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Tritium Removal by Membrane Separation

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University Honors 458
Senior Honors Project

Tritium Removal by Membrane Separation

Final Report

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Submitted: 4/24/01

UNIVERSITY HONORS PROGRAM

SENIOR PROJECT - APPROVAL

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College: Engineering Department: Chemical

Faculty Mentor: Dr. Paul Bienkowski

PROJECT TITLE: Tritium Removal by Membrane Separation

I have reviewed this completed senior honors thesis with this student and certify that it is a project commensurate with honors level undergraduate research in this field.

Signed: ______________________, Faculty Mentor

Date: 4/30/01

Comments (Optional):

Abstract

A 13-stage recycling cascade polyphosphazene membrane process was designed to separate tritium from water. The waste water stream is fed to the separation process at a flowrate of 1 gallon per minute and has an activity of 240,000 pCi/L (9.35e-11 g/min). This is twelve times the permissible activity level, as specified by the Environmental Protection Agency (EPA).

The membrane separation process is designed to remove 95.00% of the tritium into the permeate, with 99.90% of the flow as EPA approved clean water (483,626 gal/yr). The process is contained in a 17 m^3 glove box (operated at negative pressure) to prevent tritium contamination to the environment. The feed is filtered to remove calcium carbonate, and the solid waste is mixed with the tritium from the permeate into cement that is placed in 55 gal stainless steel barrels. Should any tritium leak into the glove box, a cryogenic cooler will condense the water from the air and return it to the cement mixer. The barrels will be shipped to a landfill (35 barrels/yr).

The net present value of for this preliminary design is -$10,113,610 (most of which is the cost of labor).

Economic Basis

Start-up Year	2002	
Project Life	20	years
MARR	12%	
Equipment Costs:		
Fixed Costs:		
Feed Pump (2.5 hp)	$352	
Vacuum Pump (1 hp)	$1,500	
Stage Pump (2.5 hp)	$352	
Cryogenic Unit	$50,000	
Solid Filter	$4,950	
Concrete Mixer	$19,000	
Glove Box	$12,950	
Variable Costs:		
Membrane Module	$97	m^2
Concrete	$0.38	gallon
55-Gallon SS Drums	$250	each
Waste Disposal (includes shipping)	$33.42	gallon
Gloves for Glove Box	$120	pair
Liquid N_2	$0.80	gallon
Electricity	$0.04	kW hr
Labor:		
Technicians	8 @ $40,000	year
Engineers	1 @ $60,000	year
Labor Overhead	100% of Salary	
Benefits	30% of Salary	
Social Security/Worker's Comp.	10% of Salary	
Supervisors	15% of Labor Cost	
Net Present Value of Project	**-$10,113,610**	

Process Design Basis (Preliminary Design)

Plant Capacity (Clean Water)	483,626	gallon/year
Operating Factor	93% (8150hrs/yr)	

Designed for:

95.00% HTO Removal
99.90% Retentate Flow

Feed:

Flow Rate	1	gallon/min
Density	1	g/cm^3
Feed Composition:		
H_2O	99.82	mol%
HTO	2.24e-12	mol%
$CaCO_3$	0.18	mol%

Solid Filter:

Removal of Solids	100%	
Life	20	years

Retentate:

HTO must be less than EPA max. of 20,000 pCi/L

Flow Rate	0.989	gallon/min
Retentate Composition:		
H_2O	99.99+	mol%
HTO	1.12e-13	mol%
$CaCO_3$	0	mol%

Permeate:

HTO must be less than 0.5% per Liter for safe handling

Flow Rate	9.90e-4	gallon/min
Permeate Composition:		
H_2O	99.99+	mol%
HTO	2.13e-9	mol%
$CaCO_3$	0	mol%

Membrane Module:

Life	2	years
Temperature	4	°C
Dimensions:		
Length	1.0160	m
Diameter	0.2002	m
Packing Density	690	m^2/m^3
Separation Factor	2.33	
Pressures:		
ΔP from Feed to Retentate	0.681	atm
Feed	3.26	atm
Retentate	2.58	atm
Permeate	1.00	atm
Thickness	1	µm
Area per Stage	248	m^2
Units per Stage	12	
# of Stages	13	
Recovery basis for water permeablility	15%	
Permeance	7.68	mol/(min m^2 atm)

Process Design Basis (cont.)

Cryogenic Condenser:

Liquid N_2 Consumption	1	gal/day
Air Flow Rate	0.17	m^3/day
Air Humidity	20%	
Life	20	years

Pumps:

Vacuum:		
Life	5	years
Efficiency	90%	
Capacity	1	hp
Centrifugal:		
Life	5	years
Efficiency	90%	
Capacity	2.5	hp

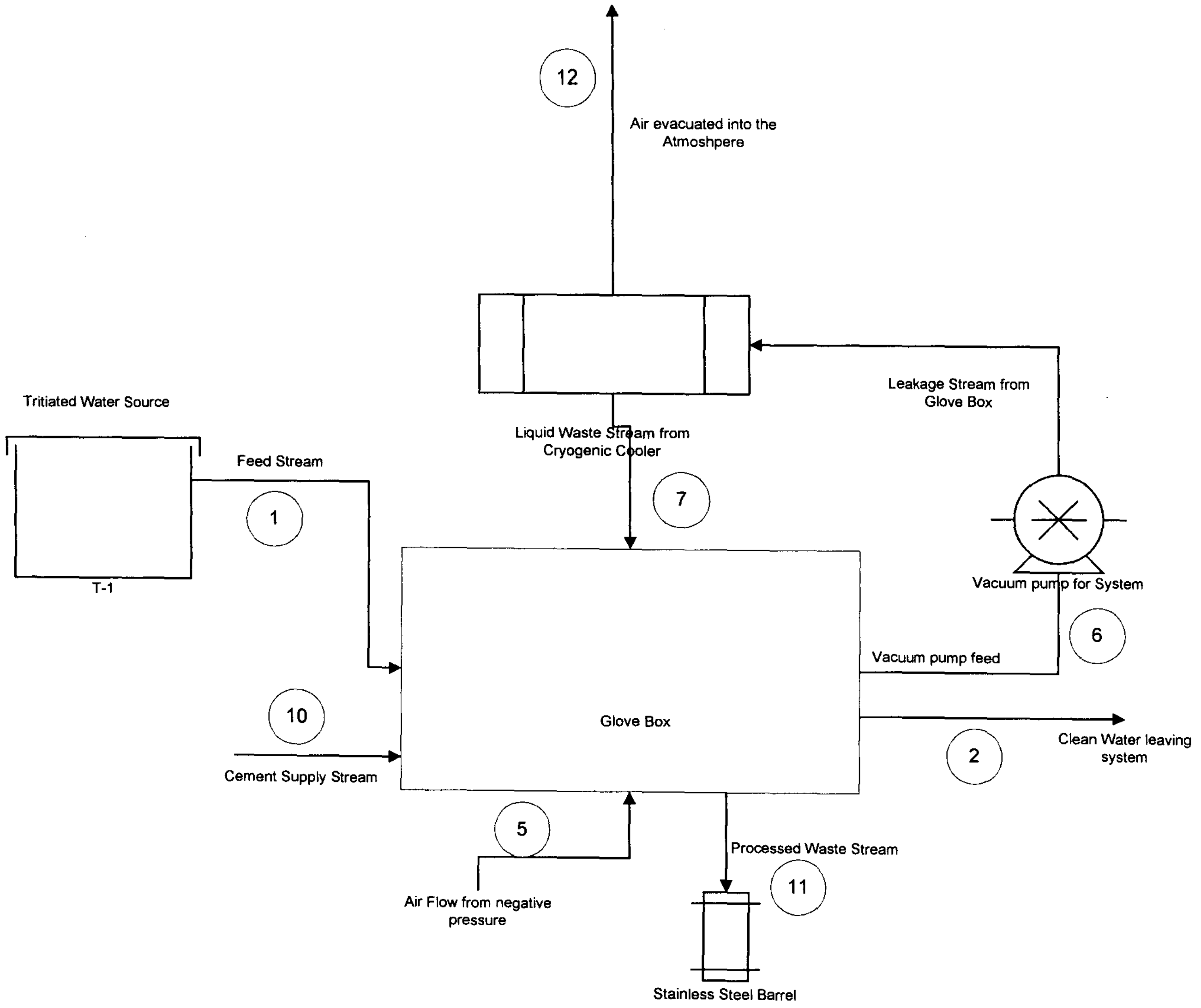
Glove Box:

Size	17	m^3
Box Life	20	years
Glove Life	1	year
# of glove ports	10	

Concrete Mixer:

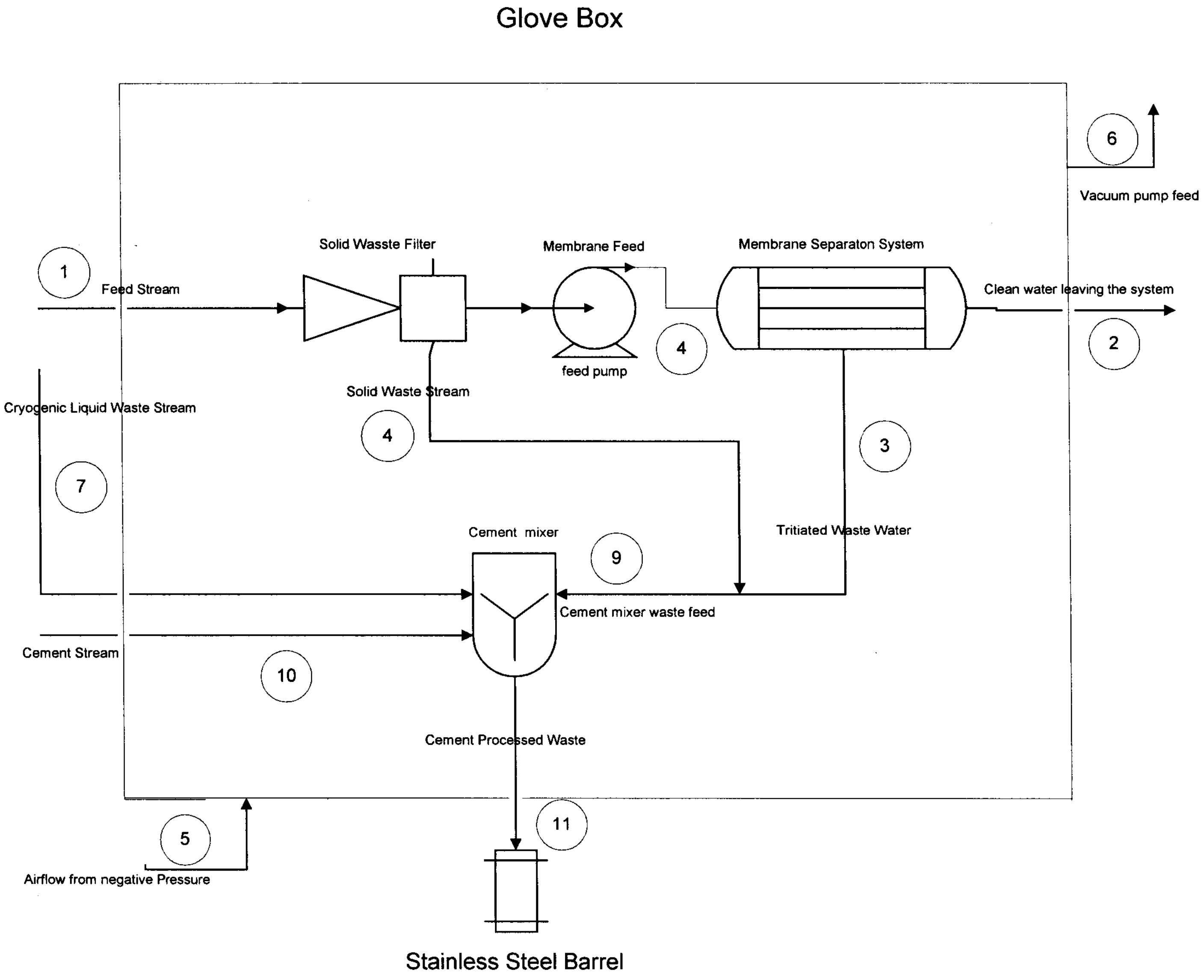
Capacity	0.59	m^3/min
Life	20	years

Process Flow Sheet (Overall System)



Outside glove box Flow Diagram of Titriated Water System

Process Flow Sheet (Inside Glove Box)



Material Balance

Stream	1 Feed Stream	2 Retentate	3 Permeate	4 Solid Filter	5 Air in	6 Vacuum Feed	7 Cryo. Return	8 Membrane Feed	9 Mixed Waste	10 Cement Supply	11 Cement processed	12 Clean Air
Component	Mass (g)	Mass (g)	Mass (g)	Mass (g)	Mass (g)	Mass (g)	Mass (g)	Mass (g)	Mass (g)	Mass (g)	Mass (g)	Mass (g)
HTO	9.35E-11	4.68E-12	8.80E-11			1.05E-18	1.05E-18	9.35E-11	8.80E-11		8.80E-11	
H_2O	3747.15	3743.4	3.75		5.53E-4	5.53E-4	5.53E-4	3747.15	3.75		3.75	
$CaCO_3$	37.85			37.85					37.85		37.85	
Air					1.42E-01	1.42E-01						1.42E-01
Cement										156	156	
Totals	**3785.00**	**3743.40**	**3.75**	**37.85**	**1.42E-01**	**1.42E-01**	**5.53E-04**	**3747.15**	**41.60**	**156.00**	**197.60**	**0.14**

Mass In (g)	Mass Out (g)
3941.14	3941.14

List of Assumptions

Feed

- Flow of 1 gal/min
- Tritiated Water enters from reservoir at 4°C
- HTO is 12 times EPA minimum of 20,000 pCi/L
- Solids ($CaCO_3$only) make up 1% by weight of feed stream and are 50-200μm in diameter and density of 2g/L
- Feed density is that of H_2O
- Cesium is removed upstream of tritium removal process
- Solids filter removes 100% of solid particles

Effluent

- HTO leaves retentate at EPA max of 20,000 pCi/L
- HTO leaves permeate at max of 0.5% per Liter

Membrane

- Membrane life of 2 years
- Packing density is 690 m^2/m^3 from Seader
- Model membrane from Hwang and Kennemeyer using constant cut and constant overflow
- Separation factor is 2.33 (best case from paper 43% reject)
- Operating pressures from paper - P_P = 1atm, P_F = 3.26atm, P_R = 2.58atm
- ΔP = 0.681 atm = PF-PR from vendor, over operating range of 4.76 to 27.22 atm and PP is 1 atm
- 15% recovery through membrane yields permeance of 7.68 (mol/m^2 min atm) from Osmonics (using $MgSO_4$ also used for water)
- Membrane Thickness is 1 micron
- Each stage consists of 12 units

Misc

- Process life of 20 years
- 1% of glove box volume leaks into system (requires vacuum pump)
- 1% of HTO leaks into glove box during maintenance periods, there are 11maintenance periods in 20 years, thus 11% leakage over 20 years, therefore 0.55%/year
- All equipment except membranes and pumps last entire project life
- Assume plant up time of 8150 hours per year (~93%)
- All pumps operate at 90% efficiency
- Unit to be installed in pre-existing facility

Economics

- Net Present Value
- Private consultant with 12% interest rate
- Disposal cost is \$100-\$500/ft^3, assume \$250 (includes shipping)
- No depreciation

Labor

- 8 operators on 4 rotating shifts (2 per shift)
- 1 engineer and 1 supervisor
- Employee benefits = 30% of salary
- Employee overhead = Employee salary
- Worker's Comp & S.S. = 10% of salary
- Supervisor Salary = 15% (Technician + Engineer Salary)

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Introduction and Background

Problem Statement

A wastewater stream from a nuclear plant has been identified as containing tritium along with some solid waste particles. Tritium is a radioactive isotope of water, ^3H, which causes little hazard or risk at low concentrations and in gaseous form. At higher concentrations, tritium poses a more serious threat, especially when suspended in water, because it can be ingested or absorbed into the body easier.

The Environmental Protection Agency (EPA) requires that all waste streams released to the environment must meet drinking water standards (see Appendix B). The maximum contaminant level as specified by EPA is 20,000 picocuries per liter. This limit corresponds to the maximum permissible body burden and maximum permissible concentration of radionuclides in water for occupational exposure. Therefore, the waste stream from the nuclear plant must be purified to comply with these standards. Currently the tritium concentration in the stream is 240,000 picocuries per liter, which is twelve times that of the (EPA). A separation process is to be designed that will effectively remove tritium from the water.

The entire operation will be housed in a large glove box that will allow the operators protected access to the process. The glove box will be operated at negative pressure so that tritium does not leak into air outside the glove box. A technique for recovery of the tritiated air must be identified. The vacuum causes a small air leak into the box, through apertures around the gloves. This too must also be accounted for in the process design. The size, cost and optimum operating conditions of this process are to be evaluated for a project life of 20 years.

Background

Because tritiated water (HTO) differs from normal water only in ways affected by mass, there are a limited number of separation techniques available that yield high separation factors. Some that have been identified are electrolysis-catalytic exchange and water distillation, but require very high concentrations of tritium, and high capital and operating costs.

Some work using polyphosphazene membranes for tritium separation has already been done, and much of the results of our research is based on these studies. Nanoporous membranes were employed, which suggests that polyphosphazene membranes are perm-selective to tritium. This suggests that most of the tritiated water passes through the membrane as the permeate, while most of the normal water passes over the membrane as the retentate (or reject). Figure 1 shows a schematic of a membrane separation system. To determine the separation that can be achieved, the system must be mathematically modeled. Through mathematical modeling, one can determine the size of membrane process required to achieve the desired separation.

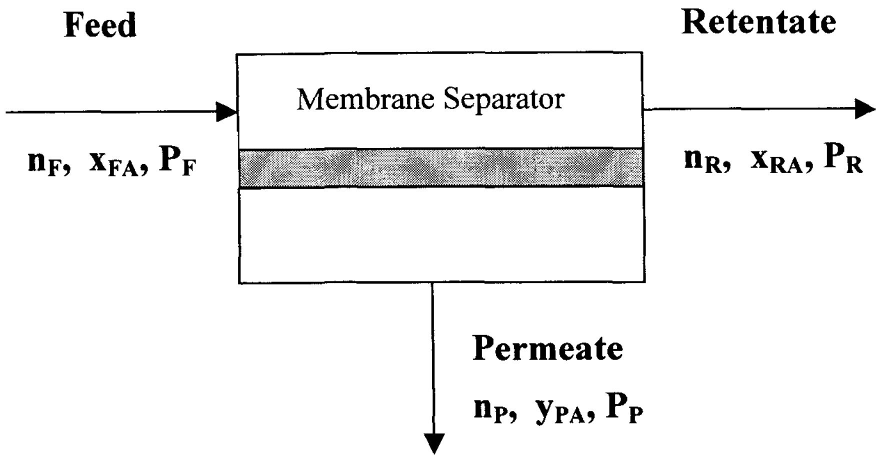

Figure 1. Schematic of a membrane separator.

Mathematical Model

The previous research cited in The Journal of Membrane Science, provided a basis for the initial calculations. Also, membrane specifications were derived from data obtained from Osmonics ™. Given the permeate flow rate and membrane size, the flux, N, of tritium can be calculated by

$$N = \frac{V_p}{A} \tag{1}$$

where V_p is the molar flow rate and A is the area of the membrane. Because the membrane thickness is known, as well as the range of pressures at which the system can be operated, the permeability is determined using the following equation (Seader and Henley)

$$P_{MA} = \frac{Nl_m}{\Delta P} \tag{2}$$

where ΔP is the change in pressure and l_m is the membrane thickness. Permeability is used to calculate the area of the membrane used in each stage.

Cascading operations are often used when a high degree of separation (that cannot be achieved in a single stage process) is desired. Figure 2 is a schematic of a countercurrent recycle cascade of membrane and illustrates the flow from each stage.

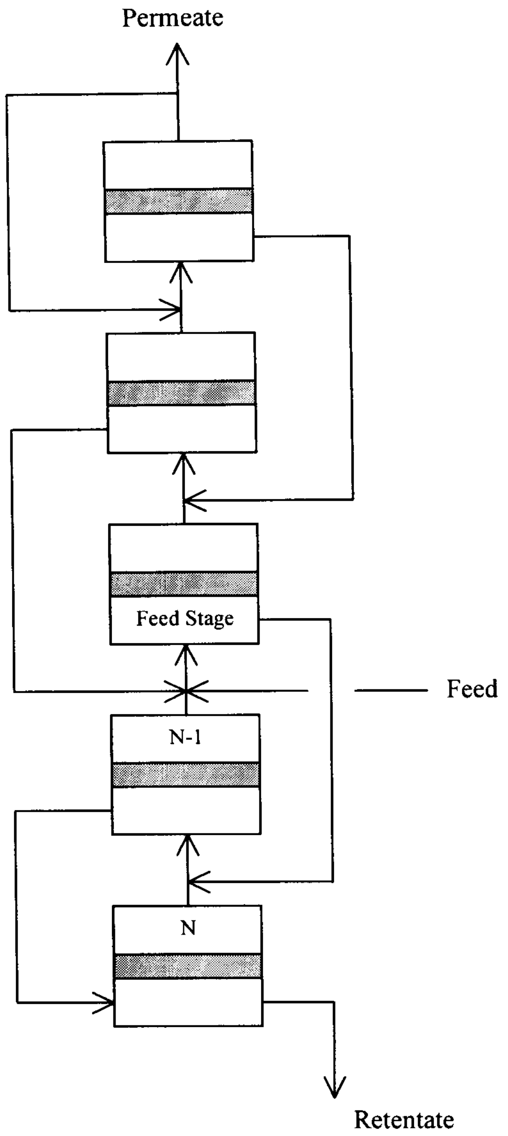

Figure 2. Schematic of a countercurrent cascade membrane separation process

The retentate and permeate of each stage are recycled in countercurrent flow. The feed stage is usually positioned in the middle of the operation. An enriching section concentrates the highly permeable component, while the stripping section concentrates the component with the low permeability. The permeate of the first stage and retentate of the last are withdrawn.

There are three methods of operating cascade membrane processes:

1. Constant cut, θ, constant overflow
2. Constant cut, varying overflow
3. Varying cut, varying overflow.

The cut, Θ, is the ratio of permeate flow rate to feed flow rate:

$$cut, \Theta = \frac{n_p}{n_F} \tag{3}$$

By keeping the cut and the overflow constant, all calculations are greatly simplified. In addition, fewer stages are required for a given separation. For this reason, the first case was selected over cases two and three. A disadvantage to using case one is that a larger amount of membrane is required because each stream remains constant. Also, the permissible range of cut values is ½ ≤ Θ ≤. 0 (Hwang and Kammermeyer).

The number of stages required to achieve the desired separation, can be determined graphically using the McCabe-Thiele method. This method involves stepping off between the equilibrium line and the operating line. The equilibrium line is obtained using the following equation (Porter):

$$y_n = \frac{x_n}{\alpha + (1-\alpha)x_n} \tag{4}$$

The operating line for the enriching section is (Hwang and Kammermeyer):

$$\text{Enriching:} \quad y_{n+1} = \gamma x_n (1-\gamma) x_D \tag{5}$$

where

$$\gamma = \frac{1-\Theta}{\Theta} \tag{6}$$

To obtain the operating line for the stripping section, a value for the mole fractions of the bottoms is plotted. Another point is plotted where the enriching section line intersects with the mole fraction in the retentate of the feed stage. A line drawn between these two points represents the stripping section.

A 45° line is plotted and equilibrium stages are stepped off a staircase from the top (x_D, y_D) to the bottom (x_B, y_B). One stage is represented by a step on the staircase, which goes from the operating line to the equilibrium line, then back down to the operating line.

The membrane area is related to the mole fractions, pressures and permeability by the following equation:

$$A = \frac{y_{PA} n_P}{\overline{P}_{MA} \left(x_{RA} P_R - y_{PA} P_P \right)} \tag{7}$$

This area is calculated for the feed stage only. To obtain the membrane area for the entire process, simply multiply by the number of stages. This will be used to size the process and generate some capital and operating cost estimates for the process.

Process Equipment and Operating Conditions

Thorough assessment of the type of equipment necessary and operating conditions has been conducted. Along with the necessary components for the separation process, auxiliary equipment has also been identified. Appendix provides schematics, product specifications and detailed descriptions.

Membrane

A polyphosphazene membrane is examined for use in separating tritium from water. Several membranes, configured into hollow cylinders, are placed into modules. This configuration offers a large surface area for separation in a small volume. The modules are representative of one stage. Therefore, the total number of modules, and thus the size of the process, is largely dependent on the number of stages. The modules will operate much like heat exchangers, in that the streams are fed to each module shell side. The retentate also leaves shell side, while the permeate leaves tube side.

Based on the research conducted by Nelson, the membrane thickness is about 1 micron. The feed, retentate, and permeate are operated at pressures of 3.26, 2.58 and 1.00 atm respectively. The optimum operating temperature is about 4°C but should not exceed 10°C (Nelson).

Product specifications for nanoporous membranes were obtained from Osmonics ™ and used for the process design. Their membranes offer 15% recovery (in the permeate) of the $MgSO_4$ feed solution, which is used to determine the permeability for water. The membranes are 8 inches in diameter with a length of 40 inches.

Filter

Removal of any solid particles from the stream will be necessary to uphold the life of the membrane. A filter will be necessary to remove solid $CaCO_3$ particles from the waste stream. These particles constitute 1% (by weight) of the stream content. The filter will be located just upstream of the membrane system. Orival, Inc., a manufacturer of automatic water filters has been identified. These filters have the capability to remove particles as small as 10 microns.

Upon removal of solids, the filtered water will be sent to the membrane. The $CaCO_3$ particles will be mixed with the cement for disposal.

Cryogenic Condenser

A cryogenic condenser will be used to condense the tritiated water vapor from the air. This technique is often used for recovery of VOCs that are emitted during manufacturing process. Liquid nitrogen is the most commonly used as a low-temperature refrigerant for the system. Because the water vapor will account for such a small volume, it will not be purified, and rather routed for disposal.

Auxiliary Equipment

A glove box with air lock chambers and 10 sets of glove ports will encase the entire process. The volume of the box is most dependent on the number of stages for the separation operation. In addition enough space must be incorporated to allow removal of membrane units and other equipment.

Several centrifugal pumps will be needed for transporting streams from each stage of the separation system. A vacuum pump will also be required to maintain a negative pressure in the encasement.

A cement mixer and 55 gallon steel drums will be used for disposal. The concentrated tritiated water will be used to make the cement. Drums will be filled and disposed of in a landfill.

Results and Discussion

Membrane Process

Using the Nelson paper as a guide, the maximum reject conditions were used to establish the separation factor of 2.33 (this corresponds to 43% reject). With this separation factor, an equilibrium curve was constructed using equation 4 as shown in Figure 4 (located in Appendix). We decided to remove 95.00% of the tritium from the feed and have 99.9% of the flow leave the bottom (retentate). With these conditions, the retentate is just below the EPA regulations. The first approach attempted to complete this separation using one stage, with the given tritium feed fraction, the desired retentate fraction is impossible to obtain using just one stage. This was determined by using equation 8 (Hwang and Kammermeyer):

$$x_{oM} = \frac{x_f\left[1+(\alpha-1)\Pr(1-x_f)\right]}{\alpha(1-x_f)+x_f} \tag{8}$$

The minimum reject fraction in one stage was determined to be 1.46e-14, which is too large. Thus, multiple stages would be required to obtain the desired separation. As discussed earlier, a Case 1 recycle cascade was used to perform the separation. Since the feed contains an extremely small amount of tritium (x_f=2.25e-14), a different approach would be needed to use the McCabe-Thiele method on this equilibrium, thus the equilibrium curve was plotted using log-log axis. This allows several orders of magnitude to be displayed over a small area. Table 1 shows the results of for the design conditions.

Table 1. Summary of Membrane Conditions (per min basis)

α	yf	xb	yd		HTO Balance	%HTO Removed	%Flow Bottom
2.33	2.25E-14	1.12E-15	2.13E-11		0	95.00%	99.90%

Θ	γ	F (mol/min)	D (mol/min)	B (mol/min)	xf
0.50025	0.999	208.18	0.208	207.97	2.03E-14

Moles HTO Feed	Moles HTO Bottom	Mass HTO Bottom	Moles H_2O Bottom	Mass H_2O Bottom	vol (L/min) Bottom	Ci Bottom	Ci/L Bottom
4.68E-12	2.34E-13	4.68E-12	207.97	3743.40	3.74	4.54E-08	1.21E-08

Area	Moles HTO Top	Mass HTO Top	Moles H_2O Top	Mass H_2O Top	vol (L/min) Top	Ci Top	Ci/L Top
248.60	4.44E-12	8.88E-11	0.208	3.75	3.75E-3	8.63E-07	2.30E-4

The results of the McCabe-Thiele method for determining the number of stages required are shown in Figure 5 (located in Appendix).

Economic Analysis

The net present worth of the project was determined in order to estimate its economic scope. Figure 3 is a cash flow diagram for the project. It is important to note that there is no income associated with the project; therefore all cash flows are negative. These costs were divided into several different categories in order to calculate the net present worth (Tables 2 through 8 located in the Appendix summarize each cost). These categories include initial equipment costs, labor costs, membrane costs, pump costs, variable costs, and electricity cost. Evaluating separate categories is useful because different pieces of the process equipment have different lives and must be replaced at different times. For example, the glove box is assumed to last the entire project life of 20 years, while the pumps are expected to last for only 5 years. There is no discount factor applied to the

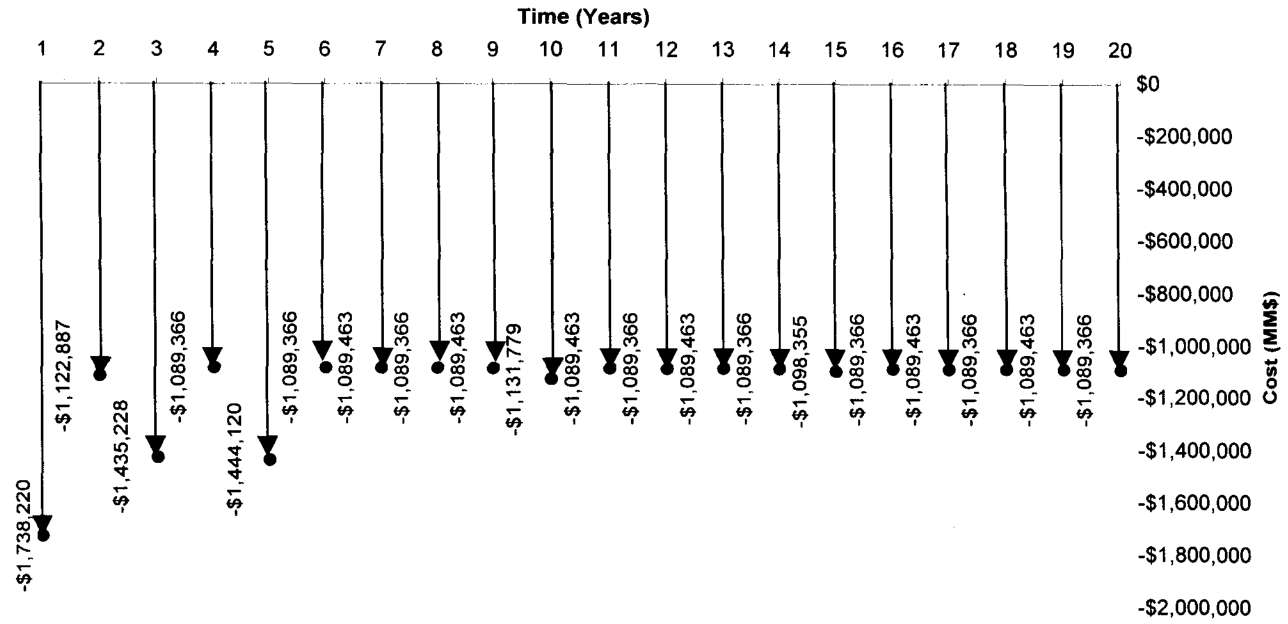

Figure 3. Cash Flow Diagram

costs that are incurred in year 1 because they are at their present value. Two different discount factors were applied to determine the present worth for costs not applied in year 1. For costs that are applied every year (labor, electricity, and variable costs) the present value of an annual cost discount factor was used. Equation 9 shows how these annual costs were calculated:

$$P = A\left(\frac{(i+1)^N - 1}{i(i+1)^N}\right) \tag{9}$$

where P is the present worth, A is the annual payment, N is the number of years, and i is the interest rate. For costs with unequal lives, a table was generated which gave the present worth of the future cost. Equation 10 shows how the present worth of a future cost is generated:

$$P = F(1+i)^{-N} \tag{10}$$

where P, N, and i are as above and F is the future cost. The Net Present Value is determined by the sum of each present value and is -\$10,113,610.

There really is no need to optimize the design because as bottoms (retentate) flow increases, the top (permeate) flow decreases, causing the disposal costs to decrease. Also, as the cut is lowered (closer to total reflux), the area decreases, as do the number of stages required, thereby reducing costs. The major cost associated with the design is the labor cost.

Conclusions

For the design specifications of 95.00% removal of tritium with 99.90% flow of EPA approved water, the following conclusions can be made:

- 13 stage recycling cascade membrane process
- 483,626 gal/yr of clean water produced
- 35 barrels/yr of waste produced
- NPV of -$10,113,610
- Majority of cost due to labor
- As cut decreases, area, stages, and cost decrease

Nomenclature

n_P	permeate flow rate, mol/min
n_F	feed flow rate, mol/min
x_D, y_D	ml fraction of tritium at the top of the cascade
x_{FA}	mol fraction of trituim in feed
x_{RA}	mol fraction of tritium in the retentate
x_{oM}	minimum mol fraction in the retentate using one stage
y_{PA}	mol fraction of tritium in the permeate
P_P	permeate pressure, atm
P_R	retentate pressure, atm
P_{MA}	Permeability,mol-m/min-m^2-atm
$\overline{P}_{MA}$	Permeance, mol/min-m^2-atm
ΔP	difference in retentate and permeate pressures, atm
N	flux through membrane, mol/m^2-s
l_m	membrane thickness, μm
V_p	molar permeate flow rate, mol/min
A	membrane area, m^2

Greek

α	separation factor
Θ	cut, ratio permeate flow to feed flow
γ	ratio of retentate flow to permeate flow

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http://www.orival.com/water.shtml. Automatic Water Filters

http://www.osmonics.com

APPENDIX A. Membrane Specification Sheet from Osmonics

OSMONICS®

Nanofiltration Membrane Elements - DS-5, Standard Flux

Product Information

These elements are used for **dye removal/concentration**, and **sodium chloride diafiltration**. They feature a fiberglass outerwrap and standard feed spacers. Other materials of construction and special feed spacers are available.

The proprietary DS-5 thin-film nanofiltration membrane is characterized by an approximate molecular weight cut-off of 150-300 daltons for uncharged organic molecules. Divalent and multivalent anions are preferentially rejected by the membrane while monovalent ion rejection is dependent upon feed concentration and composition. Since monovalent ions pass through the membrane, they do not contribute to the osmotic pressure thus enabling DS-5 nanofiltration systems to operate at feed pressures below those of RO systems.

Membrane Specification - DS-5

Membrane: Proprietary nanofiltration thin-film membrane (TFM®).

Applications: Dye removal/concentration, heavy metals removal, acid purification, and sodium chloride diafiltration.

Rejection characteristics: Divalent and multivalent anions are preferentially rejected by the membrane while monovalent ion rejection is dependent upon feed concentration and composition. The membrane is characterized by a molecular weight cutoff of 150-300 daltons for uncharged organic molecules.

Recommended pH: 2.0-11.0 operating range and 1.0-11.5 cleaning range for standard construction elements.

Chlorine tolerance: 1,000 ppm-hours, dechlorination recommended.

Maximum temperature: 122°F (50°C) with standard element construction and up to 158°F (70°C) with special element construction.

Element Series Designation: DK, DL

Element Specifications

Model	GPD (m³/d)	MgSO4 Rejection	ActiveArea ft² (m²)
DK4040F	2,000 (7.56)	98%	90 (8.36)
DK8040F	8,000 (30.24)	98%	350 (32.52)

Specifications are based on a 2,000 mg/L MgSO4 solution at 100 psig (690 kPa) net pressure, 77°F (25°C), 10% recovery, after 24 hours. Individual element flux may vary ± 15%.

Operating and Design Parameters

Membrane: Thin film membrane (TFM)

Typical operating pressure: 70-400 psig (483-2,758 kPa).

Maximum pressure: 500 psig (3,448 kPa).

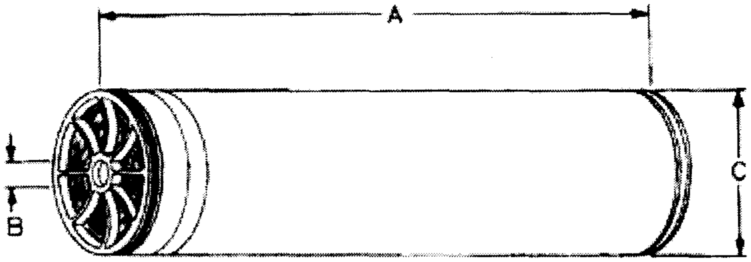
Maximum temperature: 122°F (50°C).

Recommended pH: Operating range 2-11 (pH<1 with special element construction), cleaning range 1-11.5.

Chlorine tolerance: 1,000 ppm-hours, dechlorination recommended.

Recommended per vessel in a system	**Elements per pressure vessel**					
	1	**2**	**3**	**4**	**5**	**6**
delta P - psig (kPa)	10 (69)	20 (138)	30 (207)	38 (262)	45 (310)	50 (345)
% Recovery	15	25	35	45	53	53

Element Dimensions and Weight



Model	Dimensions, inches (cm)			Dry boxed weight lbs. (kg)
	A	B	C	
DK4040F	40.00 (101.6)	0.625 (1.59)	3.88 (9.86)	12 (5.45)
DK8040F	40.00 (101.6)	1.187 (3.01)	7.88 (20.02)	32 (14.53)

Length includes ATD's. All elements are shipped dry.

APPENDIX B. EPA Regulations for Drinking Water

[Code of Federal Regulations]
[Title 40, Volume 15, Parts 136 to 149]
[Revised as of July 1, 2000]
From the U.S. Government Printing Office via GPO Access
[**CITE: 40CFR141.16**]

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TITLE 40--PROTECTION OF ENVIRONMENT

CHAPTER I--ENVIRONMENTAL PROTECTION AGENCY (CONTINUED)

PART 141--NATIONAL PRIMARY DRINKING WATER REGULATIONS--Table of Contents

Subpart B--Maximum Contaminant Levels

Sec. 141.**16** Maximum contaminant levels for beta particle and photon radioactivity from man-made radionuclides in community water systems.

(a) The average annual concentration of beta particle and photon radioactivity from man-made radionuclides in drinking water shall not produce an annual dose equivalent to the total body or any internal organ greater than 4 millirem/year.

(b) Except for the radionuclides listed in Table A, the concentration of man-made radionuclides causing 4 mrem total body or organ dose equivalents shall be calculated on the basis of a 2 liter per day drinking water intake using the 168 hour data listed in ``Maximum Permissible Body Burdens and Maximum Permissible Concentration of Radionuclides in Air or Water for Occupational Exposure,'' NBS Handbook 69 as amended August 1963, U.S. Department of Commerce. If two or more radionuclides are present, the sum of their annual dose equivalent to the total body or to any organ shall not exceed 4 millirem/year.

Table A--Average Annual Concentrations Assumed to Produce a Total Body or Organ Dose of 4 mrem/yr

Radionuclide	Critical organ	pCi per liter
Tritium............................	Total body.............	20,000
Strontium-90........................	Bone marrow............	8

[41 FR 28404, July 9, 1976]

APPENDIX C. Membrane Curves

Figure 4. Tritium Equilibrium Curve

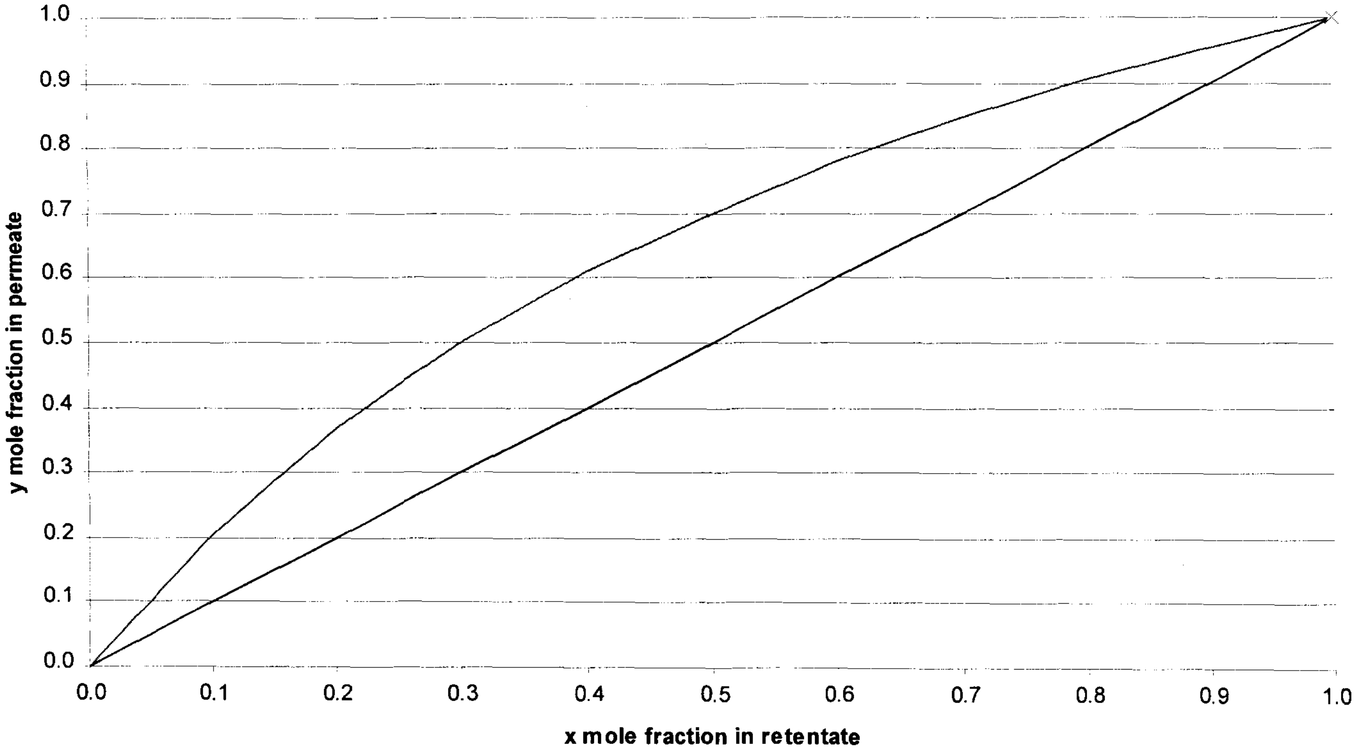
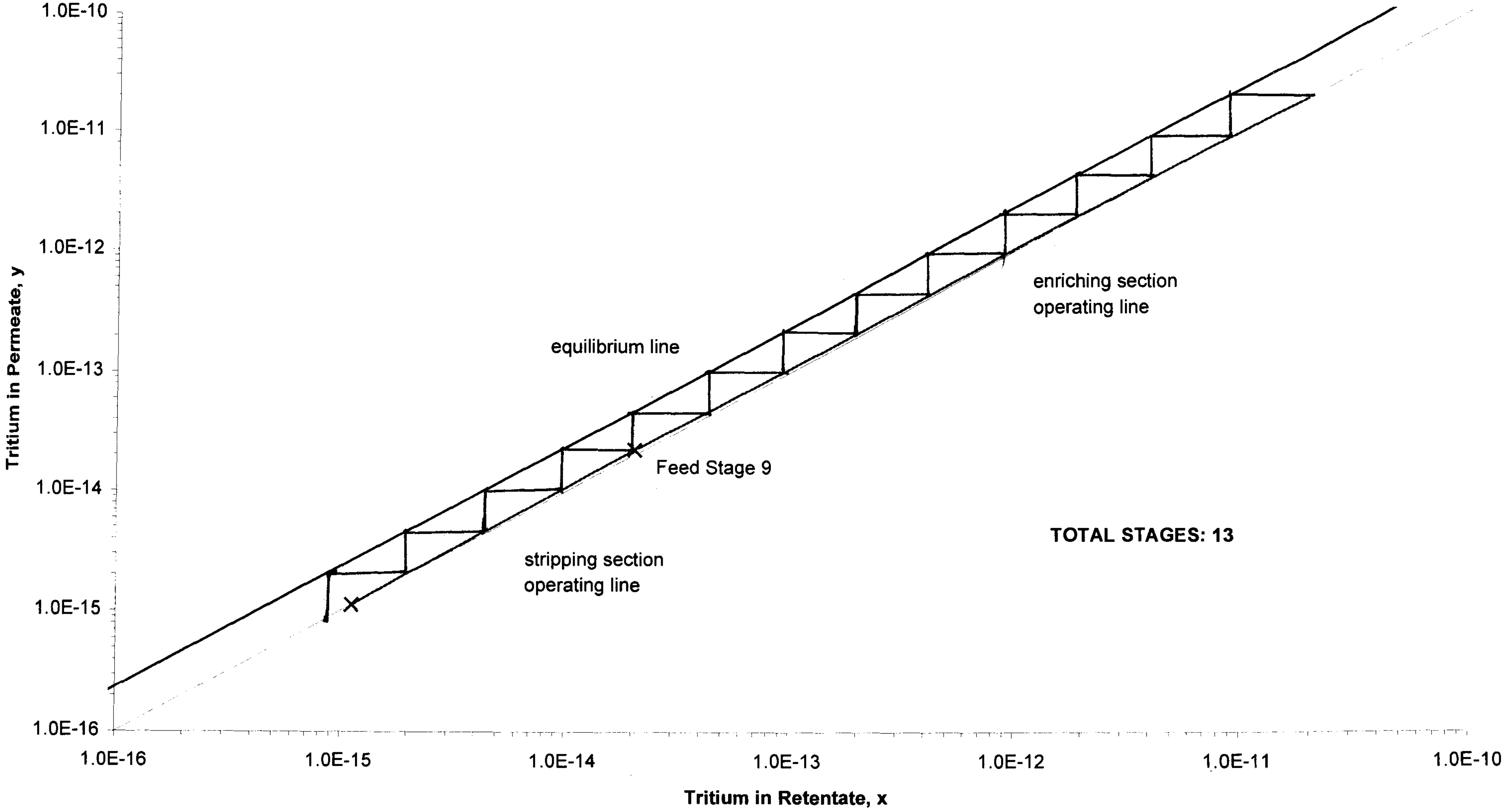


Figure 5. McCabe-Thiele Method to Determine Number of Stages



APPENDIX D. Cost Tables

Table 2. Initial Equipment Cost

Equipment	CAPACITY	UNITS	QUANTITY	TOTAL SIZE	COST	TOTAL COST
Water Feed Pump	2.5	hp	1	1	$352	$352
Stage Pump	2.5	hp	20	20	$352	$7,040
Cryogenic Unit	50	ft^3/min	1	50	$50,000	$50,000
Solid Filter	125	in^3	1	125	$4,950	$4,950
Membrane Module	1	m^2	3224	3224	$97	$312,341
Concrete Mixer	21	ft^3/min	1		$19,000	$19,000
Vacuum Pump	1	hp	1		$1,500	$1,500
Glove Box	0.89	m^3	15	17	$12,950	$220,150
					EQUIPMENT COST	$615,333

Table 3. Labor Costs

Labor	Salary	Benefits	S.S./Worker's Comp.	Overhead	Cost per worker	# of Workers	Labor Cost ($/year)
Technicians	$40,000	$12,000	$4,000	$40,000	$96,000	8	$768,000
Engineer	$60,000	$18,000	$6,000	$60,000	$144,000	1	$144,000
Supervisors	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	1	$136,800
						Total	$1,048,800

Table 4. Annual Variable Costs

Variables	Cost	Unit	Yearly Usage	Unit	Total Cost
Concrete	$0.38	gal	1444	gal/yr	$551.25
55-Gallon Drums	$4.55	gal	35	barrels/yr	$159.09
Disposal Cost	$33.42	gal	1925	gallons/yr	$64,333.50
Gloves	$120.00	pair	10	pair/yr	$1,200.00
Liquid N_2	$0.80	gal	365	gal/yr	$292.00
				Total	$65,335.84

Table 5. Annual Electricity Costs

Electricity	$0.04	kW hr			
	hp	kW/hr	#		Cost
Feed Pump	2.5	15194	14	212711	$8,508.44
Vacuum Pump	1	6077	1	6077	$243.10
				Total	$8,751.54

Table 6. Membrane Cost (2 year life)

	Year	DCFR	Cost
	2	0.797	$248,996
	4	0.636	$198,498
	6	0.507	$158,242
	8	0.404	$126,149
	10	0.322	$100,565
	12	0.257	$80,170
	14	0.205	$63,911
	16	0.163	$50,950
	18	0.130	$40,617
	20	0.104	$32,379
Membrane Cost	$312,341	2 year life	$1,100,478

Table 7. Pump Cost (5 year life)

	Year	DCFR	Cost
	5	0.567	$ 5,045.56
	10	0.322	$ 2,862.99
	15	0.183	$ 1,624.54
	20	0.104	$ 921.80
Pump Cost	$8,892	5 year life	$10,455

Table 8. Net Present Value Analysis

n	20	i	12%
P/A(i, n)	7.4694		PW
Initial Equipment Cost	$615,333	1st year	$615,333
Labor Cost	$1,048,800	yearly	$7,833,952
Variable Cost	$65,336	yearly	$488,022
Energy Cost	$8,752	yearly	$65,369
Membrane Cost	$312,341	every 2 years	$1,100,478
Pump Cost	$8,892	every 5 years	$10,455
		Net Present Worth	-$10,113,610