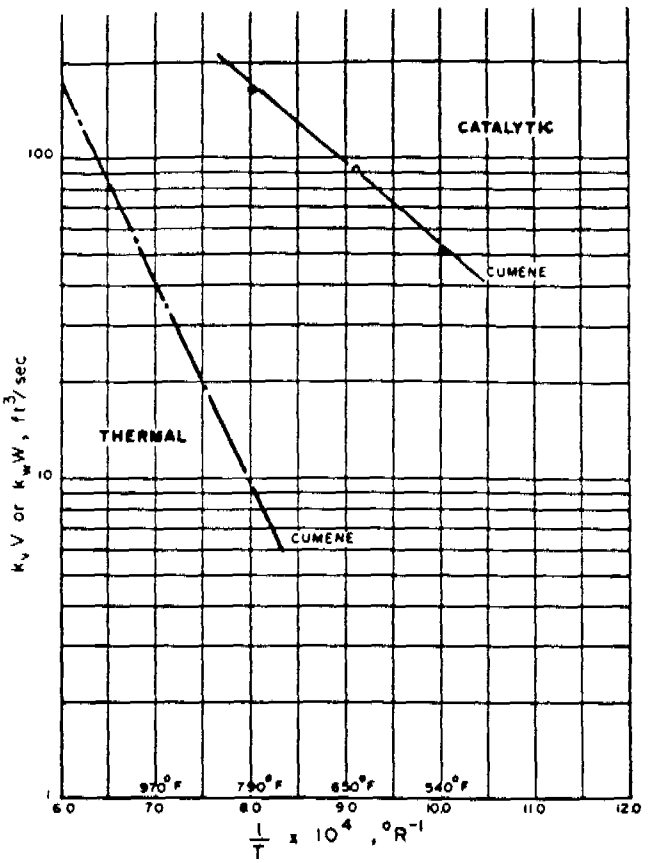

Figure 1. Plot of reaction rate constant vs. reciprocal temperature.

AIR POLLUTION CONTROL:

Emission Control Via Fluidized Bed Oxidation

A catalytic incineration system has been developed for treating organic emissions that permits lower temperatures and fuel costs than do thermal or flame incineration techniques.

L. C. Hardison and E. J. Dowd, Air Resources, Inc., Palatine, Ill.

The control of organic emissions from industrial processes was first accomplished by catalytic incineration in the late 1940s. (*1*) Fixed-element catalysts prepared by depositing precious metals on heat resisting alloy ribbons were prepared according to a patent granted to Russell Suter and Richard Ruff, who founded Catalytic Combustion Co. (2)

In general, the catalysts were applied only in circumstances where an extremely serious odor problem was associated with an industrial process, or where the concentration of organic solvents in the gases to be discharged to the air was high enough that they could be burned to utilize the heat in the process. Successful applications included treatments of effluents from magnet wire coating, metal lithographing ovens and the manufacturer of phthalic anhydride. (*1*)

Fixed bed catalysts manufactured on ceramic support materials were introduced by Oxy-catalyst, Inc., and by Englehardt Industries, Inc. Catalytic incineration utilizing precious metal catalysts supported on both metal and ceramic supports was applied successfully in the three main areas of stationary organic emission 1) food processing, 2) coating and other solvent handling processes, and 3) chemical manufacture.

While many successful applications were found for catalytic oxidation of organic materials, many potential applications involving serious odor, eye irritation, or visible organic emission problems produced unsatisfactory results. This led to installations that either did not perform the emission control function for which they were intended, or ones which required frequent and costly replacement of the catalyst elements and interruption of the process operation. (*3*) Some of these failures came about due to fouling of the catalyst surface. Others occurred because materials such as halogens in the gas stream interfered with or suppressed the activity of the catalyst, or because many elements reacted with the precious metals and rendered them permanently inactive. (*4*) Finally, all catalysts eventually deteriorate by aging or thermal processes. (*5*)

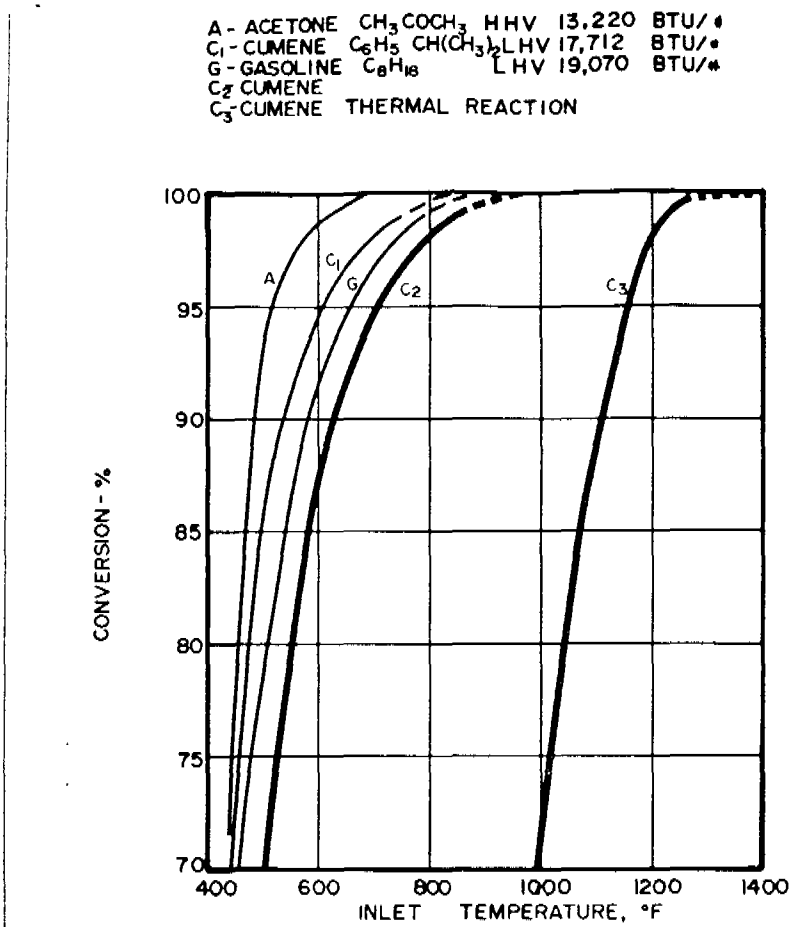

Figure 2. Calculated conversion efficiencies for several organic materials.

As the requirements for odor control, and later for control of reactive hydrocarbon emissions, became more stringent, the uncertainty as to whether or not a catalytic unit would be suitable became intolerable. A more reliable method of incinerating organic materials was necessary, and this need was fulfilled during the 1960s with the development of thermal incineration.

Thermal incinerators provided a higher degree of assurance that the process would oxidize the organic material, but required a higher fuel cost because of the considerably higher temperatures necessary. The high temperatures entailed serious mechanical design problems as a result of operating in the temperature range in which metal creep takes place.

Although catalytic incineration continued to be used for some of the more favorable applications, it was virtually replaced by thermal incineration for metal lithographing, phthalic anhydride manufacture, metal strip coating, and many other common industrial applications.

The inherent advantages of catalytic incineration with respect to fuel cost and materials of construction led Air Resources, Inc. and the Harshaw Chemical Co. to undertake a joint development program aimed at overcoming the disadvantages of catalysts. This work was initiated in 1972, and successfully completed with installation of the first commercial unit in 1975.

Theory

Organic materials can ordinarily be burned if they are mixed with air to provide an oxygen content above some minimum value in the 10 to 15% range, have a hydrocarbon concentration above the lower combustible limit, and are heated above a spontaneous ignition temperature. The resulting flame combustion can produce substantially complete oxidation of the combustible mixture. The lower combustible limit is the concentration of an organic material that will produce temperatures high enough to sustain flame reactions which depend on forming extremely reactive free radicals. Flame combustion or flame incineration is often used for the abatement of organic emissions when the organic materials are present at high concentration such as might be produced by a closed chemical reactor. (5)

Oxidation of organic materials also takes place when the concentrations are well below the lower combustible limit, provided that the temperature is high enough and the gases are maintained at the temperature long enough. This slow thermal reaction is the basis for thermal incineration of organic contaminants, and there appears to be no lower limit for either the concentration of the combustible material, or for the concentration of oxygen. (6)

For combustion of simple hydrocarbons, the thermal oxidation reactions appear to follow classical first order reaction kinetics sufficiently closely that practical incinerator designs can be established by application of the theory The reaction:

$$C_xH_y + (x + y/2)O_2 \rightarrow xCO_2 + (y/2)H_2O \quad (1)$$

can be represented by the rate equation:

$$R = (dC/dt) = -kC \quad (2)$$

where: C = organic concentration, cu. ft./cu. ft.
R = rate of change of contaminant concentration, cu. ft./cu. ft.-sec.
t = time, sec.
k = reaction rate constant, sec.$^{-1}$

This differential equation can be solved to yield an expression for the efficiency of combustion as a function of time at a given temperature as follows:

$$E = 100 \times (1 - C/C_0) = 100 \times (1 - e^{-kt}) \quad (3)$$

where: E = conversion efficiency, %
C_0 = initial organic concentration, cu. ft./cu. ft.

The rate constant, k, must be determined experimentally for the burning of various organic materials. However, the pattern of variation with temperature is predictable from kinetic theory, and follows the Arrhenius equation:

$$k = Ae^{-(\Delta E/RT)} \quad (4)$$

where: A = collision coefficient, sec.
E = energy of activation, B.t.u./lb. mol.
R = universal gas constant, B.t.u./lb. mol. °R.
T = absolute temperature, °R.

Thus, in order to establish a relatively complete pattern of behavior for a particular organic material with regard to thermal incineration, it is only necessary to make two careful determinations of the efficiency of conversion at known temperatures and residence times. The value of the reaction rate constant, k, can then be calculated as a function of temperature. The relationship between conversion and combustion efficiency for any combination of residence time and temperature can then be predicted. Figure 1 illustrates a plot of the reaction rate constant as a function of temperature for several organic materials, and Figure 2 is a plot of calculated conversion efficiencies for a one second residence time.

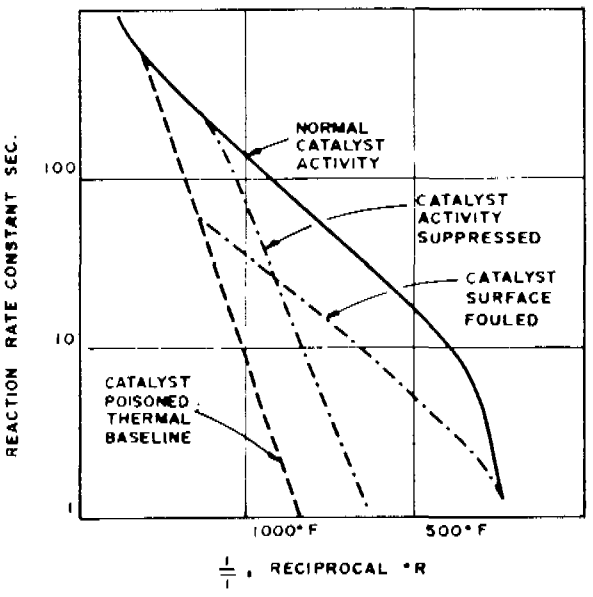

Figure 3. Effect of fouling materials, poisons, and suppressants on catalyst activity.

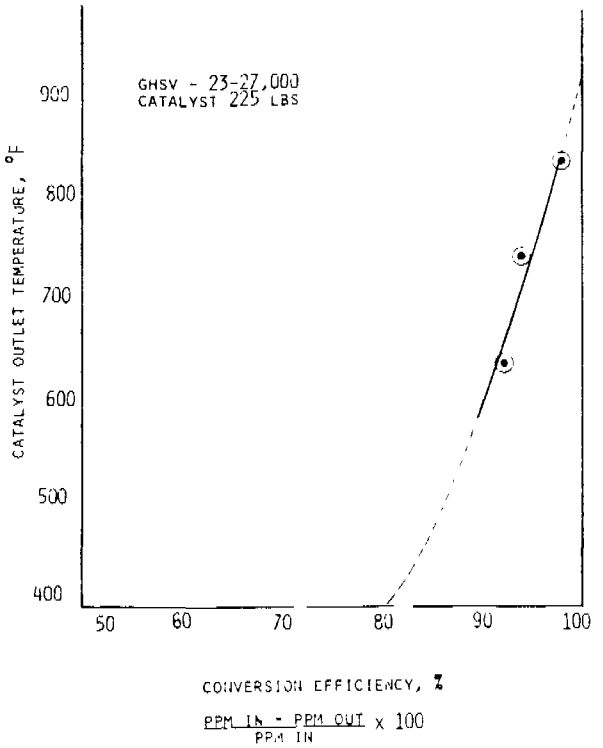

Figure 4. Measured conversion efficiencies.

How the reaction rate is increased

Catalysts increase the rate of reaction by adsorbing gas molecules on catalytically active sites. Most metals have a film of adsorbed gases on the surface that, at room temperature, completely covers the exposed metal. This film is quite stable because of the strength with which the molecules of metal hold the adsorbed gases. At elevated temperatures, gases are desorbed from the metal surface leaving a partial film, and it is in this that the catalytic oxidation reactions take place.

The catalyst may function simply to bring about a higher concentration of reactive materials at the surface than is present in the bulk gas phase, which has the effect of increasing the collision constant, A, in the Arrhenius equation. Or, the catalyst may modify a molecule of adsorbed gas by adding or removing an electron or by physically opening a bond. This has the effect of decreasing the activation energy, ΔE, in the Arrhenius equation. In either circumstance, it is necessary for the reactive materials to reach the active catalyst surface by diffusion through the gas phase, and for the reaction products to leave the surface. Materials that mechanically coat the catalyst surfaces with films that prevent or slow down the diffusion of reactants are referred to as fouling agents. Examples of such materials are:

1. Ordinary airborne dust and dirt.
2. Silicon dioxide ash remaining when silicone compounds are oxidized.
3. Organic liquids or tars condensed at temperatures too low for catalytic oxidation, and retained on the surface as the catalyst is heated.

Most oxidation reactions are extremely rapid and take place principally on the outer surfaces of the catalyst support materials. Catalysts with a very great area of in-

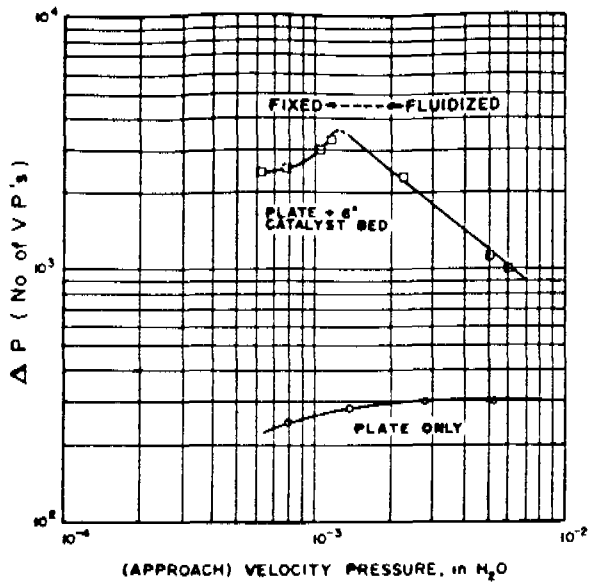

Figure 5. Pressure drop curve for the Econ-Abator catalyst.

ternal micro-surface do not show much advantage over low-surface catalysts. For this reason, the fouling of the external surface of porous catalysts is important, whereas fouling of internal pores is not. It is often possible to remove the fouling materials from the external surface of fixed element catalysts and restore the catalyst to normal activity by procedures such as washing, mechanical brushing, or blowing with high pressure air or steam. However, catalysts consisting of precious metals coated on metal or ceramic supports must be treated very gently to avoid removing this coating when the fouling materials are removed.

Halogens, sulfur dioxide, nitrogen dioxide, and numerous other materials act as catalyst suppressants for precious metal oxidation catalysts. These compounds tend to adsorb strongly on the catalytic surface, and prevent the reactants from finding unoccupied sites. The strength of adsorption is ordinarily such that the suppressant materials will gradually be stripped off after there is no longer a concentration of suppressant materials in the gas stream passing through the catalyst. (*1*) For this reason, the catalyst appears to lose activity during periods when halogens, halogenated hydrocarbons, etc., are present in the gas stream, but regains normal activity slowly after the suppressant materials are removed.

Catalyst poisons are materials such as metallic lead, zinc, and arsenic that react permanently with the precious metals and make the surface unavailable for catalytic reactions. Poisoning by heavy metals ordinarily destroys the activity of precious metal catalysts. They must be removed from the incineration unit and discarded, or subjected to complex stripping and re-refining of the precious metals. (*3*)

Figure 3 illustrates the effects of fouling materials, suppressants, and poisons on the activity of precious metal catalysts.

Development program

Early in 1972, Air Resources recognized that significant advantages in fuel economy and in simplicity of mechanical design could be achieved if the deficiencies in catalytic incineration systems could be overcome. It was reasoned that the ultimate problem, that of catalyst deactivation, could not always be entirely eliminated, but that continuous replacement of a portion of the catalyst bed during normal operation would allow continuing operation at high efficiency even in the presence of poisoning agents. This criterion could be satisfied by a fluidized bed unit into which catalyst could be added, and from which withdrawals could be made during operation.

The use of a fluidized bed would also provide protection against fouling because the self-abrading action of the catalyst particles in the fluidized bed would keep the surfaces clean. In fact, the abrasion of the catalyst posed a major problem in the development program, because minute amounts of wear would result in significant rates of catalyst loss and, potentially, in a particulate emission problem. Therefore the catalyst would necessarily be very hard and abrasion resistant.

One of the requirements established for the proposed development program was to produce a catalyst with an abrasion resistance so high that the catalyst losses would amount to less than 1 lb./million std. cu. ft. of gas treated.

Another goal involved the development of a practical non-precious metal catalyst of sufficient activity to compete favorably with the best commercially available precious metal catalysts, but with a price low enough that poisoned or deactivated catalyst might be discarded without costly and cumbersome precious metal recovery procedures.

Air Resources, Inc. and Harshaw Chemical Co. started working on the development program simultaneously. The Harshaw work was carried out in a bench-scale pilot plant in Cleveland at the company's research and development laboratories. Catalyst performance with respect to oxidation effectiveness, resistance to poisons and suppressants, etc., was measured in other laboratory equipment.

Air Resources undertook the development of a suitable fluidized bed apparatus that could be operated without skilled personnel and which would adequately demonstrate the performance of the catalyst with respect to efficiency and abrasion resistance under field conditions. Extensive plastic model studies and trial component designs were made prior to completion of the first demonstration unit.

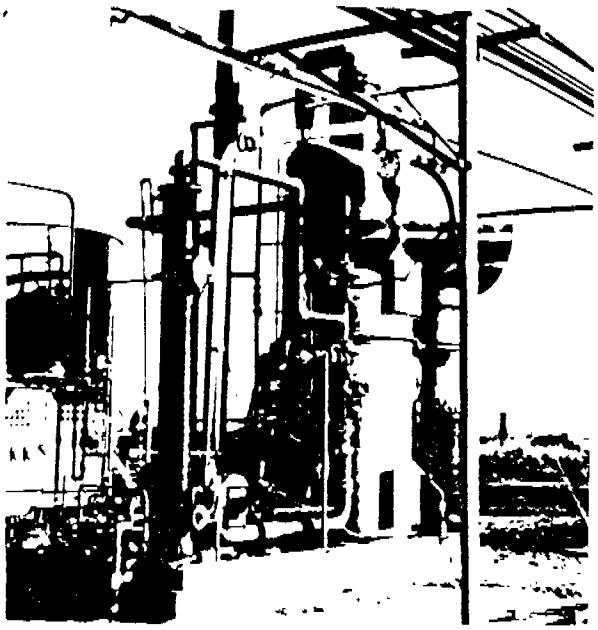

Figure 6. A commercial scale Econ-Abator unit.

A pilot unit was constructed with a nominal reactor diameter of 3 ft., and a gas flow capacity of 950 std. cu. ft./min. Catalyst was supported on a specially designed grid that insures good flow distribution over a wide range of flow rates. The bed depth was planned at 6 in., but more or less catalyst could be used if desired.

Initial operation of the pilot unit was conducted on the emissions from a series of carbon baking furnaces operated by Great Lakes Carbon Corp. at Morganton, N.C. This operation demonstrated satisfactory catalyst activity and catalyst losses below the initial specification of 1 lb./million std. cu. ft. Particulate carbon produced by the furnaces interfered with the production of a color-free discharge and rendered the measurement of minute amounts of catalyst loss difficult. It was possible to establish that the original criteria had all been met and that the fluidized bed was not restricted to operation on clean gas streams.

Phenol plant performance

Subsequent pilot operations were carried out on a phenol production plant operated by Clark Chemical Corp. at Blue Island, Ill. These tests demonstrated good catalyst activity on a variety of organic compounds, as illustrated in Figure 4. Further it was established that the catalyst deactivation rate was very low in this application, and no new catalyst was added to the reactor during an extensive testing program. In addition to activity, the pressure drop characteristics of the grid-catalyst bed combination were established as shown in Figure 5.

It was also demonstrated that the pilot unit could be operated successfully as a thermal incinerator. In fact, the results were superior to conventional thermal incineration units due to improved residence time, velocity distribution, and temperature uniformity. The catalyst support grid proved to be an excellent gas distribution device. Residence time at operating temperatures exceeded 1 sec.

A final field test was conducted at the Fleischmann Div. of Standard Brands, Inc., in Baltimore, Md. This application required oxidation of ethyl alcohol, acetic acid, and related chemicals produced by a vinegar plant. Extensive testing was carried out to demonstrate high efficiencies of destruction of the organic compounds. Odor panel testing using a syringe dilution method demonstrated that equivalent reductions in odor level were being achieved. Dispersion modeling was used to substantiate that odor complaints should be eliminated by treatment of the entire gas discharge.

The development program in the laboratory continued during these early field tests, and revealed several startling properties of the catalysts. The first of these involved the ability to oxidize halogenated organics without loss of catalyst activity. The Econ-Acat* catalyst is immune to halogen suppression. Also, tests with gas streams containing phosphates suggest that the catalyst is highly resistant to this form of poisoning.

The initial development phase has been completed, and the results are embodied in a commercial line of incinerators, marketed by Air Resources, Inc. under the trade name Econ-Abator.* The catalyst is supplied by Harshaw Chemical Co. under the trade name Econ-Acat,* by ARI International.

Two commercial units have been designed and placed in operation. One served temporarily a barge loading operation in which noxious organic materials including tertiary butyl mercaptan were released. The second unit, shown in Figure 6, is destroying chlorinated organics produced in a chemical manufacturing operation and includes provisions for by-product HCL recovery. #

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L. C. Hardison, vice president of Air Resources, Inc., holds a B.S. in mechanical engineering from the Illinois Institute of Technology. Prior to co-founding Air Resources, Inc., he worked for Universal Oil Products for 17 years. He has been involved with a number of diverse projects such as the reduction of ice fog from internal combustion engines, the design of flue gas cleaning systems for utility power stations, and the development of improved techniques for particulate collection by electrostatic precipitation, wet scrubbing, and filtration.

E. J. Dowd, general manager of ARI International, earned his B.S.Ch.E. from the Univ. of Detroit. Before joining Air Resources as chief engineer he worked for Universal Oil Products for 12 years, which he left as chief engineer of its Air Correction Div. Dowd has had extensive experience in the design of systems incorporating a variety of air pollution control equipment, particularly in the field of incineration. In addition, he has a broad background in petroleum refinery technology and in automotive fuel production.

AIR POLLUTION CONTROL:

Hydrocarbon Emission Reduction Systems

Here's a look at the design and operation of three incineration, four condensation, and two absorption systems for controlling emissions.

R. D. Pruessner and L. D. Broz, Petro-Tex Chemical Corp., Houston, Tex.

Petro-Tex Chemical Corp., located in Houston, Texas, in the midst of the highly industrialized ship channel complex, is a major producer of butadiene, butene, maleic anhydride, and neoprene rubber. Most of these products originate from the basic raw material, butane, a saturated paraffinic hydrocarbon containing four carbon atoms. Butane is dehydrogenated by the use of air to produce butene and butadiene, which is chlorinated and polymerized to make neoprene rubber.

Because hydrocarbons are the major feed, products, and byproducts of Petro-Tex, we have designed and operated a variety of air emission control devices. The design and operation of 10 such systems, three incineration, five condensation, and two absorption, are discussed in this article.

Incineration systems

Incineration treatment of waste air from the three air oxidation processes provides a method to achieve high removals of hydrocarbon contamination. Contamination level ranges from 150 parts/million in a 1 million lb./hr. gas flow to 4,000 parts/million in a 235,000 lb./hr. gas flow. Treatment of these large flows at high temperatures requires large volumes of natural gas; therefore, energy recovery is an integral part of each incinerator. The basic information on the three incinerators is summarized in Table 1.

The incinerator, Figure 1, installed on the maleic anhydride process burns a waste gas stream of 220,000 lb./hr., containing 2,500 parts/million hydrocarbon and 1.8% carbon monoxide. Waste gas is essentially air used to oxidize benzene to maleic anhydride through a catalytic production reactor. The maleic anhydride is then scrubbed in water. The water-scrubbed waste gas is fed to the incinerator. Because the waste gas is corrosive, corrosion-resistant 316 stainless steel is used up to the fires in the incinerator. A 27,000 gal. knock-out pot is installed ahead of the incinerator to catch large liquid carryover from the water

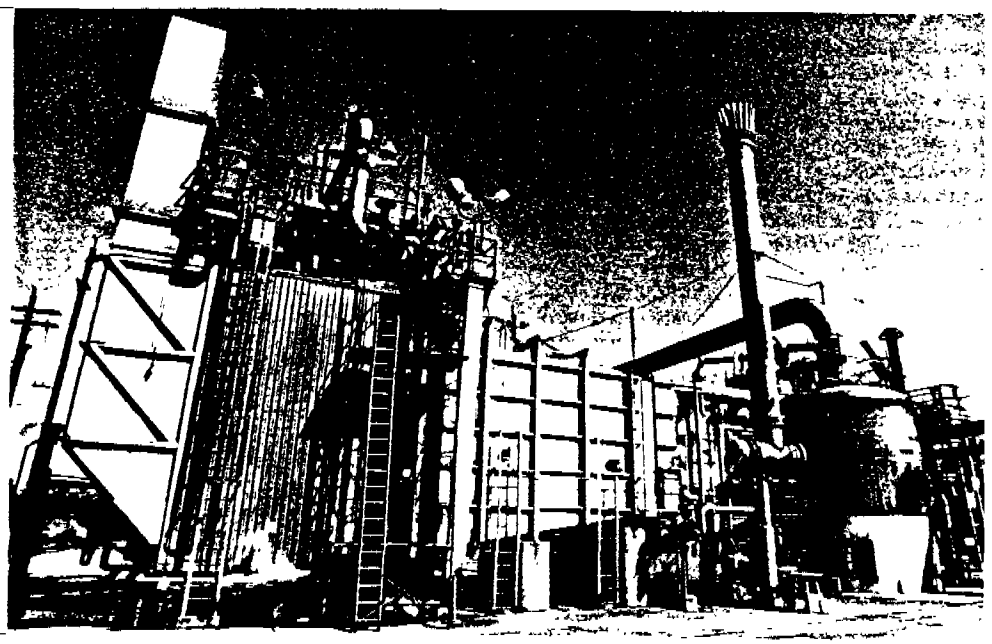

Figure 1. Maleic anhydride waste gas incinerator.

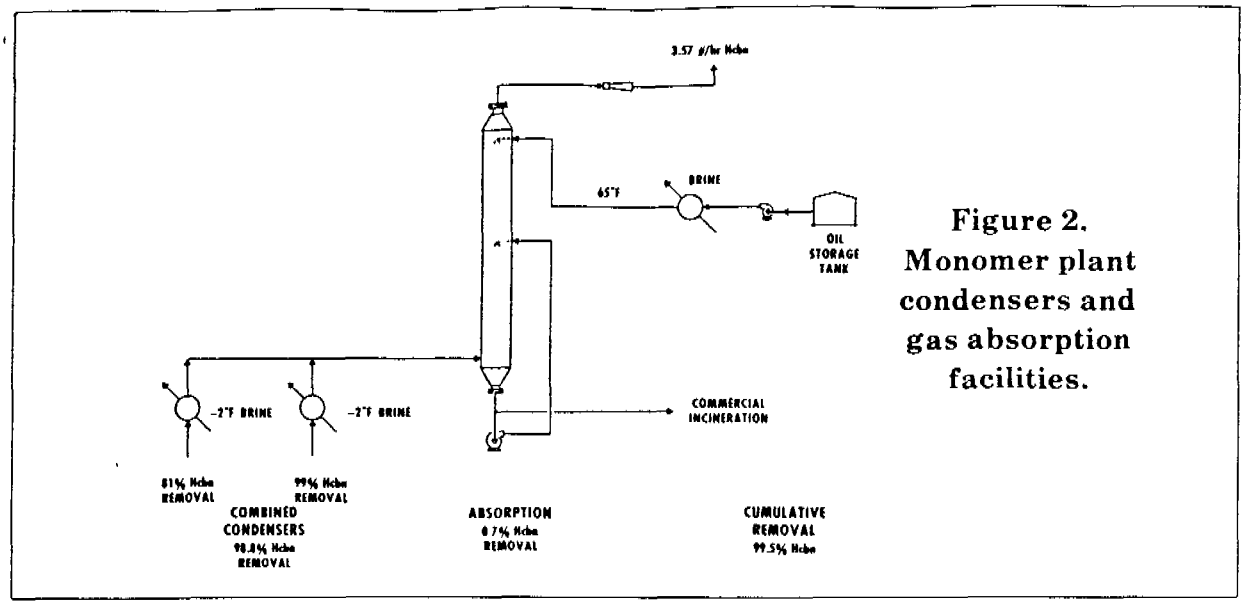

Figure 2. Monomer plant condensers and gas absorption facilities.

scrubbed tower caused by upsets. A mist eliminator pad is installed in the discharge of the knock-out pot to keep water droplets from entering the incinerator.

The incinerator was built by Henry Vogt Machine Co. and designed to operate at 1,400°F with a residence time of 0.7 sec. The heat recovery section generates 70,000 lb./hr. of 750-lb./sq. in. steam. The natural gas requirement for this unit is about 80,000 std. cu. ft./hr. Waste gas entering the incinerator flows between five vertical Coen Co. natural gas-fired, duct burner assemblies, resulting in a fire and waste gas sandwich; five layers of fire and six layers of waste gas. The fire sandwich extends 10 to 12 ft. from the burner face. The maleic incinerator system cost $1.75 million in 1975 and is the only incinerator system known on this process. The hydrocarbon destruction is at least 93% and carbon monoxide destruction is at least 95%.

The incinerator for the "Oxo" butadiene process is designed to treat 235,000 lb./hr. waste gas containing about 4,000 parts/million hydrocarbon and 7,000 parts/million carbon monoxide. Installed cost of this incineration system was $2.5 million. Waste gas treated in this system results from air used to oxidize butene to butadiene. The waste gas, after butadiene has been recovered in an oil absorption system, is combined with other process waste gas and fed to the incinerator. Waste gas enters the incinerator between seven vertical Coen duct burner assemblies to be incinerated at 1,300°F with a retention time of 0.6 sec. This incinerator, also designed by Henry Vogt Machine Co., utilizes flue gas circulation. The benefit achieved by recirculating flue gas is to incorporate the ability to generate a constant 100,000 lb./hr. of 750 lb./sq. in. steam with variable waste gas flow. Waste gas flow can range from 10% to 100% of design production rate.

The waste gas stream contains essentially no oxygen; therefore, significant combustion air must be supplied. Fresh air is supplied to the Coen duct burners by a separate blower system. This system is designed to supply up to 185,000 lb./hr. of new air. Since the air requirement is quite variable and a significant boiler efficiency penalty is incurred by supplying excess air, a continuous stack oxygen analyzer was installed. This instrument, Westinghouse Probe Type Model 218, continuously measures the oxygen in the 400°F exhaust stack.

The oxygen level in the stack is used to control the air supply. The design oxygen level is 6%; however, the plan is to incrementally reduce the oxygen to approach 2% if acceptable pollutant removal is achieved. This incinerator is fired with natural gas consuming about 130,000 std. cu. ft./hr. Natural gas supplies 84% of the firing energy. The additional 16% energy is supplied by the hydrocarbon contamination of the waste gas stream. The incinerator was recently completed and performance analyses have not

Table 1. Incinerator summary.

	Maleic Incinerator	Oxo Incinerator	Houdry "Puff" Reactor
Waste Gas Flow, lb./hr.	220,000	235,000	900,000 (total) 13,000 (Puff reactor)
Contaminants, wt.-%:			
Hydrocarbon	0.25	0.4	0.5
Carbon Monoxide	1.8	0.7	—
Removal Efficiency, %			
Hydrocarbon	93	93	92
Carbon Monoxide	95	95	—
Construction:			
Year	1975	1976	1975
Cost, $	1,750,000	2,500,000	725,000
Heat Efficiency, %	85	82	80
Natural Gas Added, Std. cu. ft./hr.	80,000	130,000	0
Retention Time, Sec.	0.7	0.5	0.3

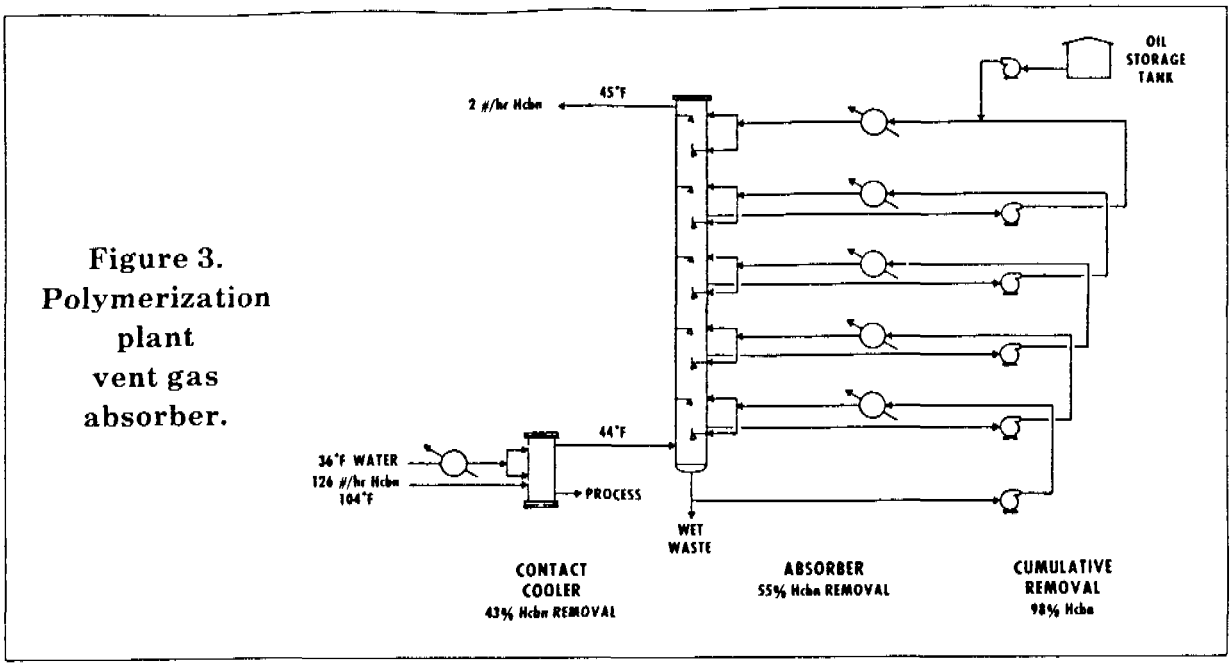

Figure 3. Polymerization plant vent gas absorber.

been completed. It is anticipated that the hydrocarbon and carbon monoxide destruction will be around 90%.

Waste gas from the Houdry butane dehydrogenation process is cleaned by what we call the "Puff" reactor. The Houdry process dehydrogenates butane through an 1,100°F, fixed bed reactor until the bed cools about 50°F. The reactor is then taken out of production service, reheated by hot air to 1,100°F, and then placed back in service. The Houdries at Petro-Tex contain seven of these reactors, which are alternately dehydrogenating and being reheated. Pollutants are emitted during the transition from production to reheat. Petro-Tex collects the high concentration of pollutants occurring during this transition and passes them over a fixed bed oxidation reactor.

A detailed process and waste gas stream analysis show that most of the hydrocarbon pollutants can be concentrated in about 1% of the regeneration (reheat) air flow. This 13,000 lb./hr. of contaminated regeneration air, which we call the "puff," is diverted to a catalytic packed bed reactor, which achieves a 92% hydrocarbon destruction. The puff reactor has a continuous 13,000-lb./hr. combustion air flow at 1,150°F. Hydrocarbon contamination is burned in 33 sec. during every 3 min. period, thereby elevating this packed bed temperature 100°F. Constant air flow in the remaining 2-1/2 min. removes the stored heat from the puff reactor and carries it to the waste heat boiler for recovery by steam generation. This hydrocarbon stream is burned without the addition of new fuel, allowing for a recovery of 3 million B.t.u./hr. of energy.

The puff process and equipment were designed by Petro-Tex and installed at a cost of $725,000 for each Houdry. The first installed puff reactor has been in service now for about a year and a half, and to date there has been no decline in efficiency of pollutant removal.

Condensation systems

Petro-Tex utilizes condensation to remove a relatively high (up to 98%) hydrocarbon concentration from small noncondensable waste gas flows. Five such systems will be described. The main advantages of condensation are that the product may be recovered and the relatively low energy requirement to remove the contamination. The basic information on the condensation system is summarized in Table 2.

The first system is a shell-and-tube exchanger installed on our neoprene monomer isomerization tower, Figure 2. A shell-and-tube exchanger cooled by −2°F brine solution

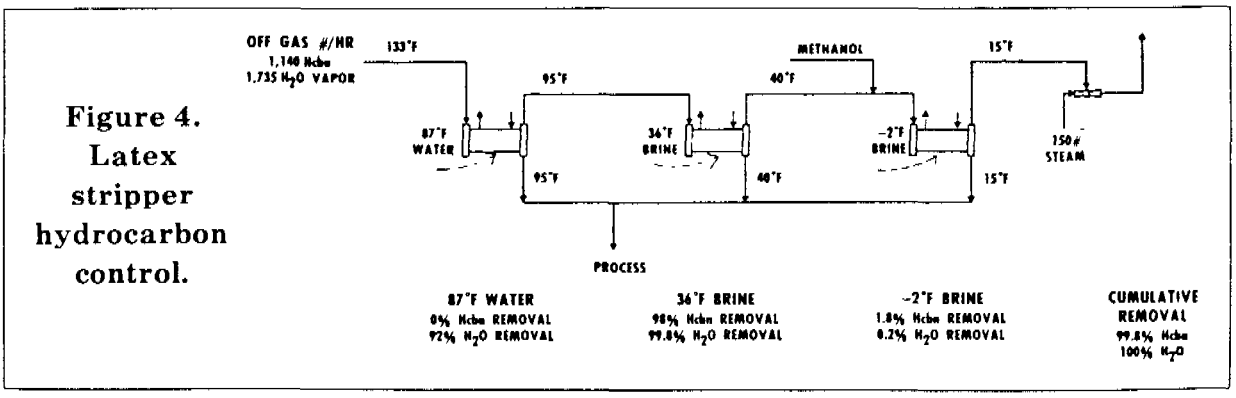

Figure 4. Latex stripper hydrocarbon control.

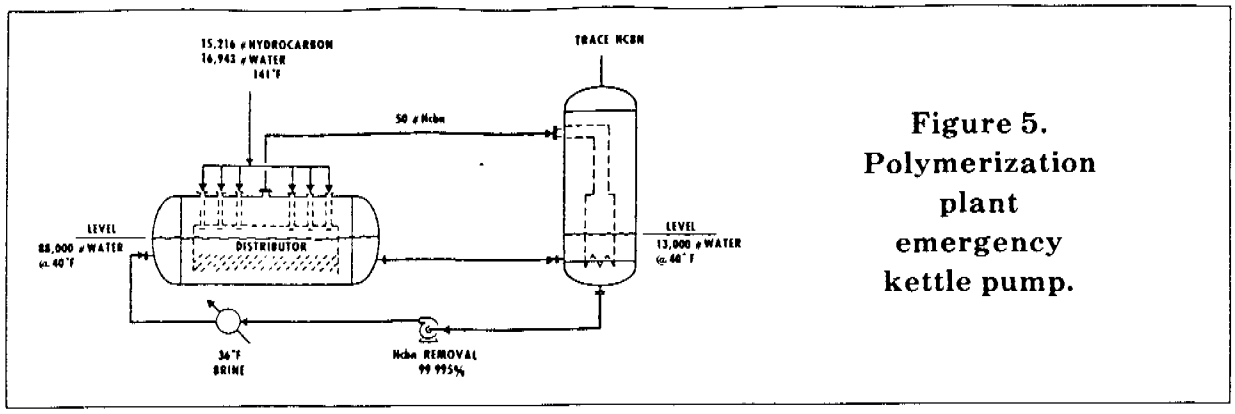

Figure 5. Polymerization plant emergency kettle pump.

was installed to treat a total waste gas flow of 331 lb./hr. This exchanger condenses 81% of the contained hydrocarbon with an energy requirement of 22,000 B.t.u./hr. The hydrocarbon recovered is returned to the process for utilization.

The second condensation system, also a shell-and-tube exchanger, was installed on the waste gas from the neoprene monomer topping column shown in Figure 2. Total waste gas flow of 550 lb./hr. passes over a shell-and-tube exchanger cooled by a −2°F brine, showing an overall hydrocarbon removal efficiency of 99%. The heat load on this exchanger is 93,000 B.t.u./hr. Recovered hydrocarbons, in this case unusable by-products, are collected in a waste organic tank and incinerated.

The third condensation system is a direct-contact cooler in which chilled water is sprayed into the waste gas flow to wash and condense about 43% of the contained hydrocarbons. It is seen in Figure 3. This system is installed in the neoprene polymer plant where neoprene latex is manufactured in batch reactors. Upon each charging and emptying of a polymerization reactor some waste gas must be displaced. A waste gas vent collection system was installed tying all the vessels in the process together. Collected waste gas is discharged into the direct-contact cooler.

One of the particular problems faced in this area was the rather regular carryover of the latex solution in the vent gas header system. Presence of the latex precluded the use of a shell-and-tube exchanger to cool the waste gas stream due to the expected high rate of fouling. Fouling is the principal reason a direct-contact cooler was chosen. About 3 gal./min. of chilled water is sprayed into the waste gas flow of 275 lb./hr. The temperature of the waste gas is reduced to about 44°F, resulting in the condensation of 43% of the contained hydrocarbon. Exit gas from the direct-contact cooler is then treated in an absorption system, which is described later in this article. The heat load on this direct contact cooler is 110,000 B.t.u./hr.

A series of three shell-and-tube heat exchangers is installed to treat the waste gas produced as a result of steam-stripping latex to remove any unreacted monomer, as shown in Figure 4. This series of exchangers treats a total gas flow of 2,875 lb./hr., which is mostly steam and chloroprene. The first exchanger in the series operates with cooling tower water which is at about 85°F. This exchanger removes most of the steam but does not remove any of the monomer. The second exchanger is cooled with 36°F brine and removes most of the remaining water and about 98% of the contained monomer. The final exchanger is cooled with −2°F brine solution and removes most of

Table 2. Condensation systems.

	Neoprene Monomer** Isomerization Tower	Neoprene Monomer* Topping Column	Neoprene Polymer* Vessel Vents	Neoprene Latex Stripper Vent	Neoprene Polymer Emergency Dump System
Type of Heat Exchanger	S&T**	S&T	DC†	S&T	DC
Waste Gas Flow, lb./hr.					
Hydrocarbon	159	—	126	1,140	15,200 Hcb. Total
Waste Gas Flow, lb./hr. Total	331	542	275	2,875	32,000 total Dump
Hydrocarbon Removal Efficiency, %	81	99	43	99.8	99.995
Heat Load, B.t.u./hr.	22,000	93,000	110,000	1.2 Million	10,000 Steady State 3 Million Heat Sink/Dump
Operating Temperature, °F	−2	−2	36	−2	40 to 75
Construction:					
Year	1973	1973	1974	1969	1974
Cost, $	20,000	30,000	40,000	120,000	250,000

*Waste gas exit this system is further treated in absorption system
**Shell-and-Tube
†Direct contact with water

efficiency of the system is 99.8%.

One of the significant operating problems that occurred on this system was water freezing in the −2°F condenser. Enough water vapor passed through the 32°F condenser to form ice crystals on the −2°F condenser and eventually plug it. Plugging of this exchanger of course substantially reduced its efficiency as a pollution control device and upset the stripping operation, which is under a partial vacuum. The solution finally developed to cope with the icing problem was the injection of a very small flow of methanol vapor into the inlet of the −2°F condensers. A nitrogen stream is bubbled through a drum of methanol to become saturated with methanol vapors. The flow of nitrogen is controlled to carry the needed 2 lb./hr. of methanol to each exchanger to keep them de-iced. Methanol de-icing has been in use for about two years and has corrected an extremely difficult operating problem. Heat load on this series of exchangers is 1.2 million B.t.u./hr. with 90% of the heat load being used to remove the contained steam.

The neoprene polymerization kettle emergency dump system is installed to prevent hydrocarbon emission to the air when a hot polymerization charge occurs, Figure 5. Polymerization of chloroprene to neoprene is a controlled exothermic reaction. Infrequently the temperature gets out of control, causing overpressuring in the polymerization vessel. The pressure is relieved at 15 lb./sq. in. gauge, for safety reasons, by a rupture disc. When the rupture disc relieves, about 15,000 lb. of hydrocarbon can be vented to the atmosphere.

The installed quench vessel contains about 12,000 gal. of 40°F water, which serves as an instantaneous 3 million B.t.u. heat sink to cool the ejected "hot charge." The "hot charge" enters the main, 11 × 25 ft. quench vessel through a 20 ft. inverted weir trough, which is partially submerged in the cool water. The weir serves to provide intimate contact between the hot liquid/vapor and the cool water. The quench system will receive the entire heat load in about 30 sec. Any vapor not condensed in the main vessel passes to a smaller, second vessel where it is again contacted with 40°F water. It was calculated that the hydrocarbon loss from a "hot charge" is about 0.005%. Water in the quench vessel is cooled to 40°F by pumping through an external shell-and-tube heat exchanger. It takes 3-1/2 days to cool the inventory of water and 10,000 B.t.u./hr. to keep it cool.

Absorption systems

Petro-Tex has installed two absorption systems utilizing oil to absorb hydrocarbon from nitrogen waste gas streams. The basic information on these absorption systems is summarized in Table 3. The first of these is installed in the neoprene rubber polymer section, Figure 4. In this part of the plant, chloroprene is polymerized in batch reactors to make a neoprene latex. Since this is a batch operation, there are contaminated vapors displaced from the reactors each time they are filled. The displaced gas containing up to 50% hydrocarbon is collected in a vent header system and cleaned by the direct-contact cooler previously described.

After passing through the direct-contact cooler, the waste gas enters the five-stage oil absorption tower. The hydrocarbons are absorbed in the 45°F oil. The oil is a mixture of paraffinic and aromatic oils, which reduces the tendency for polymerization of the contained hydrocarbons within the tower. This is a spray tower with five independent oil circulating, cooling, and spray systems. Fresh oil starts at the top of the tower with no absorbed hydrocarbon and works down through each of the five stages until the oil at the bottom of the tower contains about 3% hydrocarbon.

	Neoprene Monomer Absorber	Neoprene Polymer Vent Absorber
Spray Tower Stages	2	5
Waste Gas Flow to ABS*		
Hydrocarbon, lb./hr.	31	72
Waste Gas Flow, Total, lb./hr.	36	187
Absorber Efficiency, %	90	97
Heat Load, B.t.u./hr.	13,000	330,000*
Operating Temperature, °F	65	45
System Efficiency Including Condensation, %	99.5	98.4
Construction;		
Year	1975	1974
Cost, $	60,000	300,000

*Includes heat load for recovery of hydrocarbons.
**Absorber

Total waste gas flow to the contact cooler/absorption system is 275 lb./hr., containing about 125 lb./hr. hydrocarbon contamination. The designed discharge for this system is 2 lb./hr. hydrocarbon. Tests on this system have shown less than 1 lb./hr. discharged to the atmosphere. This results in an overall efficiency in excess of 98% hydrocarbon removal. The recovered hydrocarbon is stripped from the oil and returned to our process for further utilization. The cost of this installed system, which has been in service for two years, amounted to $300,000.

The second absorption system is installed in the neoprene monomer plant, Figure 3. This system treats waste gases generated by our installed vacuum system. Several distillation towers are operated under a partial vacuum induced by steam jets. The absorption tower treats a waste gas stream of 36 lb./hr. containing 31 lb./hr. of hydrocarbon contamination. It flows into a two-stage oil spray tower that operates under 100 mm. mercury pressure. About 90% of the contained hydrocarbon is absorbed in the spray tower. Absorption fluid is a process heavy ends stream that is chilled to 65°F. The lean oil is sprayed into the top of the tower. The bottoms circulating system recirculates the oil to the second-stage spray and feeds 850 lb./hr. hydrocarbon containing oil to the waste storage tanks. The cost of this installation was $60,000. It has been in service for two years. #

R. D. Pruessner, superintendent of environmental control for Petro-Tex Chemical Corp., is an active member of the Air Pollution Control Assn., the Water Pollution Control Federation, and the American Chemical Society. He earned his B.S. at Texas A&I Univ.

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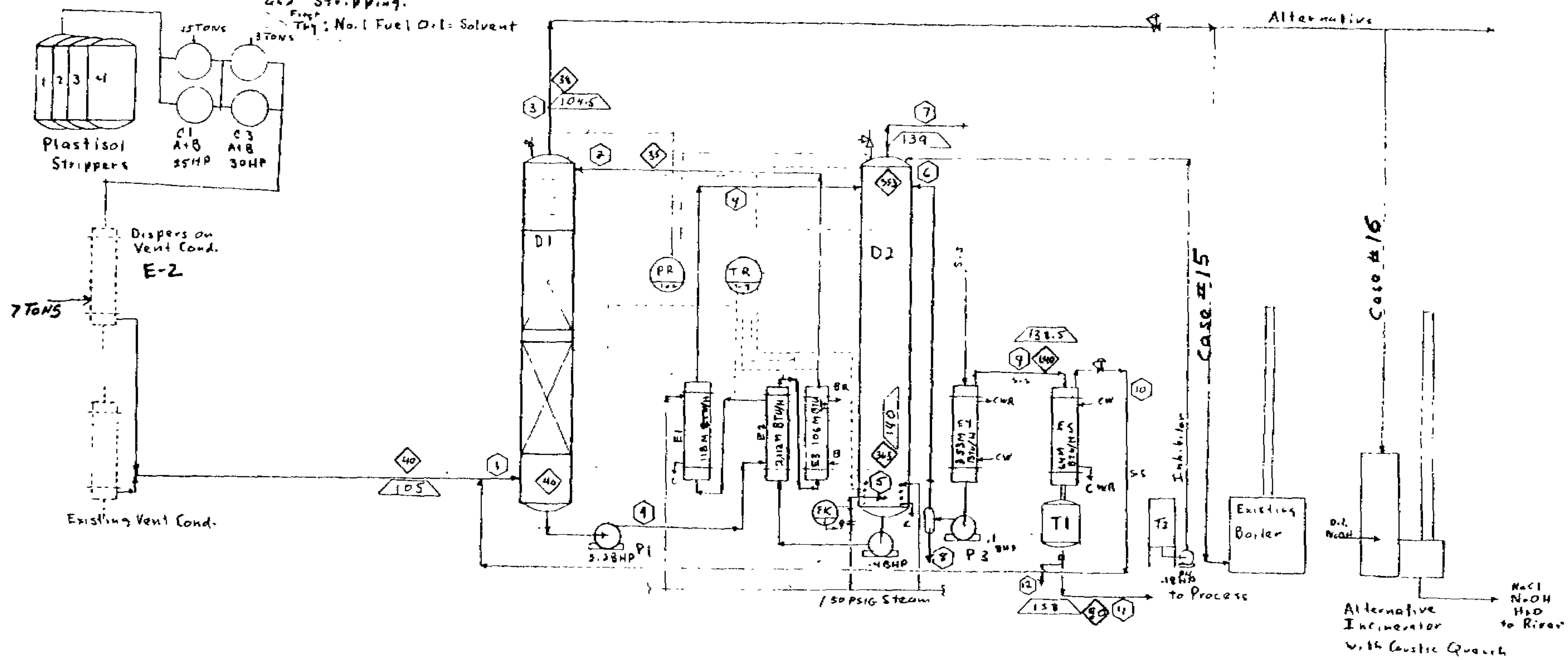
Drawer #8

— Absorption / Stripping — VCM Rates

CHKD. BY______ DATE______ VCM Recovery from Vent Gas via Refrigerated Solvent Absorbing and Stripping. JOB NO.______

First Try: No. 1 Fuel Oil: Solvent



S-2

LV = 1.3

FUEL OIL = 210 GPD

STm. = 1980 #/DAY

HP = 5HP

Pounds Per Hour: Max. Rate Conditions for Design. See Fig. 1 for Rate Fluctuation.

	MW	1 Peak	2	3	4	5	6	7	8	9	10	11	12
VCM	62.5	473	6.04	2.4	476.69		.08	470.68	.0001	470.68	0.01	470.60	
N_2	28	561	0	561	0		0	0		..	0.01		
Water	18	0	0	0	0	250	++	250	247	2.8	0....	.05	2.8
Solvent	175	0	12012	0.0000	12012		85	85	.2	....	.		
Inhibitor	-	0	?	0	?		0	0			.		
Total		1034	12018	563.4	12488.6	250	85.0	805.7	247.2	473.5	0.07	470.65	

(T) PSIA

COLORITE 006500

Tenneco Chemicals

ENGINEERING ESTIMATE
SUMMARY SHEET

ESTIMATE NO. _____ LOCATION Burlington

ESR NO. 509 DATE 3/24/77

PROJECT TITLE VCM Recovery – Case XVI – REV. 2

ASSET CAT.	DESCRIPTION	MATERIAL $	LABOR $	SUB-CONTRACT $	TOTAL $M
A.	Refrigerated Absorbing & Stripping Equipment				50
	Installation Factor				x 6
	Subtotal				300
	Other Equipment				
	C-1 A&B				42
	C-3 A&B				30
	E-2				11
	S-2 A&B				170
	Fuel Oil & Caustic Storage				15
	Subtotal				268
	Installation Factor				x 5
	Subtotal				1,340
	Escalation				160
	TOTAL - Process Equipment				1,800
B.	Utilities – same as Case XV				150
C.	Methyl Chloride – Not Applicable				—
	GRAND TOTAL				1,950

APPROVALS

PREPARED BY W. G. Wermuth PROJECT MGR. _____

COST ENGINEER _____ DIR. OF PROJECTS _____

MGR. OF GEN. ENGR. _____ DIR. OF ENGINEERING _____

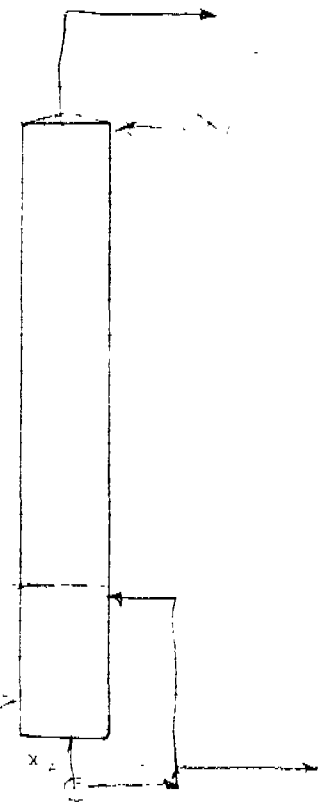
	g/g.	MW	B.P.	lb/g.	VCM Solv.	Note
Methanol	46.	32	148.1	6.63		
Isopropanol	88.	60.1	180.2	6.57		
Isobutylene	75.	56.1	20.7	5.0		
Lacquer Diluent	410.	(80)	140-200			
Naphtha VM&P	410					
Rubber Solvent	415	(80)	115-245			
Solv. Naphtha, Pet.						
Straight Arom.	64	(140)	311-344			

1.1
7

Solvents for VCM Recy Abs.

	$/g.	
Methanol	.46	
Isopropanol	.88	
Isobutylene	.75	
Lacquer Diluent Petr., 140°F-200°F	.40	——
Naphtha VM&P Petr.	.40	—
n Heptane	.99	—
n Octane	.49	—
Rubber Solv. Petr., 115°F-245°F	.45	
Solvent Naphtha Petr. - Straight Aromatic 311-340°F	.64	
Toluene	.608	
111 TCE	2.51	
TCE	2.38	
Xylene	.57	

MEK . 72.1 MW



105 PSIA

Total Film

99.9 %

99.9 %

	H							
N_2	561	20.05		561			.473	.00758	.000378
							561	20.05	.999622
	1034							

			X_i					
V	76.37	7627	.10	15.07			3.0	.048	.0007
L	V 12,012.	68.643	.90					
		76.27						

12.

$$\rho_V = \frac{1034}{27.63} \times \frac{105}{14.7} \times \frac{492}{520 \times 359} = .7045 \text{ lb/ft}^3$$

$$\rho_L = \frac{6.879^{\#}}{gal} \times 7.48 \frac{gal}{ft^3} = 51.45 \ ^{\#}/ft^3 \quad - \ 60°F$$

$$V_a \doteq .15 \left(\frac{51.45}{.7045}\right)^{1/2} = 1.28 \ ft/s.$$

$$A = \frac{1034 \ \#/h}{.17045 \ \#/ft^3 \times 3600 \ sec/h \times 1.282 \ ft/sec} = .318 \ ft^2$$

allow 20% $1.2 \times .318 = .3816$

$$D = \left(\frac{.3816}{.785}\right)^{1/2} = .697 \ ft = 8.36''$$

.000012 Atm = V. P. Kero 40°F $\frac{.04 mm}{760}$ =

Drickbach

$$\log P_{vcm} = 6.49712 - \frac{783.4}{234.44} = \quad ?$$

P = 1430 mm = 1.882 Atm

$$\alpha = \frac{1.882}{.000053} = 35,509$$

MW Kero - 175

$$y_2 = \frac{\alpha x}{(\alpha - 1)x + 1} = \frac{35,509 \times .1}{35,508 \times .1 + 1} = .99975$$

$$\pi = P_1^\circ x_1 + P_2^\circ x^2$$

$$\pi = \frac{P_1^\circ x}{y} = \frac{1.882 \times .1}{.99975} =$$

$\frac{.00005}{7} \times 20.05 =$

.025 #/H

7 Atm × .2743 = 1.9201

.1882 Atm

$P_{VCM} = .7 \times .001887 = .013209$

$P^* = .5 \times .0132 = .0066$ Atm.

$.0066 = P' x$

$x_1 = \frac{.0066}{1.882} = .0035$

$$\frac{(1.9201 - .1882) - (.0132 - .0066)}{\ln \frac{1.7319}{.0066}} = \frac{1.7253}{5.5699} = .3098 \text{ Atm.}$$

12" ϕ Column.

$h = \frac{\text{[illegible]}}{.785 \times 2 \times .3098} = 15.5'$

Note: Above is conservative because stripper is designed for $x_1 = .0014$ (vs .0035)

$P^* \cdot \pi_f = .0014 \times 1.882 = .00263$

$$\frac{(1.9201 - .1882) - (.0132 - .00263)}{\ln \frac{1.7319}{.01057}} = \frac{1.72133}{5.099} = .3374 \text{ Atm}$$

$h = 14.2'$

99.9% Rec'y $P^*_{VCM} = .000378 \times 7 = .00265$ Atm.

50% Appr. $P^* = .00132$ $x_1 = \frac{.00132}{1.882} = .000701$

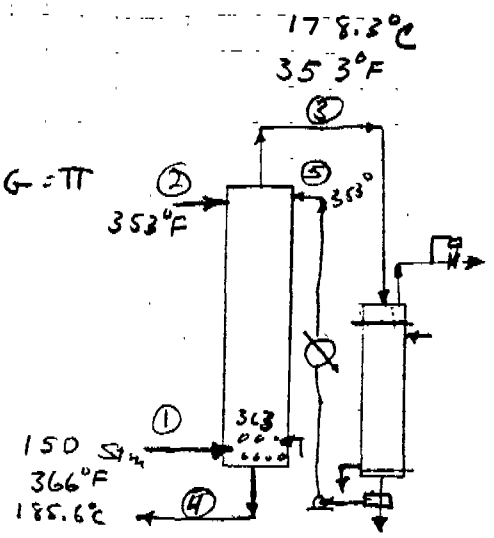
$$\frac{(1.9201 - .1882) - (.00265 - .00132)}{\ln \frac{1.7319}{.00132}} = \frac{1.72133}{7.179} = .2398 \text{ Atm}$$

$h = 20.0$ ft

(4)

Stripper

Assume 125 PSIG = π



$\frac{195}{760} = .256$ Atm

VP Kero 178.3°C

Try 250 lb./H Open Stm. - 150 PSIG Satd.

Material Balance — #/H

	①	②	③	④	⑤	⑥
VCM		476.7	475.63	6.037	5	tr
Kero		12,012.	105.4	12012	1054	tr
Water	250	0	250			250
		12,488	831.03			

$$\text{Kero Orifice} = \frac{\left(\frac{475}{68.5} + \frac{955}{18}\right) \times .}{(9.5 - 3.5)} + 175 = 105.4 \text{ #/H}$$

M/H.

	①	②	③	④	⑤
VCM	—	7.6272	7.610	.0966	.08
Kero	—	68.64	.602	68.64	.602
Wtr.	13.89	0	13.89	0	
	13.89	76.2672	22.102	68.7366	.682

.0014

B1 183.9°C π = 9.50 Atm

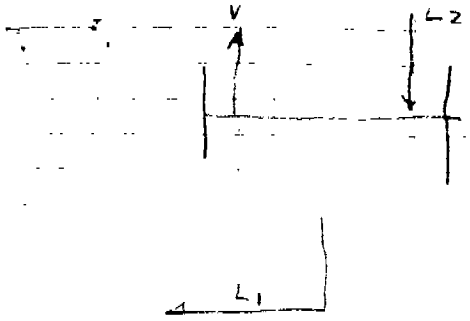
VCM P° = 52.918 Atm. $\ln P = 6.49712 - \frac{785.4}{250+t°}$

$P°x = 52.918 \times \frac{.0966}{68.7} = .0744$ Atm $y_1 = .00783$ .1137

Kero $P°x = \frac{260}{760} \times \frac{68.64}{68.7366} = .3416$ Atm $y_2 = .03596$ .522

$y_3 = .9562$ $\frac{13.89}{14.526}$

M



			M	x_2
L_2 VCM	= .1137 + .0966	=	.2103	.00303
Kero	= 68.64 + .522	=	69.162 / 69.5773	.99697

B 2 T = 181°C Π = 9.5 Atm.

VCM P° = 51.31 $\quad \log P = 6.45712 - \frac{783.4}{930+t}$

$y_2 = \frac{P^\circ x}{\Pi} = \frac{51.31 \times .00303}{9.5} = .01637$ $\quad$ M .2398

Kero $P' = \frac{243}{760} = .3197$

$y_2' = \frac{.3197 \times .99697}{9.5} = .03365$ $\quad$ .4915

$y_3' =$ .94813 $\quad$ 13.89 / 14.6499

			M	x_3
L_3 VCM	= .2398 + .0966	=	.3364	.00484
Kero	= 68.64 + .4915	=	69.1315 / 69.4679	.99516

B3 T = 179° TI = 9.5 ATM

VCM P° = 50.223 Atm log P = MF M

$y_2' = \frac{50.223}{9.5} \times .00484 =$ .02558 .3768

Kenn $P' = \frac{230}{760} = .302$ Atm.

$y_3 = \frac{.302}{9.5} \times .99516 =$ .031631 .4661

y_3' = .942781 13.89

14.733

Fenske: $S_m = \frac{\log \frac{7.61}{.1602} \times \frac{68.64}{.0966}}{\log 166.3} = \frac{9.103}{5.118} = 1.78$

			M	MF
L_{11}	VCM = .3768 + .0966	=	.4734	.00680
	K = 68.64 + .4661	= 69.1061		.993196
		69.5795		

B4 T = 179 TI = 9.5

	MF	M
$y'_4 = \frac{50.223}{9.5} \times .0068 =$	.03595	.5355
$y'_4 = \frac{.302}{9.5} \times .9932 =$	.03157	.4702
y''_4 =	.9324	13.89
		14.896

			M	MF
L_5	VCM = .5355 + .0966	=	.6321	.00906
	K = 68.64 + .4702	= 68.8567		.99094
		69.7423		

B5

		ME	M
VCM	$\eta'_5 = \frac{50.22}{9.5} \times .00906 =$	.0473	.7132
K	$\eta'_5 = \frac{.302}{9.5} \times .99094 =$	.0815	.475
W		.9212	13.89
			15.078

		M	MF x_6
L6	.7132 + .0966	= .8098	.01158
	68.64 + .475	= 69.115	.98842
		69.9248	

B6

		MR	M
	$\eta_6 = \frac{50.22}{9.5} \times .01158 =$	.0612	.9368
	$\frac{.302}{9.5} \times .98842 =$	.0314	.4806
		.90738	13.89
			15.308

		M	MF x_7
Ln	.9368 + .0966	= 1.0334	.01473
	68.60 + .4806	= 69.12	.98527
		70.154	

B7

$$\eta_7 = \frac{50.22}{9.5} \times .01473 = .07787$$

$$\frac{.302}{9.5} \times .98527 = .03132$$

B13

$$\eta'_{13} = \frac{50.22}{9.5} \times .07 =$$

η'_{13} =

Size of Absorber

Try 12" ϕ 1" Polypropylene Pall Rings → Packing Factor: 52

$$\frac{L}{G}\left(\frac{\rho_V}{\rho_L-\rho_V}\right)^{\frac{1}{2}} = \frac{12012}{1034}\left(\frac{.7045}{(51.45-.7)}\right)^{\frac{1}{2}}$$

$$= 1.369$$

$$G = \frac{1034}{3600 \times .781} = .3659$$

$$\frac{G^2 F \mu^{.1}}{\rho_G(\rho_L-\rho_G)g_c} = \frac{(.3659)^{2.0} \times 52 \times 2^{1.07}}{.7045(51.45-.7)32.2}$$

$$= .0065 \rightarrow .3''/ft$$ 1" Pall Rings OK

$$= .005$$ General Chart: 1½" Pall Rings → .25"/ft

12" ϕ with 1" Pall Rings is Good Design.

Size of Stripper

$$\rho_V = \frac{831}{22.1 \times 359} \times \frac{139.7}{14.7} \times \frac{492}{818} = .602$$

$$\rho_L = \frac{51.45}{1.18} = 43.6 \text{ lb/ft}^3$$

$$\frac{L}{G}\left(\frac{\rho_V}{\rho_L-\rho_V}\right)^{\frac{1}{2}} = \frac{12489}{831}\left(\frac{.602}{43.0}\right)^{.5}$$

$$= 1.778$$

$$\frac{G^2 F \mu^{.1}}{\rho_G(\rho_L-\rho_G)g_c} = \frac{G^2 \cdot 52 \times (.4)^{.1}}{.602\,(43)\,32.2} = .006 \quad @\ \Delta P = .4$$

General Chart.

$$G^2 = .1054$$

$$G = .3247$$

$$G = \frac{831}{3600 \times A}\ ;\ A = \frac{831}{3600 \times .3247} = .7108\ ft^2$$

$\Phi = .95$ ft Say 12" ϕ

Water Content of Recd. VCM

Assume 140 °F leaving water cooled cond

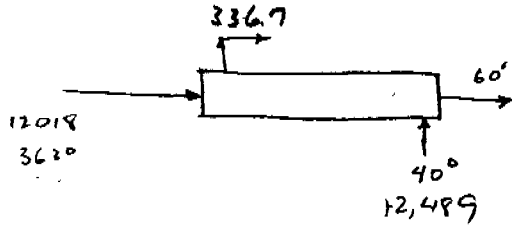
$P^{\circ}_{H_2O}$ = 149.4 mm →

π = 138.5 PSIA = 7160 mm

$$\frac{470.7}{62.5} \times \frac{149.4}{7010} = .16 \text{ cm } H_2O$$

$$= .28 \text{ \#/\#}$$

E 2 - Stripper BoHs / Absorber BoHs



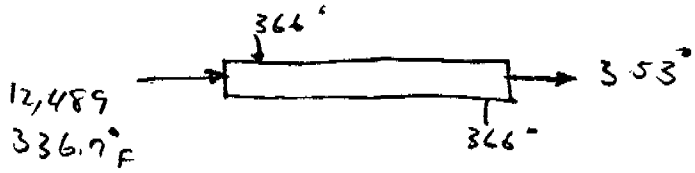
12018 x .58 (363 − 60) = 2,112,643 BTU/H

12,489 x .57 (336.7 − 40) = 2,112,043

$$\Delta T_m = \frac{26.3 - 20}{\ln \frac{26.3}{20}} = \frac{6.3}{.2738} = 23.0$$

$$A = \frac{2,112,043}{150 \times 23} = 612 \text{ sq. ft.}$$

E 1 - Feed to D2 Htr.

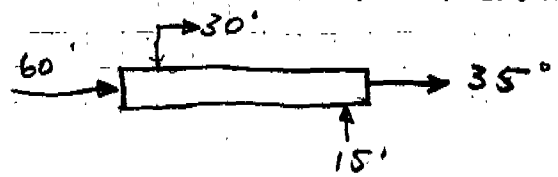


12,489 x .58 (353 − 336.7) = 118,071 BTU/Hr.

$$\Delta T_m = \frac{29.3 - 13}{\ln \frac{29.3}{13}} = \frac{16.3}{.812} = 20.0$$

$$A = \frac{118,071}{100 \times 20} = 59 \text{ ft}^2$$

E 3 Brine cooler oil to Absorber



$\frac{30-10}{\ln \frac{30}{10}} = \frac{20}{1.0} = 18.2$

$1201F \times .35 (60-35) = 105,158$ BTU/H

8.76 Tons

$A = \frac{105,158}{100 \times 18.2} = 57.8$

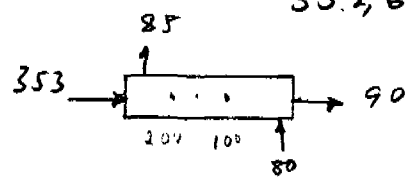
E 4 First Cond. - Stripper

$825.7 (353-90).4 = 86863$ BTU/H Sensible

$250 \times 1000 = 250,000$ BTU/H Wt. L.H.

$105 \times 150 = 15,750$

352,650

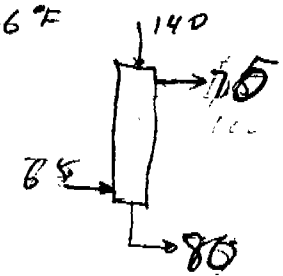


$\frac{352,650}{6 \times 500} = 117.5$

$A = \frac{352,650}{75 \times 70} = 67$ sq. ft. say 100 ft²

E 5 VCM Cond.

Cond. at 57°C = 134.6°F 138 PSIA



$\frac{79,100}{10 \times 500} = 15.8$

$470 \times 135 = 63,450$ BTU/H latent htg.

$470 \times (135-40).35 = 15,630$

$A = \frac{79,100}{100 \times 70} = 11$ ft² say 20 ft²

$\frac{63,450}{200 \times 70} = 4.5$ Cond. $\frac{15,630}{100 \times 1.5} = 10$ subcool

Say 20'

Equipment

			Mat'l.	Purchased + Del'd
D-1	Packed column 12" Dia 2 sections 8' each 1" Polypropylene Pall rings		C.S	4 000
D-2	Packed Column 12" Dia. 3 sections 8' each 1" 304 S.S. Pall Rings		C.S.	5000
E-1	Feed Heater	60 sq. ft. ¾" - 16 BWG - 8' Tubes	C.S	1000
E-2	Stripper Bottoms/Absorber Bottoms	650 sq. ft.	C.S	8000
E-3	Solvent Cooler	60 sq. ft.	C.S	1000
E-4	Water Cond.	100 sq. ft.	316 S.S.	2500
E-5	VCM Cond.	50 sq. ft	316 S.S.	1500
P 1a,b	D 1 Bottoms Pump	40 g.p.m. 105 PSIA inlet 160 PSIA out Sp. G. .8 40°F	C.I.	2000
P-2,a,b	D 2 Bottoms Pump.	40 g.p.m. 140 PSIA inlet 150 PSIA out	C.I	2000
P-3,a,b	D 2 Ovhd.	2 g.p.m. 140 PSIG. inlet 160 PSIG out	C.I.	1000
P-4	Inhibitor Pump	1 g.p.h. P.D.	S.S	500
T-1	VCM Receiver	1200 gal. 150 PSIG WP	304 SS.	5000
T2	Inhibitor Feed Tank	50g. S.S.		1000
T3	KO Tank	1000g. 150 PSIG WP	CS	2000
	Instruments			8500
				45,000
	Foundations & Supports			9,000
	Rigging & Scaffolding			6,000
	Pipe Labor & Mat'l.			50000
	Insulation			23,000
	Electrical & Instr.			17,000
	Paint & Cleaning			5000
				110,000
	Engr.			50,000
	Contgcy. (34.1%)			70,000
				275,000

Operating Costs — 330 On-stream Days/Yr. (90%)

		$/Yr.
Raw Materials		
VCM Credit	3250 #/S.D. x .995 x 330 x .15	-(160,000)
Inhibitor	(Guess)	- 2,000
#1 Fuel Oil	2500 g./Yr. + Purchase + Disposal	- 2,500
		(155,500)
Operating Supplies	(charts, Misc.)	2000
Utilities		
Steam	500 #/H. x 8400 Hr. x $3.8/M	- 15,960
Electrical	4 KW x 8400 x .028	- 940
Brine @ 15°F	10T. x 8400 x $.10/T	- 8,400
Cooling Water	50 g.p.m x 60 x 8400 x $1/M	- 2,500
Condensate Return	250 #H. @ 366°F = 83 M BTU/H; 83 x .9 x $3./MM x 8400 = $1680/Yr.	- NIC
		27,800
Maintenance		15,000
Labor + Supervision - Benefits	Allocate	20,000
Ins. + Taxes -	$50,000 x .015	3750
Depreciation	250,000 x .0625	15625
Plant Ovh. — NIC.		
	(credit)	(73,000)

Steam Requirements

E 1 $\frac{118,071}{858} = 137.6$

D 2 Coil - Assume coil supplies heat for VCM, Keep

12012 x .4 x 10 = 48,048 BTU from ΔT on Feed.

470 x 135 = 63,450

85 x 140 = 11,900

75,350

$\frac{75,350}{858} = 87.8$

E 1	137.6
D 2 Coil	87.8
Feed	250.0
	475.4 #/H
loss	24.6
	500.0 #/Hr

Cooling Water Req'd (80°F)

E 4 $\frac{353,000}{20 \times 500} = 35.3$ g.p.m.

E 5 $\frac{64,000}{10 \times 500} = 12.8$

48.1 say 50 g.p.m.

Brine @ 15°F

$\frac{106,000}{12000} = 8.83$ Tons say 10 Tons

Electrical

	Motor HP	BHP	KW
P 1	3.0	2.2	
P 2	.5	.4	
P 3	.25	.1	
P 4	.25	.1	2.08
	4.0	2.8	
Lights + Instr.			1.92
			4.00

Fuel Oil

Assume 1 charge of 2000 g. replaced per year -

Assume 500 g./Yr. loss

2500 g. x .45 = 1125
Disposal

Case 15.-

(a) Plastisol Strippers (4) with Vacuum Pumps, Compresors, Vent Cond.

(b) Refrigerated Solvent Scrubbing

(c) Vent from (b) to Existing Boiler

Steam

(a)+(c) = 0

(b) = 12000 lb./S.D. = $3.45/M - 350 Days/Yr = $/Yr. 14,490

Electrical

(a) 55 HP Conn., Fully Loaded : 984 KWH/Day .0276×350 = 9505

(b) 4. HP Conn., 2.1KW Motors, 1.9 KW Lights, etc. = 96 KWH/Day = 927

(c) None additional

1080 KWH/Day 10,432

No power incl. for Refrigeration

Cooling Water.-

(a) Cooling water for Refrigeration not included

(b) 50 g.p.m. @ 10¢/M gal. = 2500
No C.W. incl. for refrig.

(c) 0

Chilled Water
(a). 18 Tons

(b)+c = 0

Brine (15°F)

(a)+(c) = 0

(b) 10 Ton Demand 8.8 Ton Reqd.

Nitrogen Available

(b) $\frac{3250}{24 \times 473} \times 561$ = 160 pounds/H = 2180 SCF/H ← 100% utilization
2180 × .26 × .01 × 8400 = $47,600

COLORITE 006518

Case 16. (a)+(b) above + (c) Vent from (b) to Incinerator

Steam

(a) = 0
(b) = 12000 #/SD
(c) = 1980 #/SD
13,980 #/SD × 3.45/M - 3505.0 = 16,850

Electrical

(a) + (b) = 1080 KWH/DAY

(c) 5 HP × .85 PF × .746 = 3.17 KW ; 76 KWH/D

1156 KWH × .0276 × 350 = 11,167

Fuel Oil

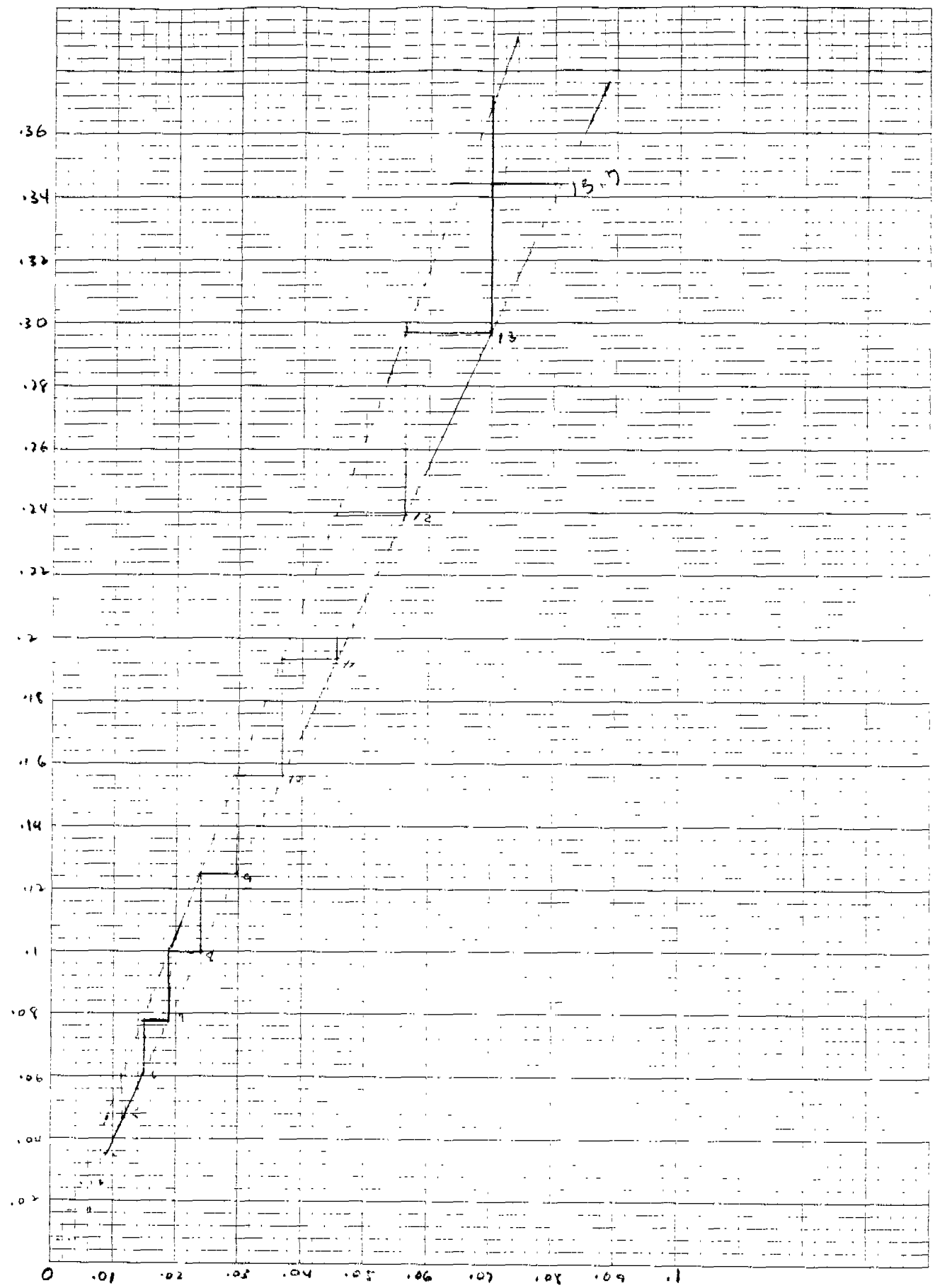
a + b = 0

(c) = 710 g.p.d. × 350 × 14,000 / 42 = 25,288

Caustic Soda

3250 × .005 = 16.25 #/S.D. VCM

$\frac{16.25 \times 40}{62.5 \times .5}$ = 20.8 # 50% NaOH/SD



46 0860

5 X 5 TO 1/2 INCH
KEUFFEL & ESSER CO

COLORITE 006520

enneco Chemicals
Tenneco Company

TENNECO

DI _____
PAGE _____

DATE:

PUMP CALCULATION SHEET P1

Indicate pressure and elevations for each equipment item.

SUCTION PRESSURE		
Origin Pressure	psia	
Static Head	psi	
Line Loss P	psi	
PUMP SUCTION PRESSURE	psia	

NET POSITIVE SUCTION HEAD		
Static Head	feet	
-Line Loss	feet	
+Vapor Press. Correction*	feet	
AVAILABLE NPSH	feet	

DISCHARGE PRESSURE		
Delivery Pressure	psia	
Static Head	psi	
P Control Valve(s)	psi	
P Exchanger(s)	psi	
P Furnace(s)	psi	
P Orifice(s)	psi	
Line Loss P	psi	
Other	psi	
PUMP DISCHARGE PRESS.	psia	

MAXIMUM SUCTION PRESSURE		
Vessel S. V. Setting	psia	
Static Head	psi	
NET	psia	

MAXIMUM DISCHARGE PRESSURE		
Max. Suction Press.	psia	
Normal Pump P	psi	
NET	psia	

*Use for liquids NOT at their bubble point.

LIQUID

PUMPING TEMP.		°F	=
VISCOSITY	@	°F cp	=
VAPOR PRESS.	@	°F psia	=
SPECIFIC GRAV.	@	°F	=
#/GALLON	@	°F	=
GPM	@	60°F	=
GPM	@	°F	=

DIFFERENTIAL PRESSURE

Discharge Press. psia =
Suction Press. psia =

TOTAL PUMP P psi =
feet =

GPM DESIGN =

BRAKE HORSEPOWER

BHP = $\frac{40 \text{ gpm} \times 55 \text{ psi}}{1715 \times 0.6 \text{ Pump Eff.}} = 2.13$

Pump Name:
Pump No.:

COLORITE 006521

Tenneco Chemicals
Tenneco Company

BY _____
PAGE _____

DATE:

PUMP CALCULATION SHEET

Indicate pressure and elevations for each equipment item.

SUCTION PRESSURE		
Origin Pressure	psia	
Static Head	psi	
Line Loss P	psi	
PUMP SUCTION PRESSURE	psia	

NET POSITIVE SUCTION HEAD		
Static Head	feet	
-Line Loss	feet	
+Vapor Press. Correction*	feet	
AVAILABLE NPSH	feet	

DISCHARGE PRESSURE		
Delivery Pressure	psia	
Static Head	psi	
P Control Valve(s)	psi	
P Exchanger(s)	psi	
P Furnace(s)	psi	
P Orifice(s)	psi	
Line Loss P	psi	
Other	psi	
PUMP DISCHARGE PRESS.	psia	

MAXIMUM SUCTION PRESSURE		
Vessel S. V. Setting	psia	
Static Head	psi	
NET	psia	

MAXIMUM DISCHARGE PRESSURE		
Max. Suction Press.	psia	
Normal Pump P	psi	
NET	psia	

*Use for liquids NOT at their bubble point.

LIQUID.			
PUMPING TEMP.		°F	=
VISCOSITY	@	°F cp	=
VAPOR PRESS.	@	°F psia	=
SPECIFIC GRAV.	@	°F	=
#/GALLON	@	°F	=
GPM	@	60°F	=
GPM	@	°F	=

DIFFERENTIAL PRESSURE		
Discharge Press.	psia	=
Suction Press.	psia	=
TOTAL PUMP P	psi	=
	feet	=
GPM DESIGN		=

BRAKE HORSEPOWER

$$BHP = \frac{2 \text{ gpm} \times 10 \text{ psi}}{1715 \times 0.5 \text{ Pump Eff.}} = .02$$

Pump Name:

Pump No.:

Tenneco Chemicals
A Tenneco Company

BY _____
PAGE _____

DATE:	PUMP CALCULATION SHEET P 2

Indicate pressure and elevations for each equipment item.

SUCTION PRESSURE		
Origin Pressure	psia	
Static Head	psi	
Line Loss P	psi	
PUMP SUCTION PRESSURE	psia	

NET POSITIVE SUCTION HEAD		
Static Head	feet	
-Line Loss	feet	
+Vapor Press. Correction*	feet	
AVAILABLE NPSH	feet	

DISCHARGE PRESSURE		
Delivery Pressure	psia	
Static Head	psi	
P Control Valve(s)	psi	
P Exchanger(s)	psi	
P Furnace(s)	psi	
P Orifice(s)	psi	
Line Loss P	psi	
Other	psi	
PUMP DISCHARGE PRESS.	psia	

MAXIMUM SUCTION PRESSURE		
Vessel S. V. Setting	psia	
Static Head	psi	
NET	psia	

MAXIMUM DISCHARGE PRESSURE		
Max. Suction Press.	psia	
Normal Pump P	psi	
NET	psia	

*Use for liquids NOT at their bubble point.

LIQUID			
PUMPING TEMP.		°F	=
VISCOSITY	@	°F cp	=
VAPOR PRESS.	@	°F psia	=
SPECIFIC GRAV.	@	°F	=
#/GALLON	@	°F	=
GPM	@	60°F	=
GPM	@	°F	=

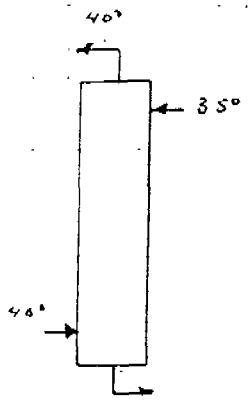
DIFFERENTIAL PRESSURE		
Discharge Press. psia	=	
Suction Press. psia	=	
TOTAL PUMP P	psi	=
	feet	=
GPM DESIGN		=

BRAKE HORSEPOWER

$$BHP = \frac{40 \text{ gpm} \times 10 \text{ psi}}{1715 \times 0.6 \text{ Pump Eff.}} = .39$$

Pump Name:
Pump No.:

Heat Balance D-1 Abs.



Heat In 32°F Datum

VCM (H) 475 × .19 (8) = 722
VCM (L) 475 × 1.8 × 76 = 64,980
N_2 562.8 × .24 (40-32) = 1080
Hexane 59.7 × .522 (35-32) = 24,888
91,670

Heat Out 475 × .19 (8)
VCM 475 × 1.8 × 76 = 722
N_2 562.8 × .24 (40-32) = 65
Hexane 18.85 (275.0 - 108) = 1080
3148

VCM 477.8 × .19 (ΔT) = 5015
5937.6 × .525 (ΔT) = 3 3208.0 ΔT = 24522
84203
5015 + 3208ΔT = 91,670

3208 ΔT = 91,670 - 5015

ΔT = 27,012
32
59°F

N Hexane C_4H_{14} 86.2

$$\log P_{mm} = 6.87776 - \frac{1171.530}{224.366 + t}$$

Hexane Loss

40°F. = 4.444°C P = 57.23 mm Hg.

20°F. = -6.666°C P = 31.36

10°F. = -12.222°C P = 22.67

0°F. = -17.777 P = 16.10

$\frac{57.23}{5320}$ = .01076 mf. Hexane 40°F

.2184 moles/Hr.

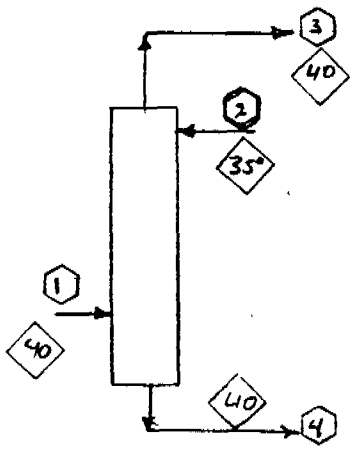
18.83 #/Hr Peak 5.39 Avg.

5.53 #/gal. $\frac{45^{\#}}{5.53} \times 5.39 \times 8400 = 3685$ #/Yr.

Scrubber Material Balance. — 99.9% VCM Recovery
40°F. gas Inlet.

Assume VCM in (2) is $x = \frac{y}{\pi}$

Assume π = 7 Atm. = 102.9 PSIA
= 5320 mm Hg.



Moles/Hr. Basis

		(1) M/H	(1) MF	(2)		(3)		(4)	
VCM	62.5	7.60	.2744	.0526	.000762	.0076	.000374	7.645	.1
N_2	28	20.10	.7256	0		20.10	.988866	0	0
Hexane	86.2	0	0	69.023	.999238	.2187	.01076	68.805	.9
		27.70		69.0756		20.3263		76.45	

At 35° F (1.667°C)

$P^{\circ}_{VCM} = 1304.8$ mm Hg. $= 1.717$ Atm.

Gas out $y_1 = .000374$ $P = .000374 \times 7 = .002618$

Choose $y^* = .000187$

$\frac{\pi y^*}{P^{\circ}_{VCM}} = x_2 = \frac{.000187 \times 7}{1.717} = .000762$ $m = \frac{1.717}{7} = .2453$

Mat'l. Bal. lb/H $mx_2 = .000187$

	MW	①		②		③		④	
VCM	62.5	475		3.28		.475		477.81	
N_2	28	562.8		0		562.8		0	
Hexane	86.2	0		5949.84		18.852		5930.99	
		1037.8		5953.12		582.127		6408.8	

Bottom of Column. 40°C

$P^{\circ}_{VCM} = 1430.7$ mm Hg $= 1.882$ Atm.

$y^* = \frac{.1 \times 1.882}{7} = .02688$

$m = \frac{1.882}{7} = .2688$ Bot

$$N_p = \frac{\log\left[\left(1 - m\frac{G}{L}\right)\left(\frac{y_1 - mx_2}{y_2 - mx_2}\right) + \frac{mG}{L}\right]}{\log \frac{L}{mG}}$$

Top $m\frac{G}{L} = \frac{.2453 \times 20.33}{69.076} = .07219$

Bot $m\frac{G}{L} = \frac{.2688 \times 27.7}{76.45} = .09739$

$\log m\frac{G}{L} = .08385$

$$N_p = \frac{\log\left[(1-.08385)\left(\frac{.2744-.000187}{.000374-.000187}\right)+.08385\right]}{\log \frac{1}{.08388}}$$

$$= \frac{\log[(.91615)(1457.38)+.08385]}{11.926}$$

$$= \frac{7.1969}{2.4787} = 2.903 \text{ Theo. Plates}$$

log mean ΔP driving force, Atm.

Top Bot$_q$ = .000374 × 7.0 = .002618 Atm.

Interface $\Pi_q = P^\circ x = 1.717 \times .000762 = 7 \times .000187$

$= .001308$ Atm

Bot. $.2744 \times 7 = 1.9208$

$.1 \times 1.882 = .1882$

$$\Delta P_m = \frac{(1.9208-.1882)-(.002618-.001308)}{\ln \frac{1.7326}{.001309}} = \frac{1.7313}{7.1881} = .2408 \text{ Atm}$$

$K_{qa} \doteq 2$ $\quad h \doteq \frac{7.6}{.785 \times 2 \times .2408} \sim 20.1$ ft. packing

1½" P.11 R.-PJ

Assuming $K_q a = 2.0$ M/H·Atm.·ft^3 1'ø

Column Pressure Drop + Size

Top

$$\rho_V = \frac{582.13}{20.326 \times 359} \times \frac{492}{500} \times \frac{7}{1} = .5494 \text{ lb/ft}^3$$

$$\rho_L = .666 \times 62.41 = 41.52 \text{ lb/ft}^3$$

$$\frac{L}{G}\left(\frac{\rho_G}{\rho_L - \rho_G}\right)^{1/2} = \frac{5953.1}{582.1}\left(\frac{.5494}{41.01}\right)^{1/2}$$

$$= 1.157$$

$$\frac{G^2 F \mu^{.1}}{\rho_G(\rho_L - \rho_G) g_c} = .01 \quad \text{at } \Delta P = .4" \text{ wc/ft.}$$

F. 40 $\mu^{.1} = .54$ 1½" Pall Rings
52 = .94

$$G^2 = \frac{.01 \times .5494 \times 41.01 \times 32.2}{\frac{40}{52} \times .94} = .1929 \quad 1\tfrac{1}{2}"$$
$$= .1250$$

$$G = .43926 \text{ \#/sec. ft}^2 = 1581 \text{ \#/Hr ft}^2$$
$$.2508 = 903$$

$$\text{Area} = \frac{582.1}{1581} = .368 \text{ ft}^2 \quad 1\tfrac{1}{2}" \text{ or } .644 \text{ ft}^2 \; 1"$$
$$\times 1.3 = .4785 \rightarrow .781 \text{ ft } \Phi \quad \times 1.3 \quad 1.03'\Phi$$
$$9.4"$$

Bottom

$$\rho_V = \frac{1037.8}{27.7 \times 359} \times \frac{492}{500} \times 7 = .7188$$

$$\rho_L = \quad = 42.0$$

$$L/G\left(\frac{\rho_G}{\rho_L - \rho_G}\right)^{1/2} = \frac{6408}{1037.8}\left(\frac{.7188}{41.3}\right)^{1/2} = .8147$$

$$\frac{G^2 F \mu^{.1}}{\rho_G(\rho_L - \rho_G) g_c} = .014$$

$$G^2 = .35592 \qquad G = .5966 \text{ \#/sec ft}^2 \quad 2148$$

$$\text{Area} = \frac{1037.8}{2148} = 0.4831 \text{ ft}^2$$

$$\times 1.3 = .628 \text{ ft}^2 \rightarrow 0.894 \text{ ft}$$

$$G^2 = \frac{.014 \times .7188 \times 41.3 \times 32.2}{\frac{40}{52} \times .94} = .35592$$

$$= .9739$$

$$G = .5966 = 2148 \text{ lb/ft}^2 \quad 1\tfrac{1}{2}''$$

$$= .5732 = 1883.7 \text{ lb/ft}^2$$

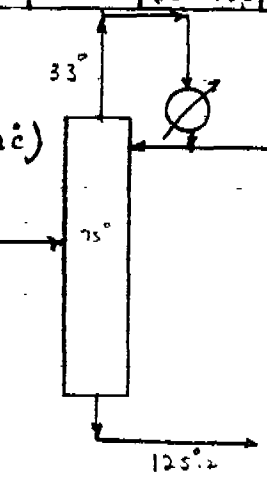
$$\frac{1037.8}{1883.7} = .551 \text{ ft}^2$$

$$\times 1.3 = .716 \text{ ft}^2 \rightarrow .955 \text{ ft}$$

say 12''

Stripper Material Balance - Mole Basis

		Feed		Bottoms		Ovhd Prod.		Total Vap Out	Reflux
		M	MF						
VCM	62.5	7.645	.1	.0525	.000762	7.5925	.9999	25.2830	17.6905
Hexane	86.2	68.805	.9	68.804	.999238	.00076	.0001	.002531	.00177
		76.45		68.8567		7.59326		25.2805	17.6922



1. Assume condensing of VCM at 90°F (32.2°C)

   $P°_{VCM}$ @ 32.2 = 3283 mm Hg.
   = 62.5 PSIA (47.8 PSIG)

2. ΔP cond = 25 mm Hg → 3258 mm
   Col. Top.

3. 30 Trays @ 3.33/tray → 100 mm Hg

   Bottom Pres = 3358 mm Hg.

Bottom Temp — temp pure hexane at P_b

$$\log P = 6.87776 - \frac{1171.53}{224.366 + t} = 3.52608$$

$$\frac{1171.53}{224.366 + t} = 3.35167$$

$$224.366 + t = 349.535$$

$$t = 125.17\ ^\circ C$$

317.1°F crit temp. 527 Atm. crit press.
158.4°C

$P°_{VCM}$ @ 125.17 = 19,562 mm. α = 5.82 at Bot.

P_{Hex} @ 32.2° = 204.9 mm α = 15.78 at top.

Overall avg α = 9.58

$$\text{Min } R = \frac{.9959}{8.58 \times .1} = 1.165$$

Use $2 \times R_m = R$ for Column Dia., Cond. & Reb. Duty

Use $1.5 R_m = R$ for Determining No. Plates.

α at Feed Plate - Assume 75°C

$P_{vap} = 8483.9$

$P_{liq} = 921.28$ $\qquad \alpha = 9.21$

α avg above feed = 12.05

$$\text{Above } S_m = \frac{\log \frac{.9999}{.0001} \times \frac{.9}{.1}}{\log 9.21} = \frac{11.4075}{2.22} = 5.14$$

α avg below = 7.32

$$\text{Below } S_m = \frac{\log \frac{.1}{.9} \times \frac{68.804}{.0525}}{\log 7.32} = \frac{4.981}{1.991} = 2.502$$

Total 7.642

at $1.5 R_m$ - Theo. Plates $= \frac{2.52}{1.5} \times 7.642$

$= 14.37$ 13.37 + Reboiler

@ 55.7% PE $\longrightarrow$ say 24 Plates

$$24 \times \frac{5.14}{6.642} = 18.6 \text{ from top}$$

Stripper. Assume Condensing at 40°F

1. Condensing Press.

$$\log P = 6.49712 - \frac{783.4}{230+t}$$

$P = 1430.9$ mm Hg 27.67 PSIA ≈ 13 PSIG

2. ΔP Cond. = 25 mm

3. ΔP Column = 100 mm

Bottom Press. = 1555 mm Hg.

Col. Top at 1456 mm.

$\frac{783.4}{230+t} = 6.49712 - 3.16316$

4.976°C

$$\log P^{\circ}_{Hexane} = 6.87776 - \frac{1171.53}{224.366+t}$$

$\frac{1171.53}{224.366+t} = 3.686$

$t = 93.46\,^{\circ}C \rightarrow 200.2\,^{\circ}F$

$P^{4.976}_{VCM} = 11,485.3$ mm $\rightarrow \alpha = 7.38$ Bottom ✓

$P^{4.44}_{Hexane} = 63.39$ mm. $\rightarrow \alpha = 22.57$ Top

Overall avg. $\alpha = 12.90$

$$R_{min} = \frac{.9999}{11.9 \times .1} = .840$$

Assume Feed Plate @ 48°C

$P_{VCM} = 4777$

$P_{Hexane} = 405.3$ $\alpha = 11.79$

Avg α above = 16.3

Avg α below = 9.32

$$\text{Min Plates Above} = \frac{\log \frac{.9999}{.0001} \times \frac{.9}{.1}}{\log 16.3} = \frac{11.4073}{2.791} = 4.09$$

$$\text{Min Plates Below} = \frac{\log \frac{.1}{.9} \times \frac{68.804}{.0525}}{\log 9.32} = \frac{4.581}{2.232} = 2.23$$

$$6.32$$

@ 1.5 R_m Theo. Plates $= \frac{2.82}{1.5} \times 6.32 = 11.88$

10.88 + Reboiler 20 Trays → 54.4 Plate Eff.

$\frac{4.09}{5.32} \times 20 = 15.32$ From top = Feed Tray

Mole Basis (2 x R_m = R

	MW	Feed		Btms		Ovhd		Vap Lng	Reflux
VCM	62.5	7.645	.1	.0525	.000762	7.5925	.9999	20.3479	12.7654
Hexane	86.2	68.805	.9	68.804	.999238	.00076	.0001	.00204	.00128
Tot		76.45		68.8567		7.59326		20.3499	12.7567

Wt. Basis

	Feed		Btms		Ovhd,		Vap	Refl.
VCM	477.81		3.28		474.53		1271.74	796.21
Hexane	5930.99		5930.93		.0655		.1755	.1100
Tot	64088		5934.21		474.596		1271.92	797.32

140°F Condensing Case —

Top. $\rho_V = \frac{474.6}{7.593 \times 359} \times \frac{273}{277.5} \times \frac{1455.9}{760} = .3281$ lb./ft³

$\rho_L = .94 \times 62.4 = 58.6$ lb./ft³

$V_a \doteq .15 \left(\frac{58.6}{.3281}\right)^{\frac{1}{2}} = 2.00$ '/sec.

$A = \frac{1271.74}{3600 \times 2.00 \times .3281} = .538$ sq. ft.

$\times 1.3 = .6998$ sq. ft. 30% Safety

Dia = .944 ft = 11.33" say 12"

90°F Condensing Case

$\rho_V = \frac{474.6}{7.593 \times 359} \times \frac{273}{305.5} \times \frac{3258}{760} = .1670$ lb./ft³

$\rho_L = .89 \times 62.4 = 55.5$

$V_a \doteq .15 \left(\frac{55.5}{.667}\right)^{\frac{1}{2}} = 1.369$ '/sec.

$A = \frac{1580.4}{3600 \times 1.369 \times .667} = .480$ sq ft

$\times 1.3 = .625$ sq ft

Dia = .892' or 10.7" — say 12"

Stripper Heat Balance- 90°F Cond. Case

Datum 32.2°C.

Feed $6408.8 \times .588 \times (115.2 - 32.2) 1.8$ = 562,995

Rebailer Duty + 235,624

798,619

Boil-up - to Cond

Latent Ht. $1580.4 \times 70 \times 1.8$ = 199,130

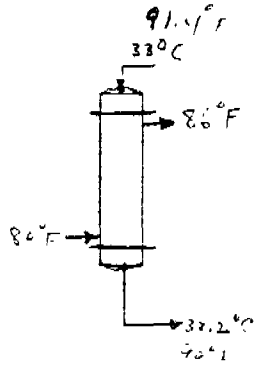
Sens. Ht. $1580.4 \times .21 \times (0.8) 1.8$ + 478

Cond. Duty = 199,608

Bo Homs $5934.2 \times .603 (125.2 - 32.2) 1.8$ = 599,011

798,619

E 1- Condenser



$\Delta T_m = \frac{10 - 5}{\ln \frac{10}{5}} = \frac{5}{.693} = 7.21$

$A = \frac{199,608}{7.21 \times 100}$

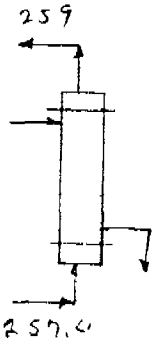
= 277 sq. ft.

× 1.26 = 350 sq. ft.

$\frac{199,608}{6 \times 500} = 66.5$ g.p.m.

two 70

E 2- Reboiler



$\Delta T_m = \frac{80.6 + 81}{2} = 80.8°F$

100 PSIG Steam Cond. Temp

338°F

$A = \frac{235,624}{5.0 \times 80.8}$

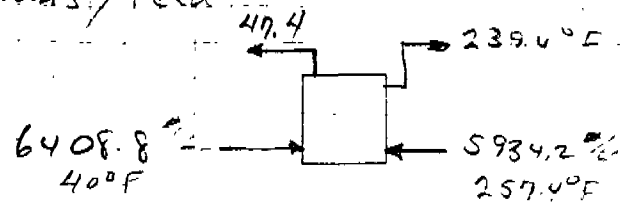
= 583 sq. ft.

× 1.28 = 75 sq ft. U = 39

$\frac{235,624}{1195 - 309} = 265.9$

× 1.2 = 320 #/M

E 3 - Bottoms / Feed



$Q = 6408.8 \times .588 \times (239.4 - 40) = 751,414$ BTU/H

$= 5934.2 \times .608 (257.4 - t) = 751,414$

$t = 47.4°F$

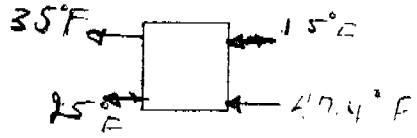
$\Delta T_m = \frac{18 - 7.4}{\ln \frac{18}{7.4}} = \frac{11.6}{.889} = 13.0$

$A = \frac{751,414}{13 \times 150} = 383.9$ sq ft. $U = 150$

$100 = 578$ sq ft. $U = 100$

Tons $U = 115.6$

E 4 Refrig.



$Q = 5934.2 \times .528 \times (47.4 - 35) = 38,852$ BTU/H

$= 3.24$ Tons

→ 5 Tons.

$\Delta T_m = \frac{20 - 22.4}{\ln \frac{22.4}{20}} = \frac{2.4}{.133} = 21.18°F$

$A = \frac{38,852}{21.18 \times 100} = 18.35$ $U = 100$

$= 24.5$ $U = 75$

say 20 ft²

Stripper Heat Balance — 40°F Cond. Temp.

— 40°R Below

Feed

6408.8 x .575 x (178 − 40) = 508,538

Reboiler Duty — 204,746

713,284

Boilup

1271.92 x 75.1 x 1.8 = 171,938

1271.92 x .195 x 2 = 496

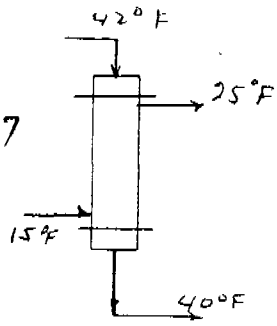
Cond. Duty = 172,434

Bottoms

5934.2 x .582 x (196.6 − 40) = 540,850

713,284

El. Cond.



$\Delta t_m = \frac{25-15}{\ln \frac{25}{15}} = \frac{10}{.51} = 19.57$

$A = \frac{172,434}{100 \times 19.57}$

= 88 U = 150

= 117 U = 75

say 120 ft²

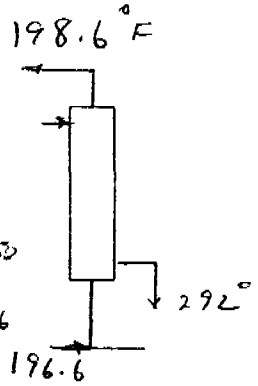
141.4 Tons

Say 20T

45 PSIG Steam Cond.

292°F

Reboiler



$\Delta t_m = 292 - 197.6 = 94.4°F$

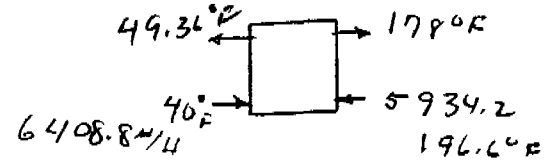
$A = \frac{204,746}{50 \times 94.4} = 43.4\ ft^2$ U=50

= 60 ft² U=36

$\frac{204,746}{1195 - 261} = 219.2$

x 1.2 = 263 ft²

E 3 – B. Homs / Feed



$Q = 6408.8 \times .575 \times (178-40) = 508,538$

$Q = 5934.2 \times .582 \times (196.6-t) =$

$t = 49.36°F$

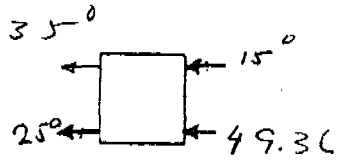
$\Delta T_m = \frac{18.6 - 9.36}{\ln \frac{18.6}{9.36}} = \frac{9.24}{.6867} = 13.45$

$A = \frac{508,534}{13.45 \times 150} = 252\ ft^2 \quad U = 150$

$100 = 378\ ft^2 \quad U = 100$

$= 400\ ft^2 \quad U = 106$

E 4 Refrigeration



$Q = 5934.2 \times .530\ (49.36-35) = 45,164$ BTU/H

3.76 Tons → 5 Tons

$\Delta T_m = \frac{-20 + 24.36}{\ln \frac{24.36}{20}} = \frac{4.36}{.1972} = 22.11°F$

$A = \frac{45,164}{22.11 \times 100} = 20.4\ ft^2 \quad U = 100$

$75 = 27.2\ ft^2 \quad U = 75$

Fancy 30 □'

	90°F, 62.5 PSIA	40°, 27.7 PSIA
Equipment - Purchased + Del'd.	74,000	55,000
Foundations, Supports, Platforms	9,000	9,000
Rigging + Millwright	6,000	6,000
Piping labor + Material C.S. (Manual)	50,000	50,000
Insulation	23,000	25,000
Electrical labor + Material	15,000	15,000
Instrument Labor + Supvn.	5,000	5,000
Paint + Clean-up	5,000	5,000
	187,000	170,000
Engineering Design, Supervision	50,000	50,000
Contingency	70,000	70,000
TOTAL	*307,000	*290,000

Utilities

Steam -- No additional capacity for	320 #/H	295 #/H
Electrical " " " "	15 KW	30
Refrigeration @ 15°F.	5 Tons	20 Tons
New Package Units	*10,000	30,000
Cooling Water @ 80°F - No additional capacity for (Including Refrig. Requirements)	85 g.p.m.	60 g.p.m.
Total Investments Incl. Refrig.	*317,000	*320,000

Preliminary Estimate - VCM Recovery - Absorber/Stripper - Hexane Solvent

Equipment	90°F 62.5 PSIA Size, Mat'l.	Purch. + Del'd.	40°F 27.7 PSIA Size, Mat'l.	Purch. + Del'd.
D-1 Absorber	12" D. 20' of 1½" Pall Rings. 1500 g. Sump, 125 PSIG W.P., C.S.	10,000	D. Ho	10,000
D-2 Stripper	12" D, 24 Seive Trays 70 PSIG W.P., C.S.	15,000	12" D, 20 Seive Trays 70 PSIG W.P., C.S.	14,000
E 1 - Condenser	350 sq. ft. 304 S.S.	10,000	120 sq.f. 304 S.S.	4,000
E 2 - Reboiler	75 sq. ft. C.S.	2,000	60 sq. ft., C.S.	1,000
E 3 - Bott's/Feed	500 sq. ft. C.S.	8.000	405 sq. ft. C.S.	6700
E 4 - Refrig.	30 sq. ft. C.S.	1.000	30 sq. ft. C.S.	1000
F 1 - Filter	R.P., GAF, Basket Type	1,000	D; Ho	1000
G 2 - Stripper Feed	40 g.p.m. 105 PSIA in. 125 PSIA Disch. Sp. G. = .675, C.S.	1000	D; Ho	1000
G-3 - Absorber Feed	40 g.p.m. 65 PSIA in 132 PSIA Disch. Sp. G. = .555, C.S.	1000	40 g.p.m. 30 PSIA in 132 PSIA Disch. Sp. G. .594, C.S.	1000
G-1 Solvent Makeup	5 g.p.m. 15 PSIA in 125 PSIA Disch. Sp. G. .66, C.S.	500	D; Ho	500
G-4 VCM Transfer	5 g.p.m. 63 PSIA in, 125 PSIA Disch Sp. G. .89, S.S.	1500	D; Ho	1500
G-5 Inhibitor	1 g.p.h Metering Pump. S.S.	1000	Ditto	1000
T-1 - Hexane Stg.	12,000 g. C.S.	4000	Ditto	4000
T-2 - KO Drum	1500 g. 125 PSIG W.P. C.S.	3000	Ditto	3000
T-3 VCM Rec.	1000 g. 70 PSIG W.P. C.S. Phenolic lined	4000	1000 g. 35 PSIG W.P. C.S. Phenolic lined.	4000
T-4 Inhibitor Feed	100 g. S.S.	1000	Ditto	1000
AG-1 Sump Agitator	1 H.P.	1500	Ditto	1000
		65,500		56,500
Instruments		8,500		8,500
		74,000		55,500

Operating Costs –

		Amount	90°F, 62.5PSIA		40°, 77.7PSIA
Raw Materials					
VCM Credit	15¢/lb.	2550 lb/D	(130,050)	2550 lb/D	(130,050)
Hexane "	$2.5/MM BTU	.104MM/H	(2,200)	.104MM BTU/H	(2,200)
Hexane Make-up	43.5¢/g.	25 g./S.D.	3,800	25 g./S.D.	3,800
Inhibitor	Guess		2,000	Guess	2,000
			(126,450)		(126,450)
Operating Supplies			3,000		3,000
Utilities					
Steam	$3.45/M lb.	820 lb/H	9,300	265 #/H	7700
Electricity	2.76¢/KWH	15KW	3,500	30KW	7000
Cooling Water 80°F	10¢/M g.	85 g.p.m	4,300	60 g.p.m	3000
Sewage		0	0	0	0
			17,100		17,700
Maintenance			16,000		16,000
Labor + Supvn	$10/Hr.	6 H/Day	21,000	6 H/Day	21,000
Insurance + Taxes	1.5% Invest.	$317M	4800	$320M	4800
Total Above			(64,550)		(63,950)
Depreciation	6.25% Invest.	$317 M	19,800	$320 M	20,000
Plant Ovhd.			2000		2000
Total			(42,750)		(41,950)

N Hexane Scrubber – 70 PSIG. operating pressure to Scrubber.
84.7 PSIA
Say 83.7 PSIA at Top = 4327 mm Hg.

$\Pi y^* = P^\circ x_2$

$\frac{4327 \times .002297}{1304.8} = x_2 = .000762$

log mean driving force – Atm

Top – $P_{Body} = .000374 \times \frac{83.7}{14.7} = .00213$ Atm

$P_{Interface} = .000762 \times \frac{1304.8}{760} = .001308$ Atm.

Bottom $.2744 \times \frac{84.7}{14.7} = 1.581$ Atm Body

$.1 \times 1.882 = .1882$ Atm. Interface

$\Delta P_{lm} = \frac{(1.581 - .1882) - (.00213 - .001308)}{\ln \frac{1.3929}{.000822}} = \frac{1.392078}{7.435} = .1872$

$h = \frac{7.6}{.785 \times 2 \times .1872} = 25.85$ ft packing.

Stripper same as previous 7 Atm. case

With Scrubber designed for 105 PSIA inlet → $\Delta P_{lm} = .2408$

$.2408 = \frac{1.3929 - (\Delta P_{Top})}{\ln \frac{1.3929}{\Delta P_{Bottom}}}$

$P_{Top} = .001 \times \frac{83.7}{14.7} = .00569$

$.2408 = \frac{1.3929 - (.00569 - .001308)}{\ln \frac{1.3929}{.000438}} = \frac{1.38851}{5.7603} = .24103$

Loss approx $.001 \times 20 \times 67.5 = 1.25$ #/H VCM

$\frac{57.23}{4327} \cdot 20.3 \times 86.2 = 23.14$ #/H Hexane Peak
6.6 #/H Avg.

Pumps-

G 1 - 103.9 PSIA - DI

9.6 = 30' lift + Friction $\times \frac{5.53}{8.33} \times \frac{14.7}{33.9}$

1.0

-12.6

99.9 PSI

$BHP = \frac{20}{5.53 \times 60} \times \frac{99.6}{1715 \times .6} = .006$

Intermittent .1 KW

G-2 125

- 2 Liq. Hd.

- 104

19

$\frac{6415 \times 19}{5.53 \times 60 \times 1715 \times .6} = .36$ BHP

.266 KW say .3 KW

G-3 $\frac{5940}{5.53 \times 60} \cdot \frac{100}{\times 1715 \times .6} = 1.739$ BHP

1.298 KW say 1.3 KW

G-4 $\frac{475 \times 100}{60 \times 8.33 \times \frac{58.7}{62.4} \times 1715 \times .6} = .0982$ BHP

1.8 KW

VCM Scrubber - TCE Solvent

Material Balance - Scrubber

	1 M/H	1 MF	2 M/H	2 MF	4 M/H	4 MF	5 M/H	5 MF
VCM	7.58	.2743	.2388	.00339	7.79[illegible]	.10	.0381	.001887
TCE	0	0	[illegible]		70.0[illegible]	.90	[illegible]	.004932
N_2	[illegible]25	.7257					20.05	.99318
			[illegible]		[illegible]		20.187	

PP TCE 4.4°C = 26.236 MM

TT = 7 × 760 = 5320 MM

MF TCE = $\frac{26.236}{5320}$ = .004932

[illegible] .01 [illegible]

13.08 × .3250 / 24 × 473 = 3.74 #/H Avg.

TCE [illegible] = 87.2°C [illegible] = 73°C

Sp. G. = 1.4[illegible]

$\log P_m = 7.02808 - \frac{1315.04}{[illegible]}$ Lange [illegible]

P = 26.236 [illegible]

Column Bottom

VCM P[illegible] [illegible] = [illegible]43 × 7 = 1.920 Atm.

" " Solvent interface = $x P^{\circ}$ = .1 × 1.882 = .1882

Column Top.

VCM PP [illegible] stream = 7 × .001887 = .0132 Atm.

VCM PP TCE [illegible] = .00339 × 1.882 = .00638 Atm.

$\Delta P_m = \frac{(1.9201 - .1882) - (.0132 - .00638)}{\ln \frac{1.73[illegible]}{[illegible]}} = \frac{1.72509}{5.537} = .3115$

$h = \frac{7.5[illegible]}{.785 \times [illegible]}$ = 15.4 ft [illegible] Polypropylene Pall Rings

VCM Stripper - TCE Solvent

Mole Basis

	Feed M	Feed MF	M	MF	Botts M	Botts MF
VCM	7.781	[illegible]	7.5[illegible]	[illegible]	[illegible]	[illegible]
TCE	70.026	.9[illegible]	[illegible]	[illegible]	70.[illegible]	[illegible]
Total	[illegible]		7.5786		70.[illegible]	

40°F. cooling wtr.
1550 mm Bott. Strip.
112.7°C
234.8°F
Reb.

90°F,
3300 mm Bott Strip
114.7°C
292.5 °F

	#1	#2	#3
VCM	486.3	471.4	14.89
TCE	9201.4	.0991	9201.3
	9687.7	471.499	9216.19

$$R_{min} = \frac{.9999}{10.93 \times .1} - \frac{11.93(.0001)}{10.93(.9)}$$

$$= .9148 - .00012 = .9147$$

$$\text{Sm Above Feed} = \frac{\log \frac{.9999}{.0001} \cdot \frac{.9}{.1}}{\log 11.93} = \frac{11.407}{2.479} = 4.60$$

Use 2 x Rm

$$\text{Actual Plates} = \frac{7.82}{2} \times 4.6 \pm 6.5 \text{ Above}$$

$$\text{Sm Below Feed} = \frac{\log \frac{.1}{.9} \cdot \frac{.99661}{.00339}}{\log 11.95} = \frac{3.486}{2.479} = 1.406$$

2 x Rm

Actual Plates = $1.41 \times 1.406 = 1.983$

α Below is less than α Above

$$\text{Feed Plate} = 52.7 + \frac{(168.4 - 52.7)}{3} = 91.3$$

$$\text{Log } P_{vcm} = 6.47972 - \frac{783.0}{230 + 91.3}$$

$P_{vcm} = 11,452$

$$\text{Log } P_{tce} = 7.02806 - \frac{1315.04}{230 + 91.3}$$

$P_{tce} = 861.4$ $\qquad \alpha = \frac{11,452}{861.4} = 13.29$

α_{avg} Above $= (22.37 \times 13.29)^{\frac{1}{2}} = 17.24$

α_{avg} Below $= 9.19$

$$\text{Sm Above Feed} = \frac{11.407}{2.847} = 4.00 \rightarrow 5.65$$

$$\text{Sm Below Feed} = \frac{3.486}{2.247} = 1.57 \rightarrow 2.22$$

7.87 T.P.

Theo

ϵ 50% $\rightarrow$ 15.7 Actual Trays.

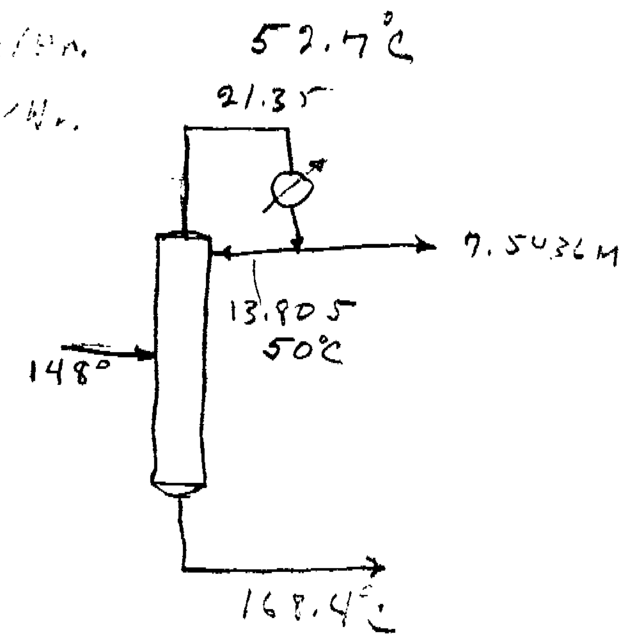
$\pm$ .5 R/M = R $\rightarrow$ 20.9 Actual

Condenser Duty:

$$\text{[illegible]-up} = (2 \times .915 + 1) \times 7.5436 \quad (1.83)$$

= 21.35 Moles/Hr.

= 1334 Pounds/Hr.



50°C [illegible]

66.2 cal./g. L.H VCM @ 50°

		BTU/H
L.H	1334 × 66.2 × 1.8 =	160,151
S.H	1334 × .23 × 2.7 × 1.8 =	1,491
	Cond. Duty =	161,642

Hea[illegible]

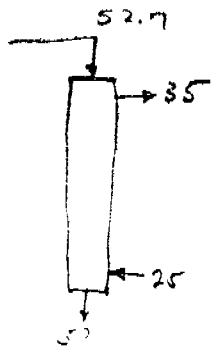
9687.7 × .45 × (148-50) 1.8 = 769,00[illegible]

[illegible]

9216.19 × .5 × (168.4-50) 1.8 = 982,0[illegible]

Reboiler Duty = 161,642 − 769,009 + 982,077 = 374,709

E 1 — Condenser



$\Delta T_m = \frac{-17.7 + 25}{\ln \frac{25}{17.7}} = \frac{7.3}{.347}$

$= 21.1$

$\frac{161,642}{18 \times 500} = 17.95$ ft²

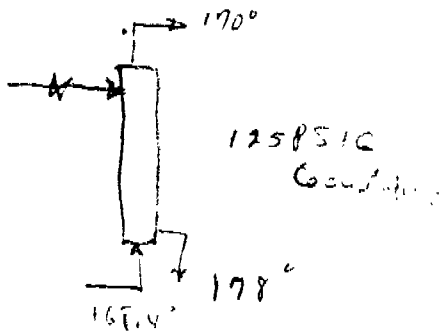
$= 53.87$ m²

6

$A = \frac{161,642}{100 \times 21.1 \times 1.8} = 42.6$ sq. ft.

Say 60 sq. ft. (4 tube passes)

E 2 Reboiler



$\Delta T_m = \frac{9.4 - 8}{\ln \frac{9.4}{8}} = \frac{1.4}{.116}$

$= 8.68$

$A = \frac{374,709}{75 \times 8.68 \times 1.8}$ = 318 sq. ft.

374,709 / 1,000 = 320.8

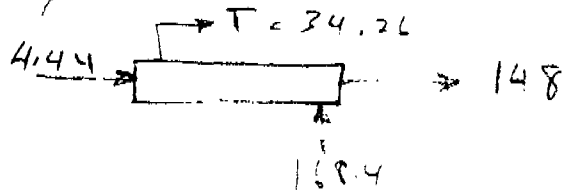
430.5 m²

95 sq. ft.

Say 100 sq. ft.

-55

E 3 Bottle/Feed Exch.



$Q = 9216.2 \times .45(168.4 - T)$ = 1,001,352 BTU/Hr

$T = 34.26$ °C

$= 134.1$

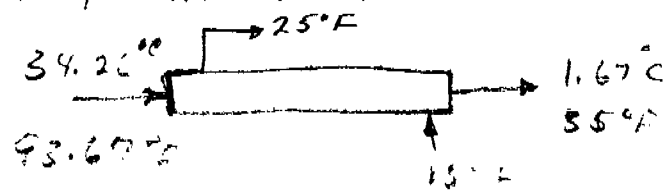
$9685.7 \times .40(148 - 4.44) .8$ = 1,001,352

$\Delta T_m = \frac{99.2 - 20.4}{\ln \frac{99.2}{20.4}} = \frac{9.42}{.579}$ = 24.8 °C

$A = \frac{1,001,352}{150 \times 24.8 \times 1.8}$ = 149.5

(224.3) sq. ft.

E4 Refrigeration Exch.



$$Q = 9216.2 \times .27\,(93.67 - 35) = 145,988$$

$$= 12.2 \text{ Tons}$$

$$\Delta T_m = \frac{68.67 - 20}{\ln \frac{68.67}{20}} = \frac{48.67}{1.233} = 39.45°F$$

$$A = \frac{145,988}{100 \times 39.45} = 37.0$$

say 50 ft^2

E3

Could make [illegible] reduce Tons Refrig.

Size [illegible] ?

[illegible] $\frac{5320}{760}$ = [illegible] lb/ft³

[illegible] 68.3

[illegible] 1,448 ft/sec.

A [illegible] $\frac{1344}{[illegible]\ 1.621[illegible]}$ [illegible]

[illegible]

Equipment

		Mat'l	Purch. Cost
D-1	Packed Col. 12" Dia. 2 Sect, 5' ea. 1" Polypropylene Pall Rings	C.S	4000
D-2	Valve Tray Column 20 Plates, 12" Dia, 12" Sp.	304 SS	10,000
E-1	Stripper Cond. 60 ft^2	316 SS.	2,000
E-2	Reboiler 500 ft^2	316 S.S.	12000
E-3	Bottoms Feed Exch. 300 ft^2	316 S.S.	8000
E-4	Refrig. Cond 50 ft^2	316 S.S.	2,000
G-1	Absorber Bottoms 20 g.p.m. Sp.G 1.48 105 PSIA inlet, 145 PSIA out	316 SS.	1,200
G-2	Stripper Bottoms 2 g.p.m. Sp.G 1.2 105 PSIA in, 145 PSIA out	316 SS.	1200
G-3	Reflux Pump 2 g.p.m. Sp.G. .85 105 PSIA in, 145 PSIA out	316 SS	800
G-4	Tower Bottoms Pump 1 g.p.m. Recirc. Pump	304 SS	500
T-1	VCM Receiver 1200 g. ISO PRESSURE	304 SS	5000
T-2	Isobutane Feed Tk 300 g	316 SS	1000
	Instruments		10,000
			$57,700

VCM Scrubber – Ethylene Dichloride

$ClH_2C\,CH_2Cl$ MW = 99

Same conditions, mole basis, as other solvents.

$P^{\circ}_{4.44°} = 31$ mm Hg.

MF MEK in Gas Out = $\frac{31}{5320} = .00583$

Dens. = 1,256 $\frac{lb}{ft^3}$

	Gas In		Solv. In		Solv. Out		Gas Out	
	M	MF	M	MF	M	MF	M	MF
VCM	7.58	.2743	.2388	.00339	7.7807	.110	.0381	.001896
EDC	0	0	70.1441	.99661	70.0263	.990	.1178	.00583
N_2	20.05	.7257					20.05	.99228
			70.3829				20.2059	

.1179 × 99 = 11.67 #/H EDC

VCM-EDC Stripper

Top Temp. = 52.69°C

P EDC = 975 mm Hg

$\alpha = \frac{5220}{975} = 19.34$

Bottom Pl. = 5340 Temp. = [illegible]

Temp = 154°C

$\log P_{VCM} = 6.49712 - \frac{783.4}{230 + 154}$

$P_{VCM} = 28,643$

$\alpha = \frac{28,643}{5340} = 5.363$

Feed Pl. Temp. Guess = 106°C

$P_{VCM} = 14,641$

$P_{EDC} = 1,425$

$\alpha = 10.27$

Avg. α above Feed = $(19.34 \times 10.27)^{1/2} = 14.1$

Avg. α below Feed = $(10.27 \times 5.363)^{1/2} = 7.42$

Avg α overall = $(19.34 \times 5.363)^{1/2} = 10.18$

$\text{Min } R = \frac{.9999}{9.18 \times .1} = 1.089$

$S_m \text{ Above Feed} = \frac{\log \frac{.9999}{.0001} \times \frac{.9}{.1}}{\log 14.1} = \frac{11.407}{2.646} = 4.31$

$S_m \text{ Below Feed} = \frac{\log \frac{.1}{.9} \times \frac{.9965}{.00350}}{\log 7.42} = \frac{3.4863}{2.0047} = 1.74$

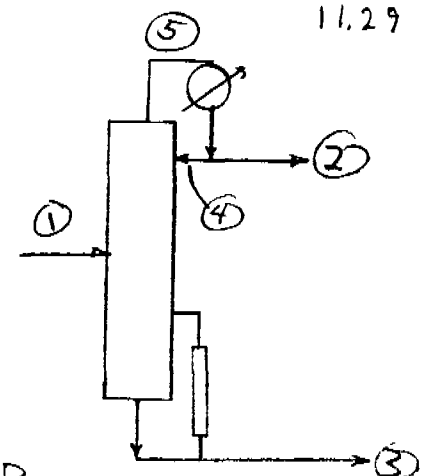
6.05

Using $R = 2 R_m$

Act. Theo Tray = 1.41×6.05 = 8.53 @ 50% → 17.06 Pl

Using $R = 1.5 R_m$

Actual TP = 11.29 @ 50% → 22.6



Material Balance - Wt. Basis

	①	②	③	④	⑤
VCM	486.31	471.41	14.89	1026.71	1498.11
EDC	6932.6	.0746	6932.5	.1623	.2371
	7418.91	471.47	6947.39	1026.87	1498.35

$\rho_V = \frac{\text{[illegible]} \times 53 \times 5320}{\text{[illegible]} \times 359 \times 326 \times 760} = 1.0215$ lb./ft.3

$\rho_L = 53.35$

[illegible]

$A = \frac{1498.35}{3600 \times 1.0215 \times [illegible]} = .202$ ft.2

$\times 1.5 = .303$

$\Rightarrow$ [illegible] . 8.5" [illegible]

Condenser Duty = $\frac{1498.4}{1636.9} \times 196,882$

= 180,223 BTU/Hr

Heat In with Feed = 7418.9 x .43 (135-50) 1.8 = 488,089

Heat Out in Bottls = 6947.4 x .46 (154-50) 1.8 =

Reboiler Duty 180,233 - 488,089 + 598,255 = 290,399

E-1- Condenser

$\frac{180,233}{100 \times 21 \times 1.8}$ = 47.45 ft²

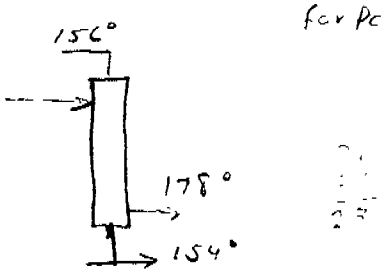
↳ 60 ft² → 26.4% for Peens

E-2- Reboiler

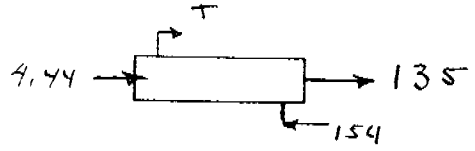
$\frac{290,397}{75 \times 23 \times 1.8}$ = 93.5 ft²

50 = 140.3 ft²

say 160 ft²



E-3 Bottls/Feed



Q = 7418.9 x .43 (135 - 4.44) 1.8 = 749,705

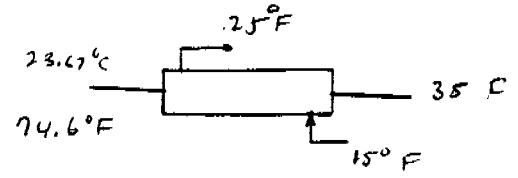
$\Delta T = \frac{749,705}{6947.4 \times .46 \times 1.8}$ = 130.3

T = 154 - 130.3 = 23.67

$\Delta T_{lm} = \frac{(23.67 - 4.44) - (154 - 135)}{\ln \frac{19.23}{19}}$ = 19.1 °C

$A = \frac{749,705}{100 \times 19.1 \times 1.8}$ = 218.06 sq. ft.

E 4 Refrig. Cond.



$Q = 7418.9 \times .31 (74.6 - 35) = 91,088.2$ BTU/H
$= 7.59$ Tons

$$\Delta T_m = \frac{(74.6 - 25) - (35 - 15)}{\ln \frac{49.6}{20}} = \frac{29.6}{.908} = 32.59^\circ F$$

$$A = \frac{91,088}{100 \times 32.59} = 27.95 \text{ ft}^2$$

VCM Scrubber-Acetone Solvent

$$\log P_m^{\circ} = 7.02447 - \frac{1161.0}{224.1 + t^{\circ}C}$$

4.44 °C

$P^{\circ} = 87.35$

$$\text{MF Acetone in Gas Out} = \frac{87.35}{5320} = .01642$$

	M	m	M	MF	M	MF	m	MF
VCM	7.58	.2743	.2388		7.7807	.10	.0381	.001866
Acetone	–	–			70.0263		.3354	.01642
N_2	20.05	.7257					20.05	.98171
							20.4235	

Acetone loss = .3354 × 58.08 = 19.48 lb/H Acetone to Incinerator

VCM- Acetone Stripper

Top Temp. = 52.69°C

$P^\circ_{Acetone}$ = 673.7 mm Hg. $\alpha_{Top} = \frac{5320}{673.7} = 7.897$

Cool 40°F, 27.7 PSIA
1550 mm → 78.8°C
3300 m → 107.15°C

Bottom π = 5340

$\log 5340 = 7.02447 - \frac{1161}{224+t} = 3.7275$

$\frac{1161}{224+t} = 3.2969$

$t = 352.1 - 224 = 128.1°C$

P°_{VCM} = 20,405 $\alpha_{Bott} = \frac{20,405}{5340} = 3.821$

Feed Plate Temp. Guess. 77.8°C

P°_{VCM} = 8952.8

$P^\circ_{Acetone}$ = 1505.0 α_{Feed} = 5.948

Avg. $\alpha_{Above} = (7.897 \times 5.948)^{1/2} = 6.85$

Avg $\alpha_{Below} = (5.948 \times 3.821)^{1/2} = 4.77$

Avg $\alpha_{overall} = (7.897 \times 3.821)^{1/2} = 5.493$

Min. Plates Above = $\frac{11.007}{1.924} = 5.928$

Min Plates Below = $\frac{3.4863}{1.562} = \frac{2.231}{8.159}$

Min Reflux = $\frac{.9999}{4.493(.1)} = 2.225$

At $R = 2 R_m$

$S = 1.41 \times 8.159 = 11.50$ Theo Pl.

23.0 Act.

At $R = 1.5 R_m$

$S = \frac{2.82}{1.5} \times 8.159 = 15.34$ Theo Pl.

30.7 Act.

Stripper Material Balance – Wt. Basis

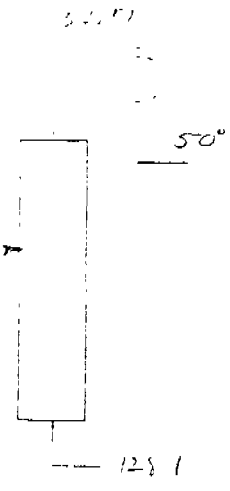
	16 #/H	#/H	B #/H	B. #/H
VCM	486.3	471.4	14.89	2569.1
Acetone	4067.1	.0437	4067.06	.2382
	4553.4	471.44	4081.95	2569.36

Area of Stripper $= \frac{2569.36}{3600 \times 1.0215 \times 1.445}$

$= .4835$ ~~ft²~~

$\times 1.5 = .7253$

$\Phi = .9612$ ~~ft.~~



Condenser Duty

$2569.4 \times 119.16 = 306,100$ BTU/H.

$2569.4 \times .23 \times 4.86 = 2,872$

$309,042$

Heat in Feed

$4553.4 \times .54 \times (106 - 50) 1.8 = 247,850$

Heat in Bolls

$40? \times .65 \times .57 \times (128.1 - 50) = 181,7??$

157 BTU/#

Reboiler Duty $309,042 - 247,850 - 181,7?? = 242,929$

Cool [illegible] 279 #/H

$11955 - 3248 = 870.7$ BTU/#

E 1- Condenser

$$A = \frac{309,042}{100 \times 21.1 \times 1.8} = 81.37$$

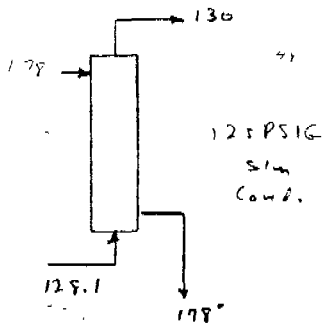
$\times 1.25 \rightarrow 100$ Sq. ft.

E. 2- Reboiler

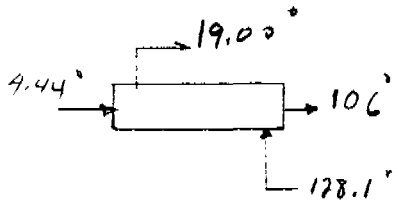
$$A = \frac{367,729}{75 \times 48.5 \times 1.8} = 56.2$$

$= 84.2$

sq/100 ft²



E-3 Bottls/Feed



$Q = 45534.52(106 - 4.44)1.8 = 432,847$ BTU/H

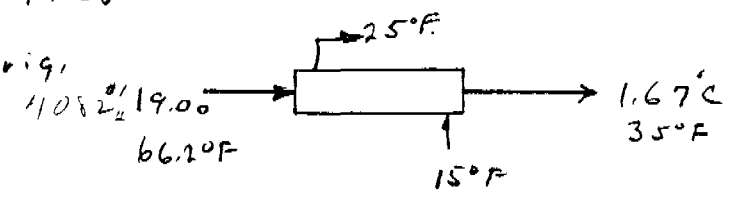
$$\Delta T_{lm} = \frac{14.57 - 22.1}{\ln \frac{14.57}{22.1}} = \frac{7.532}{.4166} = 18.08$$

$$A = \frac{432,847}{100 \times 18.08 \times 1.8} = 133.0 \text{ ft}^2$$

$$\Delta T = \frac{432,847}{4082 \times .54 \times 1.8} = 109.1^\circ C$$

$T = 19.00$

E 4 - Refrig.



$Q = 4082 \times .525(66.2 - 35) = 66,863$ BTU/H

5.57 Tons

$A = \frac{66,863}{\quad} = 22.8$

VCM Scrubber – MEK Solvent

Material Balance. – M/H

		Gas In (1) M	Gas In (1) MF	Solvent In M	Solvent In MF	Solv. Out M	Solv. Out MF	Gas Out M	Gas Out MF
	VCM	7.58	.2743	.2388	.00339	7.7807	.10	.0281	.001857
72.1	MEK	0	0	70.1429		70.0263	.9	.1166	.00577
	N_2	20.05	.7257	—	—	—	—	20.05	.99234
				70.3817				20.2047	

MEK $\log P = 6.97421 - \frac{1209.6}{216+t}$

41.44°C $P = 30.70$ mm Hg.

MF MEK in Gas Out. $\frac{30.70}{5320} = \frac{P^v}{\pi} = .00577$

.0281
20.05
20.0881 .99423
MEK .11658
20.20468

MEK MP = -85.9
BP = 79.6

Solvent Loss = $72.1 \times .1166$
= 8.41 lb./H.

Assuming Ideal Solutions using Same Absorber as TCE + Fuel Oil #1

15.4 ft. 1" P.11 Rings,

MEK-VCM Stripper

	Feed		Ovhd.		Bo Hs	
	M/H	MF	M/H	MF	M	MF
VCM	7.781	.1	7.5428	.9999	.2382	.00339
MEK	70.026	.9	.000754	.0001	70.0252	.99661
	77.807		7.5436		70.2634	

Assume Stripping at 78 lbs, 102.9 PSIA, 5320 mm Hg.

Top Temp. = 52.65°C

P_{MEK} = 296.73 mm Hg. α top = $\frac{5320}{296.73}$

= 17.93

Bo Hom π = 5340

$$\text{Log } 5340 = 6.97421 - \frac{1209.6}{t + 216} = 3.7275$$

$$\frac{1209.6}{t + 216} = 3.24671$$

$$t = 156.56\,^{\circ}C$$

$$\text{Log } P_{VCM} = 6.49712 - \frac{783.4}{230 + 156.56}$$

90°F, 62.5 PSIA Cond.
3300 mm Bo Hg
134.0°C

P_{VCM} = 29,549

α Bo Hom = $\frac{29,549}{5340}$

= 5.533

Feed Plate temp. Guess 104°C.

P_{MEK} = 1,563.9

P_{VCM} = 14,178 $\frac{14,178}{1563.9} = 9.066$

Avg α above = $(17.93 \times 9.060)^{\frac{1}{2}} = 12.745$

Avg α below = $(9.060 \times 5.533) = 7.080$

overall α = $(17.73 \times 5.533)^{\frac{1}{2}} = 9.90$

Stripper Mat'l Balance - Wt Basis

	F	O	B
VCM	486.3	471.4	14.89
MEK	5048.9	.054	5048.8
	5535.2	471.454	5063.7

Top Conditions

$\rho_V = 1.0215$ lb/ft³

P_L MEK = .805 $\frac{\#}{ft^3}$

$\rho_L = 53.35$

$V_a = .2\left(\frac{53.35}{1.0215}\right)^{\frac{1}{2}} = 1.445$ ft./sec.

7.5436 = 26.19 M/H

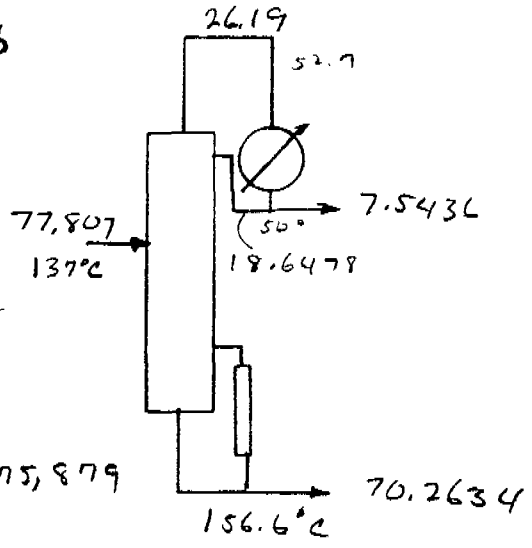
Boil-up = 3.4712 x 471.454 = 1636.9 #/Hr.

@ 2 x R min

$A = \frac{1636.9}{3600 \times 1.0215 \times 1.445} = .308$ ft²

x 1.5 = .462

$\phi = .767$ say 12" ϕ



Condenser Duty. — 50°C Datum

		BTU/H
LH	1636.9 x 66.2 x 1.8 =	195,053
	1636.9 x .23 x 2.7 x 1.8 =	1,827
		196,882

Heat In with Feed

5535.2 x .549 (137 − 50) x 1.8 = 475,879

Heat Out in BoHs

5063.7 x .58 (156.6 − 50) 1.8 = 563,541

Reboiler Duty = 196,882 − 475,879 + 563,541 = 284,544 BTU/Hr.

326.8 #/H Stm

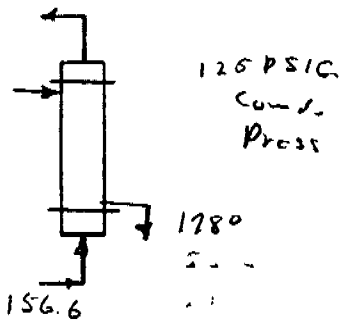
E1 - Condenser

$$A = \frac{196,882}{100 \times 21.1 \times 1.8} = 51.8 \text{ ft}^2$$

say 60 ft² ⟹ 16% for Pours

E2 - Reboiler

158°



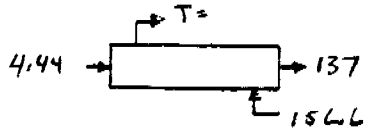
$\Delta T_m = 20.7°C$

$$A = \frac{284,544}{75 \times 20.7 \times 1.8} = 101.8 \text{ ft}^2$$

50 : 152.7 ft²

say 160 ft²

E-3 Boils/Feed HE



5535.2 × .549 (137 − 4.44) 1.8 = 725,066 BTU/H

5063.7 × .58 (156.6 − T) 1.8 = 725,061

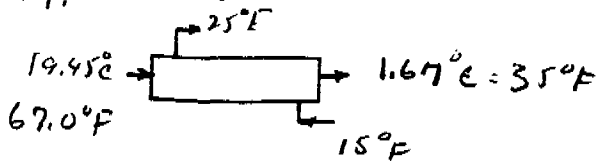
$T = 19.45°C$

$$\Delta T_m = \frac{(19.45 - 4.44) - (156.6 - 137)}{\ln \frac{15.01}{19.6}} = \frac{4.59}{.2669} = 17.20$$

$$A = \frac{725,066}{100 \times 17.20 \times 1.8} = 234 \text{ ft}^2$$ U = 100

say 250

E-4 Refrig. Exch



$Q = 5063.7 \times .54 \times (67.0 - 35) = 87,500$ = 7.29 Tons

$\Delta T_m = \frac{(67-25) - (35-15)}{\ln \frac{42}{20}} = \frac{22}{.7419}$

$= 29.65$

$A = \frac{87,500}{100 \times 29.65} = 29.5\ ft^2$

Case 4

		①		②	M%	③ M		④ M	
N_2	28	20.1	.7256			20.1			
VCM	62.5	7.6	.2744	.000031	.000072	.0004		7.6	.15
Hex	86.2	-		43.07		.1612		43.07	.85
								50.67	

Top $\frac{mG}{L} = \frac{.2588 \times 20.1}{43.07} = .1208$

Bot $= \frac{.4872 \times 17.7}{50.67} = .9664$

Avg. $= .1794$

$$Np = \frac{\log\left[(1-.1794)\left(\frac{.2744-0}{.00062-0}\right)+.1794\right]}{\log \frac{1}{.1794}}$$

$$= \frac{4.052}{.7462} = 5.43$$

Temp. at Bottom

temp. Rise $= 23.2 \times \frac{76.4}{50.67} = 34.98$

$82 + 34.98 = 54.9$ °F
(12.77 °C)

$P_{vem} = 1862.8$ mm
$= 2.451$ Atm.

$= \frac{2.451}{5.03} = .4872$

Case 4.

	Feed		Ovhd.		Bots	M%	Reflux	
VCM	7.6	.15	7.59969		.000081	.000072	8.5142	
Hexane	43.07	.85	.00079		43.06921	.999928	.000885	
							8.5141	

$$\text{Min. Plates Above} = \frac{\ln \frac{7.59969}{.00079} \times \frac{.85}{.15}}{\ln 16.3}$$

$$= \frac{10.906}{2.791} = 3.907$$

$$\text{Min Plates Below} = \frac{\ln \frac{.15}{.85} \times \frac{.999928}{.000072}}{\ln 9.33} = \frac{12.409}{2.232}$$

$$= 5.5597$$

$$9.467$$

Min R. = [illegible] = .5602

Act. = [illegible] = 1.1203

17.6 Theo. P.

16.67 = Rel.

Col. ϕ

Boil-up = 1007.15 #/H

$$A = \frac{1007.15 \times [illegible]}{3600 \times 2 \times .325} = .554219 \text{ ft}$$

D = .8403 ft = 10.09 in.

Case 2.-

Heat Balance on Stripper -

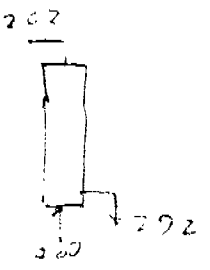
		BTU/#
Feed		
4187.6 x .575 (178-40)	=	332,289
Reboiler	=	149966
Boil up		
1007.2 x 75.1 x 1.5	=	136,146
1007.2 x .195 x 2	=	393
Conv. to Ly		136,539
Bottoms		
3712.6 x .582 (200-46)	=	345,716
		482,255

E1 Condenser

Tons of Refrig = $\frac{136,539}{12000}$ = 11.38

Area = $\frac{136,539}{75 \times 15.5}$ = 93.0 sq. ft.

E2 Reboiler



ΔT = 350 - 292 = 91

A = $\frac{149966}{40 \times 91}$

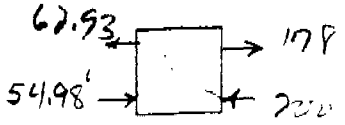
= 41.2 sq. ft.

Steam Cond. = $\frac{149,966}{1195.9}$ = 160.6 #/hr

+ loss = 9[illegible]

Unit 4

E 3



$Q = 4187.6 \times .575 (178 - 55) =$ 296,168

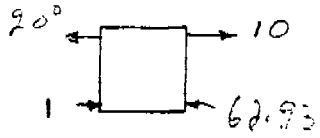
$Q = 3712.6 \times .582 (200 - t) =$ 296,168

$t = 62.93$

$\Delta T_m = \frac{22 - 7.95}{\ln \frac{22}{7.95}} = \frac{14.05}{1.018} = 13.80$

$A = \frac{296,168}{13.80 \times 100} = 214.6$ ft²

E 4. Refrigeration



$Q = 3712.6 \times .52 (62.93 - 20) = 82,878$

$\Delta T_m = \frac{52.93 - 19}{\ln \frac{52.93}{19}} = \frac{33.93}{1.02} = 33.12$

$A = \frac{82,878}{100 \times 33.12} = 25.03$ ft²

Drawer #8

VCM Rec'y Correspondence

BY ________ DATE ________ SUBJECT ________ SHEET NO. ________ OF ________
CHKD. BY ________ DATE ________ Refrigerated Hexane Scrubbing, JOB NO. ________
Stripping 27 PSIA [illegible]

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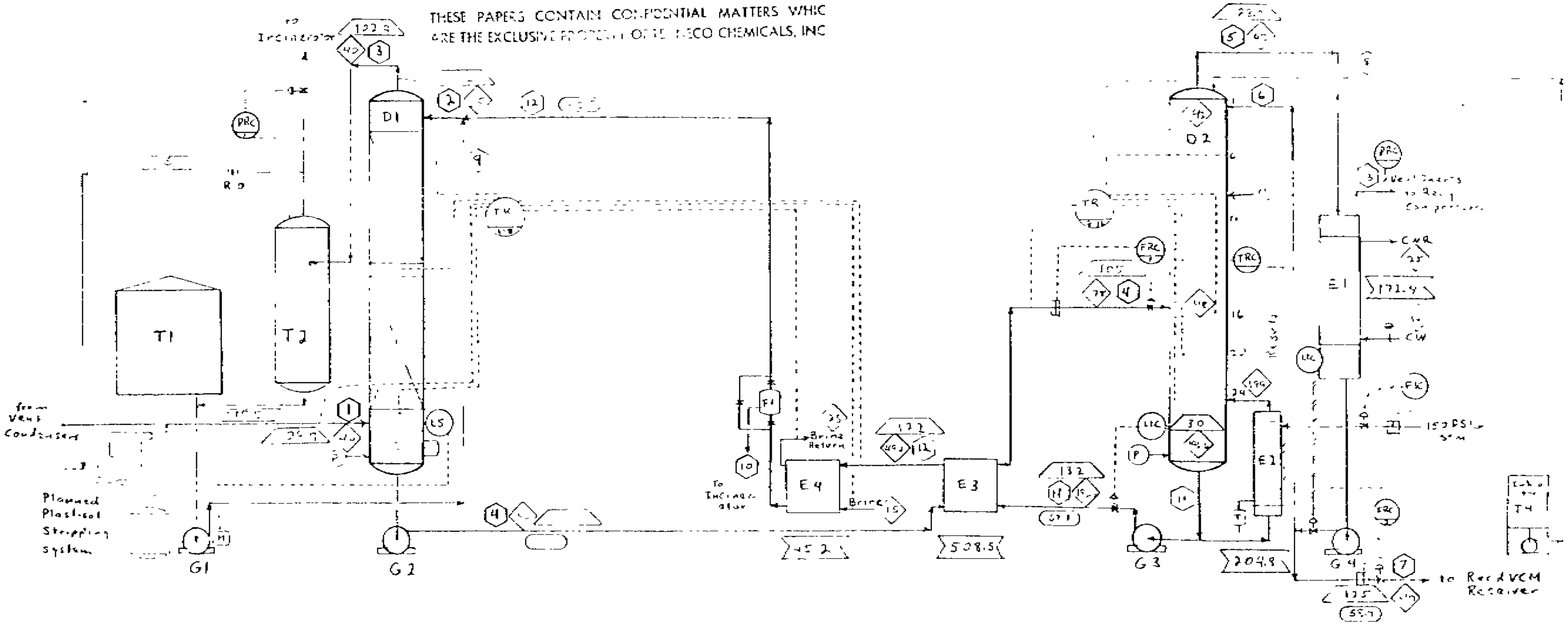


Quantities in lb/H. for Maximum Rate Conditions 8400 Hrs. Yr. Avg = 28.6% of Max

	M.W.	1	2	3	4	5	6	7	8 (1)	9	10 (2)	11	12	13 (3)
VCM	62.5	475	3.78	.475	472.8	1271.7	796.21	474.5	0		.000	3.78	3.28	[illegible]
N_2	28	562.8	0	562.8	0	0	0		0			0	0	[illegible]
n Hexane	86.2	0	5942.8	18.58	5931.0	.1775	.11	.066	.3	20	1.39	5930.5	5929.9	
Inhibitor	-	0	.6	0	.6	0			.03		.0006	.6	.6	
Misc.		0	6.	0	6.	0			-		.005	6	6.0	
Total		1037.8	5955.[illegible]	583.1	6415.4	1271.9	797.31	474.6	.33	20	1.395	[illegible]	5939.7	–

Temp °F
PSIA
lb/ft³
MBTU/H

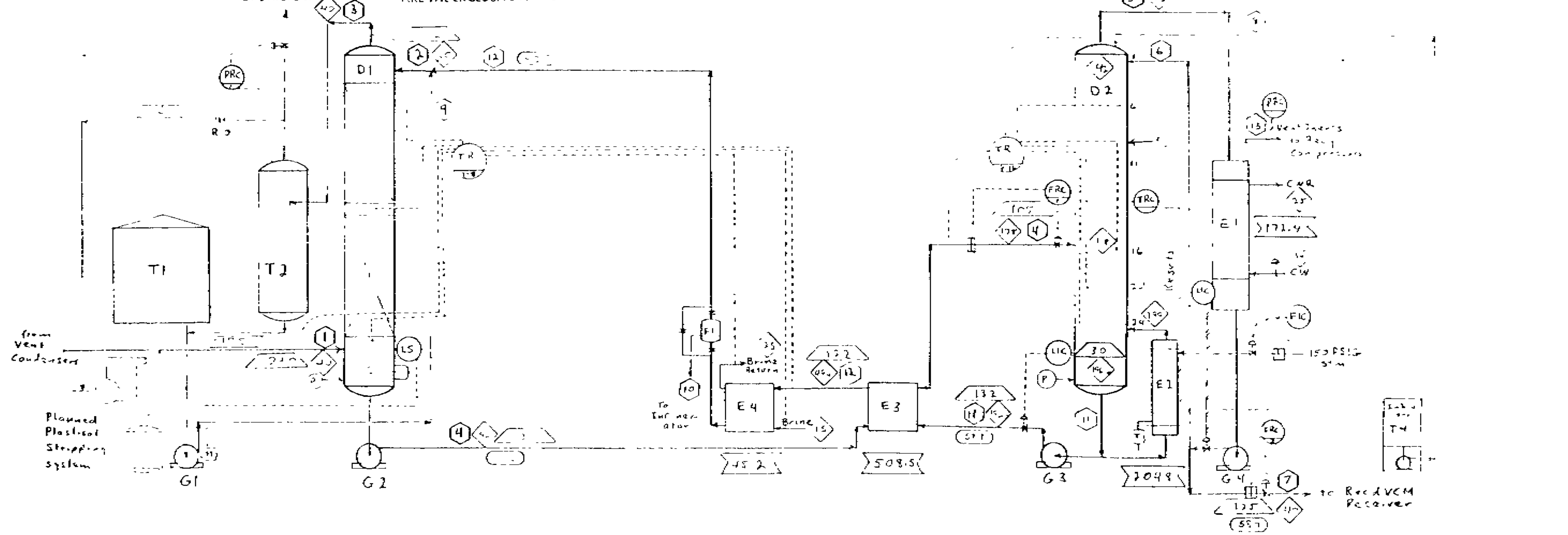
Notes:
(1) Inhibitor guess: Ional (10%) in Hexane. 100 PPM Conc. maintained by 5% replacement per hour. Corrosion inhibitors may also be [illegible].
(2) A small purge of polymers and unknown materials which may accumulate and tend to [illegible] or corrode [illegible].
(3) Except for start-up and shut-down, vent of inerts is negligible.
(4) Hexane in stream (7) [illegible] numerous recycles of [illegible] recovered [illegible], build up [illegible] bounds to [illegible] desirable level, [illegible] acceptable level by [illegible] portion [illegible] to the stripper.

COLORITE 006571

BY ______ DATE ______ SUBJECT ______
CHKD. BY ______ DATE ______ Refrigerated Hexane Scrubbing, Stripping 27 PSIA [illegible] JOB NO. ______

CONFIDENTIAL

THESE PAPERS CONTAIN CONFIDENTIAL MATTERS WHIC ARE THE EXCLUSIVE PROPERTY OF TE[illegible]ECO CHEMICALS, INC



Quantities in lb/H. for Maximum Rate Conditions 8400 Hrs. Yr Avg = 28.6% of Max

	N?	1	2	3	4	5	6	7	8 [1]	9	10 [2]	11	12	13 [3]
VCM	67.5	475	3.78	.475	477.8	1271.7	796.21	474.5	0		.00?	3.28	3.28	?
N_2	25	562.8	0	562.8	0	0	0		0			0	0	
n Hexane	562	0	5944.8	18.55	5931.0	.1775	.11	.066	.3	20	1.35	5930.5	5929.?	
Inhibitor	-	0	.6	0	.6	0			.03		[illegible]	.6	.6	
Misc.		0	6.	0	6.	0			-		.005	6	5.0	
Total		1037.8	5955.?	582.1	6415.4	1271.9	797.31	474.6	.33	20	1.355	[illegible]	5939.7	—

°F, PSIA, lb/ft³, SCFM, MBTU/H

Notes:

(1) Inhibitor guess: Ional (10%) in Hexane. 100 PPM conc. maintained by 5% replacement per hour. [illegible] inhibitors may also be [illegible].

(2) A small purge of polymers and [illegible] which may accumulate and tend to [illegible] control, [illegible].

(3) Except for start-up and shut-down, vent of inerts is negligible.

(4) Hexane in stream (7) [illegible] recovered in [illegible], build up [illegible] tends to [illegible] able [illegible] acceptable level by purging a portion back to the stripper.

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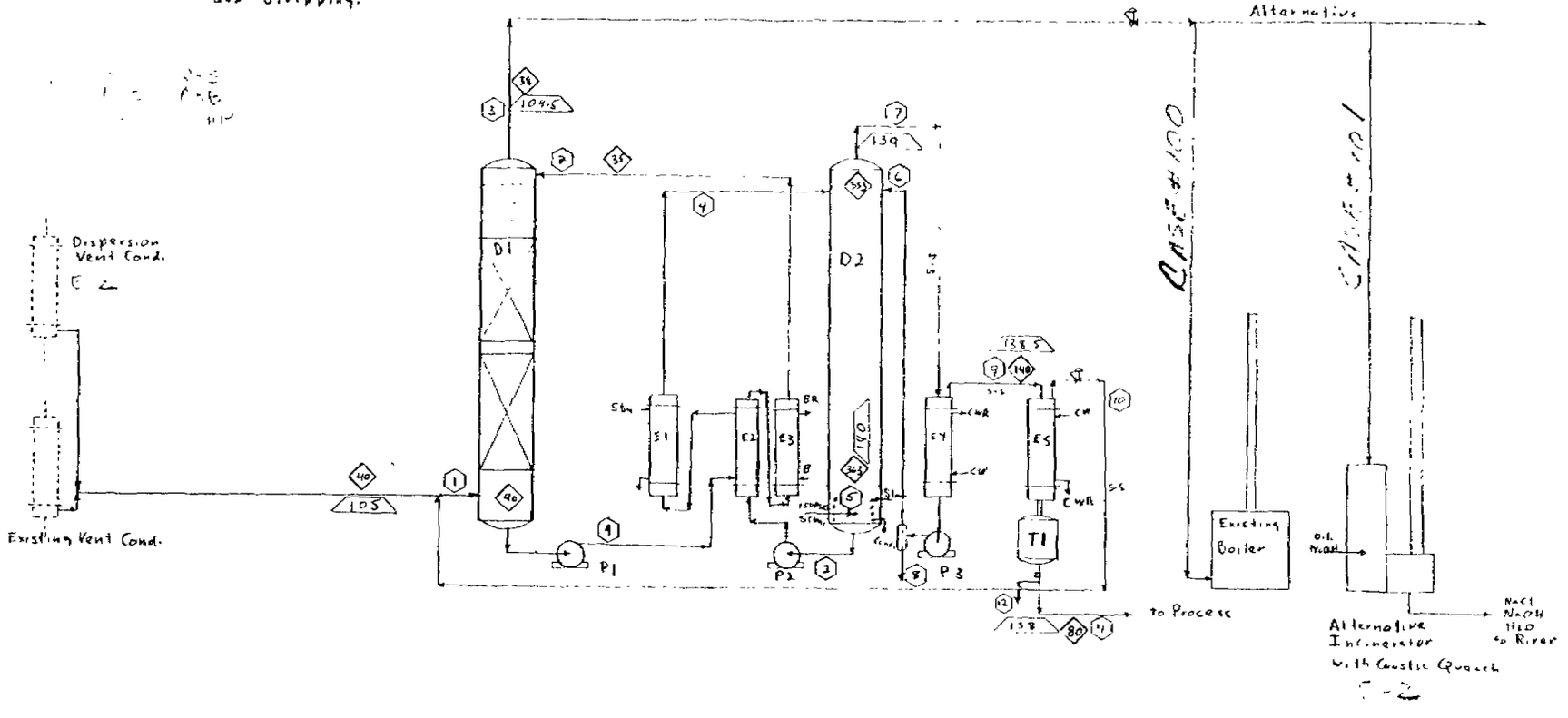
JOB NO

Refrigerated Solvent Absorbing and Stripping.



Pounds Per Hour: Max. Rate Conditions for Design. See Fig. 1 for Rate Fluctuations.

	MW	① Peak	②	③	④	⑤	⑥	⑦	⑧	⑨	⑩	⑪	⑫
VCM	62.5	493	6.04	2.4	496.64		08	470.68	.0001	470.68	0.01	470.65	
N_2	28	561	0	561	0		0	0		0	0.01		
Water	18	0	0	0	0	250	250	250	247	3.8	0....	.05	3.8
Solvent	175	0	12012	0.000	12012		105	105	0	....	0		
Inhibitor	-	0	0	0	?		0	0			0		
Total		1034	12018	563.4	12488	250	106.0	825.7	247.3	473.5	0.01	470.65	

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Case 2.- Cont

Stripper Mat'l. Bal. - Mole Basis

Comp	MW	Feed M	MF	Ovhd M	MF	Bottoms M	M%	Reflux
VCM		7.6	.05	7.599928		.000104	.000072	
Hexane		144.4	.95	.000790		144.4		
				7.600718		144.4		

$$\text{Min. Plates above} = \frac{\log \frac{7.599928}{.000790} \times \frac{.95}{.05}}{\log 163} = \frac{12.12}{2.791} = 4.341$$

$$\text{Min Plates below} = \frac{\log \frac{.05}{.95} \times \frac{144.4}{.000104}}{\log 9.32} = \frac{11.20}{2.332} = \frac{5.015}{9.356}$$

$$\frac{2.82}{1.5} \times 9.356 = 17.58 \text{ Theo}$$

16.5 + Reb.

$$\text{Min } R = \frac{.9999}{11.9 \times .05} = 1.6805$$

Reflux = 2 x R_m = 3.361

Boilup 4.361 x P = 2071.5 #/H

$$A = \frac{2071.5 \times 13}{3600 \times 9 \times .328} = .8769$$

Φ = 1.057 ft = 12.68"

$$N_{IP} = \frac{\log\left[\left(1 - \frac{mG}{L}\right)\left(\frac{\gamma_1 - mK_2}{\gamma_2 - mK_2}\right) + \frac{mG}{L}\right]}{\log \frac{L}{mG}}$$

$$= \frac{\log\left[\left(1 - .0458\right)\left(\frac{.2744 - .0437 + .0000007}{.00002 - .00000}\right) + .0458\right]}{\log \frac{1}{.0458}}$$

$$= \frac{4.117}{1.339} = \boxed{3.074}$$

Case 2-

Heat Balances

Feed $12,922.3 \times .575 (178 - 40)$ = 1,025,383

Reboiler = 414,542

1,439,926

Boil-up

$2,071.5 \times 75.1 \times 1.8$ = 280,025

$2,071.5 \times .195 \times 2$ = 808

Cond. Duty 280,833

Bottoms

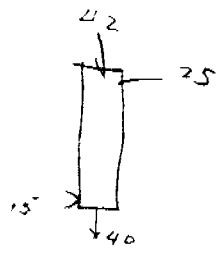
$12,447.3 \times .582 (200 - 40)$ = 1,159,093

1,439,926

E1 Cond.

Tons of Refrig. $= \frac{280,833}{12000} = 23.4$

Area $= \frac{280,833}{75 \times 19.57} = 191.3$ sq. ft.



$\frac{25 - 15}{\ln \frac{25}{15}} = 19.57$

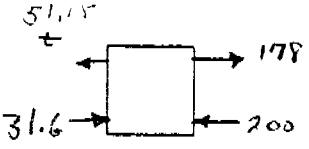
E2 Reboiler

$\Delta T = 91$

$A = \frac{414,542}{40 \times 91} = 113.9$ sq. ft.

Steam $= \frac{414,542}{1191 - 961} = 1803.8$ #/h + loss

E3



$12,922.3 \times .57 (178 - 31.6) = 1,078,340$ BTU/H

$12,447.3 \times .582 (200 - t) =$

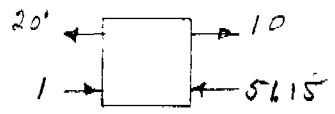
$t = 51.15$

$A = \frac{1,078,340}{100 \times 20.75} = 519.7$ sq. ft.

$\Delta t_m = \frac{22 - 19.55}{\ln \frac{22}{19.55}} = \frac{2.4530}{.11822} = 20.75$

Case 2.

E-1



$$\Delta t_m = \frac{41.15 - 19}{\ln \frac{41.15}{19}} = \frac{22.15}{.7725} = 28.66$$

$$Q = 12,000 \times .52 (51.15 - 20) = 201,616 = 16.807 \text{ tons Refrig.}$$

$$A = \frac{201,616}{100 \times 28.66} = 70.35 \text{ sq ft.}$$

G-2 Circ. Pump

$$\frac{12,417\frac{1}{2} \times 7.48}{36.8\frac{1}{2} \times 60} = 42.2 \text{ gpm.}$$

ENGINEERING ESTIMATE
SUMMARY SHEET

ESTIMATE NO. _____ LOCATION Burlington

ESR NO. 509 DATE 3/24/77

PROJECT TITLE VCM Recovery – Case XVI – Rev. 1

ASSET CAT.	DESCRIPTION	MATERIAL $	LABOR $	SUB-CONTRACT $	TOTAL $M
A.	Refrigerated Absorbing & Stripping Equipment				42
	Installation Factor				x 6
	Subtotal				252
			Esc.	252 x 159 = 25 / 1592	
	Other Equipment	$3.52M	x1.1	.75	
	C-1 A&B				42
	C-3 A&B				30
	E-2				11
	S-2 A&B				170
	Fuel Oil & Caustic Storage				15
	Subtotal				268
	Installation Factor INCINE	170/268 x 1340 = 850			x 5
	Subtotal	Esc. 850/1592 x 158 = 84		934	1,340
	Escalation				153
	Total - Process Equip.				1,750
B	Utilities – same as Case XV				150
C.	Methyl Chloride				100
	99/261 x 1340 = 490				
	490/1592 x 159 = 48.6				
	534 x 1.7 = $.614M				
	GRAND TOTAL				2,000

APPROVALS

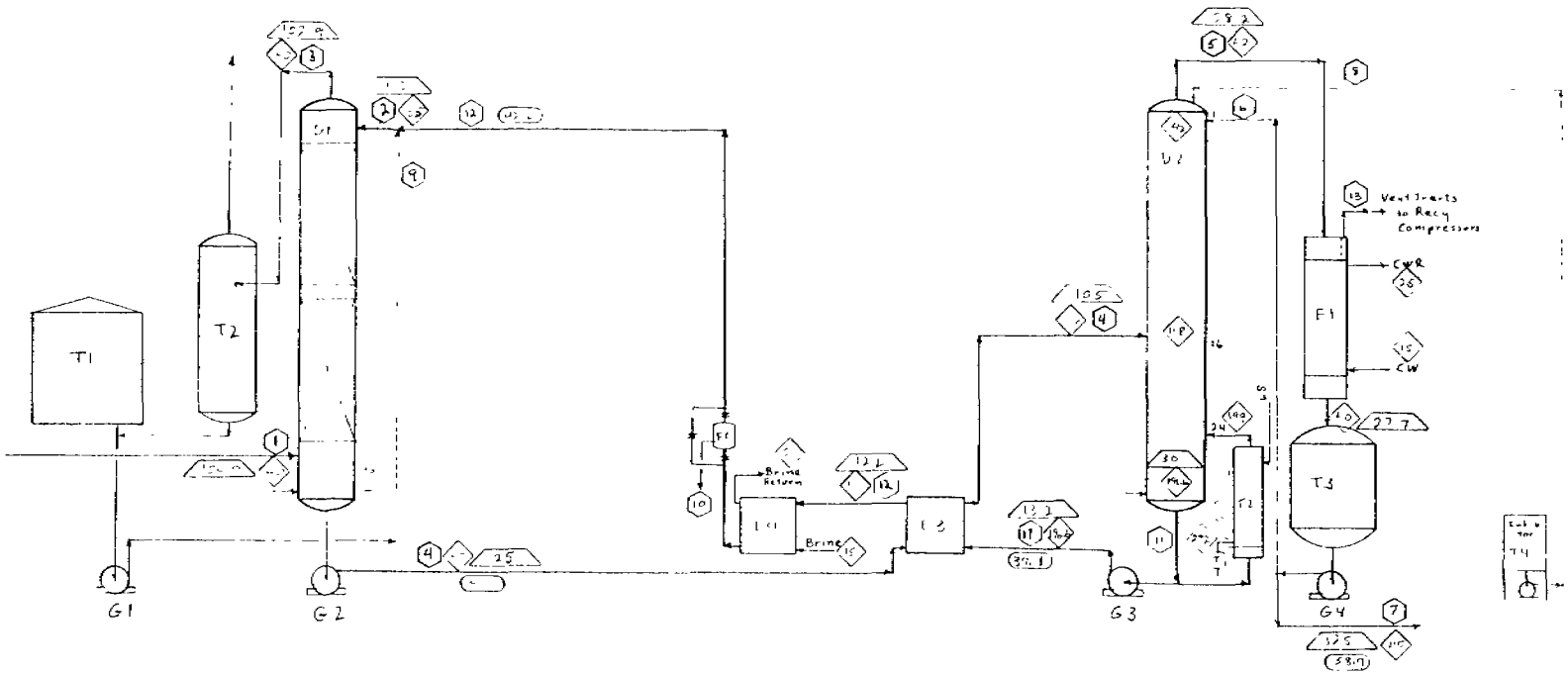
PREPARED BY W. G. Wermuth PROJECT MGR. _____

COST ENGINEER _____ DIR. OF PROJECTS _____

MGR. OF GEN. ENGR. _____ DIR. OF ENGINEERING _____

CHKD. BY_____ DATE_____ Refrigerated Isobutane Scrubbing JOB NO._____

Stripping



Quantities in lb/Hr. for Maximum Rate Conditions 8400 Hrs/Yr. Avg = 28.6% of Max

	MW	1	2	3	4	5	6	7	8 (2)	9	10 (3)	11	12	13 (3)
VCM	62.5	475		475				475						
N_2	28			562.8										
n Hexane	86			18.55	592	705			15	3				
Inhibitor	-			6.					63					
Misc.														
Total	113.8		595											

°F
PSIA
lb/ft³

Notes.

(1) Inhibitor guess: Isonel (10%) in Hexane 100 PPM conc. maintained by 5% replacement per hour. Corrosion inhibitors may also be req'd.

(2) A small purge of polymers and unknown materials which may accumulate and tend to foul or corrode equipment.

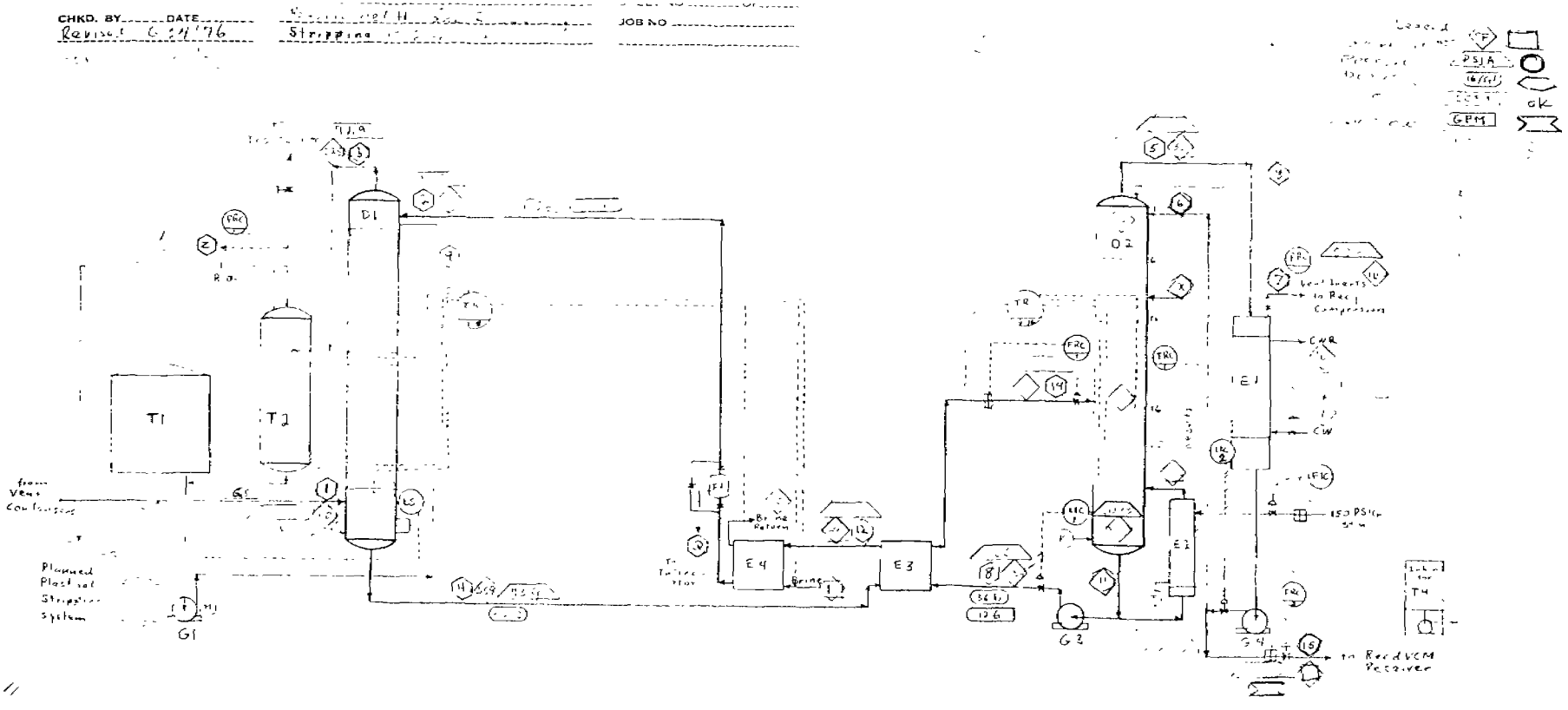
(3) Except for start-up and shut-down, vent of inerts is negligible

COLORITE 006590

CHKD. BY _____ DATE _____ JOB NO _____

Revised 6-11-76

Stripping

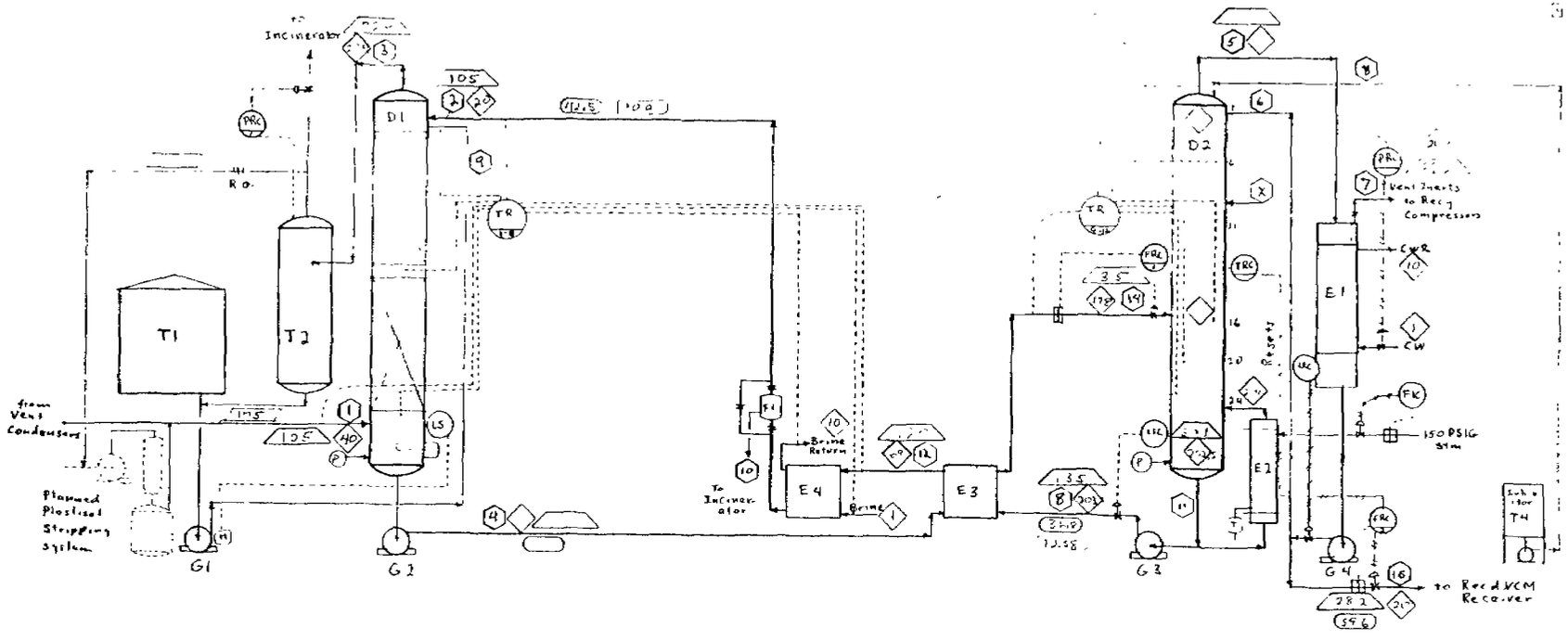


N_2	
CH_3Cl	
VCM	
VAM	
nHexane	
Water	
Total	

COLORITE 006591

CHKD BY_____ DATE_____ Refrigerated Hexane Scrubbing, Stripping JOB NO._____

Legend



Quantities in lb/H for Maximum Rate Conditions 8400 Hrs./Yr. Avg = 28.6% of Max

	MW	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)	(9)	(10)	(11)	(12)	(13)	(14)	(15)
N_2	28	562.8	0													
VCM	62.5	4175	.0063													
n Hexane	862	0														
VAM	86.09	.05														
CH_3Cl	50.5	10	0	.0001												
Water	18	.11														
Total																

Notes:

(4) Inhibitor guess: Ionol (10%) in Hexane 100 PPM conc. maintained by 5% replacement per hour. Corrosion inhibitors may also be req'd.

(2) A small purge of polymers and unknown materials which may accumulate and tend to foul or corrode equipment.

(X) Hexane in stream (15) may, after numerous recycle of VCM recovered in the condensers, build up to an undesirable level. It could be reduced by purging a portion back to the stripper.

COLORITE 006592