase 3, Technology Partnering. During Phase 1 testing conducted at the EERC using the SpinTek ST-IIL unit operating on a surrogate tank waste, a solids cake developed on the membrane surface. The solids cake was observed where linear membrane velocities were less than 17.5 ft/s and reduced the unobstructed membrane surface area up to 25%, reducing overall filtration performance.

The primary goal of the Phase 2 research effort was to enhance filtration performance through the development and testing of alternative turbulence promoter designs. The turbulence promoters were designed to generate a shear force across the entire membrane surface sufficient to maintain a self-cleaning membrane capability and improve filtration efficiency and long-term performance. Specific Phase 2 research activities included the following:

- System modifications to accommodate an 11-in.-diameter, two-disk rotating membrane assembly
- Development and fabrication of alternative turbulence promoter designs

- Testing and evaluation of the existing and alternative turbulence promoters under selected operating conditions using a statistically designed test matrix
- Data reduction and analysis

The objective of Phase 3 research was to demonstrate the effectiveness of SpinTek's centrifugal membrane filtration as a pretreatment to remove suspended solids from a liquid waste upstream of 3M's WWL cartridge technology for the selective removal of technetium (Tc).

3.0 ACCOMPLISHMENTS

Phase 2 work performed emphasized the design, testing, and evaluation of the existing and alternative turbulence promoters using a statistically designed experimental matrix. Phase 3 testing activities were performed to evaluate the effectiveness of the SpinTek system in a prefiltration application.

3.1 Phase 2 Accomplishments

3.1.1 ST-IIL System Modifications

The Phase 1 testing and process evaluation were conducted using a single rotating disk with a diameter of 8 in. and membranes mounted on both sides of the disk. The total membrane surface area of the 8-in. unit was 0.5 ft². Stationary turbulence promoters were located one on each side of a rotating membrane disk within a pressure housing.

Phase 2 testing was conducted using an 11-in.-diameter, two-disk system. This system had a total membrane surface area of 2.0 ft². Membranes were mounted on both sides of each of the rotating disks. The two-disk system had three stationary turbulence promoters, one mounted between the rotating membrane disks and the other two on the outside of the respective membranes.

3.1.2 Alternative Turbulence Promoter Development

Activities on the development of alternative turbulence promoters included membrane process analysis, a review of crossflow filtration methods, turbulence promoter design, and turbulence promoter fabrication.

3.1.2.1 Membrane Process Analysis

During the Phase 1 testing, a solids cake developed on the central portion of the membrane surface. Although the cake buildup encompassed only about 25% of the total area, its location can be expected to have had a pronounced effect on the flux, since the highest transmembrane pressure (the net filtration driving force) occurs at the membrane center. Transmembrane pressure

varies over the surface because of the membrane's rotation, which creates centrifugal forces resulting in an uneven pressure distribution.

Minimization of filter cake deposition is essential for the success of this type of filtration system. Crossflow filtration processes were studied to gather information that could indicate potential methods for reducing solids cake buildup. Crossflow, or tangential flow, filtration involves moving a fluid tangentially to a filter while simultaneously filtering it. The tangential flow creates a force parallel to the membrane that helps to wash filtered particles away, thereby keeping the filter clean and the permeate flux high. Opposing the shear generated by the crossflow is the drag force caused by water passing over the filtrate and through the membrane. The higher the flux, the faster the perpendicular flow to the membrane and the stronger the drag force. Particles become deposited on the filter when the drag forces acting upon them become higher than the shear force pulling them away. Once it is deposited, the shear required to remove the particulate must overcome the drag force as well as the friction force holding it to the filter. Thus, utilizing techniques to prohibit initial particle deposition would seem to be the most efficient method of membrane cleaning.

Particles deposited on the filter soon form a gel or filter cake layer that can be several magnitudes lower in permeability than the original filter, restricting the flow through it (1). This restriction results in a decrease in flux and its associated drag force. Eventually, an equilibrium occurs between the crossflow and the drag forces. The thickness and permeability of the gel layer when this equilibrium occurs determine the steady-state flux of the filter.

The formation of a small cake layer is almost certain in any filtration process, so it is advantageous to utilize its properties to enhance filtration. The layer can provide a barrier between the shear force and filter, protecting it from wear. It can also improve filtration by acting as a prefilter before the fluid passes through the membrane. Filtration methods should be developed to take advantage of any benefits that a gel layer may have in any crossflow application.

3.1.2.2 Crossflow Filtration Methods

The means of developing the shear force required to remove particulate from the membrane surface can be divided into two categories: low-shear crossflow filtration and high-shear crossflow filtration (2). Low-shear methods use flow velocity as the mechanism for solids cake reduction. The shear required to keep the particles suspended is a function of the feed stream pumping rate. Flow velocity downstream of the feed inlet can be reduced because of the removal of permeate as the feed passes across the membrane. The available shear is lowest near the edge of the membrane, while the solids concentration is the highest at that point due to the removal of the permeate. Low-shear methods therefore employ large pumping rates and high recirculation ratios.

The shear force in low-shear crossflow filters can be increased by utilizing unsteadiness in the crossflow. Crossflow instabilities can be a result of 1) roughness, where protuberances are placed on or near the filter surface; 2) flow pulsations in the feed stream; and 3) secondary flows

where filter geometry is chosen to help generate Taylor and Dean vortices (3). Experiments using these methods have shown increased flux and efficiency in low-shear systems.

Although the low-shear technique works well in many applications of waste treatment, it becomes ineffective for wastes containing a high percentage of solids. The high solids content overwhelms the membrane, clogging the flow passages and rendering the membrane useless. In cases such as this, high-shear filtration methods should be used. These types of systems utilize a different technique for creating shear and are designed to operate under high solids loadings. Instead of relying on high feed stream velocities to promote shear, mechanical methods are used, including moving the filter relative to the fluid (as with SpinTek) or imparting velocity to the feed water in close proximity to the filter by spinning disks or other mechanical devices. Whichever shear-promoting means is chosen, the result tends to be a much more effective filter cleaning. Since the energy is applied in the direct vicinity of the membrane, little is wasted on fluid farther away from it.

3.1.2.3 Turbulence Promoter Design

Analysis of flow patterns within the filter chamber was used to develop alternative turbulence promoter designs. The complex nature of flow within the chamber made a complete detailed analysis impractical. However, a basis for describing typical flows with the filtration chamber was found by utilizing research done on flow boundaries next to rotating disks.

Experiments have shown the velocity of a fluid within a filter chamber containing a smooth, grooveless disk (in our case the spinning filter) to comprise three regions: a boundary layer of thickness δ next to the spinning disk, a boundary layer of thickness ζ next to a fixed plate, and a core region between the two boundary layers. Suggestions for equations identifying flow velocity estimation within each of the layers for both laminar ($Re \leq 3 \times 10^5$) and turbulent flow ($Re \geq 3 \times 10^5$) where $Re = r_o \omega / \nu$ have been proposed (4).

Flows normally encountered with high-shear dynamic filtration tend to be turbulent (2, 5), due to the high energy transfer from disk to fluid. An approximation for the flow patterns of a smooth spinning disk under turbulent flow conditions was analyzed by the following equations (4):

Fixed plate boundary layer:

$$\text{Boundary layer thickness } (\zeta) = 0.309(r[1-K]^3/K)(\nu/r^2\omega_o)^{1/5}$$

$$\text{Tangential velocity } (\mu) = Kr\omega_o(s/\zeta)^{1/7}$$

$$\text{Radial velocity } (v) = -0.374Kr\omega_o(s/\zeta)^{1/7}(1-s/\zeta)$$

Spinning disk boundary layer:

$$\text{Boundary layer thickness } (\delta) = 0.526(1-K)^2 r(\nu/r^2\omega_o)^{1/5}$$

$$\text{Tangential velocity } (\mu) = r\omega_o(1-[z/\delta]^{1/7} + K[z/\delta]^{1/7})$$

$$\text{Radial velocity (v)} = 0.220(1-K)r\omega_0(z/\delta)^{1/7}(1-z/\delta)$$

Where:

- K = $\mu/r\omega$ is the ratio of tangential velocities in the core and on the spinning disk, approximately 0.4 to 0.5 for a smooth disk.
- ω_0 = the angular velocity of the disk.
- v = the kinematic velocity of the fluid.
- z = the axial distance from the spinning disk.
- r = the radial distance from the center of the disk.

The above equations were used to identify methods that could be used to enhance the flow patterns within the filter chamber, to increase both circulation and shear. Assuming the turbulence promoter to be the stationary plate, one could expect turbulent flow at much lower Reynolds numbers because of the effects of the promoter blades.

To reduce cake deposition on the membrane, the disk boundary layer thickness should be kept to a minimum. This eliminates the chance for particles to be trapped within it by the permeate flux drag force. Reducing the boundary layer size while increasing both the tangential and radial velocities of the fluid next to the disk should result in a sweeping action within the chamber that extracts particles before they are deposited on the membrane.

Experiments performed on spinning grooved disks have shown that the grooving increases the tangential and radial velocities on fluids next to them (4). The velocity increases were thought to be caused by the increased friction between the disk and fluid brought about by the grooves. Radial flow along the grooves was thought to be the cause of the increased radial velocities. Assuming the tractive force can be created by blades as well as grooves, placement of blades directly on the spinning filter disk may also help reduce filter cake buildup. Although blades on the filter disk may slightly decrease membrane area, the area lost would be small, since only a few blades should be required. To be effective, however, stationary membrane filter disks would have to replace the current turbulence promoters. The spinning filter with blades could be used to impart the shear necessary to keep the stationary filter free of deposits while increasing fluid circulation within the entire filter chamber. Drawbacks of this design could include a net decrease in shear along the membrane due to the blades increasing the amount of fluid spinning with the disk. Power consumption would be increased, as more energy is being imparted to the fluid. Finally, some system modifications would be required for such a design.

Reducing the boundary layer thickness may also be accomplished by reducing the clearance between the membrane and promoter. Though varying the clearance for filtration units with rotating shear promoters and stationary filter disks has shown few flux changes for small clearance adjustments, no evidence of the same occurrence for rotating filter disks was found. Experiments on rotating disks showed increased local shear stresses with decreased clearance for various operating conditions (2). The stress increase should result in less particle deposition and a higher flux. The increased stress may increase power consumption, but may be required only within the inner portion of the membrane where lower local velocities and highest permeation

drag forces are present. The practical limit of clearance reduction is unknown, but adequate clearance must be maintained to prevent membrane–promoter contact during operation.

The original turbulence promoter, pictured in Figure 1, looks to be well-suited for creating shear along the membrane. The stationary blades increase velocity gradients and turbulence along the membrane, resulting in a decreased solids cake. However, although turbulence and shear are high, the circulation of fluid in and out of the disks may be small. Assuming the primary flow patterns are outward fluid flow along the spinning membrane and feed inflow along the stationary blades (6), this configuration may not provide enough circulation for solutions with a high solids content. Low circulation may cause excessive solids concentrations within the filter disks, resulting in filter cake formation.

Incorporating a promoter design that increases flow circulation within the chamber may also help decrease gel layer formation. A promoter having angled blades, such as the one pictured in Figure 2, should help increase fluid circulation within the chamber. This type of design could help to remove particles from within the filtration disks by imparting a radial velocity to fluid particles having a tangential velocity component. The radial velocity increase could help expel particles from the central portion of the membrane, while helping bring fresh waste to the disks for filtration.

Creating a bevel on the leading edge of the turbulence promoter blades may also be useful in limiting filter cake buildup. Beveling the edge, alternately on top, then bottom, for consecutive blades may allow the blade to create lift and turbulence to help increase shear and reduce the

Figure 1. Original turbulence promoter.

Figure 2. Angled-blade turbulence promoter.

solids cake. The beveling effect may not be pronounced on the original promoter design, but using thicker promoters with a smaller membrane–promoter gap may magnify the beveling effect.

3.1.3 Alternative Turbulence Promoter Design Development

Development of alternative turbulence promoters included investigation of the hydrodynamics of filter cake buildup, examination of crossflow filtration methods, including high-shear crossflow filtration, and an evaluation of fluid flow in rotating-disk systems.

3.1.3.1 Hydrodynamics of Filter Cake Buildup

Classical filtration, also called dead-end filtration, has traditionally been the most widely used filtration process. This method removes suspended particulate by drawing the fluid through a filter medium that retains the suspended material. As the process continues, a cake of filtered material forms on the original filter, growing at a rate proportional to the rate of fluid filtration and concentration of the slurry. The filter may soon become clogged with filtered particulate, drastically reducing the filtration rate. The initial rate of filtration can be attained by stopping the process and removing the cake. This type of filtration works well for small filtration requirements and will always be the method of choice when recovery of the suspended material is what is sought from filtration, but it becomes impractical when continuous filtration is required.

A filtration method that continuously removes the filter cake is used when continuous filtration is required. This method, called crossflow or tangential flow filtration, involves moving a

fluid tangentially to a filter while simultaneously filtering it. The tangential flow creates a force parallel to the filter that helps to wash filtered particles away, thereby keeping the filter clean and the permeate flux high. Opposing the parallel shear force generated by the crossflow is the drag force caused by permeate passing over the filtered particles and through the membrane. The higher the flux, the faster the flow perpendicular to the membrane and the stronger the drag force. Particles become deposited on the filter when the drag force acting upon them becomes higher than the shear force pulling them away.

Figure 3, taken from Cardew and Byrne (7), gives an excellent visualization of dead-end and crossflow filtration methods. Particles deposited on the filter soon form a gel or filter cake layer that can be several magnitudes lower in permeability than the original filter, restricting the flow through it (1). This restriction results in a decrease in permeate flux and its associated drag force. Eventually, an equilibrium occurs between the crossflow and the drag forces. The thickness and permeability of the gel layer when this equilibrium occurs determines the steady-state flux of the filter.

The inevitability of filter cake development during crossflow filtration has prompted numerous studies in this area. Once formed, the properties of the cake rather than that of the original filter may control the filtration process. Since the cake permeability can be several

Figure 3. Dead-end and crossflow filtration methods (Source: Cardew and Byrne).

magnitudes lower than that of the original filter, its development can significantly decrease filtrate flux. The cake can also play a crucial role in particle rejection, due to its decreased pore size over the original filter.

Forces involved in cake formation during crossflow filtration are identified by Jiao and Sharma (1) as a hydrodynamic tangential force (shear stress) created from the crossflow, a drag force resulting from the filtrate flux, a hydrodynamic lift force, and surface forces between the particle and the filter. Surface forces, although large in some instances, act over a much shorter range than the other forces. They can also be reduced by utilization of membranes that are enhanced to treat a particular waste stream (8, 9). Lift forces have been shown to be small for laminar flow over particles attached to a smooth wall (1).

Various equations for identifying the hydrodynamic forces exerted on spherical particles during crossflow filtration have been proposed. Lu and Ju (11) identify hydrodynamic forces acting upon particles during crossflow filtration as

1. Vertical drag force pulling the particulate toward the filter

$$F_{(n+w)} = \phi 3\pi\mu d_p(q - v_l) + \pi/6(\rho_p - \rho_s)gd_p^3$$

and

2. Tangential crossflow force pushing particles parallel to the filter

$$F_t = 1.7009(3\pi\mu d_p u_p [z = d_p/2])$$

Where:

ϕ	=	correction factor of Stokes law.
μ	=	viscosity of fluid (kg/s · m).
d_p	=	diameter of particle (m).
q	=	permeate flux (m ³ /s · m ²).
v_l	=	lift velocity of sphere near bounded wall (m/s).
ρ_p, ρ_s	=	density of particle and slurry, respectively (kg/m ³).
u_p	=	undisturbed flow velocity at $z = d_p/2$.

Utilizing the above formulas and summing the moments created by each about a contact point between the particle and the filter surface, one obtains

$$M = (d_p / 2) (F_t \cos \theta - [F_n + F_w] \sin \theta)$$

Where:

M	=	moment created by hydrodynamic forces.
θ	=	the angle of repose between filter and particle (rad).

The particle may remain stable on the filter (negative moment) or be swept away (positive moment). At the equilibrium point where the net moment is zero, the above equation can be solved for the largest diameter of particle that is likely to be deposited upon the filter under the given flow conditions. By making some generalized assumptions, the authors give equations for the largest critical cutoff particle diameter for turbulent and laminar flow conditions. This diameter corresponds to the largest particle that can theoretically be deposited on the membrane under the given flow conditions. Particles smaller than the cutoff diameter can become deposited on the filter, while larger particles are swept away with the crossflow.

Jiao and Sharma (1) also offer equations for the normal drag force driving the particle to the filter (F_y) and the tangential force acting to remove the particle from the filter surface (F_x). They evaluated two different mechanisms by which particles could be released from the filter: sliding and rolling. If the particle were released from the membrane by sliding, the tangential force would have to be greater than the product of the hydrodynamic drag force and the coefficient of friction between the particle and the filter. If the particle were released by rolling, which is the case for spherical particles on a flat surface (1), a torque balance should be used to evaluate whether or not the particle becomes deposited on the membrane.

Numerous investigators (1, 10, 11) have shown that the cutoff diameter is a function of filtrate flux and crossflow velocity, with higher flux corresponding to deposition of larger particles. Development of the filter cake initially begins with the deposition of a wide variety of particle sizes on the membrane, with finer and finer particles subsequently deposited. The finer particles decrease the permeability of the filter until the flux decreases to such an extent that all of the particles present within the suspension are too large to be deposited on the filter. At this point, a steady filtration rate (or steady state) is reached (1). A possible explanation for this phenomenon is as follows.

As filtration begins on a clean membrane, filtration rates are high as a result of the high permeability of the bare membrane. The large flux values correspond to increased drag forces on the particles in the slurry. The large drag forces, in turn, correspond to an increased critical cutoff diameter for particles to be deposited. A wide variety of particle sizes can thus be initially trapped on the filter surface. As filtration continues, more and more particles become trapped on the filter, eventually restricting flow through it. The decreased flux results in less particle drag and a smaller cutoff diameter for particles deposited on the filter. This process continues until the filtration rate is low enough that all of the remaining particles in the slurry are larger than the cutoff diameter, at which time steady state is reached (1). Further investigations have shown that filtration pressure has no effect on the steady-state filtration rate, since cake permeability decreases as filtration pressure increases (1). Therefore, the filter cake acts as a secondary membrane capable of controlling the filtration process. In many cases, this layer may act as the primary filter, removing particulate from the fluid stream before it comes in contact with the original filter. Such is the case when filter aids are used. The aids develop a cake layer on the underlying filter that provides a barrier to trap unwanted particulate from going through the filter with the filtrate. These membranes are also referred to as dynamic membranes, since they develop and change continuously with the motion of the fluid being filtered.

Tanny categorized secondary membranes into three classes. These classes are summarized by Murkes and Carlsson (2) in the following manner.

Class 1 Dynamic Membranes

This class of dynamic membranes consists of those formed when a solution containing a macromolecular gel-forming solute is ultrafiltered through a membrane whose pore size is small enough to retain at least a portion of the molecule. Retention of the molecules increases their concentration from that in the bulk fluid (C_b) to an increased concentration at the membrane (C_w). As filtration continues, this concentration increases until a gel is formed, at concentration C_g , which is a constant. Mathematically, the filtrate flux (J) in this system can be expressed as

$$J = k \ln (C_w/C_b)$$

where k is a back-diffusion mass-transfer coefficient. As the concentration increases at the membrane surface, C_w approaches C_g . The flux then becomes a function of $\ln C_b$ only. Another expression for flux, written in terms of transmembrane pressure (ΔP), hydraulic resistance of the membrane (R_m), and polarization layer (R_p) is

$$J = \Delta P / (R_m + R_p)$$

By combining these flux equations, it can be shown that as long as C_w is less than that of C_g , C_w will increase with pressure. However, as soon as C_w becomes equal to C_g , increasing pressure corresponds only to an increased resistance of the gel layer and not an increase in flux.

Class 2 Dynamic Membranes

Development of this type of secondary membrane is often the case during crossflow filtration and is most probably the one formed in the SpinTek unit. This class of secondary membranes is formed from filtration of colloidal suspensions through a membrane whose pore size may be up to 1–2 μm larger than that of the colloid. Rejection of solute particles occurs through a dense layer of retained particles that form on top of the membrane. Development of the initial cake is not well understood. However, it seems that the filtered particulate first fills the membrane pores and then forms a filter cake, which subsequently rejects particles much smaller than the original filter pore size.

Mathematical models describing the filtration process where Class 2 membranes are developed include the standard law of filtration and Ruth's law. It would seem that the standard law of filtration would be valid during the initial stages of filtration, when particles smaller than the membrane pore openings become deposited within the pores at such a rate that the volume of the pores decreases in proportion to the volume filtered. This law can be expressed mathematically as

$$kt = (t / V) - (1 / J_o)$$

where J_0 is the initial flux, t is the filtration time, and V is the total filtrate quantity collected until time t .

After a certain amount of filtration, clogging of the pores causes the membrane to reject almost all colloids in the suspension, and the process becomes a cake filtration process that can be represented by Ruth's law:

$$J(t) = K / (2[V + V_f])$$

In this expression, V_f represents the volume of filtrate that produces a cake having the same hydrodynamic resistance as that of the clean filter. $J(t)$ represents the flux at time t , and K is the Ruth constant:

$$K = 2PS^2/\mu cr$$

where S is the surface area of the cake, μ is the viscosity of the fluid, c represents the concentration of the colloid, and r is the specific cake resistance (flow resistance per unit mass of solids per unit area) (2).

Class 3 Dynamic Membranes

This class of secondary membrane is formed when the solution contains molecules of a size relatively close to that of the porous support membrane. Filtration causes the molecules to enter the membrane pore, and interaction between the molecule and the membrane holds it there. The pore size of the original membrane must be very small if the held molecule is a polymer and the rejection properties of the membrane are very high. Development of this type of membrane is dependent upon the pore size of the original filter. If the pore size is too small, no particles will be trapped, while pores that are too large do not allow retention of the polymer molecules.

In almost all filtration processes, the development of a secondary or dynamic membrane is inevitable. Utilizing methods to control it may be one of the best methods to enhance filtration performance. The dynamic layer can provide a barrier between the tangential shear force and the filter, protecting the filter from wear. The secondary membrane can also improve filtration by acting as a prefilter before the fluid passes through the membrane. Filtration methods should be developed to take advantage of any benefits that a gel layer may have in any crossflow application.

3.1.3.2 Crossflow Filtration Evaluation

The means of developing the shear force required to remove particulate from the membrane surface can be divided into two categories: low-shear crossflow filtration and high-shear crossflow filtration (2). Low-shear methods use flow velocity as the mechanism for solids cake reduction. The shear force required to keep the particles suspended in the crossflow is a function of the feed stream pumping rate. Large feed streams are required because flow velocity downstream of the feed inlet can be reduced by the removal of permeate as the feed passes across

the membrane. The available shear is lowest near the downstream edge of the membrane, while the solids concentration is highest at that point because of the removal of permeate. Therefore, low-shear methods employ large pumping rates and high recirculation ratios.

The shear force in low-shear crossflow filters can be increased by utilizing unsteadiness in the crossflow. Crossflow instabilities can result from 1) roughness, where protuberances are placed on or near the filter surface; 2) flow pulsations in the feed stream; and 3) secondary flows where filter geometry is chosen to help generate Taylor and Dean vortices (3). Experiments by Mackley and Sherman (3) have shown these methods to increase flux and efficiency in low-shear systems.

Although the low-shear technique works well in many applications of waste treatment, it is ineffective for wastes containing a high percentage of solids. The high solids content overwhelms the membrane, clogging the flow passages and rendering the membrane useless. In such cases high-shear filtration methods should be used. These types of systems utilize a different technique to create shear and are designed to operate under high solids loadings with lower pumping rates than low-shear techniques. Instead of relying on high feed stream velocities to promote shear, mechanical methods are used, including moving the filter relative to the fluid (as with SpinTek) or imparting velocity to the feed water in close proximity to the filter by spinning disks or other mechanical devices. Whichever shear-promoting means is chosen, the result tends to be a much more effective filter cleaning. Since the energy is applied in the direct vicinity of the membrane, little is wasted on fluid farther away from it.

3.1.3.3 Practical High-Shear Disk Membrane System Configurations

High-shear filtration devices exist in two basic geometries: a rotating disk in a housing and a rotating cylinder in a housing (2). Of these, the rotating disk is the best known and most widely used because of its simplicity and the ability to pack large amounts of membrane area into a compact unit. The following are typical configurations used in high-shear filtration devices:

- Rotating-disk membranes consecutively layered adjacent to stationary membranes (RMSM). This system operates on the principle that the rotating disks give an angular velocity to the fluid, which imparts a shear to keep the stationary filter disks clean. The shear imparted to the stationary filter, in turn, slows the fluid, causing a return shear on the rotating disks. This system packs large amounts of membrane area into a small unit. However, experiments by Wronski et al. (5) show the filtration rates of the stationary disks to be small in comparison to that of the spinning filter disks.
- Rotating-disk membranes with opposing stationary shear devices (RMSS). This type of system packs less membrane area per chamber volume, but offers the advantage of increased shear along the rotating membrane. Whereas stationary filter disks are smooth and offer little resistance to fluid rotating within the spinning filter, the stationary shear-enhancing devices can be of varying geometry to promote shear and enhance flow characteristics within the filter chamber. An advantage of a system such as this is that energy is applied directly to the filter rather than to fluid away from the membrane.

- Stationary-disk membranes with opposing rotating shear-promoting devices (SMRS). These units utilize rotating disks placed in close proximity to stationary membranes to develop the necessary shear to help reduce membrane fouling. Although the energy is not supplied directly to the membrane in a system such as this, the filtrate does not have to overcome the centrifugal forces developed within spinning membranes.
- Oscillating-disk membranes (OM). In these units, the membrane's rotation is changed so frequently that the process fluid is not allowed to attain the speed of the membrane. Since the fluid's speed never attains that of the membrane, high-velocity gradients occur near the membrane surface. The high-velocity gradient generates considerable shear along the membrane, giving this system high fouling resistance characteristics.

3.1.3.4 Fluid Flow in Rotating-Disk Systems

To identify possible solutions to the filter cake development in the central portion of the rotating membrane within the SpinTek unit, an investigation of flow patterns within a housing containing a spinning disk was carried out. Substantial research has been conducted in this area (2, 4, 6, 12, 13), although none was found that identically matched SpinTek's case. Most research has centered on spinning disks within a housing containing a fluid without permeation, although some has been conducted with the bottom of the housing made of porous medium, and tests with permeation have been run.

To describe flow patterns within a housing containing a rotating disk with radius R and angular velocity ω , two Reynolds numbers are used (2). The first Reynolds number is dependent upon the gap width, s , between the disk and housing and the kinematic viscosity of the fluid ν .

$$Re_s = \omega s^2 / \nu$$

The second Reynolds number is dependent upon the radial distance, r , from the spinning axis.

$$Re_r = \omega r^2 / \nu$$

Spinning the disk within the housing can generate the following four different modes of fluid flow (2, 6).

Region I: Laminar Flow and Narrow Gap

$$Re_s \leq 4, \quad Re_r \leq 2 \times 10^5$$

In this flow regime, the laminar boundary layers are merged and produce a shear rate that varies inversely with the disk-housing gap, s .

Region II: Laminar Flow and Wide Gap

$$\text{Re}_s \geq 4, \quad \text{Re}_r \leq 2 \times 10^5$$

The boundary layers within this region are separated by a zone of fluid that moves as a solid unit with rotational speed of $K\omega$, where $0 < K < 1$. The shear developed within this region is independent of the gap separating the disk and housing.

Region III: Turbulent Flow and Narrow Gap

$$\text{Re}_r \geq 2 \times 10^5 \quad s/r \leq 0.05$$

The flow within this region is characterized by merged turbulent boundary layers.

Region IV: Turbulent Flow and Wide Gap

$$\text{Re}_r \geq 2 \times 10^5 \quad s/r \geq 0.05$$

Two separate turbulent boundary layers are separated by a turbulent core of fluid moving at an angular velocity of $K\omega$, where $0 < K < 1$.

The K value in these descriptions is the ratio of the tangential velocity of the core to that of the spinning disk. Experiments show this value to be 0.4–0.5 for a smooth disk and up to 0.9 for a disk having eight radial vanes (2).

Correlating to the above regions, Schiele, in Murkes and Carlsson, gives the shear stresses in each region:

$$\begin{aligned} \tau_1 &= \mu \omega r / s \\ \tau_2 &= 1.81 \rho v^{1/2} (K\omega)^{3/2} r \\ \tau_3 &= 0.008 \rho (\omega r)^{7/4} (v/s)^{1/4} \\ \tau_4 &= 0.057 \rho v^{1/5} (K\omega)^{9/5} r^{8/5} \end{aligned}$$

Where μ and ρ are the fluid dynamic viscosity and density, respectively (2).

In practice, high-shear dynamic filters operate with Reynolds numbers between 10^5 and 10^6 . Adequate clearance between the spinning disk and stationary housing or filter is maintained to prevent equipment damage, so the likely flow regime in high-shear filters is Region IV (2, 6, 14). Whereas shear stress in Region IV is independent of the gap width s , if the equipment did operate in Region III, the decrease in the disk–filter gap would increase the local shear stresses according to the above formulas.

Experiments by Shirato et al. (4) show the fluid movement within a fluid housing containing a spinning disk. Along the disk, the flow is tangential through the friction developed between the disk and fluid, and radial due to centrifugal forces throwing the fluid toward the

periphery of the disk. To compensate for the outward flow along the spinning disk, fluid flow near the housing is rotated inward toward the center of the chamber.

Assuming the spinning membrane to be the spinning disk and the turbulence promoters to represent the housing, it should be reasonable to assume outward flow along the membrane and inward flow along the turbulence promoters, as shown in Figure 4. It is important to note that the flow patterns within the housing are complicated by the variety of flows within it, including permeate entering the spinning membrane, feed inflow and concentrate outflow, and circulating flow within the membrane packs.

3.1.3.5 Turbulence Promoter Design

The analysis of flow patterns within the filter chamber was used to develop alternative turbulence promoter designs. The complex nature of flow within the chamber made a complete detailed analysis impractical. However, a basis for describing typical flows with the filtration chamber was found using research done on flow boundaries next to rotating disks. Experiments have shown the velocity of a fluid within a filter chamber containing a smooth, grooveless disk (in this case, the spinning filter) to comprise three regions: a boundary layer of thickness δ next to the spinning disk, a boundary layer of thickness ζ next to a fixed plate, and a core region between the two boundary layers. Suggestions for equations identifying flow velocity estimation within each of the layers for both laminar ($Re < 3 \times 10^5$) and turbulent flow ($Re > 3 \times 10^5$) where $Re = r_o^2 \omega / \nu$ have been proposed (4).

Figure 4. Predicted flow pattern within the pressure housing.

3.1.4 *Alternate Promoter Fabrication*

A new promoter design was created utilizing a CAD drawing of an existing turbulence promoter. Use of the existing drawing ensured a proper fit within the unit, decreasing the chance of equipment damage. Three new promoter sets were cut from stainless steel using a computer-guided plasma cutter. One set of promoters was cut utilizing the original design, but was 0.0675 in. thicker than the original. The two other sets were both of the new design, but one was equal to the original promoter thickness of 0.120 in., while the other's thickness was increased to 0.1875 in. Thickness variation allowed flux testing as a function of membrane-promoter clearance. Utilizing two thicknesses of each design allowed testing of permeate flux as a function of membrane-promoter gap as well as design efficiency.

The thicker promoters were then machined using a CNC milling machine. Removing 0.0625 in. off of each side of the promoter from the outer perimeter to a distance of 1 in. inside it increased clearance to allow for the outer adhesive bead on the membrane. All newly cut promoters were then cleaned to remove any remaining steel shavings that could possibly damage test equipment.

3.1.5 *Test Plan*

The Phase 1 testing and evaluation considered the effects of four operating variables to define system operation: temperature, pressure, membrane rotational speed, and solids loading. Phase 2 testing will consider the two variables that have the greatest influence on membrane shear forces and solids cake development: pressure and membrane rotational speed. Temperature and solids loading will be held constant. The operating temperature will be maintained at 90°F. A titanium dioxide suspension will be prepared to simulate a nominal 25 wt% solids loading. A statistical matrix test design was developed to evaluate the effects of pressure, rotational speed, and time on filtration performance using six different turbulence promoter designs. The design consists of high (+) and low (–) values for pressure and rotational speed, with the overall run time being monitored and recorded. Table 1 summarizes the general test conditions for each turbulence promoter. The run order will be randomized prior to testing. Duplicate runs will also be conducted to evaluate precision and potential membrane deterioration.

TABLE 1

Statistical Test Matrix for Each Turbulence Promoter		
Run	Pressure, psig	Rotor Speed, rpm
1	+	+
2	+	–
3	–	–
4	–	+

3.1.6 *Turbulence Promoter Testing and Evaluation*

3.1.6.1 *Test Procedure*

Testing of the turbulence promoters was performed using the statistical test matrix design shown in Table 2. Six different promoter configurations were tested: the original design in two different thicknesses (0.120 and 0.1875 in.), a new design in the forward position (0.120 and 0.1875 in.), and the new design in the reverse position (0.120 and 0.1875 in.). Figure 5 shows the original and new promoter designs. In the forward position, the membrane rotated clockwise with respect to the promoter. In the reverse position, the promoters were turned over, resulting in flow patterns similar to those that would result if the membranes were rotated counterclockwise with respect to the promoter. Each design was tested at two pressures (40 and 60 psig) and two rotor speeds (900 and 1200 rpm). Thus each turbulence promoter was tested a total of four times throughout the experiment. Feed temperature was maintained at 90°F, and a solids concentration of 25 wt% titanium dioxide suspension, although a slight increase in the solids concentration occurred because of evaporation. The tests were randomized to account for any irreversible fouling of the membrane that might occur with operating time. The first and last tests of the matrix used the original promoter operating at 60 psig and 1200 rpm. These tests were used as the baseline data for the experiment.

The membranes were conditioned by running the unit with tap water prior to performing the experimental matrix. Conditioning verified membrane integrity and helped to fill pores in the new membrane to reduce permeate flux from 1400 gal/ft²×d (gfd) on the fresh disks to about 650 gfd. If actual testing were started on fresh membrane disks, a very high solids concentration could develop within the filter housing because of such a rapid removal of permeate, causing decreased system performance and possible equipment damage. The tap water conditioning also allowed clearance testing of all promoters involved in the trials before the statistical test matrix was initiated.

Once started, the testing was performed around the clock until all tests were completed. Data for each test were logged for approximately 4 hr, giving ample time for the system to reach steady-state operation. At the end of each test, the outer surfaces of each disk were rinsed free of filter cake by water collected near the end of the trial run from the permeate line. This was done to maintain a constant solids concentration within the feed tank. The inner surfaces of each disk (i.e., the bottom of the top disk and the top of the bottom disk) were examined for filter cake development. The inner surfaces were deemed the most important, since in a full-scale unit the disks are stacked on top of one another. The fluid movement between the membrane disks on these units accounts for most of the filtration, with the contribution of the outer surfaces of the top and bottom disks being small by comparison.

TABLE 2

Turbulence Promoter Evaluation Test Matrix			
Run Number	Promoter Design	Pressure, psig	Rotor Speed, rpm
1	Original (baseline)	60	1200
2	New design, forward (0.1875 in.)	60	1200
3	New design, forward (0.120 in.)	60	900
4	New design, forward (0.120 in.)	40	1200
5	New design, forward (0.1875 in.)	40	900
6	New design, reverse (0.120 in.)	40	900
7	New design, reverse (0.120 in.)	40	1200
8	Original (baseline)	40	900
9	New design, reverse (0.1875 in.)	40	900
10	New design, forward (0.120 in.)	60	1200
11	Original design (0.1875 in.)	60	1200
12	Original design (0.1875 in.)	40	1200
13	New design, reverse (0.1875 in.)	60	1200
14	Original (baseline)	40	1200
15	New design, forward (0.120 in.)	40	900
16	New design, reverse (0.1875 in.)	40	1200
17	Original design (0.1875 in.)	60	900
18	Original design (0.1875 in.)	40	900
19	New design, reverse (0.120 in.)	60	900
20	New design, reverse (0.120 in.)	60	1200
21	New design, forward (0.1875 in.)	60	900
22	Original (baseline)	60	900
23	New design, forward (0.1875 in.)	40	1200
24	New design, reverse (0.1875 in.)	60	900
25	Original (baseline)	60	1200

Original Design

New Design,
Forward Position

New Design,
Reverse Position

EERC DS15287.CDR

Figure 5. Comparison of turbulence promoter configurations.

3.1.6.2 Power Consumption Measurement

The efficiency of the turbulence promoters was determined by measuring the power consumption of the motor rotating the filtration disks while measuring the permeate flux. A power analyzer was attached to the wires conveying power directly to the motor. This allowed measurement of power directed to the rotor motor only and decreased the possibility of measuring power fluctuations from other parts of the system. To measure the power of the three-phase motor, the analyzer measured the current in one phase and the voltage between the other two. The current was measured intermittently throughout the test runs in each of the other phases to ensure that all phases were drawing the same amount. Power consumption was logged to a computer. Although power was the only parameter that had to be logged, other parameters (including voltage, amperage, power factors, and total power usage) were logged to ensure that proper data were obtained for the power consumption calculations.

3.1.6.3 Data Collection

A link between the test unit and a computer allowed data collected from each test to be automatically logged on the computer. Test parameters such as feed pressure, temperature, and flow, as well as membrane rotational speed and permeate flow, were all measured and recorded along with the time of day and elapsed time from the start of the test. Permeate flux was determined from the measured flow rate and membrane area. Data were collected at 5-min intervals. The computer data log does not show the average value of the test parameters over the 5-min period, but rather the value of the parameters at the time the data were recorded.

3.1.6.4 Test Results

The average steady-state data collected while the test matrix was performed are presented in Table 3. The table shows the feed pressure, permeate flow rate, rotor speed, and power consumption, as well as the time of day and run time for each particular test. Data from the first hour of each test were not used in determining the averages, since, in some cases, large performance fluctuations occurred in the beginning of the test.

These values could be compared only if the membrane performance were completely restored after each test. Previous testing conducted using the test unit showed that membrane performance tends to degrade with operating time. This degradation results from irreversible fouling that happens when particles plug the pores within the membrane. Washing the membranes between test runs removes almost all of the filter cake, but cannot remove particles that are trapped deep within the membrane pores. The statistical matrix run order was randomized to take the irreversible membrane fouling into consideration. The randomization also helped to eliminate any bias that could occur while the tests were performed.

A second set of matrix tests was conducted to evaluate the effects of beveling the leading edges of the turbulence promoters. The beveling was an attempt to create increased turbulence at the membrane surface. Average steady-state data collected during these tests are presented in Table 4.

TABLE 3

Data Collected During Turbulence Promoter Evaluation Tests

Randomized Test Run Number	Feed Pressure, psi	Rotor Speed, rpm	Promoter Design ^a	Permeate Flux, gfd	Power Consumption, kW	Test Duration, hr:min	Cumulative Membrane Run Time at Start of Test, hr:min:s
1	59.42	1193.29	Orig	200.41	1.19	4:06	0:00:00
2	59.39	1198.26	NF+	161.87	1.23	4:04	4:06:00
3	59.37	909.61	NF-	98.20	0.56	4:06	8:10:33
4	39.57	1203.54	NF-	157.55	1.08	4:01	12:16:50
5	39.57	895.54	NF+	103.50	0.59	4:02	16:18:25
6	39.54	901.46	NR-	100.75	0.54	4:01	20:20:38
7	39.38	1191.68	NR-	151.22	1.06	3:59	24:21:41
8	39.54	908.95	Orig	124.14	0.63	4:01	28:20:47
9	39.55	906.16	NR+	113.70	0.61	4:06	32:22:01
10	59.42	1190.82	NF-	143.12	1.05	4:06	36:28:10
11	59.49	1201.08	Orig+	188.05	1.28	4:01	40:34:11
12	39.58	1198.39	Orig+	178.10	1.26	4:02	44:35:25
13	59.49	1201.97	NR+	169.06	1.26	4:01	48:38:08
14	39.57	1214.57	Orig	176.55	1.26	4:01	52:39:28
15	39.50	895.33	NF-	89.63	0.53	4:46	56:40:41
16	39.47	1196.66	NR+	156.97	1.26	4:06	61:27:03
17	59.55	906.37	Orig+	120.61	0.62	4:02	65:33:07
18	39.41	895.62	Orig+	120.89	0.63	4:01	69:36:01
19	59.68	914.29	NR-	95.83	0.58	4:05	73:37:05
20	59.46	1210.49	NR-	149.20	1.10	4:01	77:43:01
21	59.50	909.13	NF+	100.57	0.60	4:05	81:44:19
22	59.71	906.37	Orig	110.31	0.63	4:02	85:50:18
23	39.51	1203.89	NF+	143.18	1.28	4:01	89:52:21
24	59.38	920.81	NR+	104.50	0.66	4:01	93:53:34
25	59.59	1201.05	Orig	164.36	1.21	4:01	97:54:38

^a Orig = original design; NF = new design, forward position; NR = new design, reverse position; - = 0.120-in. thickness; + = 0.1875-in. thickness.

Statistical analyses of the steady-state, averaged values contained in Tables 3 and 4 were performed. The analyses were used to create statistical models that could account for membrane degradation with time and more accurately predict the results by which to compare turbulence promoter efficiency. The validity of these models was checked by inputting operating variables (time, rotor speed, promoter thickness and position) and comparing the resulting predicted values for permeate flux and rotor power consumption to those obtained during the test. The models fairly accurately predicted the actual values.

TABLE 4

Data Collected During Beveled-Edge Turbulence Promoter Evaluation Tests

Randomized Test Run Number	Feed Pressure, psi	Rotor Speed, rpm	Promoter Design ^a	Permeate Flux, gfd	Power Consumption, kW	Test Duration, hr:min	Cumulative Membrane Run Time at Start of Test, hr:min
1	59.41	1189.27	Orig	216.40	1.17	4:01	0:00
2	59.59	1192.96	NF+	201.97	1.14	4:09	4:01
3	59.47	908.42	NF-	123.63	0.57	3:50	8:10
4	39.55	1191.39	NF-	180.89	1.05	4:01	12:00
5	39.41	894.57	NF+	128.55	0.70	4:01	16:01
6	39.53	894.06	NR-	120.93	0.56	4:01	20:02
7	39.53	1198.59	NR-	190.17	1.09	3:58	24:03
8	39.49	896.64	Orig	151.29	0.57	3:51	28:01
9	39.50	904.68	NR+	148.32	0.59	4:05	31:52
10	59.52	1189.02	NF-	200.45	1.07	4:06	35:57
11	59.43	1208.29	NR+	222.26	1.20	4:01	40:03
12	59.68	1193.87	Orig+	239.12	1.26	3:50	44:04
13	39.49	1210.72	Orig+	224.12	1.26	3:51	47:54
14	39.65	1192.61	Orig	213.07	1.16	4:01	51:45
15	39.43	893.65	NF-	125.07	0.55	4:01	55:46
16	39.55	1187.31	NR+	209.78	1.15	4:01	59:47
17	59.29	903.73	Orig+	159.86	0.65	4:01	63:48
18	39.47	894.49	Orig+	159.30	0.63	4:06	67:49
19	59.45	896.61	NR-	130.36	0.54	4:01	71:55
20	59.36	1189.12	NR-	202.65	1.04	4:05	75:56
21	59.47	905.92	NF+	139.63	0.59	4:01	80:01
22	59.45	891.96	Orig	144.38	0.59	4:01	84:02
23	39.54	1190.33	NF+	191.20	1.19	4:13	88:03
24	39.40	888.50	NR+	130.62	0.16	3:56	92:16
25	59.37	1187.41	Orig	217.07	1.16	4:01	96:12

^a Orig = original design; NF = new design, forward position; NR = new design, reverse position; - = 0.120-in. thickness; + = 0.1875-in. thickness.

3.1.6.5 Statistical Analysis of Matrix Test Data

The best-fit models were determined using the multiple linear regression procedure of the SAS/STATTM statistical software package. Equations were formulated for the permeate flux and total power consumption as a function of system variables. Separate equations were developed for the beveled and nonbeveled turbulence promoters. In the case of the flux, the equations for the logarithm of the flux were calculated since the variance of the data was not constant. The details for this transformation were outlined in a previous report (15).

The results of the regression analyses are presented in Tables 5–8 for each of the respective models. An overall qualitative assessment of the model can be judged by the correlation coefficient, R, and the T-test value. The correlation coefficient measures the quality of fit of the model by measuring the ratio of predicted variability to the unaccounted variability, while the t-value provides a value to determine the probability that the significance of the variable resulted by chance. Several observations can be made from these models. In the case of predicting the power to provide the filtration, the original turbulence promoter position consumed the most energy, and obviously the higher rotor speed required more energy to operate. One interesting finding is that the thickness of the turbulence promoter was only significant when not beveled. In fact, when the promoter was nonbeveled, the thin promoter actually required more energy to operate. This may indicate that the beveled promoter is more aerodynamic and/or may produce less turbulence.

Overall, the permeate flux rate was greater and the power consumption was lower for the beveled promoter. This could indicate an advantage for the beveled promoter. Unfortunately, the solids concentration for the beveled tests was 7% lower, i.e., 27% reduced down to 20%. This difference in solids concentration could account for the observed benefit seen from the beveled promoter. A higher solids concentration was shown previously to decrease the flux using the older membrane design. Likewise, the friction from the additional solids could account for the higher power consumption seen with the nonbeveled promoter. Several tests were repeated using

TABLE 5

Predicting Power Consumption for Nonbeveled Promoter; Dependent Variable: kW					
Analysis of Variance					
Source	DF	Sum of Squares	Mean Square	F Value	Prob>F
Model	4	2.30653	0.57663	638.153	9.E-21
Error	20	0.01807	0.00090		
C Total	24	2.32460			
Root MSE	0.03006	R-square	0.9922		
Dep Mean	0.90800	Adj R-sq	0.9907		
C.V.	3.31056				
Parameter Estimates					
Variable	DF	Parameter Estimate	Standard Error	T for H0: Parameter=0	Prob > T
INTERCEPT	1	0.946305	0.01008453	93.837	6.E-28
rpm	1	0.331822	0.00668700	49.622	2.E-22
Thick	1	0.043958	0.00603091	7.289	5.E-07
rpm*thick	1	0.025904	0.00669148	3.871	0.00095
Position*position	1	-0.068283	0.01258260	-5.427	3.E-05

TABLE 6

Predicting Power Consumption for Beveled Promoter; Dependent Variable: kW					
Analysis of Variance					
Source	DF	Sum of Squares	Mean Square	F Value	Prob>F
Model	2	2.23877	1.11938	127.694	8.E-13
Error	22	0.19285	0.00877		
C Total	24	2.43162			
Root MSE	0.09363	R-square	0.9207		
Dep Mean	0.86484	Adj R-sq	0.9135		
	10.8260				
C.V.	0				
Parameter Estimates					
Variable	DF	Parameter Estimate	Standard Error	T for H0: Parameter=0	Prob > T
INTERCEPT	1	0.923060	0.03122220	29.564	3.E-19
rpm	1	0.325630	0.02071999	15.716	2.E-13
Position*position	1	-0.080388	0.03906697	-2.058	0.05166

TABLE 7

Predicting Flux from Nonbeveled Promoter; Dependent Variable: LFLUX					
Analysis of Variance					
Source	DF	Sum of Squares	Mean Square	F Value	Prob>F
Model	5	0.26162	0.05232	271.229	6.E-17
Error	19	0.00367	0.00019		
C Total	24	0.26528			
Root MSE	0.01389	R-square	0.9862		
Dep Mean	2.12429	Adj R-sq	0.9825		
C.V.	0.65383				
Parameter Estimates					
Variable	DF	Parameter Estimate	Standard Error	T for H0: Parameter=0	Prob > T
INTERCEPT	1	2.172800	0.00468291	463.985	6.E-40
Time	1	-0.033224	0.00497657	-6.676	2.E-06
rpm	1	0.099829	0.00310756	32.125	5.E-18
Thick	1	0.015245	0.00283069	5.385	3.E-05
Position	1	-0.013683	0.00356198	-3.841	0.00110
Position*position	1	-0.078724	0.00586715	-13.418	4.E-11

TABLE 8

Predicting Flux from Beveled Promoter; Dependent Variable: LFLUX					
Analysis of Variance					
Source	DF	Sum of Squares	Mean Square	F Value	Prob>F
Model	8	0.22712	0.02839	286.118	9.E-16
Error	16	0.00159	0.00010		
C Total	24	0.22871			
Root MSE	0.00996	R-square	0.9931		
Dep Mean	2.23455	Adj R-sq	0.9896		
C.V.	0.44578				
Parameter Estimates					
Variable	DF	Parameter Estimate	Standard Error	T for H0: Parameter=0	Prob > T
INTERCEPT	1	2.357661	0.02526225	93.327	3.E-23
rpm	1	0.086064	0.00392466	21.929	2.E-13
Pressure	1	0.008333	0.00232620	3.582	0.00249
Thick	1	0.016529	0.00206471	8.005	5.E-07
Position	1	-0.009665	0.00266180	-3.631	0.00225
rpm*pressure	1	0.008553	0.00243450	3.513	0.00288
Time*time	1	-0.032001	0.00729315	-4.388	0.00046
rpm*rpm	1	-0.089053	0.02961581	-3.007	0.00836
Position*position	1	-0.059030	0.00425851	-13.862	2.E-10

the same solids concentration (27%) with the beveled promoter, but produced very high flux rates. Since the slurry used in these new tests was from a new batch of solids and the membranes had been idle for several months, the results were believed to be in error and were not amended to the database.

The models reported in Tables 5–8 are in a coded format and not directly applicable to engineering units. After making the appropriate substitutions and mathematical transformations, the final format of the equations are given in Equations 1–4, respectively, for the tabular models.

Equation 1:

$$\text{kW, nonbeveled} = -1.1947 + 0.002029(\text{rpm}) - 0.1231(\text{thick}) + 1.584\text{E-}4(\text{rpm})(\text{thick}) - 0.06828 (\text{position})^2$$

Where:

Thick = -1 (0.120") or 1 (0.1875")
rpm = 900 to 1200
Position = -1 (reverse), 0 (original), 1 (forward)

Equation 2:

$$kW, \text{ beveled} = -1.178 + 0.001991(\text{rpm}) - 0.08039(\text{position})^2$$

Where:

rpm = 900 to 1200
position = -1 (reverse), 0 (original), 1 (forward)

Equation 3:

$$\text{Log}_{10}(\text{Flux}), \text{ nonbevel} = 1.5612 - 6.644\text{E-}4(\text{time}) + 6.115\text{E-}4(\text{rpm}) + 0.01524(\text{thick}) - 0.01368(\text{position}) - 0.07872(\text{position})^2$$

Where:

Time = 0 to 100 hr
Thick = 1 (0.120") or 1 (0.1875")
rpm = 900 to 1200
Position = -1 (reverse), 0 (original), 1 (forward)

Equation 4:

$$\begin{aligned} \text{Log}_{10}(\text{flux}), \text{ bevel} = & -4.464 + 0.007309(\text{rpm}) - 0.004462(\text{P}) + 0.01653(\text{thick}) - \\ & 0.009665(\text{position}) \\ & + 4.9821\text{E-}6(\text{rpm})(\text{P}) + 0.01193(\text{time}) - 1.280\text{E-}5(\text{time})^2 - 3.3312\text{E-} \\ & 6(\text{rpm})^2 \\ & - 0.05903(\text{position})^2 \end{aligned}$$

Where:

Time = 418 to 514 hr
Thick = -1 (0.120") or 1 (0.1875")
rpm = 900 to 1200
Pressure = 40 to 60 psig
Position = -1 (reverse), 0 (original), 1 (forward)

3.1.7 Economic Analysis

An analysis was performed to 1) determine the approximate cost of filtering 1 gal of liquid waste with the SpinTek system and 2) compare this cost to the cost of other, competing

technologies. This information can be used to identify situations in which the SpinTek technology could be most effectively applied.

The economic analysis estimated costs (excluding profit) for remediation using a SpinTek system with 10 double-sided disks producing 1,331,520 gal of permeate per year. With realistic costs and knowledge of the bases for their determination, it should be possible to estimate the economics for operating similarly sized systems at other sites. Annual costs were determined for the categories used when the Environmental Protection Agency (EPA) compares the economics of various competing technologies. These categories are as follows:

- **Site Preparation** – Site preparation responsibilities include site design and layout, surveys and site logistics, legal searches, access rights and roads, and preparations for support facilities, decontamination facilities, utility connections and auxiliary buildings. The amount of preliminary preparation depends on the site and, except for costs directly related to the SpinTek system, were considered to be performed by the responsible party or owner. The SpinTek-specific costs were assumed to include construction of a concrete pad and shelter for the equipment as well as plumbing and electrical hookups. These costs were estimated to total \$20,000.
- **Permitting and Regulatory** – Permitting and regulatory costs include actual permit costs, system health and safety monitoring, and analytical protocols. These costs can vary greatly because they are site- and waste-specific. For this analysis, the permitting and regulatory costs were considered to be too variable and were not included.
- **Equipment** – The capital cost of the equipment was amortized over the life of the unit at 6% simple interest. A 10-disk SpinTek unit is estimated to cost approximately \$150,000. Assuming a 10-year useful life, the per year equipment cost would be \$20,380.
- **Start-Up** – Start-up costs include transportation of the SpinTek unit(s) to the site, setup of the unit(s), shakedown of the system prior to site remediation, and Occupational Safety and Health Administration (OSHA) health monitoring training. Shipping costs were assumed to be \$1000; setup and shakedown lasting 3 days total for two workers at \$50/hr for a cost of \$2400; OSHA training at \$1000/person, and medical surveillance at \$500/person. The total of all start-up costs is \$8400.
- **Labor** – Labor costs include remediation labor, site supervision, and can also include hiring and relocation expenses. Assuming 24-hr operation, labor costs were estimated to total \$100,660.
- **Consumables and Supplies** – Consumables and supplies include health and safety gear for the employees (about \$1500/person) as well as maintenance supplies, which were estimated at 1% of the amortized capital cost (16). Consumables and supplies totaled \$3200.
- **Utilities** – The amount of electricity needed to operate a 10-disk SpinTek system at 60 psig and 1200 rpm is estimated to cost about \$1000/year.

- Effluent Treatment and Disposal – Treatment and disposal of effluent (i.e., concentrate) from the SpinTek system may be required, depending on the nature of the contaminants at the site. It was assumed that the SpinTek system would produce a concentrate stream that was 30% of the contaminated feed stream. For a permeate production of 1,331,520 gal, the concentrate would total 570,651 gal. At a disposal cost of \$0.82/gal, effluent treatment and disposal would total \$467,934.
- Residuals/Waste Shipping, Handling and Transport – Waste disposal costs including storage, transportation, and treatment costs are assumed to be the obligation of the responsible party or site owner. About twelve 55-gal drums of solid waste (such as contaminated health and safety gear and used materials) would be produced each year. If these were disposed of in a landfill at a cost of \$200/drum, waste disposal costs would total about \$2400.
- Analytical – It is assumed that one sample would be tested each day for contaminant content. If each analysis costs \$100/sample, the total analytical costs would be \$36,500.
- Facility Modification, Repair, and Replacement – It is assumed that the responsible party or owner would pay the costs of any site modifications or repairs. Repair and replacement as it applies to the SpinTek system is estimated to be 5% of the amortized capital cost of the equipment, or \$1020 (16).
- Demobilization – Site demobilization includes shutdown of the operation, final decontamination and removal of equipment, site cleanup and restoration, permanent storage costs, and site security. These costs were estimated to be approximately \$5000.

The estimated SpinTek cost in each category is summarized in Table 9.

The SpinTek costs were compared to competing filtration technologies, including the following:

- Membrane Filtration, SBP Technologies, Inc. – This is a membrane-based separation technology. The filtration unit consists of porous, sintered, stainless steel tubes arranged in a shell-and-tube module that is operated in crossflow mode. Multilayered inorganics and polymeric “formed-in-place” membranes are coated at microscopic thickness on the inside of the porous stainless steel tubing during recirculation of a slurry of membrane formation chemicals. The membrane functions as a hyperfiltration unit and can be readily modified to conform to waste characteristics and separation requirements. SBP Technologies’ system is applicable to groundwaters and process waters, especially those contaminated with polyaromatic hydrocarbons (PAHs) (17).

TABLE 9

Summary of Annual SpinTek Costs in Each Category	
Category	Annual SpinTek Cost, \$
Site Preparation	20,000
Permitting and Regulatory	Not calculated; site- and contaminant-dependent
Equipment (amortized over useful life at 6% interest)	20,380
Start-Up	8400
Labor	100,660
Consumables and Supplies	3200
Utilities	1000
Effluent Treatment and Disposal	467,934
Residuals/Waste Shipping, Handling, and Transport	2400
Analytical	36,500
Repair and Replacement	1020
Demobilization	5000

- Microfiltration Technology, E.I. DuPont De Nemours and Oberlin Filter Co. – The microfiltration technology combines Oberlin’s automatic pressure filter with DuPont’s microporous Tyvek® filter media. Two products are produced: filter cake and filtrate. A filter aid or filter aid/cake-stabilizing agent can be used with the system. The technology is suitable for treating landfill leachate, groundwater, and liquid industrial wastes containing metals (18).
- Rochem Disc Tube™ Module System, Rochem Separation Systems, Inc. – The Rochem Disc Tube™ Module System uses osmosis through a semipermeable membrane to separate pure water from contaminants. Waste materials that can be treated with this system include sanitary and hazardous landfill leachate containing both organic and inorganic chemical species (19).
- Colloid Polishing Filter Method, Filter Flow Technology, Inc. – In this technology, specially designed filter plates are used to support filter packs containing FF1000, an insoluble, inorganic, oxide-based, granular material that removes radionuclides and heavy metals from moderately contaminated water through a combination of chemical complexing, adsorption, absorption, and filtration (20).

Each of these technologies underwent an economic analysis according to the EPA cost categories, meaning that the results should be comparable with each other as well as the SpinTek cost analysis results. The costs for each category are summarized in Table 10.

TABLE 10

Comparison of the Costs of Filtration Using Various Competing Technologies					
EPA Cost Category	Company				
	SBP Technologies, Inc.	DuPont/Oberlin Filter Co.	Rochem Separation Systems, Inc.	Filter Flow Technology, Inc.	SpinTek, Inc.
Site Preparation ^a	\$60,000	\$209,200	\$7,020	\$15,000	\$20,000
Equipment ^b	\$53,740	\$47,800	\$41,845	\$39,605	\$20,380
Start-Up	\$5,000	\$80,000	\$30,620	\$1,000	\$8,400
Labor	\$199,080	\$133,400	\$34,700	\$28,000	\$100,660
Consumables/ Supplies	\$3,500	\$18,900	\$27,555	\$11,900	\$3,200
Utilities	\$10,600	\$5,500	\$11,225	\$800	\$1,000
Effluent Disposal	\$533,000	\$15,000	\$377,419	-- ^c	\$467,934
Residual/Waste Shipping, Handling, and Transportation	\$46,000	\$3,700	\$2,800	\$11,600	\$2,400
Analytical	\$60,000	\$36,000	-- ^d	\$24,000	\$36,500
Repair and Replacement	\$37,150	\$2,500	\$5,105	\$5,000	\$1,020
Demobilization	\$10,000	\$30,000	-- ^e	(\$20,000) ^f	\$5,000
Total	\$1,028,070	\$582,000	\$538,289 ^g	\$116,905	\$666,494
Gallons Permeate	2,600,000	525,600	1,380,800	52,400,000	1,331,520
Cost per Gallon Permeate (including all available costs)	\$0.40	\$1.11	\$0.39	\$0.002	\$0.50
Cost per Gallon Permeate (not including demobilization costs)^h	\$0.39	\$1.05	\$0.39	\$0.003	\$0.50

^a Only includes costs specifically associated with filtration equipment.

^b Amortized over the equipment life.

^c Included in the analytical costs.

^d On-line analytical instrumentation included in the equipment costs.

^e Information not available.

^f Demobilization costs would be negative because of the sale of the equipment at the end of the cleanup.

^g This value does not include demobilization costs since they were not available.

^h All values are on the same cost basis.

The “Permitting and Regulatory” category was not included in the cost-per-gallon calculations because values were not provided for all four of the competing technologies. Costs were not provided for Rochem Separation Systems technology’s demobilization needs either. Two costs per gallon of permeate produced were calculated: one including demobilization costs and the other without. The value that includes the demobilization cost is probably closer to the actual cost, but if the Rochem Separation Systems technology is included in the comparison, the costs without demobilization must be used to keep all of the values on the same basis. In a further effort to put all of the costs on the same basis, treatment for all liquid effluent was assumed to be \$0.82/gal, and solid waste disposal was assumed to cost \$200/drum.

The cost of using SpinTek technology to concentrate liquid waste compares favorably with the cost of the competing technologies. Because of the nature of the cost estimations used for each category, the actual difference in cost between the SpinTek, SBP Technologies, and Rochem Separation Systems processes is probably quite small. The magnitude of the differences between SpinTek’s cost (\$0.50/gal permeate) and that of either DuPont/Oberlin’s process (\$1.05/gal permeate) or Filter Flow Technology’s process (\$0.003/gal permeate) indicates that they are, in fact, different from SpinTek.

Variations in waste type and concentration can significantly affect the costs of analytical workups and treating/disposing of effluent and residuals. The costs of equipment setup and operation (i.e., analytical, effluent, and residual costs were not included) were determined and are compared in Table 11. These results confirm the conclusions drawn from the total cost comparison: the SpinTek, SBP Technologies, and Rochem Separation Systems technologies cost about the same; Filter Flow Technology’s process is less expensive; and DuPont/Oberlin’s process is significantly more expensive.

A final comparison of the various filtration technologies was made by comparing the percentage of total cost of each of the categories for a given technology. These values are shown in Table 12. While the table shows that by far the largest fraction of the total cost usually comes from the labor and effluent disposal categories, the comparison yielded some interesting observations. Relative to the other technologies, the DuPont/Oberlin process cost of producing permeate was very dependent on site preparation and equipment start-up costs. Labor costs made up a smaller-than-average part of the Rochem Separation Systems process. The cost of permeate production from the Filter Flow Technology process was greatly affected by the costs of equipment and sample analyses. The processes of both SBP Technologies and Filter Flow Technology had a similarly high percentage of cost associated with residual/waste disposal costs. SpinTek costs for each category were within the range of the costs for the other technologies.

The cost of using SpinTek technology compares favorably with other ultrafiltration technologies that are currently available. The choice to use SpinTek rather than another technology would depend on a number of factors, including volume of waste, contaminant types, quantity of contaminant to be removed, etc. The SBP Technologies system is applicable to ground- and process water contaminated with organics having molecular weights over 200 (such as polyaromatic hydrocarbons). The DuPont/Oberlin technology is applicable to aqueous streams that are contaminated with metals and particulate. The Rochem technology can be used to treat

TABLE 11

Comparison of the Costs of Equipment Operation for Competing Filtration Technologies

EPA Cost Category	Company				
	SBP Technologies, Inc.	DuPont/Oberl in Filter Co.	Rochem Separation Systems, Inc.	Filter Flow Technology, Inc.	SpinTek, Inc.
Site Preparation ^a	\$60,000	\$209,200	\$7,020	\$15,000	\$20,000
Equipment ^b	\$53,740	\$47,800	\$41,845	\$39,605	\$20,380
Start-Up	\$5,000	\$80,000	\$30,620	\$1,000	\$8,400
Labor	\$199,080	\$133,400	\$34,700	\$28,000	\$100,660
Consumables/Supplies	\$3,500	\$18,900	\$27,555	\$11,900	\$3,200
Utilities	\$10,600	\$5,500	\$11,225	\$800	\$1,000
Repair and Replacement	\$37,150	\$2,500	\$5,105	\$5,000	\$1,020
Total	\$369,070	\$497,300	\$158,070	\$101,305	\$154,660
Gallons Permeate	2,600,000	525,600	1,380,800	52,400,000	1,331,520
Cost per Gallon Permeate	\$0.14	\$0.95	\$0.11	\$0.002	\$0.12

^a Only includes costs specifically associated with filtration equipment.

^b Amortized over the equipment life.

sanitary or hazardous landfill leachate contaminated with both organic and inorganic chemical species. The Filter Flow Technology process removes heavy metals and radionuclides from water. The SpinTek system was designed for high-solids, high-viscosity streams. Because its use is not limited to specific contaminants, the SpinTek process is applicable to a wider variety of liquid waste streams than the other technologies. One ultrafiltration opportunity for which SpinTek is well-suited is low-level radioactive waste from decontamination activities or tank waste. According to a check of the DOE EM needs, there are 90.6 million gallons of tank waste requiring treatment and 15 weapons complex needs that require the type of polishing filtration for which SpinTek was designed, indicating a high demand for this type of filtration. SpinTek's wide applicability, effectiveness, and favorable economics make it an attractive ultrafiltration process.

TABLE 12

Percentages of the Cost Categories for Filtration Using Various Competing Technologies

EPA Cost Category	Company				
	SBP Technologies, Inc.	DuPont/Oberlin Filter Co.	Rochem Separation Systems, Inc.	Filter Flow Technology, Inc.	SpinTek, Inc.
Site Preparation ^a	5.9%	37.9%	1.3%	11.0%	3.0%
Equipment ^b	5.3%	8.7%	7.8%	28.9%	3.1%
Start-Up	0.5%	14.5%	5.7%	0.7%	1.3%
Labor	19.6%	24.2%	6.4%	20.5%	15.2%
Consumables/Supplies	0.3%	3.4%	5.1%	8.7%	0.5%
Utilities	1.0%	1.0%	2.1%	0.6%	0.2%
Effluent Disposal	52.4%	2.7%	70.1%	-- ^c	70.7%
Residual/Waste Shipping, Handling, and Transportation	4.5%	0.7%	0.5%	8.5%	0.4%
Analytical	5.9%	6.5%	-- ^d	17.5%	5.5%
Repair and Replacement	3.6%	0.5%	0.9%	3.7%	1.5%

^a Only includes costs specifically associated with filtration equipment.

^b Amortized over the equipment life.

^c Included in analytical costs.

^d Included in equipment cost.

3.2 Phase 3 Technology Partnering

SpinTek's centrifugal membrane filtration is a crosscutting technology with a number of applications for cleanup of DOE weapons complex liquid wastes. One of these applications is the enhancement of downstream unit operations such as adsorption or ion exchange processes where even low levels of suspended solids create operational problems.

3M has developed technologies that are capable of selectively removing dissolved radionuclides from liquid wastes. A limiting factor in the effectiveness of the 3M technology is the accumulation of suspended materials that decrease throughput, creating plugging well before the cartridges are completely utilized with respect to removal capability. Testing was conducted to evaluate improvements to the 3M WWL Tc removal cartridges using SpinTek's centrifugal membrane filtration as a pretreatment to removal of suspended material.

3.2.1 Test Procedures

Testing was conducted to evaluate 3M WWL Tc cartridge performance under three different conditions: 1) without prefiltration, 2) with prefiltration using conventional cartridge-style paper prefilters, and 3) with prefiltration using the SpinTek centrifugal membrane filtration technology.

Water used in the tests was collected from the English Coulee, a tributary of the Red River of the North that flows through the campus of the University of North Dakota. The water was intended to represent an impounded water at DOE facilities and had a relatively low suspended solids concentration. Total suspended solids during test trials without prefiltration were measured to be 11 mg/L. During trials with conventional prefiltration and SpinTek prefiltration, the total suspended solids concentration was 35 mg/L.

All test trials were conducted at a constant flow rate of 10 gal/hr. English Coulee water was pumped from a 1500-gal polyethylene storage tank that served as a feed storage reservoir. During trials conducted without prefiltration, the water was pumped through the 3M WWL Tc removal cartridge and recirculated back to the storage tank. Inlet pressure was monitored throughout the duration of the test trial.

Tests using conventional prefiltration and SpinTek centrifugal membrane filtration were then conducted in a parallel mode. A block flow diagram is shown in Figure 6. English Coulee water was pumped directly from the feed water storage tank through two conventional filters prior to the 3M WWL cartridges. These filters had nominal pore sizes of 5 and 0.1 μm , respectively. Filtered water was then passed through the 3M WWL cartridge and recirculated back to the feed storage tank. Pressure indicators were placed at the inlet side of each of the filter cartridges to monitor pressure increases in each filter because of accumulation of suspended solids. Filter inlet pressure readings were data-logged to a computer at 30-min intervals throughout the duration of the test runs.

Parallel test trials using SpinTek prefiltration also used English Coulee water pumped directly from the feed storage tank. Permeate from the SpinTek ST-IIL unit was directed to a 30-gal surge tank before being pumped through a 3M WWL cartridge. Concentrate from the ST-IIL unit was directed back to the feed storage tank. As with the conventional prefiltration system, inlet pressure was monitored throughout the test run and data logged at 30-min intervals. The SpinTek ST-IIL unit was operated at ambient temperatures at a feed rate of 600 L/hr, a pressure of 60 psig, and a rotor speed of 1200 rpm.

Figure 6. Prefiltration test block flow diagram.

3.2.2 Test Results

3.2.2.1 No Prefiltration

The test with no prefiltration was run for approximately 1 hr, shut down, and restarted approximately 5 days later. The test run was operated for a total of approximately 4.5 hr, at

which time the inlet pressure reached 70 psig. Total throughput at test termination was approximately 38 gal. Figure 7 shows the inlet pressure versus total throughput without prefiltration. As shown in Figure 7, the inlet pressure diminished to less than 5 psig upon test continuation.

3.2.2.2 Conventional Prefiltration

The conventional prefiltration test run was operated for approximately 102 hr, during which time the 5- μ m filter had to be replaced three times because of filter plugging. Although the inlet pressure of the 5- μ m filter reached a maximum allowable value of approximately 70 psig three times, no significant pressure increase was observed at either of the other filters.

At test conclusion, approximately 748 gal of total throughput had been achieved with an average throughput for each 5- μ m filter of approximately 250 gal. The observed inlet pressure versus total throughput volume is presented in Figure 8.

Figure 7. Pressure versus throughput – no prefiltration.

3.2.2.3 SpinTek Prefiltration

The SpinTek prefiltration test run was operated for approximately 240 hr, at which time a recycle line failure prompted the end of the test. After 240 hours of operation and approximately 2400 gal of total throughput, no significant pressure increase had been observed at the inlet of the Tc filter. Figure 9 displays the measured inlet pressure over time during the SpinTek prefiltration test run.

Based on the test run data, prefiltration was required for application with the raw water. As shown during the parallel prefiltration test run, the SpinTek prefiltration system was much more efficient at removing suspended solids prior to delivery to the Tc-specific cartridge. If the conventional prefiltration system had been operated for the 240 hr as the SpinTek prefiltration system had been operated, approximately ten 5- μ m filters would have been required to achieve a throughput of 2400 gal (one 5- μ m filter per 250 gal of throughput).

3.2.3 Cost Analysis of Conventional Prefiltration

The cost of using conventional prefiltration was calculated for comparison to the cost of using the SpinTek process. The cost categories used during the SpinTek economic analysis were also used for this analysis. As Table 13 shows, the cost of conventional prefiltration is \$0.27/gal permeate, which is approximately the same as the cost of ultrafiltration using a SpinTek system.

Figure 8. Pressure versus throughput – conventional prefiltration.

Figure 9. Pressure versus throughput – SpinTek prefiltration.

3.2.4 Conclusions

Based on the test run with no prefiltration, prefiltration was required for application with the raw water. As shown during the side-by-side prefiltration test run, the SpinTek prefiltration system was much more efficient at removing suspended solids prior to delivery to the Tc-selective filter.

Had the paper prefiltration system been operated for the 240 hr at which the SpinTek prefiltration system had operated, approximately ten 5- μ m filters would have been required to achieve a throughput of 2400 gal (one 5- μ m filter per 250 gal of throughput).

By plotting the inlet pressure versus throughput volume for the SpinTek prefiltration test run and extrapolating exponentially into the future until the inlet pressure reaches 70 psig, the following conclusions could be made:

- The test run would operate for approximately 121 days.
- At 10 gph, the total throughput would be approximately 29,000 gal.

TABLE 13

Summary of Annual Conventional Prefiltration Costs

Category	Annual Prefiltration Cost, \$
Site Preparation	15,000
Equipment (amortized over 10 yr at 6% interest)	610
Start-up	6000
Labor	50,330
Consumables and Supplies	84,223
Utilities	250
Residuals/Waste Shipping, Handling, and Transport	160,200
Analytical	36,500
Repair and Replacement	30
Demobilization	2,500
Total Cost	355,640
Cost/Gallon Permeate	0.27

Based on the exponential extrapolation (29,000 gal of total throughput) and using 250 gal of throughput per 5- μ m filter, the paper prefiltration system would require 116 5- μ m filter replacements.

Although the test run did show that the SpinTek prefiltration system is far more effective than the traditional paper prefiltration system for the supply water used, the validity of an exponential extrapolation is hypothetical and would be extremely dependent on the makeup of the supply water. In addition, the Tc-selective filter may reach breakthrough prior to prefilter plugging.

Using economic feasibility to determine which of the prefiltration systems to use would be dependent on several factors, most important of which are the volume of raw water to be treated and the physical and chemical makeup of the raw water. For small volumes of raw water or raw water containing low suspended solids, the common paper prefiltration system may be adequate. As treated volumes and suspended solids increase, the economics begin to favor the SpinTek prefiltration system.

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