

PETROLEUM CONTAMINATED AQUIFER REMEDIATION

By

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Approved by:

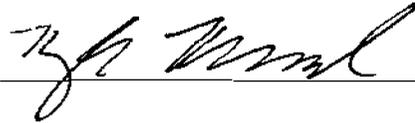


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Abstract:

In the state of Arizona there are a total of 1,134 contaminated sites caused by leaking underground storage tanks (LUST). Because a large portion of these sites are contaminated with petroleum based products, there is a large demand for petroleum contaminated water remediation systems. Many current remediation systems extract onsite but send out the product to be treated offsite. Here, we have designed a self sufficient remediation system that will remove and treat both groundwater and petroleum products. The system designed can process a maximum of 300 SCFM of contaminated vapors, 5589 gal/day of contaminated water, and 18 gallons per day of petroleum based products. The contaminated water is treated with an air stripper while the contaminated vapor is treated using granular activated carbon. The petroleum is catalytically cracked in a reactor to break down all of the hydrocarbons into methane which can then be used onsite to power the system. The economics of the design suggest that the system should not be built, however, reconsidering some of the operational cost it is thought that this system can remediate ground water while making a profit.

Group Roles and Responsibilities:

The other members of this design group are Amanda Kiesl, Christopher Lewis, and Adam Manasse. Amanda Kiesl was responsible for the safety information as well as all economical calculations. Christopher Lewis worked on the unit operation calculations for the air stripper, knockout pot, and decanter. Adam Manasse wrote a large portion of the introductory material and was responsible for all the environmental information.

The majority of the time that I spent working on this design project was put into doing unit operation calculations for the petroleum process section of the system. This included mass and energy balances for the reactor, compressor, membrane, and power generator. I also worked on utility calculations to ensure the system could power itself. For writing, I worked on the summary and conclusions and recommendations sections. Additionally, I took on the role of group leader for the semester where I was responsible for contacting our design mentor, setting up meetings, and keeping everyone on task.

Petroleum Contaminated Aquifer Remediation

THE GREEN TEAM

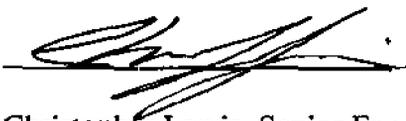
DATE SUBMITTED: May 1, 2009



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1. Introduction

1.1 Overall Goal

The goal of this project is to design a remediation process which will be used to recover spilled petroleum fuel product and clean up groundwater contaminated by such spills. The Bennett Oil facility in Prescott, Arizona is used as the model for our design since the site provided insight into a complex spill involving many points of infiltration and a variety of petroleum contamination in the form of gasoline, kerosene, and oil. Bennett Oil began operating at the site in 1984 and currently uses the facility for bulk fuel sales in addition to offices (Appendix 9.5). A convenience store operated by Bennett Oil is also located at the site. Fourteen underground storage tanks are currently registered at the site, and are used to store gasoline, diesel, kerosene, new oil and used oil. A known spill occurred in March of 1985 when approximately 920 gallons of regular gasoline leaked from a product line to the surrounding area. The leak is identified and catalogued as LUST (leaking underground storage tank) #0133.01 (Appendix 9.5).

Further spillage is believed to have occurred from other leaking underground storage tanks at the site. Free product (gasoline and kerosene) has been detected at Russell Well, located just north of the site; and at Williams well, located 700 feet northwest of the site. The kerosene came from a faulty dispenser line which is catalogued as LUST #0133.03. Diesel fuel is also detected in an analysis of free product recovered from the site, and is believed to be coming from another leaking storage tank at the site. Contamination plumes have begun to spread as evidenced by well sampling in the area, and it is our goal to remediate the water extracted from the water Table to the established standards. The main contaminants of concern are MTBE (methyl tert-butyl ether) and BTEX (the hydrocarbons benzene, toluene, ethylbenzene, and xylene). Since these contaminants are presents in the highest concentrations and have the strictest regulations, it is believed that removal of these contaminants will subsequently

remediate other less important contaminants to regulatory standards. The maximum contaminant levels (MCLs) of BTEX and MTBE can be found in section 5 along with additional environmental information. Ideally, the remediation system will be adaptable to a wide range of conditions making the system mobile and available for use at a variety of sites.

1.2 Background

Current market value does not apply in the traditional sense in this scenario. Consumers do not seek out the services this report describes. Rather, they are a necessity to avoid health problems and fines imposed by environmental damage. Additionally, the value of real estate is re-established by removing contamination, as the owner of the property is the responsible party for clean up. For these reasons, this section will focus mainly on background information rather than current market value.

The Green Team approached the problem stated above by first researching extraction methods used to recover spilled fuel product. It is found that a very common extraction method in such a situation is dual phase extraction (DPE).

A dual phase extraction system operates with the use of a high powered vacuum pump. The vacuum causes vapor phase pressure gradients to form toward the vacuum. This pressure gradient is also imposed on any subsurface liquid present and groundwater as well as free petroleum product accumulates at the vacuum. The vapor and liquid phase contaminants are then extracted using the hydraulic pressure generated by the pump. The vacuum suction pressure can be increased to increase the hydraulic pressure and subsequent recovery rates. A DPE system effectively increases the soil vapor recovery as well by lowering the water Table and increasing the air permeability in the vadose zone. DPE also stimulates biodegradation of contaminants by microorganisms as a result of introducing a greater flow of oxygen into the soil, similar to the

effects of bioventing. It should be noted that the results of DPE will vary with soil type and the best results will show in layered, fine ground soils (U.S. Environmental Protection Agency, 2008). Operating ranges of a DPE system are approximately 20 SCFM up to 4400SCFM (GeoTech).

A site visit is set up following our initial research in order to observe how the extraction process is performed in practice. The site of interest is the Thomas O'Price Service Center in south Tucson, AZ where approximately 400,000 gallons of gasoline product is spilled in 1988. Two main phases of operation existed at the site: extraction and remediation.

In the extraction process, a skimmer pump is floated on the water Table and extracted low volumes of contaminated groundwater on a semi-continuous basis. The pump is turned off when there is no space available in the storage tank that held the extracted groundwater. A separate pump is used to extract vapor phase contaminants from the soil in the area. This process is commonly known as soil vapor extraction or SVE. Both the liquid and vapor phase contaminants were then individually fed, through separate pipelines, to an enclosed area where the remediation process took place. At this site, the liquid and vapor phase contamination were dealt with in separate extraction processes, rather than combining the two processes into a single DPE process.

The groundwater is initially fed to storage tanks and then pumped to a gravity decanter when it is available for use. In this step, the free petroleum product is separated from the groundwater. The groundwater is then sent to storage tanks and shipped offsite for remediation by a separate environmental remediation company. The free petroleum product is dealt with in the same manner. The vapor phase contaminants recovered from the soil, however, were treated onsite.

The vapors were pumped to a thermal oxidizer where they were combusted and released into the atmosphere.

The O'Price site offered several valuable insights into how a petroleum product spill is handled and treated. Observing the basic equipment and operation principles at the site gave Green Team some background information concerning groundwater remediation processes as well as a basis for the design of an extraction/remediation system.

1.3 Project Premises

A dual phase extraction (DPE) process is chosen as the method to extract the free product from the water Table. It is assumed that 300 SCFM of vapor is extracted with 5607 gallons of liquid/day (United States Environmental Protection Agency, 1997). The liquid is composed of 5589 gallons of water, with the remaining 18 gallons composed of petroleum product assumed to be 50% gasoline and 50% kerosene. After extraction, the next step in our approach is to separate our extracted product. This is necessary as each of the components (vapor contaminant, free petroleum product, water) extracted will be treated individually. First, a mixture of all three components listed above is pumped from the groundwater into a vapor/liquid separator. Here the vapors are separated for later treatment. The liquid mixture is then sent to a gravity decanter where the free petroleum product is separated from the groundwater. The decanter operates on the principle of density differences between the two fluids. Since the water is denser, it will sink to the bottom of the decanter where it will then be fed to additional treatment units. The petroleum product will float on top, and is then sent to a continuous catalytic reactor.

The catalytic reactor's function is to crack the hydrocarbons present in the petroleum product and convert them to methane gas, which will be used for energy needs on site. The reactor will operate exothermically, so cooling will be required to prevent the reactor from overheating and

compromising safety. Cooling will be provided by a cooling fluid in the form of recycled water from elsewhere in the process. The catalyst used for this process is nickel oxide on silica. Research has shown that with use of a nickel oxide catalyst, nearly 100% of the petroleum is converted to natural gas in the hydrocarbon cracking reaction. The methane generated by the reactor will be captured and pumped to a power generator. The generator will utilize the methane gas to power each piece of equipment on site.

Since most components of petroleum products are soluble in water, the groundwater that is obtained from the dual phase extraction process will contain a large amount of contaminants. Particular contaminants of concern here are: methyl tert-butyl ether (MTBE), lead, and the organic hydrocarbon contaminants benzene, toluene, ethylbenzene, and xylene (collectively known as BTEX). To remediate the contaminated water, a stripping column will be used to clean the water to regulatory standards. It is determined that in order to purify the water such that all contaminants meet regulation, 3.1 SCFM of air will need to be supplied to the column, counter flow to the incoming water. The water will be passed downwards through the column, and a “dirty gas” stream containing the contaminants will exit from the top of the column. The recovered water will be used to cool the reactor, prior to being pumped back into the existing water Table.

The vapor phase contaminants pumped from the ground will be combined with the dirty gas stream from the stripping column. The vapors will adsorb onto granular activated carbon (GAC), which can then be recharged or disposed of. Initial estimates from Carbtrol indicated that about 95 pounds of activated carbon will be needed per day to remediate the vapor phase contaminants. Two GAC units will operate asynchronously to ensure that one GAC unit will be operational while the other unit is being repacked with activated carbon.

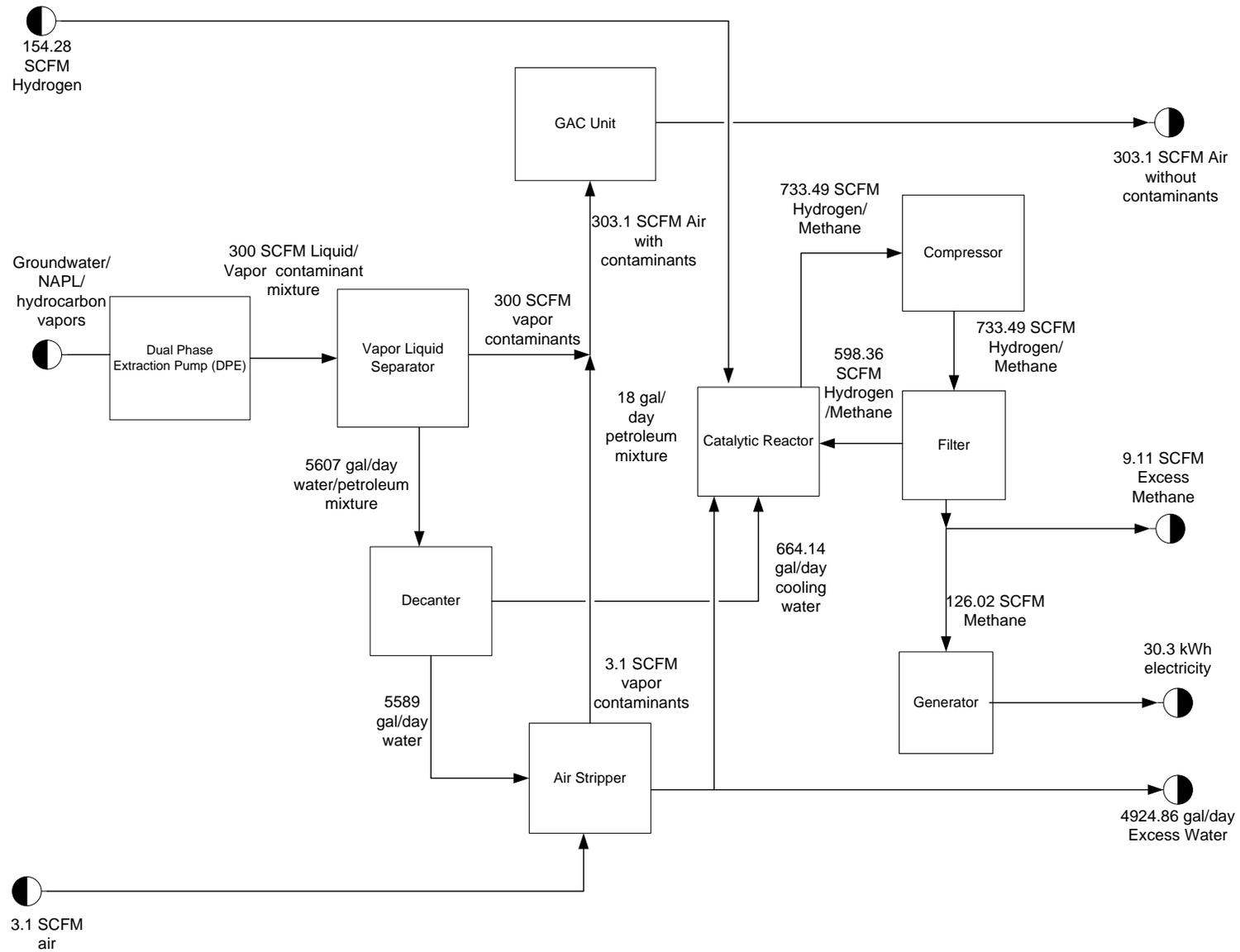
1.4 List of Assumptions

- 300 SCFM air is pumped from the DPE
- The vapors from the liquid ring pump act ideally
- V-101 is assumed to be a horizontal, cylindrical vessel with a length-to-diameter ratio of 3 to 4
- MTBE is completely soluble in water and will remain in the liquid stream in V-101
- Liquid solution in V-102 will separate into 2 fluids: gasoline/kerosene mixture and water
- Complete separation of the two liquids in V-102
- V-102 is assumed to be a horizontal, cylindrical vessel and the length of V-102 should be 5 times the diameter
- Minimum required air flow rate in S-101 is equal to the required air flow rate to remove the MTBE
- Required air flow rate is three times that of the minimum flow rate calculated
- The air used for stripping acts ideally
- Diameter of S-101 is 0.75 meters
- Composition of stream 18 is 50% gasoline and 50% kerosene
- 100% conversion of hydrocarbons to methane in R-101
- There is always twice the amount of hydrogen than needed in R-101
- The gas mixture in stream 20 mixes and acts ideally
- The pressure drop across S-101 is 5 psi
- The difference in price between an open and closed vessel is negligible

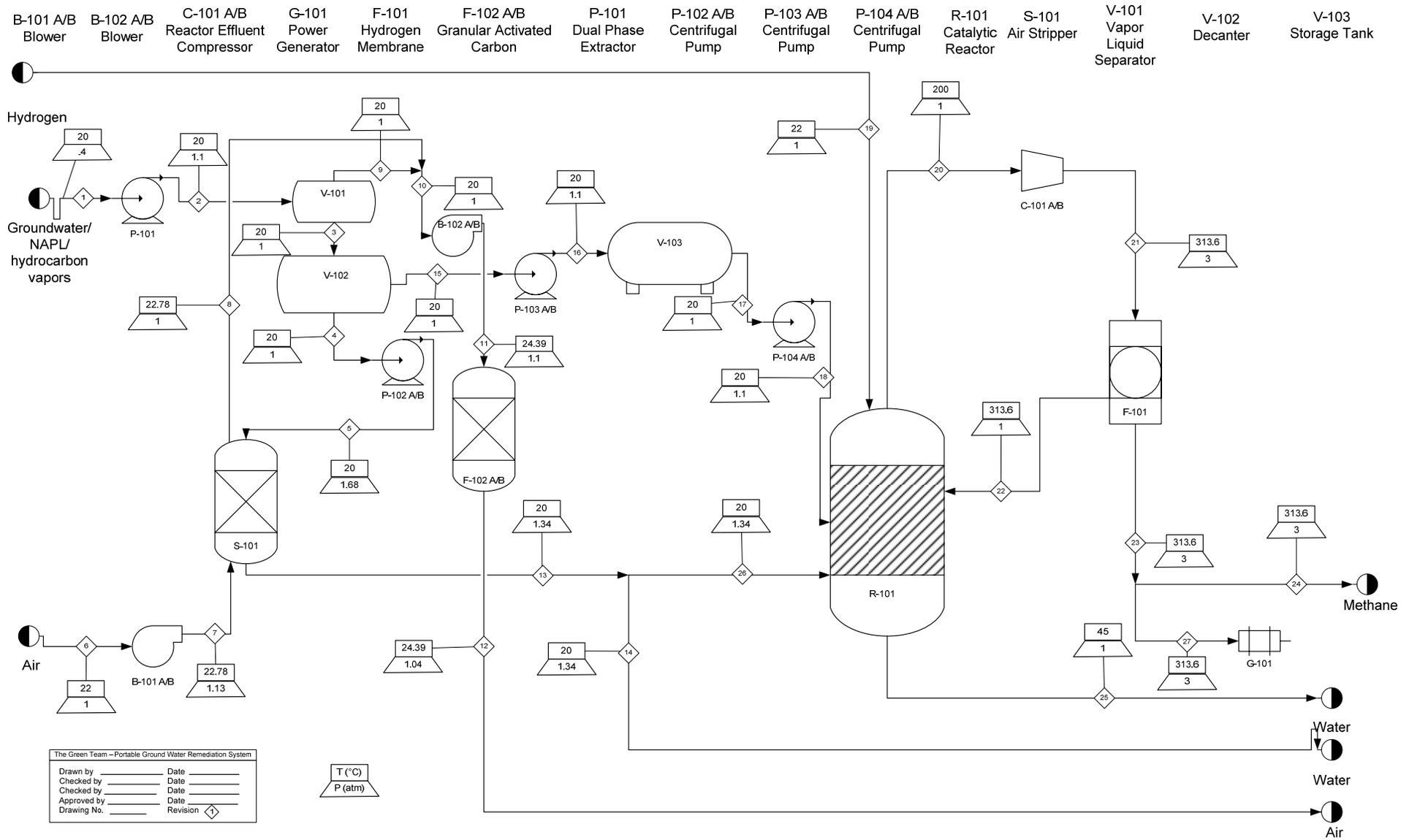
- After pricing the reactor vessel, the purchase cost will be modified to accommodate for a material different than fiberglass by multiplying by the material factor for stainless steel 316
- Only one tenth of the gas flowing through the reactor is actually residing in the reactor at any given time
- The cost of platforms and ladders for S-101 is negligible since our tower is smaller than the required parameters
- Centrifugal compressor is used to lower maintenance requirements
- Assume B-101 is a Rotary Straight-Lobe Blower because of the lower power requirements
- B-102 is priced as a Centrifugal (turbo) Blower because of the break horsepower (18.23 Hp)
- Assume each blower is constructed of stainless steel 316, and are electric driven
- Bare module factor for V-102 is one as there are no installation costs
- P-102/103/104 were adjusted to a required horsepower in order to price each pump
- The salvage value of the equipment at the end of ten years was equal to 10% of the total depreciable capital
- Since the operating pressure is less than 1.34 psig, it can be assumed that t_p equals t_v , the average wall thickness for a vertical vessel because there will be no effect of wind load
- G-101 is assumed to be 80% efficient

2. Process Description, Rationale, and Optimization

2.1 Block Flow Diagram



2.2 Process Flow Diagram



2.3 Equipment Tables

Vessels	V-101	V-102	V-103	S-101	F-102 A/B
Type	Vapor/Liquid Separator	Decanter	Storage Tank	Counter flow Air Stripper	GAC filtration unit
Orientation	Horizontal	Horizontal	Horizontal	Vertical	Vertical
Temperature (°C)	20	20	20	20	24.39
Pressure (atm)	1.1	1	1.1	1.68	2
Flow	300 SCFM*	5607 gal/day	N/A	3.1 SCFM air 3.88 gal/min water	303.1 SCFM
Volume	6.69 gal	115.54 gal	550 gal	1.26 m ³	.25 m ³
MOC	Stainless Steel	Stainless Steel	Stainless Steel	Stainless Steel	PVC
Size					
Height/Length (m)	0.263	2.52	1.83	2.86	0.86
Diameter (m)	0.35	0.5	1.22	0.75	0.61

* fluid is in both liquid and gaseous phase and is better represented in SCFM

Pumps	P-101 A/B	P-102 A/B	P-103 A/B	P-104 A/B
Type	liquid ring	centrifugal	centrifugal	centrifugal
Driver Type	electric	electric	electric	electric
Flow (gal/day)	300 SCFM*	5589	18	18
Fluid Density (lb/ft ³)	unknown**	62.4	48.54	48.54
Power (shaft) (W)	14920	16.86	0.011	0.011
Suction Pressure (atm)	0.4	1	1	1
Discharge Pressure (atm)	1.1	1.68	1.1	1.1
MOC		Stainless Steel	Stainless Steel	Stainless Steel

* fluid is in both liquid and gaseous phase and is better represented in SCFM

**because fluid is in both liquid and gaseous phase and components vary, density is difficult to determine

Blowers	B-101 A/B	B-102 A/B
Type	Rotary Straight-Lobe	Centrifugal
Driver Type	electric	electric
Flow (SCFM)	0.0245	2909.1
Power (W)	0.148	13599.98

MOC

Stainless Steel

Stainless Steel

Reactors	R-101
Temperature (°C)	200
Pressure (atm)	1
Heat Duty (kJ/day)	-263257.38
MOC	Stainless Steel
Size	
Volume (m ³)	5.37

Compressors	C-101 A/B
Type	Centrifugal
Driver Type	electric
Flow (SCFM)	654.73
Power (shaft) (W)	1193.05
MOC	Stainless Steel

Filters	F-101
Type	membrane
Surface Area (m ²)	1.673
Flow (SCFM)	654.72
Permeate Flow (SCFM)	534.1
Non-Permeate Flow (SCFM)	120.62
MOC	Stainless Steel casing
Membrane MOC	Hollow Fiber

Generators	G-101
Type	Natural Gas Standby
Power Generation	.24 kWh/SCF natural gas
Natural gas consumption	112.27 SCF
Power Output	30.3 kWh

2.4 Stream Table

Stream Number	1	2	3	4	5	6	7	8	9	10
Temperature (°C)	20	20	20	20	20	22	22.78	22.78	20	20
Pressure (atm)	0.4	1.1	1	1	1.68	1	1.13	1	1	1
Vapor Mole Fraction	0.81	0.81	0	0	0	1	1	0.79	1	1
Mole Flow Rate (mol/hr)	257703.25	257703.25	49037.47	49018.43	49018.43	1.76	1.76	235.15	208665.78	208900.93
Component Mole Flow (mol/hr)										
Gasoline content	10.76	10.76	10.76	0	0	0	0	0	0	0
Kerosene content	8.28	8.28	8.28	0	0	0	0	0	0	0
Water	48968.44	48968.44	48968.44	48968.44	48968.44	0	0	0	0	0
BTEX vapors	36.97	36.97	0	0	0	0	0	0.1	36.97	37.07
MTBE	49.99	49.99	49.99	49.99	49.99	0	0	49.99	0	49.99
Methane	0	0	0	0	0	0	0	0	0	0
Nitrogen	164816.76	164816.76	0	0	0	1.39	1.39	146.20	164816.76	164962.96
Oxygen	43812.05	43812.05	0	0	0	0.37	0.37	38.86	43812.05	43850.91
Hydrogen	0	0	0	0	0	0	0	0	0	0
Stream Number	11	12	13	14	15	16	17	18	19	20
Temperature (°C)	24.39	24.39	20	20	20	20	20	20	22	200
Pressure (atm)	1.1	1.04	1.34	1.34	1	1.1	1	1.1	1	1
Vapor Mole Fraction	1	0.999999875	0	0	0	0	0	0	1	1
Mole Flow Rate (mol/hr)	208900.93	208814.01	48968.44	43149.31	19.04	19.04	19.04	19.04	102.32	863.76
Component Mole Flow (mol/hr)										
Gasoline content	0	0	0	0	10.76	10.76	10.76	10.76	0	0
Kerosene content	0	0	0	0	8.28	8.28	8.28	8.28	0	0
Water	0	0	48968.44	43149.31	0	0	0	0	0	0
BTEX vapors	37.07	0.026	0	0	0	0	0	0	0	0
MTBE	49.99	0.117	0.000199	0.000175	0	0	0	0	0	0
Methane	0	0	0	0	0	0	0	0	0	659.13
Nitrogen	164962.96	164962.96	0	0	0	0	0	0	0	0
Oxygen	43850.91	43850.91	0	0	0	0	0	0	0	0
Hydrogen	0	0	0	0	0	0	0	0	102.32	204.63

Stream Number	21	22	23	24	25	26	27
Temperature (°C)	313.6	313.6	313.6	21	45	20	313.6
Pressure (atm)	3	1	3	3	1	1.34	3
Vapor Mole Fraction	1	1	1	1	0	0	1
Mole Flow Rate (mol/hr)	863.76	704.63	159.13	10.73	5819.13	5819.13	148.4
Component Mole Flow (mol/hr)							
Gasoline content	0	0	0	0	0	0	0
Kerosene content	0	0	0	0	0	0	0
Water	0	0	0	0	5819.13	5819.13	0
BTEX vapors	0	0	0	0	0	0	0
MTBE	0	0	0	0	0	0.000024	0
Methane	659.13	501.1	158.83	10.71	0	0	148.12
Nitrogen	0	0	0	0	0	0	0
Oxygen	0	0	0	0	0	0	0
Hydrogen	204.63	203.53	0.3	0.02	0	0	0.28

2.5 Utility Table and Raw Materials

Utility	Amount	Price	Yearly Cost
Electricity	30.03 kW	N/A	N/A
Cooling Water	242410.15 gal/year	N/A	N/A

We are able to completely avoid utility costs by recycling products from our process.

Electricity is obtained from an onsite generator which is powered by natural gas. The natural gas is produced by the hydrocarbon cracking reaction taking place in R-101 (see process description in section 2.6). The groundwater which is extracted in our process will be used as cooling water after treatment (see process description in section 2.6).

Material	Amount	Price	Yearly Cost	Cost Source
Granular Activated Carbon	34675 lbs/year	1.35[\$/lb]	\$46,800.00	(Carbtrol)
Catalyst	2000 kg	15 [\$ /kg]	\$30,000.00*	
Hydrogen	1795 kg/year	2 [\$ /kg]	\$ 3,590.00	(Doty Ph.D, 2009)
Air	13603.1 SCFM	0	0	N/A

*Catalyst is a one-time cost

2.6 Written Description of Process

The optimized remediation process begins with the extraction of contaminated groundwater from the polluted site. The contaminated groundwater is removed from the soil by a vacuum pump (P-101) and exists in both the liquid and gaseous phase. This liquid/vapor stream is defined as stream 1 on the suction side of the pump, and as stream 2 on the discharge side of the pump. The pump discharge, stream 2, is sent to a vapor liquid separation unit (V-101) in order to isolate the vapor phase contaminants from the liquid phase contaminants.

The vapor phase contaminants are discharged from the separation unit (V-101) as stream 9 and then combined with stream 8 (additional vapor phase contaminants from elsewhere in the process) to form stream 10. Stream 10 is sent to a blower, and blown into an activated carbon filtration unit (F-102) as stream 11 in order to remove the vapor contaminants. The main contaminants of concern to be filtered out are BTEX (benzene, toluene, ethyl benzene, and xylene) and MTBE (methyl tert-butyl ether). The activated carbon absorbs the vapor phase contaminants, and clean air is released into the atmosphere from the top of the filtration unit F-102.

The liquid phase contaminants are discharged from the separation unit (V-101) as stream 3. Stream 3 at this point is a mixture of groundwater and free petroleum product. The components of stream 3 are then separated by a continuous gravity decanter (V-102). Recovered free petroleum product, the less dense component of stream 3, exits the top of unit V-102 as stream 15 and is pumped to a storage tank (V-103). The suction side of the pump is defined as stream 15 and the pump discharge is defined as stream 16. The denser component in the decanter, the contaminated water, exits the bottom of the decanter as stream 4.

The contaminated water exiting the decanter (stream 4) is cleaned in an air stripper (S-101) and recycled as cooling water later in the process. Stream 4 exits the decanter and is pumped as stream 5 into the top of the stripper where it comes into contact with a counter current flow of air (stream 7). The air is captured from the atmosphere as stream 6 and blown into the air stripper as stream 7. Stream 7 passes upwards through the column, stripping the water of the contaminants BTEX and MTBE to regulatory standards. The clean water, stream 13, is used as cooling water for the exothermic reactor, R-101. Only 12% of the water in stream 13 is necessary for the cooling requirement; the excess water is diverted as stream 14, and the remaining water is

sent to the reactor to be used for cooling purposes as stream 26. The contaminated gas stream, stream 8, exits the top of the stripper and is combined with the other vapor phase contaminant stream (stream 9) and treated in F-102 as described previously.

The recovered free petroleum product stored in unit V-103 is discharged from the storage tank as stream 17 and pumped to the exothermic reactor R-101 as stream 18. The mixture of petroleum products is combined with stream 19 in R-101, and subjected to a hydrocarbon cracking process with the aim of converting all the hydrocarbons in the mixture into methane gas. Cooling water is used to regulate the temperature of the reactor during the hydrocarbon cracking process. To ensure that all hydrocarbons are broken down completely to methane gas, a selective membrane is used to feed all larger hydrocarbons back to the reactor for further cracking. The reactor effluent, stream 20, is fed to the compressor unit C-101 where the gas is pressurized to 3 atm, and passed through the hydrogen membrane F-101. Methane gas will pass through the membrane as stream 23, while hydrogen (H_2) will be returned to R-101 in stream 22. 93.3% of the methane gas in stream 23 will then be used for energy to power the generator (G-101), while the unused methane (stream 24) will be sold.

2.7 Rationale for Process Choice

Green Team's approach used the O'Price site as a design model, optimizing several of the steps involved in the process. The most significant optimization was the decision to treat nearly all contaminants onsite rather than shipping them off for treatment. It is believed that this will significantly lower the cost of the remediation process, in addition to decreasing the impact on the site (e.g. re-implementing the groundwater, aerating the soil).

DPE was chosen as the extraction process because of its adaptability and high volume yields. It is often important, especially in large spills, to remove the spilled product in a timely fashion

in order to prevent a contaminant plume from spreading – this is accomplished by the high volumes extracted using DPE. Furthermore, a DPE process simplifies the extraction step of the remediation process, since only one pipeline is required to remove both liquid and vapor phase contaminants, keeping the impact on the site low in comparison to a liquid extraction coupled with a SVE system. The groundwater extracted by this process will be separated in the same manner as that at the O’Price site by use of a liquid/vapor separator followed by a continuous gravity decanter.

At the O’Price site, the contaminated groundwater was shipped offsite to be treated – a common practice for such an operation. However, Team Green wanted the process to be completely onsite in order to increase the effectiveness of the remediation process, as well as lower offsite costs such as transportation and other environmental services. For this reason a stripping column was chosen to remediate the contaminated groundwater. A stripping column can be maintained onsite and the treated water can be recycled and used for utility requirements onsite.

Free petroleum product is also commonly shipped offsite for treatment in a groundwater recovery operation. In keeping with the decision to remediate the contamination onsite, it was decided that the free petroleum product be converted to a product that can be used on site or sold. This is accomplished by a hydrocarbon cracking process in which a catalytic reactor converts the hydrocarbons present in the petroleum product mixture into methane gas. A selective membrane will also be added to the process. The rationale here is to remove hydrogen from the methane product and ensure that it meets specifications to be sold. A compressor will pressurize the reactor effluent so that the membrane has a large enough pressure differential for efficient separation. The methane gas leaving the membrane will then be compressed into liquefied

natural gas where it can be used as fuel for a generator and sold as a product. The decision to implement such a reactor into the process lowers offsite costs such as transportation and environmental services, as well as providing a resource for the unit's overall utility requirements in the form of natural gas.

The decision to introduce an air stripper into the process brought up the issue of how to deal with vapor phase contaminants. In our design, the vapor phase contaminants represent nearly all of the pollutants extracted from the site since the air stripper removed the contamination from the water and into a "dirty gas" stream. A thermal oxidizer would not be a good process choice to deal with the vapor phase contaminants since they are now in high concentration and would require a high energy input to combust as well as raise environmental concerns. For this reason, a filtration unit using granular activated carbon will be used to capture the vapor phase contaminants. Two units will operate asynchronously to ensure that one unit will be operational while the other is being repacked. The loaded carbon will be shipped offsite to be disposed of, and new unloaded carbon will replenish the filtration unit.

3. Equipment Description, Rationale, and Optimization

Dual Phase Extractor (P-101)

The dual phase extraction system (P-101) pumps a vapor/liquid mixture from the ground in a single pipeline (see stream Tables in section 2.4). This high powered vacuum pump generates hydraulic pressure that causes vapor phase pressure gradients to form towards the well. The hydraulic pressure created by the pump is also imposed on any subsurface liquid, causing groundwater and free petroleum product to be pumped from the water Table as well. Not only will P-101 remove a vapor/liquid mixture, this process will introduce a greater flow of oxygen into the soil, stimulating further biodegradation of contaminants in the soil. The use of DPE

systems have proven to be effective, and tend to have shorter treatment times (6 months – 2 years) in comparison to other remediation techniques (bio remediation etc.) (U.S. Environmental Protection Agency, 2008).

Dual pump DPE systems are also available and have been proven to be more effective than single pump systems. In fact, dual pump systems work over a larger variety of ranges, including fluctuating water Tables and permeability of the soil (U.S. Environmental Protection Agency, 2008). However, the economic impact of each piece of equipment is one of The Green Team’s largest concerns. With this in mind, the single pump system was chosen as it is less expensive, and we are not anticipating fluctuations in the water Table (U.S. Environmental Protection Agency, 2008).

Knockout Vessel (V-101)

In order to remediate the vapor/liquid solution entering the system from P-101, they must be separated. Using a knockout vessel is the most common vapor/liquid separation technique, and allows for a continuous process (Iranian Ministry of Petroleum, 1997). This is done using with a de-entrainment mesh pad near the top that will allow the vapor to pass through, while the more dense liquid will remain at the bottom of the unit. The outlet at the top of the vessel will allow the vapor to escape to B-102, while the outlet at the bottom will send the liquid to V-102.

Decanter (V-102)

As seen in the stream Tables in section 2.4, the liquid mixture separated from V-101 is composed of free petroleum product and contaminated water. In order to separate these two liquids a continuous gravity decanter (V-102) is employed. Decanters separate immiscible liquids into two fluids based on their corresponding densities, with the lighter component atop the heavier one. These fluids discharge through overflow lines at the opposing end of the

separator (McCabe, 2005). The free petroleum product is composed of a gasoline/kerosene mixture (see section 2.4), and these two components have extremely similar densities (see Table 9.1.1). Therefore, this mixture will be isolated from the water, and each of these streams will be separately treated (see PFD in section 2.2).

Air Stripper (S-101)

Air stripping is commonly used as a remediation technique to remove Volatile Organic Components (VOC) from contaminated water (Mead & Leibbert, 1998). The function of a stripping column is to extend the surface area of the water, while passing a stripping medium through the water. This induces mass transfer from the water to the air, removing the contamination from the water (Mead & Leibbert, 1998). The two general types of stripping columns are packed columns and low-profile sieve tray columns. Packed columns generally offer greater efficiency, however these columns are designed to operate in the range of 50 gpm, whereas our water flow rate is 3.9 gpm (see section 9.1) (Jaeger Products, Inc.). In addition, stripping columns foul over time due to all of the contaminants in the water, however sieve trays are easier to maintain and clean when fouling occurs. In this remediation process, we chose to employ an air stripping column in order to remove MTBE and BTEX from the groundwater that is pumped from the water Table. This column will utilize low-profile sieve trays as they have demonstrated good efficiency for removing such contaminants (Mead & Leibbert, 1998).

Other techniques for removing VOCs from groundwater include utilizing bio-reactors in a pump-and-treat system. This is performed by pumping the contaminated water out of the water Table, sending it to a bio-reactor to treat the water, and re-injecting it back to the water Table. Pump-and-treat systems have been proven inefficient due to slow desorption, diffusion and dissolution of contaminants (Langwaldt & Puhakka, 2000). In addition, bio-reactors generally

require pretreatment of the water (i.e. heating the water) in order to improve efficiency, but this requires further utilities and raw materials. Therefore, total clean up of a contaminated aquifer requires an extended amount of time resulting in high overall operation costs (Langwaldt & Puhakka, 2000). Considering these facts, air stripping was the better overall economical choice.

Catalytic Reactor (R-101):

In attempt to reuse or recycle the petroleum product being pumped up with the contaminated groundwater several options were considered. For instance, the petroleum product could be passed through filters and other clean up operations and sold as is. The problem with this process, however, would be that the product is a mixture of gasoline and kerosene. Also, since the petroleum has been in the ground for an extended period of time, some of the hydrocarbons have either evaporated or have biodegraded, changing its composition so that it will no longer meet the specifications needed to be sold.

Additionally, the petroleum product could be reprocessed through the use of distillation towers and reactors to reproduce kerosene and gasoline. Unfortunately, since the system being designed is intended to be easily transportable, the size of this operation would be too large. The little volumetric quantity of petroleum product being processed also makes this operation unfeasible.

The last option being considered was breaking the gasoline and kerosene mixture down into a product that could still be used. After extensive research it was found that through the use of a nickel oxide catalyst supported on a silica gel (NiO/SiO_2), light crude oil could be cracked into 99+% methane over a 24 hour period (Mango, 1997). Since gasoline and kerosene are lighter than crude oil, it is assumed that the petroleum product being recovered will also be capable of being cracked into 99+% methane over a 24 hour period. The bulk price of NiO/SiO_2

could not be found. However, it was found that SiO₂ and AlO₂ could be used interchangeably when solely being used to expand the surface area of a catalyst (Speight, 1997). With the bulk price of NiO/SiO₂ known, it was the catalyst chosen for the process. Also, unlike the reaction explained in the laboratory experiment, the reaction in the process will be continuous. With this alteration, it cannot be guaranteed that the product leaving the reactor is completely methane. It is assumed that the hydrocarbons are completely cracked into methane since the reaction kinetics are unknown. However, the reaction does require hydrogen gas so the exiting stream from the reactor will contain a mixture of methane and hydrogen.

GAC Filtration Unit (F-102 A/B)

After using an air stripper to remediate the contaminated groundwater, the remaining contaminants of concern were present in the gaseous phase. In petroleum product spills such as this one, vapor phase contaminants are typically dealt with by either the use of a thermal oxidizer or GAC adsorption (Council, 2005). In choosing between the two, considerations were made for the two main contaminants of concern: MTBE and BTEX vapors. Using the BTEX concentrations as the criteria for a decision, vapor phase GAC is the best choice. At concentrations less than 200 ppm, vapor phase GAC generally incurs the lowest cost (Klemme, Romstad, & Moyer, 2007). Since the BTEX vapor phase concentration leaving S-101 is 1.84 ppm, GAC vapor phase adsorption is the most economical choice. The disadvantage of using vapor phase GAC is that the GAC has a low affinity for MTBE (Council, 2005). However, treating MTBE vapors by thermal oxidation leaves the potential for formation of other undesirable by-products (Council, 2005). Rather than creating new contamination, vapor phase GAC was used. Additionally, no EPA standard exists for the regulation of MTBE. Because of this, the GAC unit is focused on the effective removal of BTEX vapors (EPA, 2009). With GAC

as our vapor phase treatment we will subsequently release some MTBE vapors to the atmosphere; however this is of low concern as the process still meets all environmental standards.

Hydrogen Membrane (F-101)

Because the hydrogen needs to be removed from the methane product to meet natural gas specifications, the only logical operation was membrane separation. From research it was found that the best membrane to use was a hollow fiber permeator and the best feed to permeate pressure ratio was 3 (McCabe, 2005). The use of a membrane also allows the hydrogen to be recycled back to the reactor, lowering the cost of raw materials.

Compressor (C-101)

The hydrogen and methane gas mixture leaving the reactor must be compressed from 1 atm to 3 atm prior to the membrane to create a pressure differential for the membrane to operate on. For compression ratios larger than 2 a compressor must be used (Seider, 2003). For C-101 a centrifugal compressor was chosen due to its small foundations and low maintenance requirement (Seider, 2003). The compressor is made of stainless steel for compatibility with the hydrocarbons (Chemical Compatibility).

Storage Tank (V-103)

Because the catalytic reaction is a sensitive process, the flow rate of the gasoline and kerosene mixture into the reactor needs to be constant. However, the quantity of the petroleum product being removed from the ground can fluctuate. To ensure that the amount being introduced to the reactor maintains constant, the petroleum will pass through a storage tank prior to entering the reactor. With the use of this storage tank, a pump can draw liquid out at a constant rate. Additionally, if the flow rate of the petroleum product being removed from the

ground drops below the 18 gallons/day that the reactor will process, methane can still be produced assuming there is additional petroleum hold-up in the storage tank.

Pumps (P-102, P-103, P-104)

To transport liquid throughout the system pumps were required. For P-102, the liquid leaving passes through the air stripper (S-101) which experiences a 0.34 atm pressure drop, and then passes through the cooling coils in the reactor which also experiences a 0.34 atm pressure drop. Therefore, P-102 will increase the pressure of the water to 1.68 atm to ensure the liquid can pass through the system.

P-103 and P-104 are solely used to transport the liquid to the next piece of equipment. Therefore, each pump increases the pressure of the fluid to 1.1 atm to ensure there is a large enough pressure differential so that the liquid moves downstream.

All pumps are less than 1 Hp so centrifugal pumps were used since costing was most readily available. Again, stainless steel was used in order to handle the hydrocarbons (reference)

Blowers (B-101, B-102)

Similarly to the pumps, to transport gases throughout the system compression operations were needed. Since the pressure increases were used only for gas transport, the compression ratios were small enough that only blowers were needed. For B-101 the air leaving passes through S-101 where it experiences a 0.13 atm pressure drop. Therefore the blower increases the pressure to 1.13 atm. Since the blower size was so small a rotary straight-lobe blower was used to accommodate for this low size factor (Seider, 2003).

For B-102 the air leaving passes through the granular activated carbon (F-102) where it experiences a 0.06 atm pressure drop (Carbtrol). This differential appeared to be rather low. Therefore, the blower will compress the air to 1.1 atm to ensure the air passes through the tower.

For B-102 a centrifugal blower was chosen because they are the most commonly used in chemical processing plants when supplying air to packed beds (Seider, 2003).

Power Generator (G-101)

The methane being produced could be valuable in multiple ways. First, it could be sold as a product to increase profits and thus reduce the payback period for the system. If it were to be sold it could either be connected to a direct line on site or it could be compressed into liquefied natural gas and stored onsite until pickup. Second, the energy from the methane could be used to power the pieces of equipment in the system reducing utility costs. To convert the energy from the methane into electric energy needed to power the equipment a power generator is required. Because the system could be used in remote locations making it hard to connect to power lines, the power generator route was chosen. The little natural gas remaining after powering the entire process does not justify purchasing a liquefied natural gas compressor however if a connection to a direct line is possible the natural gas can be sold.

4. Safety Statement

In order to ensure the process is as safe as possible, it is important to consider even the most extreme cases that could compromise the system. There are many risks associated with the processing of hydrocarbons to produce methane. The presence of hydrocarbons in the system is the most important consideration due to the fact that these materials are flammable (Hess Corporation, 2007). If the equipment or piping fails in any part of the process, either a small-scale fire or an explosion is of great concern. In the event that either a small scale or large scale fire occurs, it is important for each company to have a fire extinguisher on site and it is recommended that each company considers a deluge system for sites processing/extracting large amounts of fuel products. It is also important that all electrical circuits/controls are explosion

proof according to National Fire Protection Association (NFPA), International Building Code (IBC), and the 2006 fire code regulations (Appendix 9.5).

A process hazard analysis (PHA) is performed on each piece of equipment in use that posed a potential threat in the event of a malfunction (see PHAs below). Although the proposed process is running at a maximum pressure of 3 atm (see section 2.3 for the operational pressures of each piece of equipment), pressure is still a great concern for the system. Since each piece of equipment is designed for the pressure needed to accommodate that particular operation, any overpressure could lead to a fire or explosion. In the proposed design, the piece of equipment that has the greatest potential to experience over pressurization is the catalytic reactor. The hydrocarbon cracking process takes place at roughly 200°C which poses a particular threat in the event of a pressure build up due to the explosive nature of gasoline. Another particular point of concern deals with the compromise of any piece of equipment's construction (i.e. a crack in the piping or any of the vessels). In the event of cracking or rupture of the vessel, V-103, the possibility of sparking, fire, or explosion increases as air enters the system. The "What if" analysis performed for the PHA details each way the system could be compromised resulting in fire or explosion. R-101 is the most hazardous piece of equipment, and the hazards associated with this unit include over pressurization, fire, and explosion (as displayed in the PHAs below) and also burns due to the high temperature the reactor operates at.

Focusing specifically on the reactor and its operating conditions, temperature is another point of interest. The reactor, R-101, runs at roughly 200°C which creates a potential hazard since this is a high temperature that can cause harm to the workers upon direct contact. When working with piping or controls on or near the reactor, one should wear protective gloves to

prevent burns. The remainder of the system runs at or near ambient temperature and therefore is not a problematic part of the process.

Another concern associated with the system is the presence of benzene in the fuel products. Benzene is a known carcinogen (Hess Corporation, 2007). In the event that there is a spike in the fuel being extracted from the water Table, the hazards associated with any leaks in the system increase greatly. An increase in the fuel products extracted leads to an increase in hazardous materials being stored in the vessel (V-103) before being processed. As a result, the odds of any given individual being exposed to gasoline (a known irritant), which can lead to benzene exposure, increase as well (Hess Corporation, 2007). Other chemicals that may be present in the fuel product include kerosene, diesel, acetone, and MTBE. All of these chemicals are known irritants as well and contact should be avoided, particularly for prolonged periods (which can range from weeks to months depending on each individual's immune system) (Hess Corporation, 2007).

These basic hazards give reason to require all employees wear goggles and lab coats while working on or near the system. Access to eye washes, showers, and sinks with clean water are critical to ensure proper care in the event of contact with the aforementioned chemicals. Because the system is transportable, each company intending to use this equipment will need to ensure their site has clean water and the proper stations available for the care of their employees in the case of a spill or leak. If the companies renting the equipment indicate that it would be beneficial to add these stations to the system with an option to hook up a freshwater line, further research would have to be done to optimize placement of these safety stations. At this time, however, the Green Team does not recommend adding the safety stations to the system. This is

due to the fact that further exposure could increase the adverse effects associated with gasoline/benzene contact.

In determining how hazardous a given process will be, it is important to also consider how each piece of equipment will react if one of the other pieces malfunctions. For example, if the Vapor-Liquid Separator, V-101, does not separate excess liquid out of the vapor stream before it enters the GAC unit, F-102, there is a possibility that liquid could ultimately foul this unit. Although V-101 could ideally handle a larger quantity of liquid than the design requires, if liquid does pass through the vessel in the gas stream, the GAC unit will require new activated carbon.

This big picture “what if...?” analysis indicates that it is critical that each piece of equipment function as designed. If one piece of equipment is not working properly, the entire vapor process is compromised until the problem is fixed. The part of the process that is not as readily affected is the reactor process because it is only run once a day. The vessel storing the fuel products, V-103, is larger than needed in order to accommodate fluctuating collection/extraction rates and therefore will not overflow if the reactor needs to be shut down for maintenance.

Company: Zelen Environmental	Plant: University of Arizona- Spring 2009 Chemical Engineering Department	Site: To be determined	Unit: R-101	System: Groundwater/fuel product extraction to purify H2O and produce CNG
Method: What-if	Type: Reactor	Design Intent: hydrocarbon cracking to produce methane		
Number: PFD Drawing 1				
Team Members: Kyle Heckel, Mandy Kiesel, Chris Lewis, Adam Manasse				

No.: 1	Description: Inlet: hydrocarbons liquid/gas mixture and gaseous hydrogen, Outlet: methane gas and hydrogen gas				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	What if overpressure in reactor?	Outlet pipeline blockage Compromise in reactor construction (i.e. corrosion of vessel/ cracks in pipelines)	Loss of regulation in amount of hydrocarbons entering the reactor Fire or explosion Leakage of air into reactor causing fire or explosion	Pressure release valve or rupture disk Fail shut for all inlet and outlet valves to prevent fire/explosion deflagration Note: Operation of pressure is low (roughly 1 atm) and would only become a hazard in the event of an extreme over pressure situation	Rec 1. Use fire fighting foam for large scale fires and fire extinguisher suitable for Class B fires for small scale fires (Hess Corporation, 2007)
1.2	What if external fire/explosion?	External fire Lightning strike	See above	See above	See action items for 1.1
1.3	What if external force compromises the construction of the reactor?	Foreign object strikes reactor Possible ignition source: spark from electrical control system	Fire or explosion Leakage of air into reactor causing fire or explosion	Concrete posts around reactor to act as barrier between foreign objects and reactor vessel Electrical circuits/controls explosion proof	See action items for 1.1

Company: Zelen Environmental	Plant: University of Arizona- Spring 2009 Chemical Engineering Department	Site: To be determined	Unit: V-103	System: Groundwater/fuel product extraction to purify H2O and produce CNG
Method: What-if	Type: Vessel	Design Intent: storage of hydrocarbons		
Number: PFD Drawing 1				
Team Members: Kyle Heckel, Mandy Kiesl, Chris Lewis, Adam Manasse				

No.: 2	Description: Hydrocarbon Tank				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
2.1	What if tank leaks or ruptures?	Supplier error Struck by vehicle or fork truck	Hydrocarbons leak back into water Tables (possibly extending region needing DPE) Air flow into tank causing ignition of hydrocarbons	Impermeable catch underneath vessel to add barrier between ground and leakage/spillage Pressure relief valves- gaseous materials, vent storage tanks (liquid) to prevent pressure build-up.	Isolate tanks from surrounding buildings and equipment

Company: Zelen Environmental	Plant: University of Arizona- Spring 2009 Chemical Engineering Department	Site: To be determined	Unit: N/A	System: Groundwater/fuel product extraction to purify H2O and produce CNG
Method: What-if	Type: Line/Piping	Design Intent: transport fuel product out of ground to end of process		
Number: PFD Drawing 1				
Team Members: Kyle Heckel, Mandy Kiesel, Chris Lewis, Adam Manasse				

No.: 3	Description: Lines transporting hydrocarbon liquid and vapor throughout process				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
3.1	What if crack or rupture in pipeline?	Corrosion Pressure increase in pipe Lightning strike or spark from electrical circuit	Air into pipeline causing sparking, fire, or explosion	See items 1.1-1.3	See action items for item 1.1
3.2	Flow rate is excessive	N/A	High pressure downstream	Specify check valves, pressure relief, flow regulators	Safe guards act as action items
3.3	Flow rate too low	N/A	Operability	N/A	Rec 1. Shut down system until flow is back to design specifications

Company: Zelen Environmental	Plant: University of Arizona- Spring 2009 Chemical Engineering Department	Site: To be determined	Unit: V-101	System: Groundwater/fuel product extraction to purify H2O and produce CNG
Method: What-if	Type: Knockout Vessel	Design Intent: separate gaseous fuel products from any remaining liquid		
Number: PFD Drawing 1				
Team Members: Kyle Heckel, Mandy Kiesl, Chris Lewis, Adam Manasse				

No.: 4	Description: Knockout Vessel (KOV) to remove water from fuel product vapors				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
4.1	What if Vapor-Liquid Separator does not remove all liquid from gas stream?	Product malfunction/supplier error	Liquid travels into GAC unit leading to deactivation of the carbon	Fail shut on outlet stream to prevent liquid from reaching the GAC	Rec 1. Consider having a sensor to determine whether materials leaving KOV are hydrated

Company: Zelen Environmental	Plant: University of Arizona- Spring 2009 Chemical Engineering Department	Site: To be determined	Unit: F-102	System: Groundwater/fuel product extraction to purify H2O and produce CNG
Method: What-if	Type: GAC Unit	Design Intent: disposal of vapor phase contaminants		
Number: PFD Drawing 1				
Team Members: Kyle Heckel, Mandy Kiesel, Chris Lewis, Adam Manasse				

No.: 5	Description: Granular Activated Carbon (GAC) Unit for removal of soil vapor contaminants				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
5.1	What if liquid leaks into vessel through cracks?	Supplier error Weathering compromises construction	Deactivation of GAC – unit no longer operational	Welding checks to ensure vessel is air tight	N/A
5.2	What if air leaks into vessel through cracks?	Construction of vessel compromised by weather/lightning strikes	Possible fire or explosion of hydrocarbon vapors	See above	N/A
5.3	What if over pressure in vessel?	Blockage in outlet piping Inlet feed exceeds design specifications	Possible fire or explosion	See page 1	See page 1

Company: Zelen Environmental	Plant: University of Arizona- Spring 2009 Chemical Engineering Department	Site: To be determined	Unit: C-101/B-101,102	System: Groundwater/fuel product extraction to purify H2O and produce CNG
Method: What-if	Type: Compressor/Blower	Design Intent: See process description Section XX		
Number: PFD Drawing 1				
Team Members: Kyle Heckel, Mandy Kiesl, Chris Lewis, Adam Manasse				

No.: 6	Description: Any compressor/blower in the system that is operating at a pressure above 5 psi				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
6.1	What if rupture in compressor/blower causes explosion?	Supplier error Weathering/wear down of equipment	Over pressure burst/explosion that could potentially harm anyone working at or around the piece of equipment	Pressure release for worst case scenario. Pressure release valve or rupture disk Note: Maximum pressure at 3 atm for C-101 (above this pressure, unit shuts down)	Flame arrester after C-101 to prevent membrane fouling in the event of a fire/explosion

Company: Zelen Environmental		Plant: University of Arizona- Spring 2009 Chemical Engineering Department	Site: To be determined	Unit: N/A	System: Groundwater/fuel product extraction to purify H2O and produce CNG
Method: What-if	Type:		Design Intent: Provide electric power, control system, plant air/steam/refrigeration. vacuum, fuel oil, natural gas, HVAC service, fire and cooling water, etc. to the facility		
Number:					
Team Members:					

No.:	Description: Utilities and plant services				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
	What if electric power were lost momentarily or longer?	Storm related incident	Shut down of equipment Possible overpressure in reactor due to continued reaction	Existing safeguards adequate	N/A
	What if the control system (DCS, PLC, etc.) were lost?	See above	Inability to regulate flow rates throughout the system	Discontinued flow to entire system Existing safeguards adequate	N/A
	What if cooling water was lost?	Feed from air stripper lost	Possible overheating of reactor	Control system set to turn off reactor in the event of lost cooling stream	N/A
	What if fire suppression (water, carbon dioxide, Halon, etc.) were lost?	N/A	Possible inability to suppress large scale fires	Keep fire fighting foam on hand in case water, etc. unavailable in the event of a large scale fire	N/A
	What if nighttime lighting were lost?	N/A	Possible personnel safety hazard: slip, trip, fall, other injury	N/A	Specify emergency power generator
	Plant people and contractors are not adequately trained?	N/A	Contractor or employee injury	N/A	Engineering to help draft contractor safety procedures and process operating procedures, including emergency shut-down and temporary operating procedures.

5. Environmental Impact

The remediation system outlined in this report is designed to have an overall positive environmental impact. The environmental advantages are clear for the system; however, this section analyzes the potential negative impacts, such as utility and raw material requirements. A comparison is made between the positive and negative impacts of the remediation system on the environment. Additional analysis is presented concerning regulatory standards for groundwater and the impacts of contaminated groundwater as it relates to human and ecological toxicity.

Regulatory Standards

To ensure quality in drinking water, the EPA has established permissible chemical levels allowed in public water systems. This threshold is referred to as the maximum contaminant level (MCL), and is derived on the basis that the MCL results in no more than 1 to 100 excess cancer cases (allowable excess cancer cases depends on the chemical) (EPA, 2009). Table 5.1 presents groundwater MCL information on some components present in various fuel types, as well as detected groundwater contaminant levels at the Bennett Bulk Fuel site. (Appendix 9.5) Vapor phase contamination is not a concern here since infiltration rates for vapor contamination are much lower and since regular human exposure to such vapors is not expected.

	Contaminant of Concern				
	Benzene	Toluene	Ethyl Benzene	Xylene	MTBE*
MCL	5 µg/L	1 mg/L	7 mg/L	10 mg/L	35 µg/L
Bennett	1900 µg/L	1500 µg/L	1000 µg/L	6600 µg/L	5000 µg/L

Table 5.1-EPA established MCL (U.S. Environmental Protection Agency, 2009) and concentration information for main groundwater contaminants at Bennett site (* no EPA established MCL) (Martinson, 2002)

The above Table shows MCL data that is enforced on a national level. However, it should be noted that currently EPA does not have an established MCL for MTBE. The state of Arizona

also does not have an enforceable regulation concerning MTBE, but has established a health based guidance level of 35 µg/L for MTBE in groundwater.

The main concern at the Bennett site is MTBE contamination, as it is present in the highest concentration above established drinking water levels. For this reason, our groundwater remediation actions place a high emphasis on the removal of MTBE. Although the guidance level is 35 ppb in Arizona, 20 ppb has more commonly been used as a guideline on a site specific basis and the air stripper used for MTBE removal in this design used a 20 ppb limit as its design parameter.

Human and Ecological Toxicity

Human toxicity information for various fuel product components is obtained through the use of material and safety data sheets (MSDS). Additional toxicity information regarding the ecological impacts of said components is obtained from the US EPA. In the analysis, each component is evaluated for: human health effects, solubility, soil effects, ozone layer depletion, and global warming potential. Results are shown below in Table 5.2.

	Human Health Effects	Solubility	Soil Effects	Ozone Layer Depletion	Global Warming Potential
Hydrocarbons (BTEX, n-butane, n-hexane, etc.)	possible carcinogen; damage to liver, kidneys and central nervous system	100-500 mg/L average; increases with ethanol cosolvent	biodegrades into less toxic products	no	no
benzene	carcinogenic; damage to blood system	~1800 mg/L	biodegrades into less toxic products	no	no
ethanol	mutagenic, teratogenic; damage to heart, kidneys, liver, central nervous system	miscible	very minimal, readily biodegrades	no	no
acetone	eye and skin irritant; toxic to central nervous system, kidneys, liver, and reproductive system	easily soluble	biodegrades into non-toxic products	no	no
lead	carcinogenic; damage to blood, kidneys, and central nervous system	insoluble	biodegrades into less toxic products	no	no
naphthalene	possible carcinogen	very slightly soluble	biodegrades into more toxic products	no	no
MTBE	eye, skin, and respiratory tract irritation; central nervous system depression and neurotoxicity	very soluble	does not readily adsorb on soil	no	no

Table 5.2 - Human and Ecological Toxicity information. Soil effects found in (Jacobs, Guertin, & Herron, 2000). Ozone layer depletion and global warming potential found in (U.S. Environmental Protection Agency, 2008). It is evident from the data that our main environmental concerns are BTEX and MTBE.

These contaminants have the most severe health implications and are the most likely to be present in groundwater due to their solubility. Benzene and other aromatic rings are a high cancer risk, while both benzene and MTBE cause considerable damage to vital organs. Although some of the other compounds such as ethanol and lead also have severe health implications, they are present in much smaller concentrations and are more susceptible to biodegradation.

Utility and Raw Materials Impact

Utility and raw material requirements account for the entire burden imposed on the environment by our remediation system. Since the system is designed to remove environmental pollution, the only negative impact of the system on the environment is present in energy and resources consumed. The sole utility requirement in this case is electricity. The generation of electricity by a power plant releases greenhouse gases into the atmosphere. For the operation of the remediation system, a generator will be fueled with the methane produced in the reactor. In turn, this generator will produce enough electricity to power all of the equipment in the system (see Section 2.5: Utility Tables). Using the generator effectively eliminates any electricity requirements from the grid and thus makes greenhouse gas emissions from utility requirements negligible.

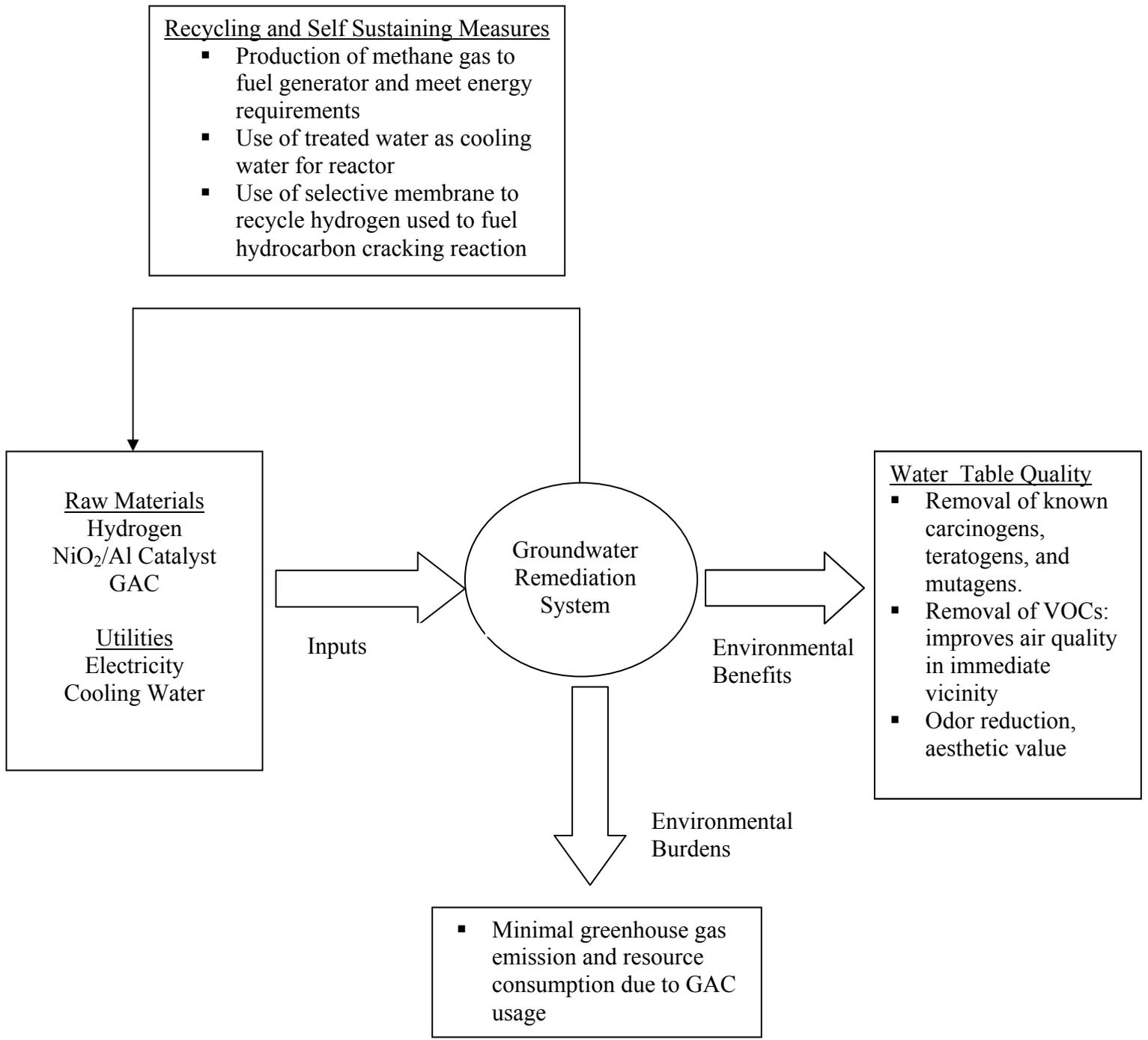
The only raw material requirement for the remediation system is granular activated carbon (GAC), which will be used to capture vapor phase contaminants. This requirement does impose some environmental burden as resources will be consumed to produce and recharge the GAC. Table 5.3 shows the environmental impact of GAC.

Production (.36 kg)		Reactivation (3.6 kg)	
Hard Coal (kg)	0.36	Transport by 32 ton lorry (kg*km)	9639
Hard Coal Combustion (MJ)	22	Hard Coal Combustion (MJ)	11
Electricity (kWh)	0.58	Electricity (kWh)	0.12
Natural Gas Combustion (MJ)	4.8	Steam (kg)	2.2
Deionized Water (kg)	4.4	Natural Gas Combustion (MJ)	38

Table 5.3 - Environmental Impact of GAC production and reactivation (Muñoz, Peral, & Ayllón, 2007)

Overall Impact

The overall environmental impact of our system is very low. The largest negative impact is present in the form of electricity and resource consumption due to the production and reactivation the GAC, but it should be noted that even this impact is very minimal (see Table 5.3 above). The environmental benefit of the system is greater than the environmental burden imposed of the system. From the above data, it is quite clear that the remediation system has numerous environmental benefits. The system, when fully functional, is able to remove a number of contaminants from groundwater where they would eventually come into contact with human, animal, and plant life. The benefit of contamination removal is nearly invaluable, as it greatly raises the quality of life and avoids unnecessary deaths due to health issues such as cancer. The flow chart below represents the positive and negative impacts of the system and show measures which were taken to reduce the negative environmental impact.



6. Economic Analysis

The remediation design proposed by The Green Team is designed to produce methane gas from free petroleum product, as well as remediate the contaminated groundwater. The groundwater that is pumped from the water Table makes up 93% of the liquid removed (5589 gallons of water/day of 6007 total gallons of liquid/day), and this water is then re-injected back into the water Table after remediation. Because of this, there is no economical benefit captured from remediating the water, yet there are significant costs associated with the overall remediation process. For this reason, the design process will have zero return, and The Green Team must cut costs in every aspect of the design.

The equipment cost is critical to assessing the economic viability of the design. Table 6.1 includes a full account of the cost of each piece of equipment as calculated in section 9.2.

Equipment	Bare Module Cost (C_{BM})
B-101	\$1.70
B-102	\$17,600
C-101	\$20,500
F-101	\$2,988
F-102	\$1,544
G-101	\$32,300
P-101	\$165,000
P-102	\$496
P-103	\$496
P-104	\$496
R-101	\$34,200
S-101	\$273,000
V-101	\$77,200
V-102	\$122,300
V-103	\$1,100
Total (C_{TBM})	\$749,000

Table 6.1 - Bare module costs for each piece of equipment

In order to perform the economic analysis on the proposed process, it was first necessary to determine the economic costs associated with the design. Table 6.2 outlines the method for calculating the total depreciable capital, C_{TDC} , which is needed to complete the cost factor outline (Table 6.3).

Cost Variable	Calculation Method	Value [\$]
Bare Module Cost (C_{BM}) *for all equipment	Sum of the C_{BM} for each piece of equipment	749,000
Total cost of spares (C_{Tspare})	Sum of spare costs for compressors, blowers, & pumps	38,000
Cost of the catalyst ($C_{catalyst}$)	See Appendix 9.2	30,000
Total Bare-Module Investment (C_{TBM})	$C_{TBM} = C_{BM} + C_{Tspare} + C_{catalyst}$	817,000
Cost for site preparation (C_{site})	4-6 % of C_{TBM}	N/A
Cost for service facilities (C_{serv})		N/A
Allocated Costs (C_{alloc})		N/A
Direct Permanent Investment (C_{DPI})	$C_{DPI} = C_{TBM} + C_{site} + C_{serv} + C_{alloc}$	817,000
Cost of contingencies (C_{cont})	18% of C_{DPI}	147,000
Total Depreciable Capital (C_{TDC})	$C_{TDC} = C_{DPI} + C_{cont}$	964,000

Table 6.2- Costs associated with determining the total depreciable capital.

The labor-related operations were based upon an assumption that only one operator per shift was needed because the proposed process is a continuous operation. Specifically, the calculation for direct wages and benefits (DW&B) is also dependent upon the basis of 5 shifts, where each worker is compensated for 2,080 hr/yr (Seider, 2003). In Table 6.3, the cost of manufacture and total sales are shown. The sale of natural gas is based upon the gas that is not consumed by the generator (G-101), which powers the entire remediation system. The molar amount of methane that will be sold, 10.73 mol/hr, (see stream Table in section 2.4) was converted to standard cubic feet (SCF) on a yearly basis for the calculation in Table 6.3. It is important to note that the only raw material required is the hydrogen, and is priced in Table 6.3. Performing a sensitivity analysis demonstrates that the price of the hydrogen does not greatly affect the overall cost of the system. For instance, if the price of hydrogen were increased to \$10.00/kg, the expected expense

of hydrogen yearly is still less than \$20,000. Further economic hazards will be discussed later in this section.

Cost Factor	Typical Factor	Value [\$/yr]
Feedstocks (raw materials)		
Hydrogen	\$2.00/kg (Doty Ph.D, 2009)	3,590
Granular Activated Carbon	\$1.35/lb	46,800
Operations (labor-related) (O)		
DW&B	\$30/operator-hr	312,000
Direct salaries & benefits	15% of DW&B	46,800
Operating supplies & services	6% of DW&B	18,720
Maintenance (M)		
MW&B		
Fluid handling process	3.5% of CTDC	33,700
Salaries and benefits	25% of MW&B	8,430
Maintenance Overhead	5% of MW&B	1,690
Property taxes and insurance	2% of CTDC	19,270
COST OF MANUFACTURE (COM)	Sum of above	491,000
General Expenses		
Selling (or transfer) expense	3% (1%) of sales	3,620
Direct research	4.8% of sales	5,790
Allocated research	0.5% of sales	600
Administrative expense	2.0% of sales	2,420
Management incentive compensation	1.25% of sales	1,510
TOTAL GENERAL EXPENSES (GE)	Sum of general expenses	14,000
TOTAL PRODUCTION COST - C	COM + GE	505,000
Sales		
Natural Gas	\$8.07/1,000 SCF (Natural Gas Prices, 2009)	633
Portable Unit Services	\$10,000/month (Appendix 9.5)	120000
TOTAL SALES, S		120633

Table 6.3 - Cost Factor Outline for the cash flow analysis (Seider, 2003)

Figure 6.1: Method for Calculating NPV & IRR (t = income tax rate, i = interest rate, n = year) (Seider, 2003)

Figure 6.1 outlines the equations and variables from the process design book that are necessary for calculating the net present value (NPV) and investor's rate of return (IRR). In order to calculate C_{WC} , the following equations were utilized (Seider, 2003).

$$C_{WC} = \text{cash reserves} + \text{inventory} + \text{accounts receivable} - \text{accounts payable}$$

$$\text{Cash reserves} = 8.33\% \text{ of COM (see Table 9-3 for COM)}$$

$$\text{Inventory} = 1.92\% \text{ of natural gas sales (see Table 9-3 for natural gas sales)}$$

$$\text{Accounts receivable} = 8.33\% \text{ of S (see Table 9-3 for S)}$$

$$\text{Accounts payable} = 8.33\% \text{ of Feedstocks (see Table 9-3)}$$

$$C_{WC} = \$15,566$$

From there, the depreciation over a ten year period was calculated using the Sum-of-the-Years-Digits Method (SYD). It was assumed that the salvage value of the equipment at the end of ten

years was equal to 10% of the total depreciable capital. The equations below illustrate the method for calculating depreciation (Seider, 2003).

$$\text{SUM} = \sum_{j=1}^n j = \frac{n(n+1)}{2}$$

n = number of years of depreciation

Because the depreciation is occurring over a ten year period, $n = 10$ and $\text{SUM} = 55$. From there the depreciation can be calculated on a yearly basis (Seider, 2003).

$$D_t = \frac{\text{depreciable years remaining}}{\text{SUM}} (B - S_{\text{equip}})$$

D_t = depreciation

B = original cost of the asset (C_{TDC})

S_{equip} = salvage value of equipment = 10% of C_{TDC}

The values for each year's associated depreciation are in Table 6.4 where the depreciable years remaining starts at a value of 10. The cost flow analysis was completed using the rest of the equations mentioned in Figure 6.1 above. As noted in Table 9.4, the net present value (NPV) at the end of 20 years is a little over -\$1.5 million. Using the solver function in excel to calculate the IRR such that the NPV is equal to zero at the end of the 20 year period, it was determined that the rate of return was less than zero for the investor. With this being the case, The Green Team then back-calculated the amount that must be charged for the remediation services (originally estimated at \$10,000/month – Table 6.3) in order to achieve a 20% IRR.

Year	Investment		D	C _{Excl. Dep.}	S	Net Earn	Cash Flow	Cum PV i = 15%	IRR
	fC _{TDC}	C _{WC}							
0	-321260						-321260	-321260	
1	-321260						-321260	-600617	
2	-321260	-15566					-336826	-855306	
3			157710	333335	120633	-233359	-75650	-905047	
4			141939	349106	120633	-233359	-91421	-957317	
5			126168	364877	120633	-233359	-107192	-1010610	
6			110397	380647	120633	-233359	-122963	-1063770	
7			94626	396418	120633	-233359	-138734	-1115925	
8			78855	412189	120633	-233359	-154504	-1166433	
9			63084	427960	120633	-233359	-170275	-1214836	
10			47313	443731	120633	-233359	-186046	-1260824	
11			31542	459502	120633	-233359	-201817	-1304203	
12			15771	475273	120633	-233359	-217588	-1344872	
13				491044	120633	-233359	-233359	-1382799	
14				491044	120633	-233359	-233359	-1415780	
15				491044	120633	-233359	-233359	-1444458	
16				491044	120633	-233359	-233359	-1469396	
17				491044	120633	-233359	-233359	-1491081	
18				491044	120633	-233359	-233359	-1509938	
19				491044	120633	-233359	-233359	-1526335	
20		15566		491044	120633	-233359	-217793	-1539642	0%

Table 6.4 - Cost Flow Sheet [\$]

This was done by varying the sales associated with the portable unit services such that the CUM PV at the end of 20 years is equal to zero, using an interest rate, $i = 20\%$ instead of the original 15%. Table 6.5 shows the sales required in order to achieve an IRR of 20%.

Sales	Value [\$ /yr]
Natural Gas	633
PorTable Unit Services	730,020
Total Sales	730,653

Table 6.5 - Projected Sales for an IRR of 20%

In order for this design to be profitable, it would require the unit be rented out for roughly \$60,840 per month. Should this be the case, the NPV at the end of 20 years is equal to \$241,118. Table 6.6 displays the cost flow analysis with an interest rate of 15%, IRR of 20% with the sales adjusted such that this rate of return holds true.

Another important aspect of an economic analysis deals with the hazards associated with the process. For this process, a hazard assessment for a jump in the price of hydrogen, as discussed above, does not greatly impact the costs associated with design. Because the system produces enough methane to power the process, a rise in the cost of energy is of no concern. The two largest concerns associated with the economics of this design lie within the cost of both the GAC and the catalyst. Ideally, the catalyst will be regenerated such that the process maintains efficiency throughout the entire life cycle of the plant (Zhang & Amiridis, 1998). If this is the case, an increase in the price of the catalyst above \$15/kg will not significantly increase the expenses of the manufacturing of the plant. However, if the catalyst needs to be replaced on a more frequent basis than anticipated, the costs will become far greater than economically acceptable.

Year	Investment		D	C _{Excl. Dep.}	S	Net Earn	Cash Flow	Cum PV	IRR
	fC _{TDC}	C _{WC}						15%	
0	-321260						-321260	-321260	
1	-321260						-321260	-600617	
2	-321260	-15566					-336826	-855306	
3			157710	333335	730653	150953	308663	-652355	
4			141939	349106	730653	150953	292892	-484893	
5			126168	364877	730653	150953	277121	-347115	
6			110397	380647	730653	150953	261350	-234126	
7			94626	396418	730653	150953	245579	-141804	
8			78855	412189	730653	150953	229808	-66679	
9			63084	427960	730653	150953	214037	-5836	
10			47313	443731	730653	150953	198266	43172	
11			31542	459502	730653	150953	182495	82398	
12			15771	475273	730653	150953	166724	113560	
13				491044	730653	150953	150953	138094	
14				491044	730653	150953	150953	159428	
15				491044	730653	150953	150953	177980	
16				491044	730653	150953	150953	194111	
17				491044	730653	150953	150953	208139	
18				491044	730653	150953	150953	220337	
19				491044	730653	150953	150953	230943	
20		15566		491044	730653	150953	166519	241118	20%

Table 6.6 - Cost Flow Sheet for a 20% IRR [\$]

Another economic hazard associated with the remediation system is the \$47,000 per year investment in GAC. This is a significant yearly expense and an increase in price per lb, even by as little as a single dollar, would increase the yearly expenses greatly. For example, if the purchase price increased by one dollar (a new price of \$2.35/lb), the purchase price of the GAC would nearly double, totaling \$81,400/year. Therefore, the price of the GAC is a major concern for The Green Team's remediation system.

7. Conclusions and Recommendations

When looking at the economic aspects of the design, it is clear that the price of remediating the environment is large. While the equipment costs are relatively small at \$749,000 the calculated cost of operating the plant is very large at 458,200 per year. With only making \$120,633 annually, it becomes impossible to payback the investment of the system or turn a profit in the end. For the investment to have a reasonable investor's rate of return (IRR of 20%), the monthly rate for the unit services would have to increase from \$10,000 a month to \$60,840 (see section 6).

The equations used to price the operations are for a large scale plant and therefore are not completely accurate for our particular design. It is known that a full time operator is not always required to be at current ground water remediation facilities (Appendix 9.5). Since there may be multiple facilities in reasonable distances from each other, a better option may be to split a single operator amongst multiple sites to reducing operating costs. Once a more accurate operating cost is determined, a better economic analysis can be performed.

While the system may not seem economical, in the end, there are still several contaminated ground water sites that need to be remediated. The system designed is considered to be an

improvement over current remediation systems because it treats all contaminants on site. Unlike other current processes, this system was designed to be mobile. In addition, this system is self sufficient, and the raw material costs are minimized as well. Therefore, The Green Team's final recommendation is to build the proposed system.

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9. Appendices

9.1 Process Calculations

Decanter (V-102)

As shown in the PFD in section 2.2, the liquid in stream 3 is pumped from the ground and will be held in a gravity decanter (V-102) prior to being sent to R-101 or S-101. A gravity decanter is used as it enables the continuous separation of immiscible liquids based on the corresponding densities of the materials (McCabe, 2005). The liquid solution in the decanter will separate into a gasoline/kerosene mixture, settling on top of the contaminated water. Table 9.1.1 depicts the flow rate and composition of the contaminated water extracted from the ground (United States Environmental Protection Agency, 1997). The corresponding densities for

gasoline and kerosene are very similar and smaller than that of water, thus verifying our assumed separation in the decanter.

Component	V [gal/day]	Density [kg/m ³]	m [kg/day]	Viscosity [cSt]
Gasoline	9	803	25.1	0.429
Kerosene	9	810	27.8	2.39
Water	5589	1000	21154.365	0.891

Table 9.1.1 - Inlet stream to the decanter. Volumetric flow rates were taken from (United States Environmental Protection Agency, 1997). Densities were taken from (Elert, 2008). Viscosities found in (National Institute of Justice, 2001).

The size of the decanter is dependent upon the time required for separation of the liquid mixture. The separation of the mixture is dependent upon the viscosity of the continuous phase (McCabe, 2005). In order to predict the continuous phase viscosity, the Refutas method for predicting the overall viscosity of a mixture will be used (Al-Besharah, Mumford, Akashah, & Salman, 1989). This method computes a viscosity blending index (VBI) number for each individual component based on an empirical correlation using the viscosity of the substance. The overall viscosity is then determined using a weighted average of the VBI's of the mixture.

$$VBI_x = 14.534 \times \ln(\ln(\mu_x + 0.8)) + 10.975$$

$$VBI_{mix} = VBI_1x_1 + VBI_2x_2 + VBI_3x_3$$

Where,

x_n - mass fraction of component n in the mixture.

The computed overall viscosity is determined to be 0.891 centistokes (cSt). This value is identical to the viscosity of water, and this is expected as the mass fraction of water is 0.99.

Using this viscosity, the required separation time can now be calculated for V-102 (McCabe, 2005).

$$t = \frac{100 \times \mu}{\rho_a - \rho_b}$$

Where,

ρ_n – density of component n

Component b is assumed to be the mixture of gasoline and kerosene, so a weighted average is used to compute the density of this mixture (806.68 kg/m^3). The separation time is calculated to be 27.7 minutes for our mixture. Lastly, we will calculate the required size of the decanter by multiplying the separation time by the flow rate of the mixture.

$$V_{\text{decanter}} = \dot{V} \times t = \frac{6017 \text{ gal}}{\text{day}} \times \frac{1 \text{ day}}{24 \text{ hours}} \times \frac{1 \text{ hour}}{60 \text{ mins}} \times 27.7 \text{ mins}$$

$$V_{\text{decanter}} = 115.54 \text{ gallons}$$

This value represents the total liquid holdup in the decanter. The decanter should be sized to operate at 95% capacity, so the actual size becomes $115.54/0.95$ or 122 gallons (0.5 m^3). The length of the tank should be 5 times the diameter (McCabe, 2005). Based on this assumption, the length of the decanter should be 2.52 m, and the diameter is 0.5 m.

Air Stripper (S-101)

Stream 4 represents the contaminated water that is separated from the gasoline/kerosene mixture in V-101, and it is pumped to an air stripping column (S-101). We will utilize an air stripper to remove the remaining contaminants before re-injecting the water back into the water Table. Table 9.1.2 displays the concentration levels of the contaminants of concern in the water.

Notice Henry's law coefficient for MTBE is one order of magnitude smaller than the rest.

Converting this number to the slope of the equilibrium line (m) in mole-fraction units, we see the slope for MTBE is significantly less than the other contaminants. This will cause the MTBE to require the most air off all the components, and we will assume this calculated value for the minimum air flow required for the stripping column.

$$m = \frac{6.25 \times 10^{-4} \text{ atm} \times \text{m}^3}{\text{mol}} \div 1 \text{ atm} \times \frac{1000 \text{ kg}}{\text{m}^3} \times \frac{\text{mol}}{0.018 \text{ kg}} = 34.72$$

	MTBE	Benzene	Ethylbenzene	Toluene	Xylene
Initial Concentrations [ppm]	5000	1900	1000	1500	6600
Final Concentrations [ppm]	0.02	5	700	1000	1000
Henry's constant [m ³ *atm/mol]	6.25E-04	4.76E-03	6.67E-03	6.25E-03	5.88E-03
Equilibrium line slope (m)	34.72	264.55	370.37	347.22	326.80

Table 9.1.2 - Contaminants to be removed from groundwater. Henry's law coefficients for the corresponding material in water at 20C (Sander, 1999).

The exit gas will be in equilibrium with the incoming solution at the minimum air flow rate (McCabe, 2005). The mole fraction (x_a) of MTBE at the inlet of S-101 can be calculated using the molecular weight of MTBE, which equals 88.15 g/mol.

$$x_a = \frac{5000 \text{ g MTBE}}{10^6 \text{ g H}_2\text{O}} \times \frac{1 \text{ mol MTBE}}{88.15 \text{ g MTBE}} \times \frac{18 \text{ g H}_2\text{O}}{1 \text{ mol H}_2\text{O}} = 1.021 \times 10^{-3}$$

Using the slope of the equilibrium line, the concentration of MTBE in the air at the outlet can be calculated.

$$y_a = 34.72 \times 1.021 \times 10^{-3} = 0.0355$$

The total volume of MTBE removed per cubic meter of solution fed is determined using the initial and final concentrations of MTBE in the water.

$$V_{MTBE} = \frac{(5000 \times 10^{-6} - 20 \times 10^{-9}) \text{ mol}}{\frac{88.15 \text{ g}}{\text{mol}}} \times \frac{10^6 \text{ g}}{\text{mol}} = \frac{2.78 \text{ mol MTBE}}{\text{m}^3 \text{ solution}}$$

The total amount of gas leaving is

$$V = \frac{V_{MTBE}}{y_a} = \frac{2.78}{.0355} = 78.35 \text{ moles}$$

Ideally, 1 g mol = 0.0224 std m³, and we know the change in gas flow rate is minimal.

$$F_{min} = 78.35 \text{ moles} \times \frac{0.0224 \text{ std m}^3}{1 \text{ mol}} = 1.755 \text{ std m}^3$$

We will assume the air flow rate will be three times the minimum calculated value. This will ensure enough air is supplied to remove all of the contaminants.

$$F_{air} = \frac{1.755 \text{ m}^3}{\text{hour}} \times 3 \times \frac{3.28^3 \text{ ft}^3}{1 \text{ m}^3} \times \frac{1 \text{ hr}}{60 \text{ mins}} = 3.1 \text{ scfm}$$

Table 9.1.3 depicts the calculated minimum required flow rates of air for each component, verifying the assumption that MTBE will require the most air.

Component	Minimum Air Flow [SCFM]
MTBE	3.1
Benzene	0.4
Toluene	0.1
Ethylbenzene	0.09
Xylene	0.3

Table 9.1.3 - Required air flow rates to remove contaminants in S-101.

Because we are using an air flow rate of three times the minimum, we will calculate new equilibrium mole fractions.

$$y_a = \frac{0.0355}{3} = 0.012$$

$$x_a^* = \frac{0.012}{34.72} = 3.4 \times 10^{-4}$$

The final concentrations of the contaminants in stream 8 leaving S-101 are displayed in Table 9.1.4.

Component	Concentration [mol/L]	Concentration [ppm]
MTBE	0.0149	1311.76
BTEX	1.88E-05	1.84

Table 9.1.4 - Concentration of contaminants in stream 8

We will need to determine the overall height of the stripping column in order to cost the unit.

This requires determining the number of transfer units (N_{ox}) which is based on the concentrations at the inlet and outlet streams (McCabe, 2005).

$$N_{ox} = \frac{C_a - C_b}{(C - C^*)_L}$$

Where,

$$\overline{(C - C^*)_L} = \frac{\Delta C_a - \Delta C_b}{\ln \frac{\Delta C_a}{\Delta C_b}}$$

And,

$$\Delta C = C_n - C_n^*$$

The equilibrium concentration at the liquid inlet is determined using the calculated equilibrium mole fractions.

$$C_a^* = 3.4 \times 10^{-4} \times \frac{88.15g}{18g} \times 10^6 = 1666.67 \text{ ppm}$$

$$\Delta C_a = 5000 \text{ ppm} - 1666.67 \text{ ppm} = 3333.33 \text{ ppm}$$

At the bottom of the column, the exit concentration (C_b) equals 20 ppb. The concentration at the liquid outlet (C_b^*) equals zero. Therefore, $\Delta C_b = 0.02$ ppm. We can now calculate the log mean concentration difference as follows.

$$\overline{(C - C^*)_L} = \frac{3333.33 - 0.02}{\ln \frac{3333.33}{0.02}} = 277.23 \text{ ppm}$$

The number of transfer units is now calculated using the equation mentioned earlier.

$$N_{ox} = \frac{5000 - 0.02}{277.23} = 18.0$$

To calculate the overall height of the stripping column we will use the following relationship, where Z_T is the overall height, and HTU is the height of each transfer unit (McCabe, 2005).

$$Z_T = N_{ox} \times HTU$$

The liquid mass transfer coefficient (K_{la}) for MTBE in water at 20°C and 1 atmosphere is found to be 0.0035 sec⁻¹ (Ramakrishnan, et al., 2004). We will assume a tower diameter of 0.75 meters, and the following equation applies (Ramakrishnan, et al., 2004).

$$HTU = \frac{Q_{H2O}}{A_x \times K_{la}} = \frac{0.881 \frac{m^3}{hr} \times \frac{1 hr}{3600 sec}}{0.442 m^2 \times 0.0035 sec^{-1}} = 0.158 meters$$

Note,

$$5589 \frac{gal H2O}{day} \times \frac{3.785 L}{1 gal} \times \frac{1 m^3}{1000 L} \times \frac{1 day}{24 hr} = 0.881 \frac{m^3}{hr}$$

$$A_x = \left(\frac{D}{2}\right)^2 \times \pi = 0.442 m^2$$

Finally, the overall height of the stripping column is calculated, $Z_T = 2.86$ meters.

Knockout Vessel (V-101)

As seen in the PFD (section 2.4), the liquid-vapor mixture in stream 1 is immediately pumped to a knock out vessel (V-101) in order to separate the vapors from the liquid. This is done with a de-entrainment mesh pad near the top of the vessel that will allow the vapor to flow through, while restricting any liquid particles suspended in the vapor. In order to calculate the size requirement for V-101, we will use the Souders-Brown equation to calculate the maximum allowable vapor velocity (Austrheim, Gjertsen, & Hoffmann, 2007).

$$V = k \times \left[\frac{\rho_L - \rho_v}{\rho_v} \right]^{\frac{1}{2}}$$

Where,

ρ_n – density of component n

The density of the vapors is calculated to be 5.18 kg/m^3 using the ideal gas law, knowing the operating conditions are 1.05 atm and 20°C . The density of the liquid mixture is determined to be 997.7 kg/m^3 using a weighted average based on the composition of stream 2. The k constant has been experimentally determined to be 0.107 m/s when incorporating a de-entrainment mesh pad in the vessel (Austrheim, Gjertsen, & Hoffmann, 2007). Plugging these values into the Souders-Brown equation returns a value of 1.48 m/sec as the maximum allowable vapor velocity.

We calculate the size of the vessel by first dividing the vapor volumetric flow rate (300 SCFM) by the vapor velocity to determine the cross sectional area (A_x) of the knock out pot. The cross sectional area is 0.1 m^2 . Assuming the knock out pot is cylindrical, we calculate diameter to be 0.35 m, and assuming the length-to-diameter ration is 3 to 4, the length equals 0.263 m (Austrheim, Gjertsen, & Hoffmann, 2007).

Catalytic Reactor Calculations

Mole Balance

The liquid input entering the reactor, stream 18, is the petroleum product from the decanter (V-102). As previously stated, there are 18 gallons per day of petroleum product. It is known that there are both gasoline and kerosene contamination at the site however the composition is not exactly known and may vary. For these calculations it is assumed that the composition is half gasoline and half kerosene. If these compositions were to change, the calculations could easily be adjusted to fit the specifications. To determine how much methane would be produced a mole balance is performed on the reactor. It is known that all the

hydrocarbons completely break down into methane over a 24 hour period (Mango, 1997). Since both gasoline and kerosene are composed of many different hydrocarbons the easiest way to perform the mole balance is to find the average carbon number for each fluid. The average carbon numbers for gasoline and kerosene were found to be 7 and 10 respectively (Shephard, 2008). This means that each mole of gasoline would produce 7 moles of methane and each mole of kerosene would produce 10 moles methane. To convert the volumetric flow rates to mass flow rates the gasoline and kerosene densities were found to be 0.737 g/cm³ and 0.817 g/cm³ respectively (Specific Gravity of Liquids, 2007). With this info the mass flow rates, \dot{m} , were calculated.

$$\dot{m} = \dot{V} * \rho$$

Therefore,

$$\dot{m}_{gasoline} = 9 \frac{gal}{day} * \frac{3.785 L}{gal} * \frac{0.737 g}{cm^3} * \frac{1000 cm^3}{L}$$

$$\dot{m}_{gasoline} = 25106 \frac{g}{day}$$

$$\dot{m}_{kerosene} = 9 \frac{gal}{day} * \frac{3.785 L}{gal} * \frac{0.817 g}{cm^3} * \frac{1000 cm^3}{L}$$

$$\dot{m}_{kerosene} = 27831 \frac{g}{day}$$

The molar flow rate, \dot{n} , can be calculated using the molecular weights for gasoline and kerosene of 97.25 g/mol and 140 g/mol respectively (EPA, 2009).

$$\dot{n} = \frac{\dot{m}}{MW}$$

Therefore,

$$\dot{n}_{gasoline} = 25106 \frac{g}{day} * \frac{mol}{97.25 g}$$

$$\dot{n}_{gasoline} = 258.16 \frac{mol}{day}$$

$$\dot{n}_{kerosene} = 27831 \frac{g}{day} * \frac{mol}{140 g}$$

$$\dot{n}_{kerosene} = 198.79 \frac{mol}{day}$$

The methane production can then be calculated using the average carbon numbers.

$$\dot{n}_{methane} = 258.16 \frac{mol gasoline}{day} * 7 \frac{mol carbon}{mol gasoline} + 198.79 \frac{mol kerosene}{day} * 10 \frac{mol carbon}{mol kerosene}$$

$$\dot{n}_{methane} = 3795.0 \frac{mol}{day}$$

From the molar flow rate of methane and knowing that 0.03 g CH₄ can be produced per gram of catalyst (Mango, 1997), the amount of catalyst needed can be determined.

$$m_{catalyst} = 3795.0 \frac{mol CH_4}{day} * 16 \frac{g CH_4}{(mol CH_4)} * \frac{g catalyst}{0.03 g CH_4}$$

$$m_{catalyst} = 2024.24 kg catalyst$$

Also, using the molar flow rate of methane and knowing that 0.647 moles of H₂ is needed for every mole of CH₄ produced (Mango, 1997), that amount of H₂ needed can be calculated.

$$\dot{n}_{H_2} = 3795.0 \frac{mol CH_4}{day} * 0.647 \frac{mol H_2}{mol CH_4}$$

$$\dot{n}_{H_2} = 2455.6 \frac{mol H_2}{day}$$

Energy Balance

To perform an energy balance on the reactor modeling is done to determine the enthalpies of formation for the various molecules in the reactor. First, the mass compositions were found for both gasoline and kerosene (ATSDR, 1998). The gasoline mass percentages did not total to 100% so those compositions were adjusted using the following equation:

$$\text{Adjusted Mass \%} = \frac{\text{Mass \% of compound}}{\text{Total mass \%}}$$

The compositions and mass flow rates can be found in Tables 9.1.5, 6 below.

Gasoline	Mass Comp.	Adjusted Mass Comp.	Mass Flow Rate (kg/day)
Alkanes	0.191	0.2203	5.53082798
Branched Alkanes	0.321	0.370242	9.29526587
Cyclo alkanes	0.05	0.05767	1.44786073
Aromatics	0.305	0.351788	8.83195043
Total	0.867	1	

Table 9.1.5 - Depicts mass composition of gasoline in stream 18.

Kerosene	Mass Comp.	Adjusted Mass Comp.	Mass Flow Rate (kg/day)
Alkanes	0.6771	0.6771	18.8444412
Branched Alkanes	0.0366	0.0366	1.01861844
Cyclo alkanes	0.0805	0.0805	2.24040395
Aromatics	0.2058	0.2058	5.72764141
Total	1	1	

Table 9.1.6 - Depicts mass composition of kerosene in stream 18.

For each hydrocarbon group, a model compound is selected using the appropriate average carbon number. From these model compounds, the MW is found and then the molar flow rates were calculated. The model compounds, MW, and molar flow rates can be seen in Tables 9.1.7, 8 below.

Gasoline	Mass Comp.	Adjusted Mass Comp.	Mass Flow Rate (kg/day)	Model Compound	MW	Mol Flow Rate (kmol/day)
Alkanes	0.191	0.2203	5.53082798	Heptane	100.2	0.0551979
Branched Alkanes	0.321	0.370242	9.29526587	2,4-dimethyl pentane	100.2	0.0927671
Cyclo alkanes	0.05	0.05767	1.44786073	Methyl-cyclohexane	98.2	0.014744
Aromatics	0.305	0.351788	8.83195043	Toluene	92.14	0.0958536
Total	0.867	1				0.2585626

Table 9.1.7 - Molar flow rates of model compounds for gasoline in stream 18.

Kerosene	Mass Comp.	Adjusted Mass Comp.	Mass Flow Rate (kg/day)	Model Compound	MW	Mol Flow Rate (kmol/day)
Alkanes	0.6771	0.6771	18.8444412	decane	142.3	0.1324276
Branched Alkanes	0.0366	0.0366	1.01861844	2,3 - dimethyl octane	142.3	0.0071582
Cyclo alkanes	0.0805	0.0805	2.24040395	2-methylpropyl cyclohexane	140.3	0.0159687
Aromatics	0.2058	0.2058	5.72764141	tetralin	132.2	0.0433256
Total	1	1				0.19888

Table 9.1.8 - Molar flow rates of model compounds for kerosene in stream 18.

A chemical compound modeling program called Pirika is used to determine the heat of formation for each compound at 20 °C (Pirika). The total enthalpy of reactants is then calculated by using the following equation:

$$\dot{H} = \bar{H} * \dot{n}$$

These enthalpies can be found in Tables 9.1.9, 10.

Gasoline	Model Compound	MW	Mol Flow Rate (kmol/day)	dH formation (kJ/mol)	Enthalpy Reactants (kJ/day)
Alkanes	Heptane	100.2	0.0551979	-225	-12419.5
Branched Alkanes	2,4-dimethyl pentane	100.2	0.0927671	-235	-21800.3
Cyclo alkanes	Methyl-cyclohexane	98.2	0.014744	-195	-2875.08
Aromatics	Toluene	92.14	0.0958536	12	1150.243
Total			0.2585626		-35944.6

Table 9.1.9 - Displays enthalpies of model compounds of gasoline in stream 18.

Kerosene	Model Compound	MW	Mol Flow Rate (kmol/day)	dH formation (kJ/mol)	Enthalpy Reactants (kJ/day)
Alkanes	decane	142.3	0.1324276	-301	-39860.7
Branched Alkanes	2,3 - dimethyl octane	142.3	0.0071582	-270	-1932.73
Cyclo alkanes	2-methylpropyl cyclohexane	140.3	0.0159687	-230	-3672.79
Aromatics	tetralin	132.2	0.0433256	126.7	5489.351
Total			0.19888		-39976.9

Table 9.1.10 - Displays enthalpies of model compounds of kerosene in stream 18.

From these Tables, the total enthalpy of formation for the gasoline and kerosene is -35944.6 kJ/day and -39976.9 kJ/day respectively. These enthalpies were then summed to find the total enthalpy of the reactants.

$$\bar{H}_{reactants} = -35944.6 \frac{kJ}{day} - 39976.9 \frac{kJ}{day}$$

$$\bar{H}_{reactants} = -75921.5 \frac{kJ}{day}$$

With the enthalpies of the reactants calculated, the molar enthalpy of the product, methane, is needed to do an overall energy balance. Since the products leave the reactor at 200 °C the following equation is used to calculate the molar enthalpy of the methane product.

$$\bar{H}_{methane} = \Delta H_{f,298} + \int_{298.15}^{473.15} C_p dT$$

Where,

$$C_p = R * (A + BT + CT^2)$$

For methane, the following values were found (Koretsky, 2004)

$$A = 1.702$$

$$B = 9.081 * 10^{-3}$$

$$C = -2.164 * 10^{-6}$$

$$\Delta H_{f,298} = -74.81 \text{ kJ/mol}$$

Therefore,

$$\bar{H}_{methane} = -74.81 \frac{\text{kJ}}{\text{mol}} + \int_{298.15}^{473.15} 0.008314 \frac{\text{kJ}}{\text{mol K}} (1.702 + 9.081 * 10^{-3} T \pm 2.164 * 10^{-6} T^2) dT$$

$$\bar{H}_{methane} = -67.7146 \frac{\text{kJ}}{\text{mol}}$$

$$\dot{H}_{methane} = -67.7146 \frac{\text{kJ}}{\text{mol}} * 3795.0 \frac{\text{mol}}{\text{day}}$$

$$\dot{H}_{methane} = -256976.9 \frac{\text{kJ}}{\text{day}}$$

Now, the overall energy balance can be calculated.

$$\Delta \dot{H} = \dot{H}_{products} - \dot{H}_{reactants}$$

$$\Delta \dot{H} = -256976.9 \frac{\text{kJ}}{\text{day}} - \left(-75921.5 \frac{\text{kJ}}{\text{day}} \right)$$

$$\Delta \dot{H}_{reaction} = -181055.4 \frac{\text{kJ}}{\text{day}}$$

Additionally, hydrogen and methane will be recycled back into the reactor from the membrane (stream 22) which will need to be cooled down in the reactor. Stream 22 is found to have 203.53

mols/hr of hydrogen and 501.10 mols/hr of methane and is at 313.6 °C. The process for which these flow rates and temperature were found can be seen in the membrane calculations further ahead in this section. To determine the enthalpy change in cooling down this stream to 200 °C, enthalpy values were taken from NIST Chemistry Webbook and can be seen in Tables 9.1.11, 12 (NIST, 2008).

	T (c)	P (ATM)	Entropy (J/mol K)	Enthalpy (kJ/mol)	Flow Rate (mols/hr)
Hydrogen	200	1	121.04	13.024	203.53
Methane	200	1	125.74	21.603	501.10

Table 9.1.11 - Displays enthalpy values for components in stream 22 at 200 C (NIST, 2008)

	T (c)	P (ATM)	Entropy (J/mol K)	Enthalpy (kJ/mol)	Flow Rate (mols/hr)
Hydrogen	313.6	3	118.21	16.353	203.53
Methane	313.6	3	126.62	27.086	501.10

Table 9.1.12 - Displays enthalpy values for components in stream 22 at 331.6 C (NIST, 2008)

From this information the enthalpy change for cooling the stream is calculated.

$$\Delta\dot{H} = 203.53 \frac{\text{mols}}{\text{hr}} * (13.024 - 16.353) \frac{\text{kJ}}{\text{mol}} + 501.10 \frac{\text{mols}}{\text{hr}} * (21.603 - 27.086) \frac{\text{kJ}}{\text{mol}}$$

$$\Delta\dot{H} = -3425.08 \frac{\text{kJ}}{\text{hr}}$$

Or,

$$\Delta\dot{H} = -82201.98 \frac{\text{kJ}}{\text{day}}$$

Summing this with the enthalpy of reaction, we can obtain the total enthalpy change for the reactor.

$$\Delta\dot{H}_{\text{reactor}} = -181055.4 \frac{\text{kJ}}{\text{day}} + -82201.98 \frac{\text{kJ}}{\text{day}}$$

$$\Delta\dot{H}_{\text{reactor}} = -263257.38 \frac{\text{kJ}}{\text{day}}$$

Catalyst

From research on the catalytic decomposition of petroleum into natural gas, it is determined that 0.03 g CH₄ can be produced per gram NiO per day (Mango, 1997). From this, the amount of catalyst needed is found.

$$m_{catalyst} = 3.795 \frac{kmol}{day} * 16 \frac{kg}{kmol} * \frac{g NiO day}{0.03 g CH_4}$$

$$m_{catalyst} = 2024 kg NiO$$

Membrane

Because the kinetics of the reaction are unknown, it is impossible to determine what the composition leaving the reactor in stream 20 would be. To simplify calculations it is assumed that all the hydrocarbons are completely converted into methane. This does not necessarily have to be the case since regulated natural gas is allowed to be composed of up to 5.1% ethane and 1.5% propane (NAESB, 2008). The real concern is eliminating the hydrogen that will be leaving the reactor with the methane. Through the use of a membrane, the hydrogen in stream 20 must be reduced to 0.2% (NAESB, 2008). To begin calculations we must determine what the composition of stream 20 is, leaving R-101. To ensure that the reaction goes to completion it is assumed that there will always be twice the amount of hydrogen that is needed in the reactor. Therefore, 204.64 mols/hr of H₂ will be leaving the reactor along with a minimum of 158.13 mols/hr of CH₄. From this, the mol fraction, x_a , of the more permeable species, H₂ can be calculated.

$$x_a = \frac{204.64 \frac{mols}{hr}}{(204.64 + 158.13) \frac{mols}{hr}}$$

$$x_a = 0.564$$

And x_b , the mole fraction of CH₄ can be found.

$$x_b = 1 - x_a$$

$$x_b = 0.436$$

The pressures for R-101 can be chosen in order to optimize the reactor operations. For this membrane the gas will be pressurized to 3 atm, three times the pressure of the permeate gas at 1 atm which will be fed back into the reactor to recycle the hydrogen. This gives a pressure ratio, R, of 0.3333. For this membrane, we will be using a hollow-fiber permeator which has a selectivity, α , of 100 for H₂ and CH₄ (McCabe, 2005). With this information, we can begin calculating local values for the concentration of H₂ in the permeate, y. This can be done using the following equation.

$$(\alpha - 1)(y)^2 + \left[1 - \alpha - \frac{1}{R} - \frac{x_a(\alpha - 1)}{R} \right] y + \frac{\alpha x_a}{R} = 0$$

By plugging in all the know information, we can solve for y.

$$(100 - 1)(y)^2 + \left[1 - 100 - \frac{1}{0.333} - \frac{0.564(100 - 1)}{0.333} \right] y + \frac{100 * 0.564}{0.3333} = 0$$

$$y = 0.982$$

This represents the concentration of hydrogen in the permeate leaving the membrane.

Further calculations can be done as the hydrogen concentration decreases throughout the membrane. For this case, we will use increments of 0.05 for the decrease in composition.

Again, another y value is calculated.

$$(100 - 1)(y)^2 + \left[1 - 100 - \frac{1}{0.333} - \frac{0.514(100 - 1)}{0.333} \right] y + \frac{100 * 0.514}{0.3333} = 0$$

$$y = 0.975$$

These calculations were continued in excel using solver until the composition of the hydrogen in the membrane is negligible. The results can be seen in Table 9.1.13 below.

Xa	Eq	y
0.56410256	-3.51725E-07	0.982161
0.51410256	-3.69355E-07	0.975347
0.46410256	-2.45248E-07	0.964548
0.41410256	-2.15382E-07	0.945692
0.36410256	-2.46333E-08	0.909556
0.31410256	-1.96413E-07	0.841487
0.26410256	-7.51627E-08	0.73762
0.21410256	-2.00656E-07	0.611336
0.16410256	-9.83378E-07	0.474416
0.11410256	-5.31679E-07	0.332399
0.06410256	-3.49386E-08	0.187697
0.01410256	-1.98045E-07	0.041443

Table 9.1.13 - Final solver calculations for membrane

From this we want to determine the volumetric flow rate of permeate. To do so, we must calculate the local volume, V , leaving in between each increment previously chosen by using the following equation.

$$\Delta V = L_i * \frac{x_i - x_j}{\bar{y} - x_j}$$

Where L is the total flow rate in the membrane, \bar{y} is the average permeate composition in between each increment, and x_i and x_j are the compositions of hydrogen in the membrane at each increment. For the first increment,

$$\bar{y} = \frac{0.983 - 0.975}{2}$$

$$\bar{y} = 0.979$$

$$L_i = (204.64 + 158.13) \frac{\text{mols}}{\text{hr}}$$

$$L_i = 362.76 \frac{\text{mols}}{\text{hr}}$$

And,

$$\Delta V = 362.76 \frac{\text{mols}}{\text{hr}} * \frac{0.564 - 0.514}{0.979 - 0.514}$$

$$\Delta V = 39.04 \frac{\text{mols}}{\text{hr}}$$

From this the hydrogen flow rate in permeate and hydrogen recovery % can be calculated as follows.

$$H_2 \text{ in permeate} = \Delta V * \bar{y}$$

$$H_2 \text{ in permeate} = 39.04 \frac{\text{mols}}{\text{hr}} * 0.979$$

$$H_2 \text{ in permeate} = 38.21 \frac{\text{mols}}{\text{hr}}$$

$$H_2 \text{ recovery} = \frac{H_2 \text{ in permeate}}{X_a}$$

$$H_2 \text{ recovery} = \frac{38.21 \frac{\text{mols}}{\text{hr}}}{204.64 \frac{\text{mols}}{\text{hr}}}$$

$$H_2 \text{ recovery} = 0.187$$

These calculations were done for all of the incremental changes and can be seen in Table 9.1.14 below.

Xa	Eq	y	y avg	L (mol/hr)	ΔV (mol/hr)	H2 in perm (mol/hr)	H2 recover
0.56410256	-3.51725E-07	0.982161		362.7616			
0.51410256	-3.69355E-07	0.975347	0.978754	323.7258	39.03588	38.2065307	0.186706
0.46410256	-2.45248E-07	0.964548	0.969948	291.7272	31.99851	69.2434108	0.3383756
0.41410256	-2.15382E-07	0.945692	0.95512	264.7663	26.961	94.9943918	0.4642143
0.36410256	-2.46333E-08	0.909556	0.927624	241.2741	23.49213	116.786251	0.5707058
0.31410256	-1.96413E-07	0.841487	0.875521	219.7862	21.48789	135.599357	0.6626408
0.26410256	-7.51627E-08	0.73762	0.789553	198.8722	20.91406	152.112125	0.7433347
0.21410256	-2.00656E-07	0.611336	0.674478	177.2733	21.5989	166.680114	0.8145249
0.16410256	-9.83378E-07	0.474416	0.542876	153.8723	23.40094	179.383932	0.8766053
0.11410256	-5.31679E-07	0.332399	0.403408	127.2789	26.59344	190.111927	0.9290304
0.06410256	-3.49386E-08	0.187697	0.260048	94.80074	32.47814	198.557804	0.9703033
0.01410256	-1.98045E-07	0.041443	0.11457	47.62099	47.17975	203.963197	0.9967182

Table 9.1.14 - Calculations of hydrogen in permeate (stream 22)

From the information in the Table, the total flow rate and compositions of both the permeate and non-permeate streams can be calculated. The total permeate flow rate, V_{perm} , can be found by summing all the local ΔV s for the membrane.

$$V_{perm} = \sum \Delta V_s = 315.14 \frac{mols}{hr}$$

And,

$$V_{non-perm} = L - V_{perm}$$

$$V_{non-perm} = 362.76 \frac{mols}{hr} - 315.14 \frac{mols}{hr}$$

$$V_{non-perm} = 47.62 \frac{mols}{hr}$$

Knowing the final H_2 in the permeate and these flow rates, the compositions for the hydrogen and methane in the permeate and non-permeate can be found with.

$$y_a = H_2 \text{ composition in perm}$$

$$y_b = CH_4 \text{ composition in perm}$$

$$x_a = H_2 \text{ composition in non - perm}$$

$$x_b = CH_4 \text{ composition in non - perm}$$

$$y_a = \frac{H_2 \text{ in perm}}{V_{perm}}$$

$$y_a = \frac{203.96 \frac{mols}{hr}}{315.14 \frac{mols}{hr}}$$

$$y_a = 0.65$$

$$y_b = 1 - y_a$$

$$y_b = 0.35$$

$$x_a = \frac{H_2 \text{ initial} - H_2 \text{ in perm}}{V_{\text{non-perm}}}$$

$$x_a = \frac{204.64 \frac{\text{mols}}{\text{hr}} - 203.96 \frac{\text{mols}}{\text{hr}}}{47.62 \frac{\text{mols}}{\text{hr}}}$$

$$x_a = 0.014$$

$$x_b = 1 - x_a$$

$$x_b = 0.986$$

From these the molar flow rates can be calculated for each stream. These values have been calculated and are in Table 9.1.15.

	Perm	Non-perm
Flow	315.1406551	47.62099
H2 comp	0.65	0.014103
CH4 comp	0.35	0.985897
Mol H2 (mol/hr)	203.9631967	0.671578
Mol CH4 (mol/hr)	111.1774584	46.94941

Table 9.1.15 – Initial molar flow rates around the membrane

From Table 9.1.15 you can see that the H₂ in the non-permeate stream does not quite meet the specifications of being below 0.2%. More importantly is that 111.18 mols/hr of CH₄ is being recycled into the reactor and only 46.95 mols/hr of CH₄ leaving the system. Since this number is much smaller than the methane production rate of 158.13 mols/hr the system has to be reworked. All of the previous calculations were performed multiple times with adjustments made to the methane inlet flow rate until the non-permeate stream had approximately 158.13 mols/hr leaving and the compositions of hydrogen in the non-permeate stream is below 0.2%. In the end, 501 mols/hr had to be recycled throughout the system to meet these requirements. The results can be found in the following Tables (9.1.16, 17).

X	Eq	Y'	y	L(mol/h)	ΔV (mol/h)	H2 in perm (mol/hr)	H2 recover
0.23691116	-4.07356E-08	0.670795		863.7616			
0.18691116	-2.45055E-08	0.537736	0.604265	760.281	103.4807	62.5297729	0.3055677
0.13691116	-1.39368E-08	0.397619	0.467677	645.3536	114.9273	116.278657	0.5682253
0.08691116	-8.25382E-09	0.253949	0.325784	510.2703	135.0833	160.286586	0.7832813
0.03691116	-7.52743E-07	0.108307	0.181128	333.3595	176.9108	192.330088	0.93987
0.00191116	-7.55421E-08	0.020319	0.064313	159.1353	174.2242	203.534965	0.9946255

Table 9.1.16 - Final results for membrane recovery

	Perm	Non-perm
Flow	704.6263258	159.1353
H2 comp	0.29	0.001911
CH4 comp	0.711144819	0.998089
Mol H2 (mol/hr)	203.5349649	0.304133
Mol CH4 (mol/hr)	501.0913609	158.8312

Table 9.1.17 - Final molar flow rates leaving the F-101

For costing purposes the size of the membrane must be determined. The first step in determining the size of the membrane is calculating the permeability of the membrane, Q . To do so, the following equation is used (McCabe, 2005).

$$Q = \frac{J}{P_1x - P_2y}$$

Where P_1 is the inlet pressure, P_2 is the permeate pressure, x is the mole percentage of H_2 in the feed, y is the mole percentage of H_2 in the permeate, and J is the molar flux. For a similar hydrogen membrane, the molar flux is $0.00648 \text{ mol/hr} \cdot \text{cm}^2$ (Dorris, Wang, Picciolo, & Dusek, 2002). Using the data from above the permeability for each section of the membrane can be calculated. A sample can be seen below.

$$Q = \frac{0.00648 \frac{\text{mol}}{\text{h} \cdot \text{cm}^2}}{(3 \text{ atm} \cdot 0.1869 - 1 \text{ atm} \cdot 0.5377)}$$

$$Q = 0.2818 \frac{\text{mol}}{\text{h} \cdot \text{cm}^2 \cdot \text{atm}}$$

With the permeability of each section known, the area, A , for each section can be calculated using the following equation (McCabe, 2005).

$$A = \frac{\Delta Vy}{Q_A(P_1x - P_2y)_{ave}}$$

For the first section of the membrane

$$A = \frac{103.48 \frac{mol}{h} * 0.5377}{0.2818 \frac{mol}{h * cm^2 * atm} \left(\frac{(3 atm * 0.237 - 1 atm * 0.671) + (3 atm * 0.187 - 1 atm * 0.538)}{2} \right)}$$

$$A = 6275.77 cm^2$$

Or

$$A = 0.628 m^2$$

This represents the area required for the first section of the membrane. Similar calculations were done for the remainder sections and can be found in Table 9.1.18.

X	y'	y	$\Delta V(mol/h)$	Q (mol/h*cm ² atm)	A (cm ²)
0.23691116	0.670795				
0.18691116	0.537736	0.604265	103.4807	0.281766	6275.771
0.13691116	0.397619	0.467677	114.9273	0.494091	5122.147
0.08691116	0.253949	0.325784	135.0833	0.955088	3609.84
0.03691116	0.108307	0.181128	176.9108	2.67074	1557.768
0.00191116	0.020319	0.064313	174.2242	15.62512	159.4923

Table 9.1.18 - Calculations for area of membrane in F-101

The areas are then added to find the total area of the membrane.

$$A = 1.673 m^2$$

Compressor (C-101)

Before the membrane there must be a compressor to compress the gas in stream 20 from 1 atm to 3 atm so that the pressure differential forces the gas through the membrane. This compressor has a total inlet flow of 863.76 mols/hr. Of that inlet flow, 204.71 mols/hr is hydrogen and 659.05 mols/hr is methane. This compressor will operate isentropically and it is assumed that the gas mixture has ideal mixing and is an ideal gas. First, the inlet properties of

the gas are found using the NIST Chemistry WebBook at a temperature of 200 °C and a pressure of 1 atm (Table 9.1.19) (NIST, 2008).

	T (c)	P (ATM)	Entropy (J/mol K)	Enthalpy (kJ/mol)	Flow Rate (mols/hr)
Hydrogen	200	1	121.04	13.024	204.7111
Methane	200	1	125.74	21.603	659.0489

Table 9.1.19 – Entropy data for stream 20

The total entropy rate can be calculated as follows.

$$\dot{s} = \dot{n}_i * \bar{s}_i + \dot{n}_j * \bar{s}_j$$

$$\dot{s} = 204.7 \frac{\text{mols}}{\text{hr}} * 121.04 \frac{\text{J}}{\text{mol K}} + 659.05 \frac{\text{mols}}{\text{hr}} * 125.74 \frac{\text{J}}{\text{mol K}}$$

$$\dot{s} = 107647 \frac{\text{J}}{\text{hr K}}$$

The temperatures at which each gas can be compressed to 3 atm while maintaining the same entropy is determined. For hydrogen and methane these temperatures are 375 °C and 304 °C respectively. From this, it is known that the temperature at which the gas mixture would contain the same entropy once compressed would be in this range. From NIST Chemistry WebBook, this temperature is found to be 313.6 °C. The physical data for the compressed gas can be found in Table 9.1.20 (NIST, 2008).

	T (c)	P (ATM)	Entropy (J/mol K)	Enthalpy (kJ/mol)	Flow Rate (mols/hr)
Hydrogen	313.6	3	118.21	16.353	204.7111
Methane	313.6	3	126.62	27.086	659.0489

Table 9.1.20 - Physical data for stream 21

Performing an entropy balance shows that the compressed gas has the same entropy as the non-compressed gas.

$$\dot{s} = 204.7 \frac{\text{mols}}{\text{hr}} * 118.21 \frac{\text{J}}{\text{mol K}} + 659.05 \frac{\text{mols}}{\text{hr}} * 126.62 \frac{\text{J}}{\text{mol K}}$$

$$\dot{s} = 107647 \frac{J}{hr K}$$

This confirms that the temperature of stream 21 leaving the compressor is 313.6 °C. Similarly, an enthalpy balance is done on the compressor to determine how much work is necessary to power it.

$$\dot{H} = \dot{n}_i * \bar{H}_i + \dot{n}_j * \bar{H}_j$$

$$\dot{H}_{initial} = 13.024 \frac{kJ}{mol} * 204.7 \frac{mols}{hr} + 21.603 \frac{kJ}{mol} * 659.05 \frac{mols}{hr}$$

$$\dot{H}_{initial} = 16903.6 \frac{kJ}{hr}$$

$$\dot{H}_{final} = 16.353 \frac{kJ}{mol} * 204.7 \frac{mols}{hr} + 27.086 \frac{kJ}{mol} * 659.05 \frac{mols}{hr}$$

$$\dot{H}_{final} = 21198.6 \frac{kJ}{hr}$$

$$\Delta H = 21198.6 \frac{kJ}{hr} - 16903.6 \frac{kJ}{hr}$$

$$\Delta H = 4295 \frac{kJ}{hr}$$

Blowers (B-101/102)

For B-101 the inlet pressure is 1 atm and the outlet pressure is 1.13 atm. The reasoning for the outlet pressure can be seen in the blower utilities calculations in section 9.3. Using the inlet temperature of 22 °C the outlet temperature can be calculated using the following heuristic (Seider, 2003)

$$T_2 = T_1 \left(\frac{P_2}{P_1} \right)^a$$

Where

$$a = \frac{k - 1}{k}$$

$$k = \frac{c_p}{c_v}$$

The k value for air is 1.4 (The Engineering Tool Box, 2005). Therefore,

$$a = \frac{(1.4 - 1)}{1.4}$$

$$a = 0.286$$

$$T_2 = 22 \text{ }^\circ\text{C} \left(\frac{1.13}{1} \right)^{0.286}$$

$$T_2 = 22.78 \text{ }^\circ\text{C}$$

For B-102 the inlet pressure is 1atm and the outlet pressure is 1.1 atm. The reasoning for the outlet pressure is noted in the blower utilities calculations in section 9.3. Because the gas stream is greater than 99% air, the same k value will be used. Using the inlet temperature of 20 °C the outlet temperature is calculated.

$$T_2 = 20 \text{ }^\circ\text{C} \left(\frac{1.1}{1} \right)^{0.286}$$

$$T_2 = 20.55 \text{ }^\circ\text{C}$$

GAC Estimate

The vapor contaminants in stream 11 are entering the GAC unit (F-102) at 303.1 SCFM. To get an estimate for the amount of GAC needed Carbtrol Corporation was contacted. A “request for quotation” form was submitted to Carbtrol. Their estimate to reduce the MTBE concentration from 1311 ppm to 20 ppb and the BTEX concentration from 1.84 ppm to 5 ppb was 95 pounds of carbon per day. The “request for quotation” form and response can be seen in appendix 9.5.

PROJECT: Zelen Environmental

FLOW IN
CFM: 303.10
FLOW IN
CFD: 436464.00

PERFORMANCE:

<u>CONTAMINANT</u>	<u>CONC(ppmv)</u>	<u>#CONT</u> <u>/DAY</u>	<u>#CARBON</u> <u>/DAY</u>	<u>#CONT</u> <u>/100,000cf</u>	<u>#CARBON</u> <u>/100,000cf</u>
MTBE	1311	21.21	94.90	4.53	20.28
Benzene	1.84	0.00	0.03	0.00	0.01
TOTALS	1312.84	21.21	94.93	4.53	20.29

Calculation based on CARBTROL CSV carbon having a Carbon Tetrachloride number of: 60.00

9.2 Economic Calculations

In order to determine the expenses associated with this design, it is first necessary to price each piece of equipment. These pricings either came from calculations based upon information in *Product and Process Design Principles* (Seider, 2003) or from quotes given by specific companies.

Gas Permeation Membrane (F-101)

The area of the membrane directly impacts the price of the membrane itself. As described in the process calculations in section 9.1, the membrane is 1.673 m² which is roughly 18 ft². From here the calculation for the membrane is taken from Table 16.32 in the process design book (Seider, 2003). Due to the fact that streams 21, 22, and 23 are gaseous, a gas permeation membrane is priced using the following equation:

$$C_P [\text{\$}] = 35A$$

Where, C_P is the purchase cost and A is the area in ft^2 . Plugging 18 ft^2 into the equation, we get a purchase cost equal to \$630. For the purposes of calculating the bare module cost, C_{BM} , the equation below is utilized for all pieces of equipment.

$$C_{BM} [\text{\$}] = F_{BM} * C_P$$

F_{BM} is the bare module factor, which is unitless

Since the purchase costs in the process design book are for the year 2000, it is important to use a cost index, I , to determine the approximate cost today (Chemical Week Associates, 2008). This CE cost index states that I_{base} , setting our base as the year 2000 to accommodate the base for the book, is 394.1 and I_{2008} , is 583.7 (Chemical Week Associates, 2008). From this, the equation above becomes:

$$C_{BM} [\text{\$}] = \left(\frac{I_{2008}}{I_{\text{base}}} \right) * F_{BM} * C_P$$

For membrane separators, F_{BM} equals 3.2 as taken from Table 16.14 in the process description book (Seider, 2003). With all of the variables known, C_{BM} is calculated as follows.

$$C_{BM} [\text{\$}] = \left(\frac{583.7}{394.1} \right) (3.2) (\$631)$$

$$C_{BM} = \$2988$$

Catalytic Reactor (R-101)

Because the reactor is running at a pressure of 1 atm and only the volume is known, the Green Team decided it is unnecessary to price a pressurized vessel. The reasoning behind this decision is due to the fact that pricing a pressurized vessel requires a known length and diameter,

both of which are unknown (Seider, 2003). Instead, we will price an open storage vessel based upon a few assumptions. The first assumption is that the difference in price between an open and closed vessel is negligible. This assumption is made due to the fact that no vessels were available for costing that fell within our volume estimation. The second assumption we will make is that after pricing the vessel, the purchase cost will be modified to accommodate for a material different than fiberglass by multiplying by the material factor for stainless steel 316 as defined in Table 16.26 (Seider, 2003). The third assumption is that only one tenth of the gas flowing through the reactor is actually residing in the reactor at any given time. This assumption is necessary in order to accommodate for the uncertainty associated with determining the precise amount of vapor inside the reactor.

With these assumptions in place, the total volume is calculated first by summing the volume of the catalyst, the volume of the liquid in the reactor, and the volume of the gas in the reactor. The volume of the catalyst is determined based upon the requirement of 2000 kg of NiO, which is calculated as described in section 9.1. From there, the volume of catalyst needed is calculated by dividing the mass by the density, which is 1.1 kg/L (Appendix 9.5), and converting from liters to gallons. This volume, V_{cat} , amounted to 480.36 gal., and the volume of the liquid in the reactor, V_{liq} , is equal to 18 gal. The volume of the gas in the reactor, V_{gas} , is calculated assuming the ideal gas law holds true for this process. We know that the molar flow rate of the gas is 863.76 mol/hr (see stream Tables in section 2.4), and the reactor is operating at a pressure of 1 atm and a temperature of 200 °C.

$$V_{gas} = \frac{nRT}{P}$$

$$V_{gas} = \frac{\left(863.76 \frac{mol}{hr}\right) \left(0.08206 \frac{L \cdot atm}{mol \cdot K}\right) (200 + 273.15 K)}{1 atm} \cdot \frac{264.2 gal}{1000 L}$$

$$V_{gas} = 8861 \text{ gal/hr}$$

Considering the assumption that only one tenth of the gas is residing in the reactor at any given time, the V_{gas} is 886.1 gal.

$$V_{tot} = V_{cat} + V_{liq} + V_{gas}$$

$$V_{tot} = 480.36 \text{ gal} + 18 \text{ gal} + 886.1 \text{ gal}$$

$$V_{tot} = 1385 \text{ gal}$$

To calculate the purchase cost of the reactor, the equation for an open storage tank from Table 16.32 is utilized (Seider, 2003).

$$C_P [\text{\$}] = 14(V_{tot})^{0.72}$$

However, since this is the purchase price for the tank made of fiberglass we must adjust for this using the assumption made earlier.

$$C_{P,adj} [\text{\$}] = F_M * C_P$$

$$C_{P,adj} [\text{\$}] = 2.1 * 14(1385)^{0.72}$$

$$C_{P,adj} = \$ 5371$$

For reactors F_{BM} is equal to 4.3. With that and the purchase cost for the tank, the bare module cost is then calculated.

$$C_{BM} [\text{\$}] = \left(\frac{583.7}{394.1}\right)(4.3)(\$5371)$$

$$C_{BM} = \$34,200$$

Air Stripping Column (S-101)

For the purposes of costing an air stripping column, the weight of the tower must first be determined. From the process calculations, it is determined that the tower had a length, L , equal to 2.86 m (9.4 feet) and an inner diameter, D_i , equal to 0.75 m (2.5 feet). For a D_i up to 4 feet, the wall thickness, t_p , is equal to $\frac{1}{4}$ in (Seider, 2003). Since the operating pressure is less than 1.34 psig, it can be assumed that t_p equals t_v , the average wall thickness for a vertical vessel because there will be no effect of wind load (Seider, 2003). From there, the shell thickness, t_s , is calculated.

$$t_s = t_v + t_c$$

$$t_c \text{ (corrosion allowance)} = 1/8 \text{ in}$$

$$t_s = 0.25 \text{ in} + 0.13 \text{ in}$$

$$t_s = 0.38 \text{ in or } 0.03125 \text{ ft}$$

After D_i , L , t_s are determined, it is possible to calculate the weight of the tower, W , according to the equation below (Seider, 2003).

$$W = \pi(D_i + t_s)(L + 0.8D_i)t_s\rho$$

$$\rho = 490 \text{ lb/ft}^3 \text{ for stainless steel (K\&K Associates, 2008)}$$

$$W = \pi(2.5 \text{ ft} + 0.03125 \text{ ft})(9.4 \text{ ft} + 0.8*2.5 \text{ ft}) * 0.03125 \text{ ft} * 490 \text{ lb/ft}^3$$

$$W = 1361 \text{ lbs}$$

With the weight calculated, the cost of the vessel, C_V , can be calculated using the equation (Seider, 2003):

$$C_V [\text{\$}] = \exp\{7.0374 + 0.18255[\ln(W)] + 0.02297[\ln(W)]^2\}$$

$$C_V [\text{\$}] = \exp\{7.0374 + 0.18255[\ln(1361)] + 0.02297[\ln(1361)]^2\}$$

$$C_V = \$14,060$$

Another cost associated with an air stripping tower is that of the platforms and ladders, C_{PL} , required for maintenance. However, this cost is only incurred in the event that the tower is between 3 and 24 feet in diameter and 27 and 170 feet in length (Seider, 2003). Since our air stripping column has measurements smaller than these parameters, this cost is considered negligible. The third cost to be considered, is the cost of the plates for inside the column. The number of plates, N_T , is calculated to equal 18 (see section 9.1). The equations for calculating the cost of the plates are described below (Seider, 2003).

$$C_T [\text{\$}] = N_T F_{NT} F_{TT} F_{TM} C_{BT}$$

$$F_{NT} = \frac{2.25}{(1.0414)^{N_T}} = 1.084 \text{ (factor for number of trays)}$$

$$F_{TT} = 1.0 \text{ (factor for tray types: sieve)}$$

$$F_{TM} = 1.189 + 0.0477 D_i \text{ (factor for material of construction: SS316)}$$

$$F_{TM} = 1.33$$

$$C_{BT} [\text{\$}] = 369 \exp(0.1739 D_i) \text{ (base cost for sieve trays)}$$

$$C_{BT} = \$566$$

$$C_T = 18 * 1.084 * 1.0 * 1.33 * \$566$$

$$C_T = \$14,702$$

From here, the purchase cost can be calculated. For air stripping columns, this value is priced as follows.

$$C_P [\text{\$}] = F_M * C_V + C_{PL} + C_T$$

$$F_M = 2.1 \text{ for stainless steel 316 (Table 16.26) (Seider, 2003)}$$

$$C_P [\text{\$}] = 2.1 * \$14,100 + \$0 + \$14,702$$

$$C_P = \$44,300$$

With an F_{BM} equal to 4.16 for vertical pressure vessels and towers, the bare module cost is then calculated using the previously cited equation.

$$C_{BM} [\text{\$}] = \left(\frac{583.7}{394.1}\right)(4.16)(\$44,300)$$

$$C_{BM} = \$273,000$$

Compressor & Electric Motor (C-101 A/B)

The cost of the compressor is dependent on the power required by the unit. It is known that 4295 kJ/hr are required, (see utility Table, section 2.5) which can therefore be used to determine the power requirement for the compressor by converting the kJ/hr into Hp. In performing this conversion, the brake power, P_B , is found to equal 1.6 Hp. The efficiency of the compressor, η_M , is assumed to be 80% (Seider, 2003). This information is used to calculate the size factor, P_C .

$$P_C = P_B / \eta_M$$

$$P_C = 1.6 \text{ Hp} / 0.8 = 2.0 \text{ Hp}$$

Now, the base compressor purchase cost, C_B , as defined for a centrifugal compressor can be calculated using the following equation. A centrifugal compressor is chosen because it has small foundations and low maintenance requirements (Seider, 2003).

$$C_B [\text{\$}] = \exp\{7.2223 + 0.80[\ln(P_C)]\}$$

$$C_B [\text{\$}] = \exp\{7.2223 + 0.80[\ln(2.0)]\}$$

$$C_B = \$2385$$

Taking into consideration the drive factor, F_D , and the material factor, F_M , for the unit, the purchase cost can be determined using the equation below.

$$C_P [\text{\$}] = F_D F_M C_B$$

$$F_D = 2.5 \text{ for stainless steel}$$

$$F_M = 1.0 \text{ for an electric motor}$$

$$C_P [\text{\$}] = 2.5 * 1.0 * \$2385$$

$$C_{P,\text{compressor}} = \$5962$$

Before the bare module cost can be calculated, the cost of the electric motor must first be calculated. Using the graph on page 510 of the process design book, the purchase cost for an electric motor at 3600 rpm in an open, drip-proof enclosure is equal to \$250 for a $P_C = 2.0$ Hp. However, since our process is potentially explosive, the motor type factor, F_T , (as found in Table 16.22 for a explosion-proof enclosure at 3600 rpm) must be multiplied by the purchase cost, which can now be considered the base cost (C_B) (Seider, 2003).

$$C_P [\text{\$}] = F_T C_B$$

$$C_P [\text{\$}] = 1.8 * \$250$$

$$C_{P,\text{motor}} = \$450$$

Adding the purchase cost for the motor to that of the compressor, and noting that the bare module factor for a compressor and driver is 2.15 allows for the calculation of the bare module cost.

$$C_{BM} [\text{\$}] = \left(\frac{583.7}{394.1}\right)(2.15)(\$5962 + \$450)$$

$$C_{BM} = \$20,500$$

Blowers (B-101/B-102)

B-101

B-101 is priced as a Rotary Straight-Lobe Blower due to the fact that this type of blower can accommodate very small size factors (P_C). With a break horsepower (P_B) equal to 1.99×10^4 Hp and a motor efficiency (η_M) of 80%, P_C is calculated using the same equation as discussed in the compressor pricing above (Seider, 2003).

$$P_C = 2.49 \times 10^{-4} \text{ Hp}$$

From there the base cost is calculated using the following equation.

$$C_B [\$] = \exp\{7.35356 + 0.79320[\ln(P_C)] - 0.012900[\ln(P_C)]^2\}$$

$$C_B [\$] = \exp\{7.35356 + 0.79320[\ln(2.49 \times 10^{-4})] - 0.012900[\ln(2.49 \times 10^{-4})]^2\}$$

$$C_B = \$0.89$$

In order to determine C_P for a blower the equation below is used, knowing a blower has a bare module factor, F_{BM} , equal to 2.15 (Seider, 2003).

$$C_P [\$] = F_M C_B$$

$$F_M = 0.60 \text{ for cast aluminum blades}$$

$$C_P = 0.60 * \$0.89 = \$0.53$$

$$C_{BM} [\$] = \left(\frac{583.7}{394.1}\right)(2.15)(\$0.53)$$

$$C_{BM} = \$1.70$$

B-102

B-102 is priced as a Centrifugal (turbo) Blower because P_B is calculated as 18.23 Hp. The design book indicates in Figure 16.8 that this type of blower has a lower overall cost for size factors from 5 Hp to 1,000 Hp (Seider, 2003). Using the same η_M (80%) as in the calculations for B-101, P_C is equal to 22.788 Hp. From there, C_B is calculated using the following equation as given in the process design book.

$$C_B [\text{\$}] = \exp\{6.6547 + 0.7900[\ln(P_C)]\}$$

$$C_B [\text{\$}] = \exp\{6.6547 + 0.7900[\ln(22.788)]\}$$

$$C_B = \$9177$$

Using the equation above for C_P , the purchase cost is determined to be \$5506.

$$C_{BM} [\text{\$}] = \left(\frac{583.7}{394.1}\right)(2.15)(\$5506)$$

$$C_{BM} = \$17,600$$

Storage Tank (V-103)

V-103 is priced as a 550 gal vessel with a purchase cost of \$1,100. The volume of 550 gal is chosen to ensure the dual phase extractor does not need to be shut down in the event of a malfunction downstream of V-103. Because we are pumping in 18 gal/day of petroleum, a tank volume of 550 gal allows the reactor to be shut down for a little over 30 days without affecting the dual phase extractor. The bare module factor of this tank is assumed to be one due to the fact that there is no installation cost associated with this piece of equipment. Finally, $C_{BM} = \$1,100$.

Dual Phase Extractor (P-101)

The price for the dual phase extractor is \$50,000 (Appendix 9.5). For pumps, F_{BM} is equal to 3.3, which results in a $C_{BM} = \$165,000$. This price is also already at its present value, and therefore does not need to be adjusted.

Centrifugal Pumps (P-102,103,104)

Pumps 102,103 and 104 are priced based upon a quote received online. Because the required horsepower for each pump is less than 1 Hp, the pumps were chosen based upon the assumption that they could be adjusted to the required horsepower. The pump found is a centrifugal pump with Hp equal to $\frac{3}{4}$. They are made out of stainless steel 316 and cost \$150.25 each (Appendix 9.5). For three pumps, the purchase cost is equal to \$450.75. Using the same material factor as for the dual phase extractor, $C_{BM} = \$1488$.

Vapor-Liquid Separator (V-101) & Decanter (V-102)

Both the vapor-liquid separator and decanter were priced as pressurized vessels assuming that the equations for a horizontal vessel hold true at vessel weights less than 1000 lbs and vessel inside diameters less than 3 ft. An example calculation will be performed for the vapor-liquid separator. Since the same equations were used to calculate the cost of the decanter, the values will be cited without typing reiteration of the equations.

The vapor-liquid separator has D_i equal to 0.35 m (1.15 ft) and L equal to 0.263 m (0.863 ft). Using the same assumption as done with S-101, which justified how the shell thickness is determined, the shell thickness, t_s , for V-101 is found to equal 0.03125 ft (Seider, 2003). From here it is possible to calculate the weight of the separator.

$$W = \pi(D_i + t_s)(L + 0.8D_i)t_s\rho$$

$$\rho = 490 \text{ lb/ft}^3 \text{ for stainless steel (K\&K Associates, 2008)}$$

$$W = \pi(1.15 \text{ ft} + 0.03125 \text{ ft})(0.863 \text{ ft} + 0.8*1.15 \text{ ft})*0.03125 \text{ ft}*460 \text{ lb/ft}^3$$

$$W = 94.9 \text{ lbs}$$

Now the cost of the vessel, C_V , and the cost of the platforms and ladders, C_{PL} , can be calculated using the equations for horizontal vessels (Seider, 2003).

$$C_V [\text{\$}] = \exp\{8.717 - 0.2330[\ln(W)] + 0.04333[\ln(W)]^2\}$$

$$C_V [\text{\$}] = \exp\{8.717 - 0.2330[\ln(94.9)] + 0.04333[\ln(94.9)]^2\}$$

$$C_V = \$5190$$

$$C_{PL} [\text{\$}] = 1,580(D_i)^{0.20294}$$

$$C_{PL} [\text{\$}] = 1,580(1.15)^{0.20294}$$

$$C_{PL} = \$1625$$

With these costs and an F_M of 2.1 for stainless steel, C_P can be calculated using the following equation.

$$C_P [\text{\$}] = F_M C_V + C_{PL}$$

$$C_P [\text{\$}] = (2.1)(\$5190) + \$1625$$

$$C_P = \$12,530$$

Now the bare module cost can be evaluated using 4.16 as the bare module factor (Seider, 2003).

$$C_{BM} [\$] = \left(\frac{583.7}{394.1}\right)(4.16)(\$12,530)$$

$$C_{BM} = \$77,200 \text{ (V-101)}$$

The decanter has a D_i of 0.5 m (1.64 ft) and a L of 2.52 m (8.27 ft). Using these variables and a t_s of 0.03125 ft (because the inner diameter is under 4 ft), the decanter is calculated to have $W = 723.2$ lbs which leads to a $C_V = \$8614$. The diameter results in a $C_{PL} = \$1747$. Since both the decanter and separator were priced as horizontal pressurized vessels, the equations for purchase cost and bare module cost are the same. With this holding true, the purchase cost is found to equal \$19,840 which gave a $C_{BM} = \$122,300$ (V-102).

Generator (G-101)

The purchase cost of the generator is set at \$14,999 (Appendix 9.5). Assuming that the F_{BM} is equal to 2.15, due to the fact that the generator is acting as a driver for the process, the $C_{BM} = \$32,300$.

Granular Activated Carbon Unit (F-102)

The vessel size needed to contain the granular activated carbon (GAC) has a diameter of 24" and a height of 34" (Carbtrol). For the purposes of this design, it was assumed that two vessels would be needed in order to have one running while the other is regenerating. At a cost of \$772 per tank (the lids were considered to be a negligible cost compared to the tank), the total cost for the vessels is \$1,544.

A second expense is a yearly cost for purchasing the GAC itself. The price of GAC is \$1.35/lb (Muñoz, Peral, & Ayllón, 2007) and our design requires 95 lb/day. From there, it is possible to calculate the total cost per year for the GAC. This yearly expense amounts to \$46,800.

Catalyst Expenses

As discussed in the process description (section 2.6), 2000 kg of the NiO catalyst supported on Al₂O₃ are needed to crack the hydrocarbons into methane. The catalyst is priced at \$15/kg and the total estimated cost for the catalyst is then equal to \$30,000 (Appendix 9.5).

9.3 Utility Calculations

Reactor (R-101)

The enthalpy for the reactor is previously found to be -263257.38 kJ/day. From this, the flow rate of cooling water required to keep the reactor maintained at 200 °C can be determined. The cooling water will be entering the reactor at 20 °C and leaving at 45 °C, according to Heuristic 9.11-7 (Walas, 1988). Knowing these temperatures enables us to calculate the molar enthalpy that the cooling water will gain.

$$\Delta\bar{H}_{water} = \int_{293.15}^{318.15} C_p dT$$

Where,

$$C_p = 9.069 * .008314 \frac{kJ}{mol K}$$

Therefore,

$$\Delta\bar{H}_{water} = \int_{293.15}^{318.15} 9.069 * 0.008314 \frac{kJ}{mol K} dT$$

$$\Delta\bar{H}_{water} = 1.885 \frac{kJ}{mol}$$

The total moles of cooling water required can now be found.

$$\dot{n}_{water} = \frac{263257.38 \frac{kJ}{day}}{1.885 \frac{kJ}{mol}}$$

$$\dot{n}_{water} = 139659.09 \frac{mol}{day}$$

Or,

$$\dot{n}_{water} = 5819.13 \frac{mol}{hr}$$

As previously stated, the system will be treating 5589 gallons/day of contaminated groundwater.

This treated groundwater will be used as the cooling water for the reactor. To determine if any additional cooling water is needed, a few unit conversions are carried out.

$$\dot{n}_{treated\ water} = \frac{\dot{V} * \rho}{MW}$$

Where ρ is 1g/cm^3 and the MW of water is 18g/mol .

$$\dot{n}_{treated\ water} = 5589 \frac{gal}{day} * 3.785 \frac{L}{gal} * 1 \frac{g}{cm^3} * \frac{mol}{18\ g} * \frac{1000\ cm^3}{L} * \frac{kmol}{1000\ mol} * \frac{day}{24\ hr}$$

$$\dot{n}_{treated\ water} = 48968.44 \frac{mol}{hr}$$

The amount of treated water is much larger than the cooling water requirement, so no additional cooling water is needed.

Compressor (C-101)

From the process calculations in section 9.1, it is found that the enthalpy change across the compressor is 4295 kJ/hr or 1.6 Hp. Using an efficiency value, η , of 0.8 the actual Hp requirement, P_c can be calculated.

$$P_c = \frac{P}{\eta}$$

$$P_c = \frac{1.6 \text{ Hp}}{0.8}$$

$$P_c = 2.0 \text{ Hp}$$

Or

$$P_c = 1492 \text{ W}$$

Pumps (P-102/103/104)

For P-102, there is 5589 gallons/day (3.88 gallons/min) entering through stream 4 at 1 atm (14.7 psi). The water leaving the pump goes into S-101 and is then used downstream as cooling water in R-101. For the cooling coils in the reactor, there is a pressure drop of 5 psi (Seider, 2003). We will assume the same pressure drop for the air stripper, therefore the pressure increases from 14.7 psi to 24.7 psi. From this data we can calculate the theoretical horse power, T_{Hp} , required for the pump using the following equation (Seider, 2003).

$$T_{Hp} = \frac{(gpm)(\text{Pressure increase, psi})}{1714}$$

$$T_{Hp} = \frac{(3.88 \text{ gpm})(10 \text{ psi})}{1714}$$

$$T_{hp} = 0.0226 \text{ Hp}$$

Converting to power requirements

$$P = 0.0226 \text{ Hp} * \frac{746 \text{ W}}{\text{Hp}}$$

$$P = 16.86 W$$

For P-103 and P-104 there is 18 gallons/day (0.0125 gallons/min) entering at 1 atm (14.7 psi). These pumps are solely used to transport the liquid to the next piece of equipment, so the pressure will be increased to 1.1 atm (16.2 psi). Similarly the theoretical horse power can be calculated.

$$T_{Hp} = \frac{(0.0125 \text{ gpm})(1.5 \text{ psi})}{1714}$$

$$T_{Hp} = 1.5 * 10^{-5} Hp$$

Converting to power requirements

$$P = 1.5 * 10^{-5} Hp * \frac{746 W}{Hp}$$

$$P = 0.011 W$$

Blowers (B-101/102)

For B-101 there is 1.76 mol/hr of air entering at atmospheric pressure in stream 6. The outlet will be entering the air stripper (S-101) which will create a pressure drop. In order to make sure the air is pressurized enough to pass through S-101, the pressure drop across the air stripper is calculated. There is a pressure drop of 0.1 psi drop associated with each tray in S-101 (Walas, 1988). There are 19 trays in the tower, so the pressure drop is 19 psi or 0.13 atm. Therefore, the pressure leaving B-101 must be 1.13 atm. From this data the theoretical horse power can be calculated using the following equation (Seider, 2003)

$$T_{Hp} = SCFM \left(\frac{T_1}{8130a} \right) \left[\left(\frac{P_2}{P_1} \right)^a - 1 \right]$$

Where,

$$a = \frac{k - 1}{k}$$

$$k = \frac{c_p}{c_v}$$

The k value for air is found to be 1.4 (The Engineering Tool Box, 2005). Therefore,

$$a = \frac{(1.4 - 1)}{1.4}$$

$$a = 0.286$$

The inlet temperature is 22 °C which is the annual average temperature for Prescott, Arizona (The Weather Channel, 2009). To calculate the horse power required the temperature must to be converted to °R.

$$(22 + 273) * 1.8 = 531^\circ R$$

Additionally, the molar flow rate must to be converted to SCFM. Using the ratio of 379 SCF/lbmol (Seider, 2003) the volumetric flow rate is found

$$1.76 \frac{gmol}{hr} * 379 \frac{SCF}{lbmol} * \frac{hr}{60 \text{ min}} * 2.2046 \frac{lb}{kg} * \frac{kg}{1000 \text{ g}} = 0.0245 \text{ SCFM}$$

Therefore,

$$T_{Hp} = 0.0245 \left(\frac{531}{8130 * 0.286} \right) \left[\left(\frac{1.13}{1} \right)^{0.286} - 1 \right]$$

$$T_{Hp} = 1.99 * 10^{-4} Hp$$

Converting to power requirements

$$P = 1.99 * 10^{-4} Hp * \frac{746 \text{ W}}{Hp}$$

$$P = 0.148 \text{ W}$$

B-102 has an inlet flow of 208900.93 mol/hr at 1 atm and 20 °C. The outlet stream will be passing through the granular activated carbon (GAC) tower where it will experience a 0.06 atm pressure drop (Carbtrol). To ensure that the air passes through the GAC, the air will be

compressed to 1.1 atm . This stream contains some MTBE and BTEX vapors, however since the gas is greater than 99% air the same specific heat ratio, k, will be used for this blower. Similarly to B-101, the theoretical horse power is calculated.

$$(20 + 273) * 1.8 = 527.4^{\circ} R$$

$$208900.93 \frac{gmol}{hr} * 379 \frac{SCF}{lbmol} * \frac{hr}{60 \text{ min}} * 2.2046 \frac{lb}{kg} * \frac{kg}{1000 \text{ g}} = 2909.1 \text{ SCFM}$$

$$T_{Hp} = 2909.1 \left(\frac{527.4}{8130 * 0.286} \right) \left[\left(\frac{1.1}{1} \right)^{0.286} - 1 \right]$$

$$T_{Hp} = 18.23 \text{ Hp}$$

Converting to power requirements

$$P = 18.23 \text{ Hp} * \frac{746 \text{ W}}{\text{Hp}}$$

$$P = 13599.58 \text{ W}$$

Dual Phase Extractor (P-101)

The dual phase extraction system (P-101) being used is the VMX0303K by GeoTech. This model is 20 Hp or 14920 W (GeoTech). By summing up all the power requirements for each piece of equipment, the total power requirement is found to be 30.03 kW. This power is generated using a power generator (G-101) fueled by the natural gas that is produced by the system. It is found that one SCF of natural gas has 1031 BTU in energy (Administration, 2007). It is assumed that G-101 is 80% efficient, so the power generated from one SCF of natural gas is 824.8 BTU. The amount of natural gas that will be required to power the system is calculated below.

$$30.03 \text{ kW} * \frac{\text{BTU}}{2.93 * 10^{-4} \text{ kWh}} * \frac{\text{SCF}}{824 \text{ BTU}} * \frac{\text{lbmol}}{379 \text{ SCF}} * \frac{\text{kg}}{2.2046 \text{ lb}} * 1000 \frac{\text{g}}{\text{kg}} = 148.12 \frac{\text{mol}}{\text{hr}}$$

The process produces 158.83 mols/hr of natural gas; therefore the entire system can be self-powered. The remaining 10.71 mol/hr of natural gas can be sold as a product.

9.4 Future Considerations

Lead Removal from Groundwater

Prior to 1985 lead is a common additive in gasoline as it increased the octane of the gasoline, however the Environmental Protection Agency has since regulated the use of lead in gasoline. The common use of lead in gasoline caused high concentrations of lead in the environment affecting the mental and neurological development in children (Andreottola, Dallago, & Ferrarese, 2008). EPA first proposed the phase out of lead in gasoline in 1973, and by 1986 the use of lead is cut by 90 percent (United States Environmental Protection Agency, 1996). In 1996 the Clean Air Act required complete removal of lead from gasoline, completing what EPA Administrator Carol M. Browner believed “is one of the great environmental achievements of all time” (United States Environmental Protection Agency, 1996).

In our proposed system, lead contamination is not considered as our data showed there is no lead in the groundwater. In addition, lead is no longer added to gasoline, therefore lead removal from groundwater may not pertain to every site. With these thoughts in mind, we made the assumption that lead removal from groundwater is not a concern. However, for sites that require the removal of organolead compounds there are techniques such as adsorption and in situ bioremediation that could be affective. Adsorption is a common technique for treating site/ground water, and the largest cost associated with adsorption is the adsorbent itself. A common adsorption medium includes the use of granular activated carbon (GAC), however this can be quite costly. One way of maintaining low operating costs would be to consider tobacco stems as an adsorbent.

Wei Li et. al. performed batch experiments and concluded that a contact time of 120 minutes at room temperature and a pH of 5.0 would reach equilibrium. Using a dosage of 0.8 g of adsorbent and an initial lead ion concentration of 50 mg/L, roughly 90 percent of the lead ions were removed. Li et. al. also reports this method is endothermic, spontaneous and thermodynamically favorable as with a Gibb's free energy on the order of -6.2 KJ/mol (Li, Zhang, Peng, Li, Zhang, & Guo, 2008). This method appears to be the most effective in comparison to GAC or any oxidizers, as well as being the most inexpensive process.

In situ bioremediation is also a viable option for removing organolead compounds from groundwater. By injecting additives that will stabilize heavy metals, including lead, into the groundwater and filtering the water, the lead can be collected and removed (Kaluderović Radoičić & Raičević, 2008). The only set back associated with in situ bioremediation is the cost to install the filter underground, to separate the lead. A few of the sorbents that have been studied for this application include hydroxyapatite (HAP) and natural sources of apatite from phosphate ore deposits. Experimental procedures demonstrated that HAP is able to remove 90% of the lead in the sample water, while the naturally occurring apatite removed 67% of lead (Kaluderović Radoičić & Raičević, 2008). These high removal percentages demonstrate the effectiveness of in situ bioremediation.

9.5 Meeting Logs/Web Documentation

9.5 Meeting Logs/Web Documentations

Meeting/ Phone Log

Date: 4/3/09

Members Present (name of senior design group members plus name and title of person providing information)

Kyle Heckel, Mandy Kiesl, Chris Lewis, Adam Manasse
Mentor: Mike Kazz

Summary of Information, that pertains to the report (costs, flow rates, sizes, assumptions).

MTG. @ MR. KAZZ'S OFFICE

- COST OF DPE ON SITE @ BENNETT \$55,000 (PURCHASE COST)
↳ based on value in a file @ office

- Lab data for ground H₂O samples from Bennett
(see attached)



ORANGE COAST ANALYTICAL, INC.

3002 Dow Ave., Suite 532, Tustin CA 92780 (714) 832-0064 Fax (714) 832-0067
4620 E. Elwood, Suite 4, Phoenix, AZ 85040 (480) 736-0960 Fax (480) 736-0970

LABORATORY REPORT FORM

ORANGE COAST ANALYTICAL, INC.

4620 East Elwood Street, Suite 4 Phoenix, AZ 85040

(480) 736-0960

Laboratory Certification (ADHS) No.: AZ0558, AZ0646, AZM499
Expiration Date: 2009

Laboratory Director's Name:

Mark Noorani

Client: Zelen Environmental

Laboratory Reference: ZEL AZ5614

Project Name: Bennett Bulk Plant

Project Number.:

Sample Matrix: Water

Date Sampled: 01/13/09

Date Received: 01/14/09

Date Reported: 01/26/09

Chain of Custody Received: Yes

Analytical Method: 8260B


Mark Noorani, Laboratory Director

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VOLATILE ORGANICS BY GC/MS (EPA 8260B) (continued)

Laboratory Reference #: ZEL AZ5614
 Client Project ID: Bennett Bulk Plant
 Client Project #:

Sampled: --- 01/13/09 ---
 Received: --- 01/14/09 01/14/09
 Analyzed: 01/23/09 01/23/09 01/23/09
 Reported: 01/26/09 01/26/09 01/26/09

Lab Sample #: MBMN0123081 AZ5614-001 AZ5614-002
 Client Sample #: --- Cone Tank Trip Blank
 Dilution Factor: 1 100 1

ANALYTE (cont)	CAS #	µg/L	µg/L	µg/L
1,1-Dichloropropene	563-58-6	<1.0	<100	<1.0
cis-1,3-Dichloropropene	10061-01-5	<1.0	<100	<1.0
trans-1,3-Dichloropropene	10061-02-6	<1.0	<100	<1.0
Ethylbenzene	100-41-4	<1.0	1000	<1.0
Isopropylbenzene	98-82-8	<1.0	140	<1.0
4-Isopropyltoluene	99-87-6	<1.0	210	<1.0
Methyl t-butyl ether (MTBE)	1634-04-4	<1.0	5000	<1.0
Naphthalene	91-20-3	<3.0	4200	<3.0
n-Propylbenzene	103-65-1	<1.0	490	<1.0
Styrene	100-42-5	<1.0	<100	<1.0
1,1,2,2-Tetrachloroethane	79-34-5	<1.0	<100	<1.0
Tetrachloroethene	127-18-4	<1.0	<100	<1.0
Toluene	108-88-3	<1.0	1500	<1.0
1,2,3-Trichlorobenzene	87-61-6	<1.0	<100	<1.0
1,1,1-Trichloroethane	71-55-6	<1.0	<100	<1.0
1,1,2-Trichloroethane	79-00-5	<1.0	<100	<1.0
Trichloroethene	79-01-6	<1.0	<100	<1.0
Trichlorofluoromethane	75-69-4	<2.0	<200	<2.0
1,2,3-Trichloropropane	96-18-4	<1.0	<100	<1.0
1,2,4-Trimethylbenzene	95-63-6	<1.0	1900	<1.0
1,3,5-Trimethylbenzene	108-67-8	<1.0	7300	<1.0
Vinyl chloride	75-01-4	<2.0	<200	<2.0
Xylenes, Total	1330-20-7	<2.0	6600	<2.0
Acceptable Surrogate %RC		%RC	%RC	%RC
Dibromofluoromethane	48-143%	121	137	134
Toluene-d8	68-130%	105	97	109
4-Bromofluorobenzene	64-130%	97	112	90

D2= Sample required dilution due to high concentration of target analyte.

Zelen Environmental
 ATTN: Mr. Mike Kazz
 1830 E. Broadway, #124-256
 Tucson, AZ 85719

Laboratory Reference #: ZEL AZ5614
 Client Project ID: Bennett Bulk Plant
 Client Project #:

VOLATILE ORGANICS BY GC/MS (EPA 8260B)

Sample Description: Water

Sampled:	—	01/13/09	---
Received:	—	01/14/09	01/14/09
Analyzed:	01/23/09	01/23/09	01/23/09
Reported:	01/26/09	01/26/09	01/26/09

Lab Sample #:	MBMN0123081	AZ5614-001	AZ5614-002
Client Sample #:	---	Cone Tank	Trip Blank
Dilution Factor:	1	100	1
Data Qualifier:		D2	

ANALYTE	CAS #	µg/L	µg/L	µg/L
Acetone	67-64-1	<10	<1000	<10
Benzene	71-43-2	<1.0	1900	<1.0
Bromobenzene	108-86-1	<1.0	<100	<1.0
Bromochloromethane	74-97-5	<1.0	<100	<1.0
Bromodichloromethane	75-27-4	<1.0	<100	<1.0
Bromoform	75-25-2	<1.0	<100	<1.0
Bromomethane	74-83-9	<5.0	<500	<5.0
n-Butylbenzene	104-51-8	<1.0	540	<1.0
sec-Butylbenzene	135-98-8	<1.0	140	<1.0
tert-Butylbenzene	98-06-6	<1.0	<100	<1.0
Carbon tetrachloride	56-23-5	<1.0	<100	<1.0
Chlorobenzene	108-90-7	<1.0	<100	<1.0
Chlorodibromomethane	124-48-1	<1.0	<100	<1.0
Chloroethane	75-00-3	<5.0	<500	<5.0
Chloroform	67-66-3	<1.0	<100	<1.0
Chloromethane	74-87-3	<5.0	<500	<5.0
2-Chlorotoluene	95-49-8	<1.0	<100	<1.0
4-Chlorotoluene	106-43-4	<1.0	<100	<1.0
1,2-Dibromoethane	106-93-4	<1.0	<100	<1.0
1,2-Dichlorobenzene	95-50-1	<1.0	<100	<1.0
1,3-Dichlorobenzene	541-73-1	<1.0	<100	<1.0
1,4-Dichlorobenzene	106-46-7	<1.0	<100	<1.0
1,1-Dichloroethane	75-34-3	<1.0	<100	<1.0
1,2-Dichloroethane	107-06-2	<1.0	<100	<1.0
1,1-Dichloroethene	75-35-4	<1.0	<100	<1.0
cis-1,2-Dichloroethene	156-59-2	<1.0	<100	<1.0
trans-1,2-Dichloroethene	156-60-5	<1.0	<100	<1.0
Dichlorodifluoromethane	75-71-8	<2.0	<200	<2.0
1,2-Dichloropropane	78-87-5	<1.0	<100	<1.0
1,3-Dichloropropane	142-28-9	<1.0	<100	<1.0
2,2-Dichloropropane	594-20-7	<1.0	<100	<1.0

QA/QC Report
for
Volatile Organic Compounds (EPA 8260B)
Reporting Units: ppb

1. Matrix Spike (MS) / Matrix Spike Duplicate (MSD)

Date of Analysis : 01/23/09

Laboratory Sample No : AZ5609-001

Laboratory Reference No : ZEL AZ5614

ANALYTE	R1	SP CONC	MS	MSD	% MS	% MSD	RPD	ACP%	ACP RPD
1,1-Dichloroethene	0.0	50	61	60	122	120	2	41-130	23
Benzene	0.0	50	52	53	104	106	2	70-130	15
Trichloroethene	0.0	50	50	50	100	100	0	70-130	15
Toluene	0.0	50	53	53	106	106	0	70-137	17
Chlorobenzene	0.0	50	51	51	102	102	0	70-130	16

Definition of Terms :

R1	Result of Laboratory Sample Number
SP CONC	Spike Concentration Added to Sample
MS	Matrix Spike Results
MSD	Matrix Spike Duplicate Results
% MS	Percent Recovery of MS: $\{(MS-R1) / SP\} \times 100$
% MSD	Percent Recovery of MSD: $\{(MSD-R1) / SP\} \times 100$
RPD	Relative Percent Difference: $\{(MS-MSD) / (MS+MSD)\} \times 100 \times 2$
ACP%	Acceptable Range of Percent for MS/MSD
ACP RPD	Acceptable Relative Percent Difference

2. Laboratory Control Sample

Date of Analysis : 01/23/09

Laboratory Sample No : MN0123091

ANALYTE	SP CONC	RESULTS	% RECOVERY	ACCEPTABLE %
1,1-Dichloroethene	50	46	92	47-130
Benzene	50	52	104	70-130
Trichloroethene	50	51	102	70-130
Toluene	50	53	106	70-130
Chlorobenzene	50	51	102	70-130

Meeting/ Phone Log

Date: 4/24/09

Members Present (name of senior design group members plus name and title of person providing information)

Kyle Hecker, Mandy Kiesl, Chris Lewis, Adam Manasse

Summary of Information, that pertains to the report (costs, flow rates, sizes, assumptions).

Discussing economic options:

How much (if any) does Zeien charge owner of Bennett site for remediation services?

*\$10,000/month => counts as sales!!!

Safety:

Electrical circuits/controls explosion proof

- NFPA
- IBC
- '06 Fire code

Date: Wed, 1 Apr 2009 09:13:28 -0400 [Wednesday April 01, 2009 06:13:28 AM MST]

From: Charles H Powell <charles.h.powell@basf.com>

To: akiesl@email.arizona.edu

Subject: BASF Methanation Catalyst

Part(s):  2 Datasheet MET Final 2.pdf [application/octet-stream] 56 KB

 1 unnamed [text/html] 2.28 KB

Good morning Amanda,

I am the Account Manager for the Americas for BASF Syngas Catalysts. BASF has several different methanation catalysts, however, I assume the one you are referring to is our SG 9701, data sheet attached. This catalyst is designed for removal of low levels of CO and CO2 from dry process streams in ammonia and hydrogen plants. Typically the catalyst does not need regeneration. I would need further information on your projected use of the catalyst to advise what type of regeneration procedure BASF would recommend. Also, after understanding your application, we may have an alternative catalyst that would better meet your process requirements. The current price, CPT Singapore Port, is \$15.00 / kg. Each drum contains 200 kg and the average bulk density of the catalyst is 1.1 kg / liter. Delivery via standard over seas shipping from Singapore is 4 to 6 weeks after order placement and entry into the BASF SAP order entry system.

BASF looks forward to working with you to achieve optimum results from your project.

Best regards,

Charles H. Powell, III
Syngas Account Manager – Americas

Office: 732-205-5653
Mobile: 848-333-0055
Fax: 732-591-1428

Email: charles.h.powell@BASF.com

BASF Catalysts LLC
25 Middlesex-Essex Turnpike
Iselin, NJ 08830

BASF – The Chemical Company

Date: Fri, 27 Feb 2009 15:58:47 -0500 [Friday February 27, 2009 01:58:47 PM MST]

From: Sharon Pappin <s.pappin@carbtrol.com>

To: amanasse@email.arizona.edu

Subject: Carbtrol Corporation

Part(s):  2 Zelen [application/octet-stream] 10 KB

 3 Carbon Systems Table.pdf [application/pdf] 931 KB

 1 unnamed [text/html] 0.55 KB

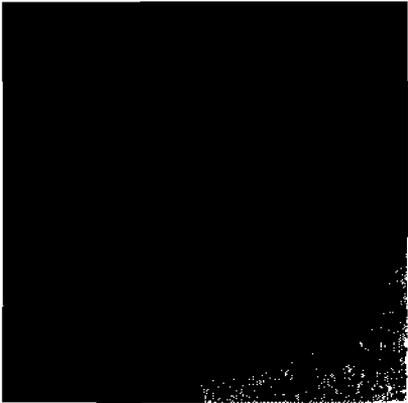
Adam:

Please see the attached carbon usage estimate (open in excel workbook), showing you will utilize an estimated 95 pound of carbon per day. Is this a long term or short term project? Please advise as I can size and price a unit appropriate to your needs.

Please feel free to contact me with questions or comments or if you need further assistance.

--
Thank You,
Sharon Pappin
CARBTROL Corporation
800-242-1150, ext.3031
s.pappin@carbtrol.com

[Home](#) << [Tanks & Accessories](#) << [PVC Tanks](#) << [Heavy Duty PVC Fabricated Tank...](#)



PVC Rectangular Tank(Two Support Flanges)

Rectangular and cylindrical self supporting tanks of welded construction for many industries such as plating,chemical processing,pulp and paper.Operating temperatures to 140 deg.F.Optional covers sold separately.Rectangular tank dimensions are ID and cylindrical tank dimensions are OD.Wall thickness is 1/4 inch,not UV resistant and not intended for fuels or oils.Custom tanks can be made, send for possible quotation.

Product Description

PVC Rectangular Tank(Two Support Flanges)
Qty 2: 5% off. Qty 4: 10% off. Qty 12: 15% off.

Product Pricing	
Price	\$771.97

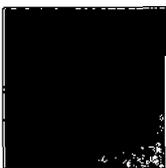
[Print This Page](#)

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UNIT
F-102

Product Information

Item No.: 9069
In Stock: Drop Ship
Capacity: 66 Gallon
Sold By: Each
Size: 24 x 18 x 36
Catalog Page Number: P-61
Manufacturer: Not Listed



Product Group This Item Belongs To: Heavy Duty PVC Fabricated Tanks

Rectangular and cylindrical self supporting tanks of welded construction for many industries such as plating,chemical processing,pulp and paper.Operating temperatures to 140 deg.F.Optional covers sold...

[See More Heavy Duty PVC Fabricated Tanks](#)

1390 Neubrecht Rd. Lima, Ohio 45801-3196 Phone: 1-800-809-4217 Fax: 1-800-854-5498

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Catalog > Home Standby Generators > 76001NG

Briggs & Stratton IntelliGEN Series 42kW Natural Gas Liquid Cooled Standby Generator
Part#: 76001NG



Quantity: 1 ~~MSRP \$15,149.00~~
 MSRP \$15,149.00
 Our Price: \$14,999.00
 You Save: \$150.00
 See this item for less?

[Installation Drawing](#)
[Operations Manual](#)
[Product Specs](#)

Unit
 G-101

Click to enlarge
Ships Factory Direct 4-5 Weeks
Free Shipping

Product Details:

Briggs & Stratton IntelliGEN Series 42kW Natural Gas Liquid Cooled Standby Generator Model 76001NG
FEATURES:

- GM Vortec Liquid Cooled Engine, GM Vortec engines provide sophisticated yet extremely durable industrial grade power
- Sound Shield Technology, engineered sound passages, an automotive-style exhaust system and acoustic foam dampeners ensure ultra-quiet, neighbor-friendly operation.
- Patented Central Air Technology, The AC power Control feature allows you to manage the high power demands of central air conditioners, in addition to essential items while protecting against AC overload.
- Coast Guard All-Climate Enclosure, Perfect for coastal climates, this premium enclosure features rounded corners, and an automotive rust proofing paint process for years of rust resistant 4 season protection. Backed by a 3 year enclosure warranty.
- Remote System Status, Transmits advanced diagnostic reporting on 8 different system functions (twice the protection of the leading competitor) to a safe location inside your home. Available on automatic units only.
- Overload Protection, When multiple high-wattage items attempt to start at the same time, the power demand can cause a generator to overload and shutdown. The exclusive AC Power Control Module's advanced computer controls automatically manage 2 central air conditioners while safeguarding the generator against power overload.
- Turnkey Extra includes a battery charger
- 4 Year Residential, 2 Year Commercial Warranty

Product Specifications:

Running Watts	42,000
Surge Watts	53,000
Engine Brand	GM
Engine Series	Vortec
Liters	5.0
Cylinders	V-8
Operation	Fully Automatic
Voltage (V)	120/240
Amps	350/220.83
Alternator Type	Brushless
Full Pressure Lubrication	Yes
Fuel Type	Natural Gas
Fuel Consumption Cu. Ft. per Hr. @ 1/2 Load	178
Weekly Exerciser	Yes
Sound Rating	65 dbA
Sound Attenuation	Sound Shield Technology
Battery Charger	Yes
Overcrank Protection	Yes
Hour Meter	Yes
Diagnostic Alerts w/Remote System Status	Low Oil Shutdown, Fail to Start, Low Frequency, Engine Overspeed, Low Voltage, Low BATTERY Voltage,
Dimensions (L x W x H) Inches	98.5 x 39.5 x 44
Weight	2001 lbs
Warranty	4 Year or 3000 Hour Limited
Battery Required	Not Included
Automatic Transfer Switch	Sold Separately

Related Products:

Part#	Name	Price	Buy Now
71036	Briggs & Stratton 200 Amp ATS w/ACCM for Liquid Cooled Generators Only	\$1,249.00	Purchase

**** Use Google Checkout and Save 1.5% on Your Total Order!! ****
No Sales Tax if Shipped Outside Of Arizona!

Here's a list of the most commonly used freight companies.



TOP THE DUES WAS GLEET DONE

printed April 26, 2009



UNITS
P-102,103,104

Pump,Centrifugal,3/4hp

Centrifugal Pump, High Head, HP 3/4, Phase 3, Volts 230/460, Amps 2.3/1.2, Motor Enclosure TEFC, Impeller Type Closed, Impeller Material 316 SS, Outlet 1 In, Inlet 1 1/4 In, Max Pressure 35.5 PSI, 316 SS, Seals Viton, Shaft Material SS, Bearings Ball, For Use With Nonflammable and Nonabrasive Liquids, Port Rotation 4 Position Increments, Replaces 4XZ02, Max Temp 200 F, GPM of Water @40 Ft of Head 32, @70 Ft of Head 15, Max Head 82 Ft

Grainger Item #	4TU40
Price (ea.)	\$150.25
Brand	DAYTON
Mfr. Model #	4TU40
Ship Qty.	1
Sell Qty. (Will-Call)	1
Ship Weight (lbs.)	43.45
Usually Ships	While Stock Lasts
Catalog Page No.	N/A

Price shown may not reflect your price. Log in or register.

Additional Info

There is currently no additional information for this item.

Tech Specs

Item: Centrifugal Pump
 Type: High Head
 HP: 3/4
 Phase: 3
 Voltage: 230/460
 Amps: 3.2/1.6
 Hz: 60
 Inlet (In.): 1 1/4
 Outlet (In.): 1
 Motor Enclosure: TEFC
 NEMA/IEC Frame: 56J with Base
 Motor RPM: 3450
 Motor Type: TEFC
 Wetted Materials: 316 Stainless Steel, Viton, Carbon and Ceramic
 Impeller Material: 316 SS
 Housing Material: 316 SS
 Volute Material: 316 SS
 Shaft Material: SS
 Seal Type: Mechanical Carbon on Ceramic
 Seal Material: Viton
 Seal Application: Long life When Not Run Dry or Pumping Abrasives
 Max. Liquid Temp. (F): 200
 GPM of Water @ 10 Ft. of Head: 42
 GPM of Water @ 20 Ft. of Head: 39

Optional Accessories

There are currently no optional accessories for this item.

Alternate Products

Pump,Centrifugal,3/4hp



Item #: 4UB58
 Brand: GOULD'S
 Usually Ships: Today
 Price (ea): \$624.00

Pump,Centrifugal,1 HP



Item #: 4XZ33
 Brand: DAYTON
 Usually Ships: Today
 Price (ea): \$583.50

Repair Parts

Repair Parts Information is available for this item.

GPM of Water @ 30 Ft. of Head: 36
GPM of Water @ 40 Ft. of Head: 32
GPM of Water @ 50 Ft. of Head: 28
GPM of Water @ 60 Ft. of Head: 23
GPM of Water @ 70 Ft. of Head: 15
GPM of Water @ 80 Ft. of Head: 2
Max. Head (Ft.): 82
Minimum GPM @ Head (Ft.): 2 @ 80
Maximum GPM @ Head (Ft.): 42 @ 10
Best Efficiency GPM @ Head (Ft.): 29 @ 47
Best Efficiency Range GPM @ Head : 14-36 GPM @ 70-30 Ft.
Max. Specific Gravity: 1.1
Max. Case Pressure (PSI): 200
Max. Fluid Viscosity: 100 SSU
Inlet Pressure (PSI): 200
Max. Pressure (PSI): 35.4978354978355
Impeller Type: Closed
Bearing Type: Ball
Port Rotation: 4 Position Increments
Drain Plug: Easily Accessible
Warranty Length: 1 Year
For Use With: Nonflammable and Nonabrasive Liquids
Length (In.): 14.25
Width (In.): 10.00
Height (In.): 8.50

Notes & Restrictions

There are currently no notes or restrictions for this item.

MSDS

This item does not require a Material Safety Data Sheet (MSDS).

Required Accessories

There are currently no required accessories for this item.

To order this item online visit: www.northerntool.com and enter Item# 109582 | [Print this page](#) | [Back to product page](#)



Fuel Storage Tank — 550 Gallon, Model# 500SK

Unit
V-103

Description

Heavy duty steel construction. Includes float gauge and locking vented fill cap. Check local ordinances regarding possession and use. U.S.A.

Key Specs

- Tank Size (gal.): 550
- Tank Dimensions L x W x H (in.): N/A
- Box Dimensions L x W x H (in.): N/A
- Overall Dimensions L x W x H (in.): N/A
- Dimensions O.D. x L (in.): 48 x 72
- DOT Approved: No
- Transfer Tank: No
- Storage Tank: Yes

General Specs

- Mount: skid
- Single wall steel: 12 gauge

Ships from Manufacturer

Delivery Time: 34 - 39 Business Days

Ship Wt. 480.0 lbs.

Item# 109582

Only \$1099.99

Available Accessories For: Fuel Storage Tank — 550 Gallon, Model# 500SK

- [1096004](#) Tuthill Diesel Fuel Transfer Pump with Suction Pipe — 12 Volt, 10 GPM, Model# FR1616
- [1096002](#) Tuthill Diesel Fuel Transfer Pump with Hoses — 12 Volt, 10 GPM, Model# FR1614
- [1096005](#) Tuthill Diesel Fuel Transfer Pump — 115 Volt AC, 18 GPM, Model# FR1618
- [1096006](#) Tuthill Cast Iron Diesel Fuel Transfer Pump — 115 Volt AC 18 GPM, Model# FR1620
- [1096007](#) Tuthill Wall Mount Cast Iron Diesel Fuel Transfer Pump — 115 Volt AC, 18 GPM, Model# FR1622
- [170305](#) Goldenrod Standard Fuel Filter & Cap — 3/4in. Fittings, Model# 495-3/4
- [10930](#) GPI Digital Flow Meter
- [109589](#) Tuthill Mechanical Fuel Meter — 3/4in., Model# 807CMK
- [8888](#) Class B & C Fire Extinguisher
- [174689](#) AO Safety Virtua Safety Eyewear, Indoor/Outdoor Lens