ENHANCED PARTICLE REMOVAL
FROM LIQUIDS

K. M. KALUMUCK, G. L. CHAHINE, P. ALEY and G.S. FREDERICK
DYNAFLOW, Inc.
7210 Pindell School Road
Fulton, Maryland 20759

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Abstract

Technical Summary: The feasibility of significantly improving solid particle removal from liquids using microporous tubes with flow swirl and flow interruption was demonstrated. Experiments were performed with laboratory scale modules using particle laden water obtained from an actual field project of high pressure water jet paint stripping and mixtures of rust and water. The new filtration concept generated flux rate increases of a factor of 20 over conventional microporous tube cross flow filtration with comparable or improved effluent quality. The flow rate, orientation, and degree of swirl were varied, and their effect on performance observed and measured as a function of time. These effects were analyzed using a mechanistic approach. Measured data using 1, 3, and 7 tube modules showed no significant effect due to an increased number of tubes. Results obtained from the Phase I research clearly demonstrate the feasibility of this new filtration technology to greatly improve the efficiency, capacity and economy of particle removal from liquids.

Commercial Potential: The new filtration technology meets the need for efficient and economical particle removal from the effluent and process streams of industrial, government, and commercial plants. It has applications in advanced processes of materials production, manufacturing, and biotechnology. It can separate particles from liquid streams for further process use or discharge to the environment and reclaim particles such as catalysts, precious metals or reaction products. It has broad application for suspended solids removal including toxic heavy metals in areas such as water jet stripping of paint, scale and coatings; battery manufacturing; metal finishing, plating, smelting and refining; and other industries with similar waste waters. The very high flux rates will make it particularly competitive and attractive for high volume applications and strongly indicate the potential for a viable commercial product.
RESEARCH OBJECTIVES FROM PHASE I PROPOSAL

“Brief preliminary experiments had indicated the promise for substantial improvements in the separation of particles from liquid streams through the use of the concepts of swirling and periodic interruption of the feed stream and change in flux direction across microporous tubes. In Phase I, these concepts will be investigated in detail to demonstrate the feasibility of using such techniques and to verify and quantify their advantages. The specific objectives of Phase I include:

1. Develop a mechanistic model to explain and predict observed performance and effects, perform parametric studies, guide selection of experimental configurations, and assist in scale up predictions.

2. Determine the conditions under which swirl of the feed flow with a single porous tube increases flux rates. Quantify the improvement. Investigate a range of governing parameters suggested by the modeling.

3. Determine the conditions under which periodic flow interruption alone and in conjunction with feed flow swirl around a single porous tube leads to increased flux rates and time between required cleanings. Quantify the improvements.

4. Using the most promising configurations from the single tube investigations, determine the effects of employing multiple tubes at laboratory scale.

5. Using the results of the single and multiple tube laboratory scale work and predictions of the modeling, estimate the performance at “full” scale.

Phase I will lay a firm base for more detailed research in Phase II by demonstrating feasibility and providing a preliminary mechanistic explanation of performance and a predictive model. Estimated scaled up performance improvements and Phase I experimental data will be useful in attracting the interest of a commercial partner for Phase III. In Phase II, a more detailed investigation, for a range of particle laden feed flows of practical interest, will be performed. The modeling will be refined and expanded with particular emphasis on practical applications and larger scale systems and inclusion of more detailed physics that may be suggested by experiment. Phase II experiments may include more detailed measurements such as velocity fields. The primary objective of Phase II then would be to obtain sufficient understanding to enable work, in Phase III, with a commercial partner towards actual application at “pilot unit” scale. As such, Phase II research will emphasize scaling, more detailed physical understanding, performance prediction, and determination and demonstration of specific practical applications.”

OVERVIEW OF RESEARCH AND RESULTS

The potential for significant improvement of solid particle filtration rate using a combination of microporous tubes, flow swirl and flow interruption was investigated in this NSF Phase I SBIR project. Experiments were performed with three different modules using 1, 3 and
7 microporous tubes per module. Results were compared to those obtained using the conventional HYDROPERM™ cross-flow filtration method. Initial experiments enabled selection of a basic module configuration and orientation. Two types of real-life waste waters were employed. Waste water from the high pressure water jet stripping of paint and rust from a bridge from an actual field project was obtained and utilized. This water contained particulate contaminants of lead, zinc, and rust among others due to the nature of the paint removed. A second waste water was obtained in the laboratory from a mixture of well water and rust from the tanks of a cavitation tunnel. The flow rate and the amount of swirl were varied and periodic flow interruption was employed. Tests included both short (2 hours) and long duration (24 hours) filtration. Flux rates were measured as a function of time. Measurements were made of the quality of the feed and effluent to assess contaminant removal. These included pH, light transmittance, total dissolved solids and suspended solid measurements. In addition, visual observations were made through the transparent experimental modules. Measurements indicate an improvement in the plateau value of the flux by a factor of 20 over conventional HYDROPERM™ with a comparable or improved effluent quality (suspended solids content). A mechanistic approach was used to analyze the data. The importance of the governing parameters on performance was analyzed based on the experimental results. Recommendation for future development of the system are given.

Results obtained from the Phase I research unequivocally demonstrate the feasibility of this new filtration technology termed DYNAPerm™ to greatly improve the efficiency, capacity and economy of particle removal from liquids. Phase II research should enable the creation of a very successful proprietary filtration system with ready commercial applications.

POTENTIAL APPLICATIONS

The new filtration technology called DYNAPerm™ meets the increasingly important need for efficient and economical particle removal from the effluent and internal process streams of industrial, government, and commercial plants. This need is driven by more stringent local and federal regulations, concerns for the environment and the need for “cleaner” process feed streams within plants. This would allow control of product quality and use of more sophisticated process technology in order to stay competitive on both national and international levels. Such capabilities are particularly important for new and advanced processes in materials production, manufacturing, and biotechnology. The DYNAPerm™ technology can be used to separate particles from liquid streams to “clean” the liquid – for further process use or for discharge into the environment; to collect or reclaim the particles – e.g., catalysts, precious metals or reaction products; and to separate the two phases. It has broad application for removal of suspended solids including toxic heavy metals in a wide range of industries, including battery manufacture, metal finishing, smelting and refining, water jet stripping of paints and coatings, and other industries with similar wastewaters.

1. Background

Filtration processes can be divided into two general categories: cross-flow and through-flow filtration. In through-flow filtration, both feed and filtrate flows are normal to the surface of
Table 1: *Results of solids removal tests by Hydromer™ cross-flow filtration [5].*

In cross-flow filtration, the feed flow is parallel to the filter surface, with the filtrate permeation occurring perpendicular to the flow. A quasi-steady operation is possible, since the continuous build-up of the separated solids on the filter surface is largely prevented by the hydrodynamic shear exerted by the cross-flow (Figure 1). Examples include microfiltration and membrane filtration such as ultrafiltration and reverse osmosis. Cross-flow microfiltration which removes primarily suspended solids, is significantly different from membrane *ultrafiltration* (UF) or *hyperfiltration* (RO) which remove substances on the molecular level in addition to suspended solids. In UF, higher filtration pressures (50 psi) are used, with even higher pressures (600-1200 psi) for RO, compared with only 15 psi for cross-flow microfiltration. Furthermore, UF and RO employ relatively *thin* membranes, compared with the in-depth, relatively thick-walled (1mm) cross-flow microfilters (such as *Hydromer™*). As a result, power requirements as well as capital and operating costs are much higher for UF and RO membrane systems. Under the relatively harsh conditions of industrial filtration applications, membrane systems are prone to clogging, fouling and leaks leading to unacceptably low filtrate flux levels. Cleaning is complicated and cost is relatively high. A major advantage of the cross-flow filtration is its high filtration rate.

In the *Hydromer™* microfiltration concept the walls are microporous, with the pore structure and sizes being controlled during the manufacturing process [1-4]. This technology has been successfully used in removal of solids. Table 1 provides results of some tests conducted on battery and electroplating wastes [5]. Because of their basic ruggedness as well as their chemical and biological inertness the tubes are not susceptible to the handling, fouling and cleaning problems associated with membrane systems. These tubes can be made from a variety of extrudable thermoplastics.

In this report, a laboratory scale investigation of novel means of increasing the filtrate flux...
rate of the conventional HYDROPERM™ tube system was investigated. Results indicate very large increases in the filtration (flux) rates and a substantially easier means of removal of particle build up on the tube walls.

2. Description of the DYNAPerm™ Concept

2.1. Novel Aspects

The DYNAPerm™ cross flow filtration concept combines the conventional HYDROPERM™ microfiltration concept with several major modifying and enhancing mechanisms. These enhancements include:

1. **Reversing the filtrate flow direction to flow across the tube from its outside to its inside.** This results in three major improvements:
   
   a) The effective area for particle deposition is significantly increased. For example, for the tubes used in this study of inside diameter 6 mm and outside diameter 8 mm, the modification in the filtration direction results in an increase of the particle capture area by more than 33%. In addition this area, as opposed to the conventional case, increases with particle deposition and “dynamic membrane” build-up.
   
   b) Cleaning or removal of the deposited particles on the tube wall is made easier.
   
   c) The problem of clogging of the inside of the tubes is eliminated.

2. **Combining swirl with cross-flow filtration.** As sketched in Figures 4a and 4b in Section 4, the flow is forced tangentially into the cylinder containing the tubes, and leaves it either through the microporous tube or through an opening in the axial direction, thus creating a cyclone type flow inside the module. As discussed in the theoretical section this has three advantages:

   a) It increases the residence time of the particles in the filtration module.
   
   b) It increases the effective shear at the tube wall which now has a tangential and an axial component (velocities $V_t$ and $V_z$), so that the effective local velocity is $\sqrt{V_t^2 + V_z^2}$.
   
   c) It retards particle motion towards the wall through centrifugal forces, therefore reducing particle concentrations near the tubes and particle deposition on the walls. This results in increased filtration rates.

3. **Adding of flow interruption to the filtration process.** This has the advantage, through sudden stopping of the flow, of enhancing tube cleaning by entrainment in the flow of particles deposited on the outside of the tube.

4. A further advantage of the DYNAPerm™ that was not investigated in this work, and could be of significant importance for some applications, is the enhanced potential for particle recovery. Cycling flow interruption would result in particles falling off the tubes. An adequate particle collection scheme would need to be devised.
2.2. “Dynamic Membrane” Concept

The filtration characteristics of the porous HYDROPERM™ / DYNAPERM™ tubes combine both the “in-depth” filtration aspects of multimedia filters and the “thin-skinned” surface filtration aspects of membrane ultrafilters. For example, while the removal of micron-sized particles and colloids is often impossible with conventional through-flow filters, microporous tubes are capable of removing such particles. On the other hand, in a manner similar to multimedia filters, the tubes will allow the smaller particles and colloids in the waste streams to actually penetrate into their wall matrix. The tube pore structure differs from those of membrane ultrafilters in that the pore sizes of the former are of the order of several microns, with the pore “length” many times their diameter.

Pore-size distributions of two typical tubes are shown in Figure 2a. Tube II has a “flat” distribution with pore size ranging from 3 to 9 microns. Tube I has a “peaked” distribution, with most of the pores being in the 2-micron range. Figure 2b shows a photograph of the pore structure of a typical transverse tube section taken with a scanning electron microscope. The open-cell, reticulated nature of the pore structure is apparent. These features are of importance in determining the performance of a given tube when used with a specific effluent.

In general, any effluent from which suspended solids removal is desired will contain a wide range of particle sizes. When such effluents are circulated through the tubular filter, the solid particles will be slowly driven, with the permeating flow, toward the wall. The particle concentration near the wall will steadily increase, this tendency being limited by the turbulent diffusion of the particles from regions of high concentration to those of lower concentration (that is, away from the tube walls toward the center of the tube for the HYDROPERM™ system and towards the module tangential injection wall for the DYNAPERM™ system). The turbulent diffusion is dependent on the shear stress exerted on the walls by the flow, and, hence, its velocity. The permeation rate tends to increase the particle concentration near the wall and depends on the pressure differential across the filter surface and the tube pore structure (Darcy’s law). As described in more detail in the following section, a quasi-steady state profile of particle concentration (“particle polarization”) will be established near the wall when these two opposing tendencies balance each other.

Because of the in-depth filtration characteristics of the tubes, other factors also come into play. Only particles which are smaller than the largest pore size can enter the wall matrix. Thus as filtration proceeds, the pore structure of the tube as well as its permeability will undergo a gradual change. The tendency of new particles to enter the tube matrix will decrease as a fine, dynamic filter cake or “dynamic membrane” forms on the walls due to the particle polarization described earlier. As a result, the initially high flux rate of filtrate across the tube walls at first rapidly decreases until attaining a slowly decreasing “plateau” due to build up of particles along and within the wall. These considerations need to be kept in mind to explain the experimental results observed later in this report.
One should also note that the filtration performance is influenced not only by such factors as the filtration pressure, circulating flow velocity and temperature (which changes the fluid viscosity and, hence, by Darcy’s law, the permeation rate), but also by the pore size distribution and structure and the particle size distribution.

3. Discussion of Physical Mechanisms

3.1. “Dynamic Membrane” modeling

The “dynamic membrane” described in Section 2.2 can be modeled as the net result of particle deposition on the tube wall due to flow through the wall, and a particle removal due to entrainment by the shear flow parallel to the wall. The permeate flux rate across the tube wall, $F$, can be expressed as a function of the pressure drop across the wall and cake, $\Delta P$, and time, $t$, as:

$$F(t) = \frac{\Delta P}{R_t + r_m \delta(t)}.$$  \hspace{1cm} (1)

Here $R_t$ is the resistance of the tube while $r_m$ is the resistance per unit thickness of the dynamic membrane. The membrane composed of highly concentrated particles starts forming when the filtration begins. Its thickness, $\delta$, grows with time, increasing the tube resistance by an amount $\delta r_m$. (In this analysis, the effects of gravity settling are assumed negligible at normal filtration rates. This may not be true at lower filtration rates and deserves further consideration in Phase II efforts.)

The net particle mass flux to the wall, $q_s$, is given by:

$$\rho_m \frac{d\delta}{dt} = q_s,$$  \hspace{1cm} (2)

where $\rho_m$ is the density of particles in the membrane. This flux arises from the difference between the particle deposition rate, $D$, and entrainment rate, $E$,

$$q_s = D - E$$  \hspace{1cm} (3)

A simple hypothesis regarding deposition rate was introduced by Monin [14]:

$$D = \beta C_b$$  \hspace{1cm} (4)

where $\beta$ is called the reflectivity parameter and $C_b$ is the bulk particle concentration. From measurements of equilibrium conditions, $q_s = 0$, Fukuda and Lick [15] measured $E$ and $\beta$ in a rotating flume for fine grained cohesive sediments 1-10 $\mu m$ in size. Measured values of $\beta$ were of the order of $10^{-2} cm/sec$. Beyond threshold condition, the entrainment rate increased approximately linearly with shear stress at the fluid/cake interface, $\tau_w$. A similar result was been obtained by Parthenaides [16] in linear flume experiments on cohesive soils.

Tulin [17] suggested the following functional form based on data of [15] and [16]:

$$E = A_E p u^*(u^*_b)^{2\gamma}$$  \hspace{1cm} (5)
with $\gamma \approx 1$. Here $u^*$ is the friction velocity defined by

$$u^* = \sqrt{\tau_w / \rho}$$

and $u_b$ is the “binding speed” corresponding to $\tau_b$, the binding stress—i.e., the threshold stress (and corresponding velocity) needed to remove particles from the membrane. For cohesionless particles, $u_b = F$ (the permeate flow holds down the particles). For cohesive particles, $u_b = \sqrt{\tau_b / \rho}$. $A_E$ is a constant of proportionality. The form of $D$ suggested was:

$$D = \eta C_s M_s F = C_b M_s F \eta \frac{C_s}{C_b} = C_b M_s F \left( \frac{F}{u^*} \right)^n A_K f \left( \frac{d}{\nu} \right).$$

Here $C_s$ is the particle concentration next to the wall, $M_s$ is the average particle mass in suspension, $d$ is the particle diffusivity, and $\eta$ and $A_K$ are proportionality constants. This assumes a simple convection, immediately above the membrane with an efficiency factor $\eta$. The concentration ratio $C_s / C_b$ is assumed to depend on a balance between convection toward the surface, and counter-diffusion. The counter-diffusion depends on turbulent processes, which scale with $u^*$, and on diffusive processes which scale with $u^*$, $d$, and $\nu$, where $\nu$ is the kinematic viscosity of the suspension. Here, $f$ is a function to be determined and $\eta$ is a power to be determined.

During the initial flux decline period in which the “dynamic membrane” is being formed, deposition dominates entrainment. Thus Tulin [17] obtained the following relation from the above equations:

$$F \approx \frac{\Delta P \rho_m}{\tau_m M_s A_K f C_b} \left[ \frac{1}{u^*} \right]^{\frac{n}{1+\gamma}} \left[ \frac{t}{\tau} \right]^{-\frac{1}{1+\gamma}}.$$  

This relation was found to be fit well by experimental data [17] with $n=3$:

$$F \sim \Delta P^{0.2} u^{0.6} t^{-0.2}.$$  

Following the initial flux decline period, a plateau of flux and membrane thickness versus time is achieved. This is due to an approximate equilibrium between particle deposition and entrainment. Combining (5,7) and taking $D = E$ and $\gamma = 1$ we obtain for cohesionless particles that the flux varies directly as the friction velocity:

$$F \approx u^* \left( \frac{A_E \rho}{C_b M_s A_K f} \right)^{\frac{1}{1+\gamma}}.$$  

### 3.2. Flow Field

Of particular importance to the performance of the proposed particle separation scheme is the flow field within the swirl chamber. Initially, consider the flow field in the absence of the centerline filtration tubes. Similar flows occur in vortex tubes and cyclone chambers (see, for example [18,19]). In general, the tangential velocity $V_\theta$ has an inner core region of solid body rotation of radius $a$ surrounded by an approximately “free vortex” region (referred to as a Rankine vortex) and finally some effects due to wall boundary layers. Neglecting boundary layer effects, the field is given by:
where $R$ is the swirl chamber radius, and $\Gamma$ is the circulation. The pressure field decreases towards the center and is given in non-dimensional form:

\begin{equation}
\overline{P}(r) = 1 - \frac{\rho}{2P_i} \left( \frac{\Gamma}{2\pi a} \right)^2 \left[ 1 - \frac{1}{2} \left( \frac{r}{a} \right)^2 \right] ; \quad r \leq a, \\
\overline{P}(r) = 1 - \frac{\rho}{2P_i} \left( \frac{\Gamma}{2\pi a} \right)^2 \left( \frac{a}{r} \right)^2 ; \quad r \geq a,
\end{equation}

where pressures are normalized with the ambient pressure outside the vortex chamber, $P_i$.

$V_t$ is the tangential injection velocity and can be varied and is directly determined by the total tangential flow rate, $Q_t$ and the total tangential slot injection area, $A_t$ by:

\begin{equation}
V_t = V_\theta |_{r=R} = \frac{\Gamma}{2\pi r} = Q_t/A_t
\end{equation}

From continuity, the average axial velocity must increase from bottom to top along the chamber since as sketched in Figures 4a and 4b the flow exits from the top and injection occurs along the entire length of the chamber. (In fact with the presence of the microporous tubes at the center of the chamber this velocity increase is damped by loss of fluid through filtration along the tube length). The average axial velocity, $V_z$ when the flux across the porous tube is neglected is then given by

\begin{equation}
V_z = \frac{1}{\pi R^2} \int_0^z V_t h dz = \frac{V_t h z}{\pi R^2}
\end{equation}

where $z = 0$ is at the bottom of the injection slot, $h$ is the total thickness of the injection slots and $V_t$ and $h$ are taken independent of $z$. The relation of the $V_\theta$ distribution with $V_t$ and chamber geometry can be found in [25] where it is shown that the maximum value of $V_\theta$ can be many times $V_t$.

A simple dimensionless “swirl parameter”, $S^*$ characterizing the overall configuration, can be defined as

\begin{equation}
S^* = \frac{V_t |_{z=t}}{V_z |_{z=t}} = \frac{\pi R^2}{hl_t} = \frac{\pi R^2}{A_t}
\end{equation}

The presence of one or more porous tubes at the center of the swirl chamber leads to a somewhat more complicated flow. At $r = r_w$, the radius of the outer tube (or tube bundle) wall/membrane combination, $V_\theta = V_z = 0$ (no slip) and a radial flow, $V_r$ exists that is related to the permeate flux rate, $F$, discussed above:

\begin{equation}
V_r |_{r=r_w} = -F,
\end{equation}

and the pressure is related to the pressure of the filtered liquid inside the porous tube, $P_f$, and the pressure drop, $\Delta P$, across the tube/membrane by

\begin{equation}
P |_{r=r_w} = P_f + \Delta P.
\end{equation}
If the axial, radial, and tangential flow fields can be decoupled, the presence of the porous tube can be considered to modify the vortex flow equation (far away, the tube will look like a sink) presented above and the radial velocity is given by:

\[ V_r = -\frac{r_w}{r} F. \]  

(19)

Since the measured permeate flow, \( Q_P \), in our tests was generally less than 20% of the feed flow \( (Q_P < 0.2Q) \), we have

\[ V_t h \gg 2\pi r_w F \]  

(20)

and thus \( V_r \) is much less than \( V_\theta \) over much of the field. This suggests a dimensionless parameter, \( \xi \),

\[ \xi = \frac{Q_P}{Q} = \frac{2\pi r_w F}{hV_t}. \]  

(21)

that measures the relative influence of radial to tangential flow as well as the fraction of permeate recovered – i.e. the permeate flow divided by the feed flow.

The effects of viscosity in vortical flow around a cavity whose size changes with time, thus creating a source/sink effect – similar to the presence of the microporous tube with permeation – and a coupling between the radial and tangential equations of motion was considered in [23]. In that case, the source strength was given by \( q \). Here, we would have

\[ q = Q_P = 2\pi r_w F \]  

(22)

The equations of motion are then given by:

\[ \frac{1}{\rho} \frac{\partial P}{\partial r} = \frac{q^2}{r^3} + \frac{V_\theta^2}{r} \]  

(23)

\[ \frac{\partial V_\theta}{\partial t} + \frac{q}{r} \frac{\partial V_\theta}{\partial r} + V_\theta \frac{q}{r^2} = \nu \frac{\partial}{\partial r} \left[ r^{-1} \frac{\partial}{\partial r} (r V_\theta) \right], \]  

(24)

which indicates that in the presence of the source term, or when permeation or filtration occur, the tangential velocities and the flux rates are coupled.

Without simplifying assumptions, the more complicated incompressible coupled momentum and continuity equations in axisymmetric geometry (assuming no circumferential variation) would need to be solved for \( V_\theta, V_r, \) and \( V_z \).

3.3. Particle Behavior

The behavior of a suspended particle can be calculated by considering the forces acting on it. The equation of motion of the particle, derived from a force balance, is:

\[ m_p \frac{d\mathbf{u}_p}{dt} = C_d \frac{1}{2} \rho_l |\mathbf{u}_l - \mathbf{u}_p| (\mathbf{u}_l - \mathbf{u}_p) \cdot \mathbf{S}_p + \mathbf{L} + \mathbf{B}, \]  

(25)

Here, \( m_p \) and \( S_p \) are the particle mass and projected area in the direction of motion, \( \mathbf{u}_p \) and \( \mathbf{u}_l \) are the particle and liquid velocities, \( \rho_l \) the liquid density, \( \mathbf{L} \) the lift force vector due to
Figure 3: a) Sketch of small scale test loop for a single Hydroperm™ tube, used in both conventional and outside-in filtration direction. b) Sketch of DynaPerm™ operation loop – also includes details shown in (a).

Figure 4: Detail sketch of the filtration modules. a) 3-tube unit. Outside diameter 6.0 cm. Length 48.3 cm. b) 7-tube unit. Outside diameter 7.6 cm. Length 47.0 cm

Shear, $C_d$ the drag coefficient, and $B$ the body force. The drag coefficient $C_d$ is a function of the particle shape and of the Reynolds number of the relative motion, $R_e$, defined as

$$R_e = \frac{|u_i - u_p| D_p}{\nu}.$$  \hspace{1cm} (26)

Here $D_p$ is the particle characteristic dimension and $\nu$ is the feed stream kinematic viscosity. Expressions for $C_d$ (Stokes, Oseen, experimental data, etc.) that depend on $R_e$ at low values of $R_e$ can be found, for example, in [10].

The body force $B$ includes gravity and centrifugal forces, $F_c$. The radial component of the latter is given by:

$$F_{c,r} = m_p \frac{u_{p,\theta}^2 \theta}{r}$$ \hspace{1cm} (27)

where $u_{p,\theta}$ is the $\theta$ component of the particle velocity. Any surface or other body forces (due, for example, to particle attraction or repulsion) could be included similarly in $B$.

4. Experimental Setup and Procedures

Experiments were conducted in three flow loops schematically depicted in Figures 3 and 4. The first loop depicted in Figure 3a was used for the conventional Hydroperm™ system tests as well as for tests where permeation from the outside of the tube to its inside was investigated in the absence of swirl. The two flow loops schematically depicted in Figure 3b with some details on each shown in Figures 4a and 4b were used to investigate the innovative DynaPerm™ system. For all systems the flow controls such as metering and temperature controls are the same even though they are not shown in all sketches.

Particulate laden feed was drawn from 2 and 5 gallon reservoirs by a circulating centrifugal pump. The flow rate is primarily controlled by the pressure drop across the system using valves. In addition, a bypass loop between the pump output and the feed tank was provided to enable additional adjustment of the flow rate. The feed then entered the filtration module, described below. Part of the flow permeated across the porous tubes leaving the particulates behind, and flowed out of the module as the permeate stream. The remaining flow was circulated back to the reservoir. The pressure drop across the module was controlled by valves at the module inlet and outlet, and pressures at these locations were monitored using gages. A rotameter type flow meter monitored the feed flow. Due to the relatively small flow
rates involved, permeate flow was periodically measured using a graduated cylinder and a stop watch.

The feed flow was periodically interrupted by closing the valve at the feed flow entrance to the filtration module. Typically, the flow was shut off for a period of 60 seconds.

4.1. Characteristics of the Three Filtration Modules

Three different tests modules were designed and fabricated. Figures 4a and 4b sketch the DynaPerm™ modules. In addition Table 2 shows the characteristics of all three modules. The chamber walls of all three modules are made of clear plastic to afford good observational opportunities. This is shown in Figures 5a-d where photographs of dye injection into the swirl module visualize the flow pattern. Actual observations with time variations show the relative importance of tangential and axial flow in the inner chamber and in between the outer cylinder and the inner one containing the slots. Strong swirl is apparent near the slots and less dominant away from the slots and near the outside cylinder.

In the DynaPerm™ units the particle laden flow entered the outer chamber from one end and was forced into the inner chamber through a set of tangential injection slots cut in the wall separating the inner and outer chambers. This introduced a swirling component to the flow about the microporous tubes as well as a longitudinal component along the tubes. The flow exited via ports in the inner chamber located on the opposite end of the module from which the flow entered. The “clean” water (or “permeate”) permeated across the tube wall and was collected as it flowed out of the test module.

Of importance are the total slot area for tangential injection, $A_t$, and the annular cross sectional area available for axial flow, $A_z$. Their importance was discussed earlier in the Section 3.2. As detailed in Table 2 the larger module (Figure 4b) has values of $A_z$ and $A_t$ twice those of the smaller module (Figure 4a). The smaller module was operated with a single porous tube at its center or a set of 3 porous tubes. The larger module set up was operated with a total of 3 or 7 tubes along its center line.

In the larger module, two drain ports were located at the bottom of the module. In some preliminary qualitative tests when the feed flow was interrupted, these ports were opened to enable removal of the accumulated particles in the module and those which “flaked” off the tubes during this flow interruption period. These particles were remixed in the feed reservoir. These “observation-type” tests are encouraging in terms of future potential studies for solid particle recovery and recycling in some applications.
4.2. Testing Parameters and Measurements Techniques

A mixture of water and contaminant particles was introduced into the system as the feed. Two types of feeds were employed:

1. A mixture of water and rust obtained in the laboratory as a mixture of well water and rust from the tanks of a cavitation tunnel.

2. Actual waste water from the high pressure water jet stripping of a bridge from an actual project in the field \(^1\). This water contained particulate contaminants of lead, zinc, and rust among others due to the nature of the paint removed.

The majority of the experiments were performed using the paint stripping waste water. Typical measured concentrations with that water after removing – using a filter mesh – particles larger that 1250 \(\mu m\) were of the order of 600 \(ppm\) (Figure 10). In some of the later tests where the availability of this water was becoming a problem we mixed the paint stripping water with well water thus lowering the particle concentration to the order of 200 \(ppm\).

The flow of “clean” water (permeate) produced by the system was monitored as a function of time, and the permeate flux rates per unit area of tube deduced. Particle concentration within this stream was periodically measured to verify and quantify particle removal. In addition to running the experiments with swirl in two different test modules, the flow rates and the slot area were varied. The latter enabled variation of the relative size of the axial and tangential velocities as described below. Due to the time limitations of this Phase I grant we could not afford to make several more modules to vary the axial to tangential velocity ratio \(S^*\). Instead this was accomplished in a less than optimum way by blocking various portions of the slot area with plastic tape. The blockage was distributed uniformly along the slot length by wrapping tape in a helicoidal fashion along the tube length, thus blocking along each of the slots a portion \(1/n\) of about one fifteenth of the total slot length while leaving the other \(1-1/n\) portion open. Blockages (values of \(1/n\)) of 1/3, 1/2, and 2/3 were employed – resulting in an increase in the average tangential velocity component by factors of 1.5, 2 and 3. The governing parameters were discussed earlier in Section 3.2.

Particle concentrations in the feed and filtrate streams were measured to quantify particle removal. This was done both with a series of fine pore sized filter papers (pore size down to 2 \(\mu m\), weighting the captures particles with a Metler precision balance) and with a spectrophotometer which uses light transmittance through a sample of the analyzed water to measure particle concentrations. In addition, pH and total dissolved solids content (TDS) were monitored using a pH meter and an electric resistivity TDS meter. Not much will be said in this report about pH and TDS since they both appeared unaffected by the filters (same values for feed and filtrate, characteristically, 7 and 160 \(ppm\))

In addition, visual observations of the flow and particle behavior were conducted, and baseline cases were run with a single conventional HYDROPERM™ tube head to head with the same water/particle mixes to provide a relative performance yardstick.

\(^1\)We are grateful to the jet cleaning contractor, Cavi-Tech, Inc. for provision of this water.
4.3. Operating procedures

Pre-treatment

In order to insure some uniformity in the tube performance, and to pack down the pore matrix of the microporous tubes, the porous tubes were pretreated prior to operation with a 3 g/liter solution of diatomaceous earth and bottled spring water to decrease the effective tube pore size. This procedure is conventionally used with the HYDROPERM™ systems and we followed it here to preserve continuity and to enable comparison at the same operating conditions. This procedure will be optimized in future studies. Preliminary tests conducted during this Phase I project have reconfirmed that pretreated tubes consistently out performed non pretreated tubes. The solution was circulated through the loop for a period of 30 min. Following pretreatment, the system was purged using bottled spring water.

Cleaning and Back Flushing

For some long duration tests, for investigation of procedures in actual field operations, and for tests run to study the possibility of reuse of the tubes after long usage and significant plugging two methods of cleaning were used: cleaning with an acid solution and back flushing. Most of these tests are not reported in the results section because we have made a deliberate effort in that section to always use a clean new tube – at the expense of added time for module refitting – for all comparative tests.

Cleaning the tubes can be achieved by operating the system as usual but by replacing the feed particle laden liquid with a dilute acid solution. In our tests, a solution of 3% hydrochloric acid, HCl, appeared to be very efficient in cleaning the tubes. This results in dissolution of a large amount of the small particles thus leading to unclogging of the tube pores. A secondary unwanted effect is a large increase of the total dissolved solid content of both the feed solution and the permeate solution, therefore necessitating very thorough flushing of the system and removal of all acid traces after cleaning. This operation is not lengthy, is very effective and had a total duration of about 20 minutes. In an actual operating system such a procedure could be repeated once a day.

Back flushing consists of operating the system in reverse flow using either clean water or an acid solution. In the DYNAPEM™ setup this amounts to operating the tube as in an conventional HYDROPERM™ case with the in-flow circulating through the tube while the outflow is forced to come out of the porous tube walls. This results in pushing the “dynamic membrane” away from the tube and in expelling some of the clogged particles out from the tube wall matrix.

4.4. Operating Modes

Three different modes of operation can be used when carrying out the laboratory tests.

The first involves “constant concentration” operation, with the permeate remixed into the feed reservoir, so that (except for evaporation losses and the particulate layer build up on the tubes) the volume of the recirculating feed, as well as its suspended-solids concentration, remain constant. This is the operation mode utilized for the majority of the Phase I experiments.
In the second “concentration” mode of operation, a batch wise process of a field system is simulated. Here the permeate is removed and collected in a separate reservoir, so that the volume of the circulating feed continuously decreases while its suspended solids concentration continuously increases. A limited number of experiments with water/rust solutions were performed in this mode. The tests can be continued until a specified feed concentration is reached or until the volume of the feed becomes so low that adequate pump suction from the reservoir can no longer be maintained.

In the third mode of operation, a “continuous” field process is simulated. As permeate is withdrawn by the filtration process, comparable amounts of new feed are added. Periodically, concentrated particles sludge is removed.

In Phase II, the second and third operating modes, representative of field operation would be emphasized more.

5. Results and Discussion

5.1. Influence of flow direction and of the initial filtering

Initial tests were conducted using the HYDROPERM™ unit and a preliminary DYNAPERM™ module in order to define the operating procedures, improve on the flow configuration and select water treatment procedures prior to actual filtration tests. This enabled us to improve on various components of the flow system (e.g. reduction of pressure losses in the valving and tubing of the system). This also enabled the definition of the parameters needed to set up a second loop including new pump characteristics, module and filtration tube characteristics.

The first solid/liquid mixture characteristic that we investigated concerned the determination of the amount of pre-filtering needed before circulation of the refuse water though the microporous tubes filtration units. Removal of all particles above a given size was achieved using one of the following sieves: 150 μm, 250 μm and 1250 μm. Contrary to what one would expect intuitively, excessive filtering can be counter-productive. As seen in Figure 6 much better filtration rates were obtained with the same initial jet paint cleaning water when the water fed to the microporous filtration unit contained initially larger particles in addition to smaller particles. This can be attributed to:

either a better build up of the dynamic membrane around the tube by generation of a particle distribution in the membrane which enables a higher level of porosity in it (as opposed to a tightly packed matrix with very fine particles),
or a better dynamic activation of the shear layer around the tube where the larger particles are more energetic in loosening any strong bonding of the particles in the dynamic membrane.

Visually the presence of the larger particles manifests itself by a very obvious flaking and fall-off of the cake formed at the tube wall during flow interruption.

Another major improvement was obtained when the flow configuration was modified to become the one shown in Figures 4 and 5. While the flow into the tube was from top to
Figure 6: Influence of initial maximum particle size on filtration rates.

Figure 7: Influence of the vertical direction of the feed flow on filtration rates. Upper curve corresponds to flow from bottom to top of the module.

bottom in the original test configuration, inverting it to bring in the feed from the bottom has a very significant effect on the flux. As explained above, this reversal of the flow appears to also be strongly connected to the loosening of the cake built up on the outside of the tube. Another explanation is that the local swirl number ($V_t/V_z$) varies along the tube. Since $V_t$ does not vary as much as $V_z$ along the tube, the relative influence of $V_t$ is largest near the inlet. Gravity will tend to settle the particles, producing higher concentrations near the bottom then the top. Thus, if the inlet is at the bottom, higher particle concentrations will beneficially coincide with higher local swirl numbers. Figure 7 illustrate this effect by showing filtration rates versus time in two selected cases where all conditions are similar except for the vertical direction of the flow. Greatly increased flux rates (more than 5 times) are seen when the feed is from the bottom.

5.2. Comparison between presence and absence of swirl

Based on the above results a long series of tests was conducted with single and three-tube modules to investigate the feasibility of enhanced filtration using the DYNAPerm™ concepts described in Section 2.1, namely outside-in flow and superposition of swirl and flow interruption. Figure 8 summarizesthe data of a large number of tests and compares them with the conventional HYDROPERM™ system tests rerun here with the same water at its optimum operating conditions.

Figure 9 isolates one set of DYNAPerm™ tests, with the test module axis oriented vertically, the flow was periodically turned off briefly and then resumed. Figure 9 presents measured flux rates as a function of time for this last test set compared to results without the use of swirl or flow interruption using the HYDROPERM™ unit. Figure 9 clearly shows substantially higher flux rates for the DYNAPerm™. Flow interruption continually reestablished the flux rate at near the original high value. As also illustrated in the more general Figure 8 the combined swirl and flow interruption resulted in an increase in flux rates of approximately a factor of 20 on average!

Observations indicated that a particle layer rapidly built up on the outside wall of the tube. Upon flow interruption, the particle layer flaked off and fell to the bottom of the module. The

Figure 8: Comparison of long term tests between the DYNAPerm™ and the HYDROPERM™ modules. Data is a collection of tests using 1 and 3-tube modules.
idea of refreshing or regenerating the filter medium is not new. Indeed, periodic cleaning of most filtration systems including HYDROPERM\textsuperscript{TM} tubes is standard practice. This is often done, however, with a frequency of the order of a day, done to maintain the plateau at a suitable level, and requires some down time – often for cleaning by backwashing. In our tests, the necessary down time would appear to be very short – of the order of a few tens of seconds without use of a back flow. This short down time enables a high frequency of flow interruption if future tests show it is needed, such that the flux rate never drops to the plateau level, but rather stays in the higher flux rate region. Since the particles are deposited on the outside of the tube, it is easier for the flaking off process to occur relative to deposition on the inside of the tube – a more open arrangement.

The use of a configuration resulting in permeate flow across the porous tube wall from outside to inside is an integral part of the swirl concept and provides for increased surface area for particle deposition since the outer wall area is larger than the inner wall area. For relatively thick walled tubes, this can be significant. In the present tests, the tubes are such that the outer surface is 33\% larger than the inner. However, this is not the most important parameter as also illustrated in Figure 9 where a test with the conventional HYDROPERM\textsuperscript{TM} system set-up was conducted after inverting the flow to outside-in. A significant improvement in the flux as expected from the surface area increase is seen, but this improvement is small compared to the one obtained with the DYNA\textsc{Perm}\textsuperscript{TM} system where in addition swirl and flow interruption are applied.

Discussion of swirling the flow

Swirling of the feed flow with the porous tube(s) at the center of rotation likely produces at least two beneficial effects. In the inner chamber, the liquid has a net flow into the porous tube wall and, without swirl, would carry particles along with it and deposit those onto the wall. The outward centrifugal body force exerted on the particles (particularly, the larger particles) retards their otherwise radially inward flow resulting in “slip” relative to the liquid flow into the wall, less large particle flow to the wall, and less deposition on the wall. This also likely modifies the concentration of particles in the bulk fluid, decreasing it near the wall. In a sense, a form of stratification occurs that would decrease particle deposition. Depending on particle characteristics and the flow configuration, particles may be forced to the outside of the inner chamber or only be retarded in their movement inward. In addition, it is the smaller particles which will move toward the tube, with the larger more likely to move outward – further reducing particle build up. The second beneficial effect is due to the shear set up on the tube wall which tends to limit particle deposition in the same manner as the shear of longitudinal flow in conventional cross-flow filtration. In tube flow, for a given tube size, the shear is essentially determined by the feed flow rate in the tube. However, in the swirling
system, that shear can be varied over some range independent of the feed flow rate by adjusting the tangential injection geometry to vary the amount of swirl. This leads to higher shears at a given feed flow rate, resulting in deposition of a thinner particle layer and higher flux rates.

5.3. Influence of the flow rate and swirl parameter

Once confidence in the feasibility and effectiveness of the DYNAPerm™ system was gained a series of tests was conducted to investigate the influence of three key parameters: the number of tubes, the flow rate, and the swirl parameter. This investigation followed the test matrix shown in Table 3. Initially the tests were conducted in parallel with the 3-tube and the 7-tube setups using two different loops. Later on, since the pump in the smaller loop is limited in power, the 3-tube module was mounted and tested in the larger loop. This enabled both modules to be tested for flow rates varying between 2 and 9 gpm. Table 1 shows the corresponding nominal axial and tangential velocities in each tube. In fact, the actual velocities depended on the filtration or flux rate in the particular test. The corresponding figures shown below were drawn using corrected values of the velocities to correct for loss of fluid through permeation along the porous tubes lengths. One key problem which affected this short duration Phase I study is that the testing of each condition of flow and swirl is time consuming, since it always involves a time dependent phenomenon associated with the increased amount of deposition of particles on the tubes. Even though meaningful data for practical applications cannot be obtained in test durations shorter than say 20 hours, we deliberately decided to restrict our test duration to two to three hours per condition. This is justified by the fact that – as known from previous tests [5] and as could be seen from Figures 8 and 9 – the first few hours give a very good indication about the behavior of the tubes over the longer time periods. For all the tests a new set of newly extruded tubes were mounted for each test. These tubes were then pre-treated in the same conditions as described in Section 4.3.1 in order to minimize as much as possible differences between the initial conditions of the tubes. These lengthy procedures prevented repetition within the constraints of Phase I of this last series of tests to increase confidence in the results. Such repetition of the tests should be conducted in Phase II of the study.

Particle size distribution

For all test cases water from the same water jet paint stripping job was used. Figure 10a shows the cumulative distribution of particle sizes in this water as deduced from weight measurements of particles captured by filter papers and screens of various sizes as described in Section 4.2. The upper curve shows the particle distribution of the water used for the 7-tube module. The lower curves show the same water after it had been diluted in well water in order to increase the amount of water available to conduct the rest of the test matrix and water used in the 3-tube module tests. In all cases the particle size distribution appears to peak between
30 and 150 $\mu m$ with very few measurable particles below 10 $\mu m$ and above 200 $\mu m$. Also shown in the bar chart of Figure 10b is the particle size distribution density obtained from the cumulative distribution which conveys the same information. No particle size distributions of the permeate from any of the tests are shown because no appreciable weight could be measured using any of the filter sizes down to 2.5 $\mu m$.

**Time variation of filtration rates**

The influence of the flow rate on the filtration rate of the DynaPerm™ system can be deduced from the results of each experimental set shown along a row in Table 3. In fact, as discussed below, conclusive results cannot be obtained without an overview of a large range of operating parameters and a more intensive set of tests than the set presented here. However, for illustrative purposes we show below some specific results.

Figures 11 considers the case of the 7-tube DynaPerm™ module in the case where the tangential injection slots are kept fully open. This is a case where the flow field is the “cleanest”, where no secondary flows in the module are created by the artificial blockage of the injection slots. Figure 11 shows the filtration rates of the module versus time for three flow rates through the 7-tube module: 2, 5 and 9 gpm. It is apparent in this case where the swirl parameter value is equal to one ($V_t = V_z$) that the flux rate increases with the flow rate through the module and that this tendency remains true for the duration of the filtration test. This tendency, however, is not true for all the cases studied here in Table 3.

Figure 12 shows the influence of the amount of slot blockage as described in Section 4.2 on the filtration rates versus time. By varying the length of the tangential injection slot that is blocked the tangential velocity or the swirl parameter, $S^*$, is varied. Here $S^*$ takes the values 1, 1.5, 2 and 3 for the same flow rate 2.5 gpm or nominal axial velocity 0.625 $ft/s$. As can be seen in the figure the overall tendency in this case is for increased filtration rate with increased tangential velocity. This tendency is conserved at all times during the filtration tests. However, small discrepancies exist in the details of the tests showing for instance higher fluxes for 1/3 blockage versus 1/2 blockage ($V_t = 0.94 ft/s$ versus 1.25 $ft/s$)
Figure 13: Contour plots of filtration rates for given axial and tangential velocities. Note that HYDROPERM™ results are about 200 gpd/ft². a) 3-Tube module b) 7-Tube module DYNAPERM™.

Comparison of all results at the end of the same filtration period

As we have mentioned in the previous section, it is difficult to draw any general conclusions about the operating ranges of the DYNAPERM™ by observing only a restricted number of tests. Indeed, based on known previous results with the HYDROPERM™, and on the mechanistic discussion in Section 3, no simple trend is expected. Instead, for a given module geometric configuration a range of flow rates (or maybe more than one range) should lead to optimum behavior or maximized flux rate. This has made the analysis of the results obtained difficult. In order to represent the data, a “three-dimensional” representation in a contour plot form is used. Figures 13a and 13b show filtration rates for given axial and tangential velocities for the 3-tube and 7-tube DYNAPERM™ modules.

It appears from the figure that in the range of variation of the parameters $V_t$ and $V_z$ the 3-tube module was most always operating in the high flux regime area. Only in a very small region of the considered map – say in the region $V_t \approx 2.5 \text{ ft/s}$ and $V_z \approx 1.2 \text{ ft/s}$ – did the flux rate drop below 1000 gpd/ft² which is still five times higher than the HYDROPERM™ tube results in the exact same conditions (200 gpd/ft²).

The range of parameters investigated for the 7-tube module appear on the other hand to have been on the edge of the region separating very good filtration rates from good rates. Figure 13b shows that the results improve with increased axial velocity (or flow rate) while the results seem to deteriorate with increased tangential velocity (in fact with increased blockage of the tangential injection slots, since this was the only way employed in this phase for increased swirl or tangential velocity).

Trends from the data and analysis

As discussed in the analysis section (Section 3) the filtration process is mainly a balance between two dynamic forces: one moving the particles perpendicular to the wall and the second moving the particles along the microporous tube wall. If such a premise is accepted then the actual velocity to consider in the analysis is neither $V_t$ or $V_z$ independently but rather the actual velocity vector along the tube wall composed of $V_t$ and $V_z$, that is a vector of magnitude $V_{res} = \sqrt{V_t^2 + V_z^2}$. Figures 14a, 14b and 15 represent the results shown above in Figures 12a and b using the above defined velocity $V_{res}$. As can be seen in Figure 14a which combines both 3-tube and 7-tube data, the curves of filtration rates versus resultant velocity along the tubes become more and more complex when the amount of blockage is increased. Figure 14b which is restricted to the fully open and 1/3 blocked slot data appears however to correspond much better to the expected trend, that is existence of a region of maximum flux rate, here $1.5 \text{ ft/s} \leq V_{res} \leq 2.5 \text{ ft/s}$. When the blockage is increased to 1/2 Figure 14a still shows the same region to be a region of maximum high fluxes. However, it also appears that this region is not unique and that the flux increases again further when
$V_{res} \geq 4 \text{ ft/s}$ for instance. This tendency is even stronger when the blockage is further increased to $2/3$. This deterioration and complication with increased blockage is probably due to the tangential injection flow becoming less and less “clean” and uniform with blockage. Additional experiments are needed with better ways to vary the swirl which conserve the uniform injection along the full length of the tube. Also a larger number of data sets is needed to enable elimination of variations in either the quality of the tubes or the water tested. Such approaches should be considered in future phases.

Another effect of imposing high tangential velocities is a decreased pressure at the chamber center along the tubes outer wall. This will decrease the pressure differential across the tube wall and adversely affects the flux rate. Thus, as the data indicates, increasing the swirl too much can be counterproductive. For example with $V_t=3.5$ ft/s and a $V_\theta$ distribution consistent with that of [25], we estimate that the pressure differential across the tube is only 3 psi much lower than the imposed 10 psi when $V_t = 0$.

Figure 15 considers the same data as in Figure 14a represented on a log-log scale to compare with the trends expected from the analysis in Section 3. Represented in this fashion the data seem to fall along the lines of $F \approx V_{res}^{7/8}$. This can be predicted from the analysis presented above. The Reynolds number, $Re$, for these flows based on hydraulic diameter and $V_z$ varies from 1,300 to 2,000. Based on $V_{res}$, it is between 2,000 and 40,000 approximately. The friction factor, $f$, for flow in this range varies as [24]:

$$f = \frac{4\tau_w}{\rho V_{res}^2} \sim Re^{-0.25}. \tag{28}$$

Using (6, 10), we obtain:

$$F \sim u^* \sim V_{res} Re^{-0.125} \sim V_{res}^{7/8}. \tag{29}$$

### 5.4. Concentration Mode and Rust Experiments

We finally present in this section some preliminary experiments which need further investigation in Phase II. Experiments were run with a rust and water mixture as described above.

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**Figure 14**: Filtration rates versus resultant velocity along the tubes. Results combining both 3-Tube and 7-Tube DYNAPerm™ module. a) All data combined. b) Full open and 1/3 blocked slot data.

**Figure 15**: Filtration rates versus resultant velocity along the tubes. Comparison with 7/8 slope theoretical predictions.

**Figure 16**: Results of filtration rates in the concentration mode with rust and water mixture.
Figure 16 presents data for two DYNAPerm™ experimental runs with the 3 tube module compared with conventional Hydroperm™. Run Rust-1 was made, as most of our work, in the "constant concentration" mode with a small reservoir. To perform a brief preliminary investigation under conditions more typical of practical field applications, a run was made in the "concentration mode" as described in Section 4.4 with a 300 gallon reservoir of rusty water. The results are presented as run Rust-2. The conventional Hydroperm™ was also run in this concentration mode with the 300 gallon reservoir and became completely clogged after three hours of operation. The large flux increase at approximately 5 hours of operation for the Rust-2 run is due to cleaning with a dilute acid solution. At the end of the conventional run (3 hours), the DYNAPerm™ fluxes are seen to be approximately 4-8 times higher than those of the conventional system. It is not clear from this limited data whether the difference in the results of the 2 DYNAPerm™ runs is due to the operation mode or to tube variation. This mode of operation needs more thorough investigation in future phases.

6. Conclusions

The Phase I investigation presented in this report has met all the objectives of Phase I proposal restated on Page 1. In addition, Phase I research results unequivocally demonstrate the feasibility of the new DYNAPerm™ filtration technology to greatly improve the efficiency, capacity and economy of particle removal from liquids. Phase II research should enable the creation of a very successful proprietary filtration system with ready commercial applications.

Using particulate laden water obtained from an actual field project of high pressure water jet paint stripping, the new filtration concept was found to result in measured flux rate increases of a factor of 20 over conventional microporous tube cross flow filtration.

The orientation of the flow was found to be significant with feed flow from the bottom of the module producing substantially higher flux rates than flow from the top. This is believed due to gravity effects redistributing the particle concentration in relation to the relative amounts of axial and tangential flow velocities. It was also found that pre-filtering with too small of a filter degrades performance contrary to what might be otherwise believed. This is believed due to a contribution of larger particles in creating a more porous dynamic membrane.

Variation of flow rates and swirl index showed that flux rates increased with moderate increases in shear exerted on the tube as controlled by increasing values of the total velocity made up of the axial and tangential components. Flux rates tended generally to increase with the amount of swirl although there was sometimes a decrease with large amounts of swirl, probably due both to the manner in which this was created - partial slot blockage - and to a decrease in the pressure drop across the tube wall due to the low pressures created at the center of the swirling flow. Measured flux data were found to reasonably follow the expected power law relation with friction velocity based on a mechanistic theory.

No significant changes in measured flux rates were observed in going from 1 to 3, and 7 tube configurations (in which case the inner tubes are shielded from the flow by the outer tubes) suggesting that, at least with the data available to date, there will not be any decrease in flux rates in scaling up to larger systems. This will, of course, need to be verified in future work at larger scales.
In Phase II, a more in depth understanding of the process will be sought and many practical issues addressed to enable work in Phase III with a commercial partner at pilot unit scale. Many scale up issues need to be addressed.

Although the results of Phase I comparing 1, 3 and 7 tube results show no significant performance differences, and thus indicate flux rates are independent of the number of tubes employed, 7 is likely a much smaller number of tubes than would be employed at full scale. For example, HYDROPERM™, utilizes modules with up to 200 tubes, probably more than would need to be used for DYNAPERM™ due to the higher flux rates. Scale effects in going to a larger number of tubes and to a larger diameter module need to be investigated. This may be particularly true concerning the beneficial effects of flow interruption on removing the particle layer which may be affected by the packing density of the tubes.

The capacity of a given system will be determined by the total filter area (number of tubes, length, and diameter) and the flux rate. Understanding how quantities such as tube number, packing density and module length affect flux rates must be determined to incorporate the resulting trade-offs in an engineering optimization of a particular system. Greater understanding of the flow field desired and the flow field generated needs to be gained. This may include velocity field measurement, flow visualization, and modeling.

In Phase I, a modest parameter space was investigated. In Phase II, this range will be broadened and reproducibility issues addressed. In Phase I, the swirl was varied by blocking the injection slots. This was found to be unsatisfactory at high blockages. In Phase II, new modules will enable introduction of a more uniform swirl. Particular emphasis will be directed at high swirl to ascertain the cause of performance decrease – which may be related to improper means of generating the swirl – and to optimize the system.

Issues of engineering importance such as tradeoffs between flux rate, module size, and required pump power (pump flow and pressure) as well as system configuration issues (multiple modules, series or parallel operation, etc.) need to be addressed. Also of importance are tests of much longer duration in modes representative of fielded batch and continuous process applications. An effective cleaning procedure at large scale should also be developed.

7. Commercial Applications

Filtration of particles from the effluent and internal process streams of industrial, government, and commercial plants and facilities is a very important practical ongoing problem. The DYNAPERM™ technology has application in a number of important process stream and effluent treatment roles such as:

- **Pretreatment**, for suspended solids removal prior to reverse osmosis, carbon adsorption, or ion exchange treatment.
- **Polishing**, for removal of fine suspended solids after chemical or biological treatment.
- **Liquid Reuse**, when this is otherwise impeded by the presence of suspended solids.
- **In-Plant Processes**, for valuable materials recovery.
- **Toxic Heavy Metals** removal when in suspended form.
• Treatment for discharge.

The DYNAPerm™ technology enables greatly increased flux rates with minimal down time for near complete removal of suspended particles down to the micron size. The radical increase in flux rate of this technology compared to conventional cross-flow filtration should prove attractive to potential Phase II partners and find application in a broad spectrum of applications, particularly those in which large volumes of liquid must be processed.

In the treatment of both domestic and industrial waste waters, the removal of suspended solids is a required operation which arises under a variety of circumstances. For example, this operation is often essential as a pretreatment step prior to undertaking further advanced treatment. Fine-solids removal is also often necessary as a “polishing” step after chemical or biological treatment processes. There are several existing physical treatment processes available for this purpose, such as gravity clarification, multimedia filtration, etc. However, most of these processes have inherent disadvantages in dealing with fine and difficult-to-settle solids. In addition, neither of these processes produces a highly dewatered sludge. DYNAPerm™ can meet this need.

One application with a very large potential market is the water jet cleaning and removal of paint and rust from bridges and other structures. Water jetting is an efficient and economical means of paint and surface scale removal. However, environmental regulations require that the substances removed - typically containing heavy metals - not be released to the environment. Thus a high flux means of filtering the water either for reuse or for discharge is required.

Battery manufacturing is a large source of lead and zinc contaminated water. In addition to effluents from such facilities, there is growing concern over the quality of the rainwater runoff from such sites. Often the site has sufficient contamination that the rainwater runoff will become contaminated with heavy metals. This is potentially a very large volume of water to process prior to its discharge into the environment.

Other commercialization potential for removal of suspended solids including toxic heavy metals in a wide range of industries, includes battery manufacture, metal finishing, plating, smelting and refining, and other industries with similar process streams and wastewaters.

8. References


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<th>TYPE</th>
<th>TS (mg/l)</th>
<th>SS (mg/l)</th>
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Nylon DYNAPERM® Tube, 65% Porosity

Polyethylene DYNAPERM® Tube, 65% Porosity
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7-TUBES DYNAPERM ASSEMBLY
TANGENTIAL INJECTION FULL OPEN

Permeate Flux, GPD/ft²

Time, minutes

Vz=1.125 ft/s
Vz=0.625 ft/s
Vz=0.25 ft/s
3-TUBES DYNAPERM SYSTEM
FLOW RATE 2.5 gpm, Va=0.625ft/s

Permeate Flux rate, GPD/ft²

Time, minutes

Vt=1.875 ft/s
Vt=0.94 ft/s
Vt=1.25 ft/s
Vt=0.5 ft/s, & 0.625 ft/s
COMBINATION OF ALL RUNS
3 and 7-tubes tests

FLUX, gpd/ft**2

sqrt(Vt**2+Vz**2)

- - No Blockage  →  1/3 blocked  -----  1/2 Blocked  ----  2/3 blocked