MALEIC ANHYDRIDE
AP-42 Section 5-24
Reference Number
2

DRAFT

EMISSIONS CONTROL OPTIONS FOR THE SYNTHETIC ORGANIC CHEMICALS MANUFACTURING INDUSTRY

EPA Contract No. 68-02-2577

Maleic Anhydride - Product Report

J. F. Lawson

Prepared for
Emission Standards and Engineering Division
Office of Air Quality Planning and Standards
ENVIRONMENTAL PROTECTION AGENCY
Research Triangle Park, North Carolina

March 1978

This report contains certain information which has been extracted from the Chemical Economics Handbook, Stanford Research Institute. Wherever used, it has been so noted. The proprietary data rights which reside with Stanford Research Institute must be recognized with any use of this material.

HYDROSCIENCE INC. • 9041 EXECUTIVE PARK DRIVE • KNOXVILLE, TENNESSEE [illegible] • [illegible]
EMERSON, NJ • WESTWOOD, NJ • WALNUT CREEK, CA • ARLINGTON, TX

TABLE OF CONTENTS

TABLE OF CONTENTS
(Continued)

LIST OF TABLES

LIST OF FIGURES

APPENDICES

ABBREVIATIONS AND CONVERSION FACTORS

EPA policy is to express all measurements in agency documents in metric units. Listed below are the International System of Units (SI) abbreviations and conversion factors for this report.

To Convert From	To	Multiply By
Pascal (Pa)	Atmosphere (760 m Hg)	9.870 x 10^{-5}
Joule (J)	British thermal unit	9.480 x 10^{-4}
Degree Celsius	Degree Fahrenheit	(C° x 9/5) + 32
Meter (m)	Foot	3.28
Meter (m^3)	ft^3	3.531 x 10^{1}
Meter (m^3)	Gallon (U.S. liquid)	2.643 x 10^{2}
Meter/second (m^3/s)	Gallon (U.S. liquid)/min	1.585 x 10^{4}
Watt (W)	Horsepower (electric)	1.340 x 10^{-3}
Meter (m)	Inch	3.937 x 10^{1}
Pascal (Pa)	Pound-force/inch^2 (psi)	1.450 x 10^{-4}
Kilogram (kg)	Pound-mass	2.205
Joule (J)	Watt-hour	2.778 x 10^{-4}

Standard Conditions

60°F = 15.6°C

1 atmosphere = 101,325 Pascal

PREFIXES

Prefix	Symbol	Multiplication Factor	Example
G	giga	10^{9}	1 GPa = 1 x 10^{9} pascals
M	mega	10^{6}	1 MW = 1 x 10^{6} watts
k	kilo	10^{3}	1 km = 1 x 10^{3} meters
m	milli	10^{-3}	1 mV = 1 x 10^{-3} volt
u	micro	10^{-6}	1 ug = 1 x 10^{-6} gram

I. PREFATORY AND INTRODUCTORY MATERIAL

(This section to be supplied by EPA)

II. INDUSTRY DESCRIPTION

A. Maleic Anhydride

Maleic anhydride (MA) production was selected for the following reasons: (1) the total estimated emissions of volatile organic compounds (VOC) are relatively high; (2) the predominant manufacturing process emits large quantities of benzene, which was listed as a hazardous pollutant by the EPA in the Federal Register on June 8, 1977; and (3) the product growth is expected to be higher than the industry average.

MA is solid at ambient conditions (see Appendix A for pertinent physical properties) although it generally exists in the process as a liquid (molten MA) or as maleic acid. The predominant emission, benzene, however, is a volatile liquid at ambient conditions, but is emitted as a gas.

B. MA Usage and Growth

Table I shows MA end uses and expected growth rate. The predominant end use is the production of unsaturated polyester resins, which go into reinforced plastic applications such as marine craft, building panels, automobiles, tanks, and pipes.

The U.S. MA production capacity for 1977 was reported to be 236,000 megagrams (Mg) with only 56% of this capacity currently being utilized.[1,2] Assuming an 11% annual growth in MA consumption, production will reach 95% of present capacity by 1982. No shortage of benzene, the major raw material, is expected during this period.[2]

C. Domestic Producers

As of 1977, there were eight producers of MA in the U.S. with ten plants. Table II lists the producers and the processes being used; Figure 1 shows the plant locations. Approximately 83% of the 236,000 Mg/yr domestic capacity is based on the oxidation of benzene. Oxidation of n-butane

TABLE I

MALEIC ANHYDRIDE USAGE AND GROWTH[1]

End Use	% Production	Average Annual % Growth (1974-1980)[(A)]
Unsaturated Polyester Resins	51.1	9.5
Fumaric Acid	6.4	2.0
Agricultural Chemicals	10.0	6.5
Alkyd Resins	1.3	0.0
Lubricating Additives	7.8	8.0
Copolymers	5.3	6.0
Reactive Plasticizers	3.6	5.0
Maleic Acid	3.8	7.0
Chlorendic Anhydride and Acid	1.1	10.0
Surface-Active Agents	2.9	5.0
Other	6.7	2.0
	100.0	

(A) These numbers include a decline in MA consumption of 27% in 1975. The annual MA growth rate from 1976-1982 is now projected to average 11%. Polyester resins are expected to continue as a growth leader.[2]

TABLE II

MALEIC ANHYDRIDE CAPACITY[1]

		Capacity - 1977 10^3 Megagrams (Mg)	Process
1.	Amoco, Joliet, IL	27	2
2.	Ashland, Neal, WV	27 ✓	1
3.	Koppers, Bridgeville, PA	15 ✓	1
4.	Koppers, Chicago, IL	5	3
5.	Monsanto, St. Louis, MO	48 ✓	(80%)1 - (20%)2
6.	Denka (Petrotex), Houston, TX	23 ✓	1
7.	Reichhold, Elizabeth, NJ	14 ✓	1
8.	Reichhold, Morris, IL	~~27~~ 20 ✓	1
9.	Tenneco, Fords, NJ	12 ✓	1
10.	U. S. Steel, Neville Island, PA	38 ✓	1
	TOTAL	236	

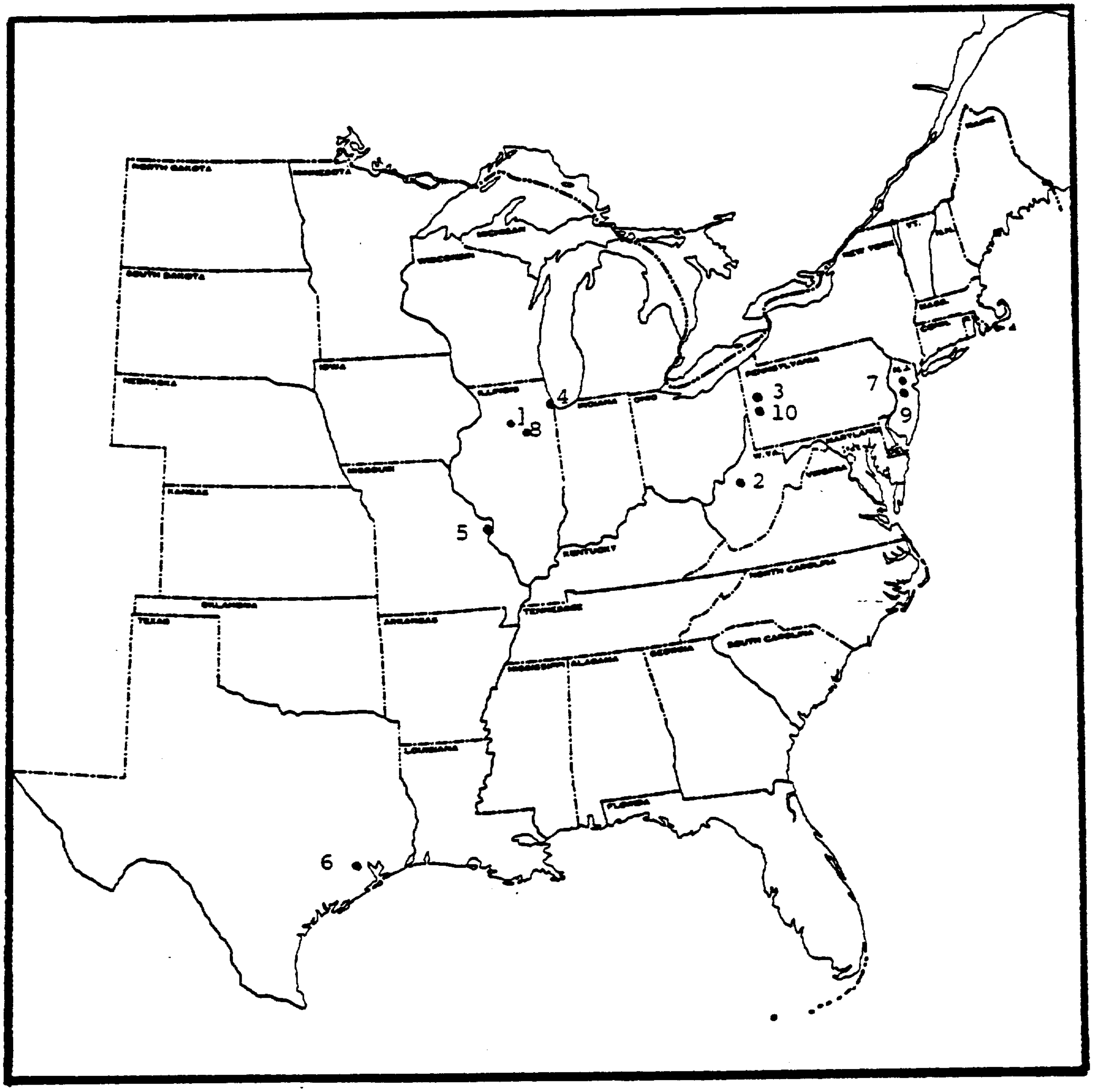
'77 by [illegible]

.72
36.0
3

Processes:

(1) Oxidation of benzene
(2) Oxidation of n-butane
(3) By-product of phthalic anhydride manufacture

lb x 454



MANUFACTURING LOCATIONS

OF

MALEIC ANHYDRIDE

Key

1. Amoco, Joliet, IL
2. Ashland, Neal, W. VA
3. Koppers, Bridgeville, PA
4. Koppers, Chicago, IL
5. Monsanto, St. Louis, MO
6. Denka (Petro-Tex), Houston, TX
7. Reichhold, Elizabeth, NJ
8. Reichhold, Morris, IL
9. Tenneco, Fords, NJ
10. U.S. Steel, Neville Island, PA

FIGURE 1

accounts for another 15% of capacity, and the remaining 2% is from phthalic anhydride production which gives MA as a by-product.[1] The projected growth rate for the n-butane oxidation process through 1982 is 24.3%, primarily through conversion, as compared to only 9.1% for the benzene oxidation process. Data regarding the economic incentives for switching to n-butane oxidation are not available. No growth in the quantity of MA recovered during phthalic anhydride production is expected.[2]

Work began in 1960 to develop a catalyst suitable for producing MA from butane/butene (C_4) streams available from naphtha cracking. This effort was curtailed during the 1961-1967 period when the MA market was depressed and low-cost benzene was available. In 1967, demand for MA increased, and work was renewed in Japan by Kasei Mizuishima. In 1974, announcements concerning the production of MA from C_4's were made by Petro-Tex, Chem Systems, BASF, Bayer, Alusuisse/UCB, and Mitsubishi.[3] Presently, Amoco and Monsanto are producing MA from a n-butane feedstock.[1] The main drawback of the n-butane process is the unavailability of a catalyst that provides competitive yields.[4]

Producing Companies

1. Amoco Chemicals Corporation
   Amoco has the only U.S. plant totally dedicated to the n-butane process.[1] Still in the start-up stage at the time of this report, it will have an annual capacity of 27,000 Mg and is expandable to 41,000 Mg.[4]

2. Ashland Chemical Company
   The Ashland facility is a new benzene based plant with an annual capacity of 27,000 Mg and is expandable to 41,000 Mg.[1] This plant can be switched from benzene to n-butane feedstocks.[4]

3. Koppers Company, Inc.
   Much of the MA from their 15,000 Mg/yr benzene oxidation plant is used captively to produce unsaturated polyester resins and alkyd

resins. Their Chicago facility can recover 5,000 Mg of MA per year from the effluent of their phthalic anhydride plant which started in 1975.[1]

4. Monsanto Company
   Monsanto at 48,000 Mg/yr is the largest producer of MA. Some MA is consumed captively to produce fumaric acid, maleate/fumerate esters, styrene copolymers, and ethylene-maleic anhydride copolymers.[1]

5. Denka USA
   Their 23,000 Mg/yr Houston facility was designed by Scientific Design Company, Inc., and was purchased from Petro-Tex Chemical Corporation July 1, 1977.[5] They have a current permit from the Texas Air Control Board to operate a n-butane reactor.[6]

6. Reichhold Chemicals, Inc.
   Reichhold's combined production from both plants is 34,000 ~~41,000~~ Mg/yr, some of which is used captively to produce unsaturated polyester resins, alkyd resins, and plasticizers.[1]

7. Tenneco Chemicals, Inc.
   A small part of their 12,000 Mg/yr MA production is used captively to produce fumaric acid, dibutyl maleate, and dodecanyl-succinic anhydride.[1]

8. United States Steel Corporation
   Their MA capacity was recently expanded to 38,000 Mg/yr.[4] Some of their MA production is used captively to produce fumaric acid, dibutyl maleate, and dioctyl maleate.[1]

9. Allied Chemicals Corporation
   Allied ceased production of MA at their Moundsville, West Virginia, 900 Mg/yr capacity plant in 1974.[1]

The expansion capabilities of 14,000 Mg each for Amoco and Ashland represent a potential nationwide capacity of 263,000 Mg.[4]

SECTION II

REFERENCES

1. Blackford, J. C. CEH Marketing Research Report on Maleic Acid. July 1976. Chemical Economics Handbook, Stanford Research Institute, Menlo Park, California.

2. Unpublished EPA Survey and Ranking Information.

3. Technical Week, European Chemical News, April 5, 1974.

4. "Maleic Makers Build on Hopes for Polyester", Markets Section, Chemical Week, February 2, 1977. pp. 37-38.

5. Personal communication between J. F. Lawson, Hydroscience, Inc., and R. E. Hinkson, Denka USA, November 17, 1977.

6. Permit Exemption Request from R. D. Pruessner, Petro-Tex Chemical to Mr. Charles Barden, Texas Air Control Board, February 23, 1976.

III. PROCESS DESCRIPTIONS

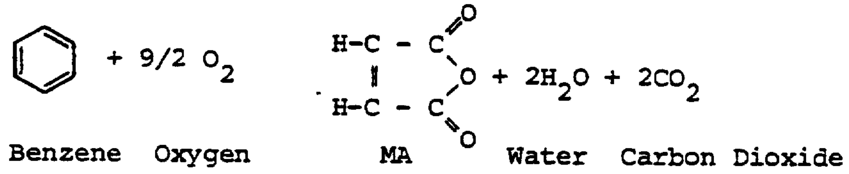
A. Introduction

As previously discussed, the two major processes used to manufacture maleic anhydride (MA) in the United States are benzene oxidation and butane oxidation. A small amount of MA is recovered as a by-product of phthalic anhydride production. The only significant foreign process for MA production not used in the U.S. starts with a butene mixture feedstock. This process is operated in France[1] and Japan.[2] There are no known plans to introduce this process domestically.

B. Benzene Oxidation Process

1. Basic Process

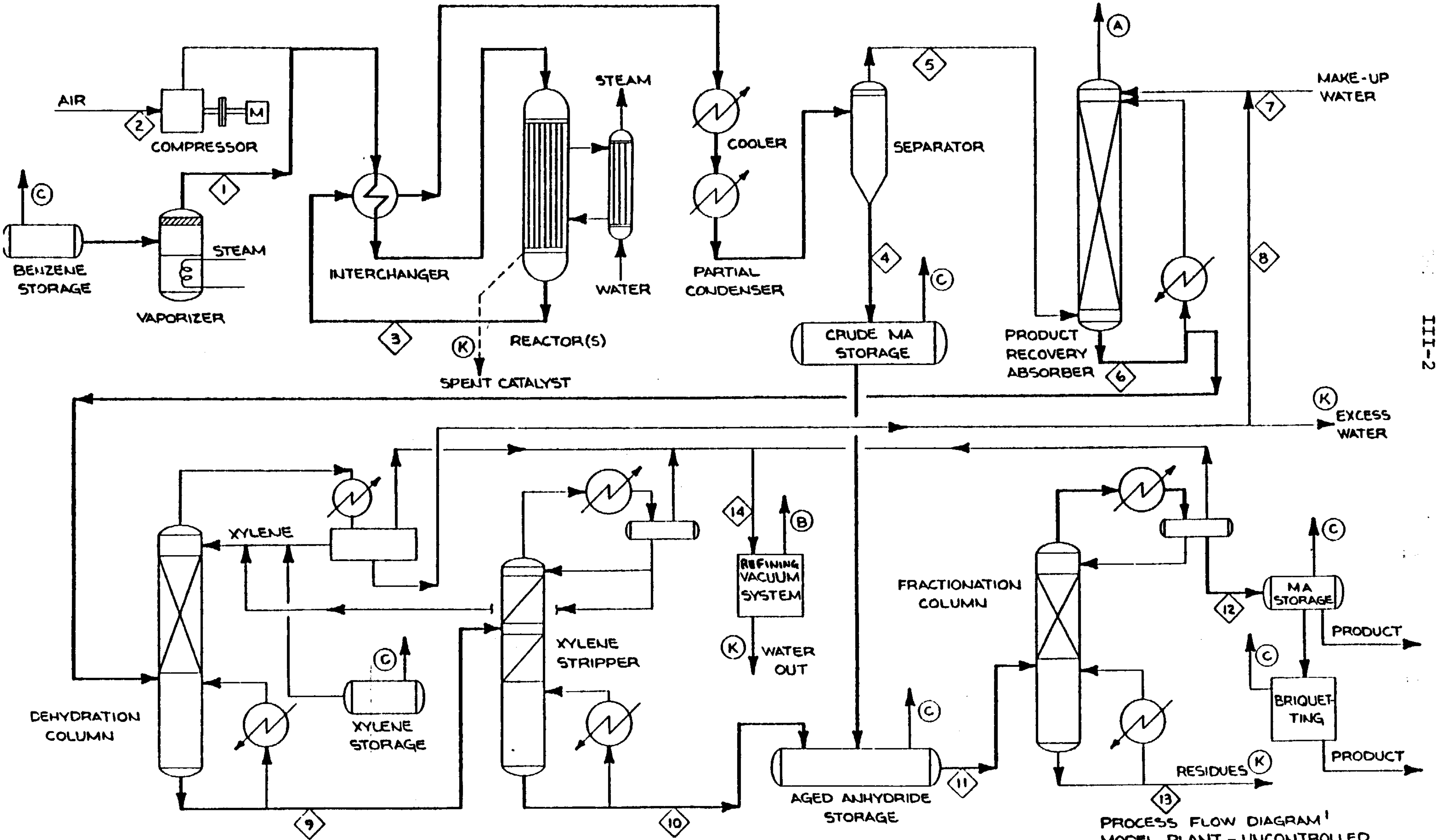
MA is produced by the following chemical reaction:



The process flow diagram shown in Figure 2 represents a typical process. The typical process is continuous, although some plants operate batchwise. The emissions in either case are judged to be equivalent.[3]

A mixture of vaporized benzene and air enters a tubular reactor where the catalytic oxidation of benzene is carried out at a temperature of 350-400°C. The catalyst contains approximately 70% vanadium pentoxide supported on an inert carrier; most of these systems also contain 25-30% molybdenum oxide. The reaction is highly exothermic, releasing 24.4 MJ/kg of benzene reacted, with the excess heat being used to generate steam. MA yields range from 60-67% of theoretical.[3]

The reactor feed mixture contains an excess of air because benzene is explosive in air at concentrations above 1.5 volume percent. The resulting large volume of reactor exhaust (Stream 3) dictates the



PROCESS FLOW DIAGRAM[1]
MODEL PLANT – UNCONTROLLED

Ⓒ - STORAGE AND HANDLING EMISSIONS
Ⓙ - FUGITIVE EMISSIONS – OVERALL PLANT
Ⓚ - SECONDARY EMISSION POTENTIAL

Hydroscience

FIGURE 2

MALEIC ANHYDRIDE-BENZENE
OXIDATION PROCESS

1 of 1

9107

size of subsequent product recovery equipment. The stream passes through a cooler, partial condenser, and separator in which 40% of the MA is condensed and separated as crude MA (Stream 4).[4] The remaining vapor (Stream 5) enters the product recovery absorber where it is contacted with water or aqueous maleic acid. The absorber product (Stream 6) is a 40 wt. % aqueous solution of maleic acid. The absorber vent (Vent A) exhausts to the atmosphere or is directed to an emission control device.[3]

The 40% maleic acid (Stream 6) is dehydrated by azeotropic distillation with xylene. Any xylene retained in the crude MA (Stream 9) is removed by the xylene stripping column, and the crude MA (Stream 10) from this column is combined with the crude MA from the separator (Stream 4).[5]

Crude MA is aged, which causes any color-forming impurities to polymerize. After aging, the crude MA (Stream 11) is fed to the fractionation column which yields molten MA as the purified overhead product (Stream 12). A small percentage is taken an additional step for sale as briquets. The fractionation column bottoms containing the color-forming impurities are removed as liquid residue waste (Stream 13). This stream either becomes part of the untreated effluent or is fed to a liquid incinerator.[3]

The vacuum lines from the dehydration column, xylene stripper, and fractionation column are joined to the vacuum system (Stream 14). The refining vacuum system vent (Vent B) can exhaust to the atmosphere, recycle to the product recovery absorber (Stream 5), or be directed to an emission control device. Water from the vacuum system can be recycled as make-up water (Stream 7) or join the liquid residue waste (Stream 13).[3]

Essentially all process emissions will exit through the product recovery absorber (Vent A). These emissions will include any unreacted benzene which can constitute 3-~~7~~10% of the total benzene feed.[6] The only other process emission source is the refining vacuum system vent (Vent B), which can contain small amounts of MA and xylene.

Fugitive emissions throughout the process can contain benzene, xylene, MA, and maleic acid. Corrosion problems due to leaks caused by maleic acid can increase fugitive emissions. As with most organic chemical processes, leaks into cooling water could occur allowing volatile organic compounds (VOC) to escape as a fugitive emission.

Storage and handling emission sources (labeled C on Figure 2) include benzene, xylene, MA, and crude MA storage, plus emissions from the briquetting operation.

There are four potential sources of secondary emissions (labeled K on Figure 2): (1) spent reactor catalyst, (2) excess water from the dehydration column, (3) vacuum system water, and (4) fractionation column residues. The small amount of residual organics in the spent catalyst after washing has low vapor pressure and produces no significant emissions. Xylene is the principal organic contamination in the excess water from the dehydration column and the vacuum system water. Residues from the fractionation column are relatively heavy organics with a molecular weight greater than 116 and produce no significant secondary emissions.

2. Process Variations

   a. In place of the partial condensation system (cooler, partial condenser, and separator) shown in Figure 2, a so-called switch condenser system can be incorporated. This utilizes a series of condensers which are alternately cooled to freeze solid MA on the surface and then heated to melt the MA for pumping to crude MA storage. Switch condensing can remove up to 60% of the MA from the process compared to 40% for the partial condensation system.[7] The removal of additional MA would allow the size of the product recovery absorber to be reduced and would slightly reduce the maleic acid loss through the product recovery absorber (Vent A).

   b. Xylene is the only known azeotropic agent currently being used for dehydration. Several other agents can be used, including isoamyl butyrate, di-isobutyl ketone, anisole, and cumene.[3]

c. A vacuum evaporation system, which replaces the dehydration column and xylene stripper, is used by at least one plant to remove water and dehydrate the maleic acid to form MA.[7] Since an azeotropic agent is no longer required, xylene is eliminated from process emissions.

C. n-Butane Oxidation

All process data concerning the n-butane oxidation process are currently proprietary and unavailable. A benzene oxidation process can be converted to n-butane oxidation by changing the catalyst system; this conversion can be done for much less than the cost of a new plant. However, the converted process would be less efficient than a new n-butane process because reaction conditions are not optimum. The lowered efficiency might be serious enough to warrant other major design changes.[8]

A major advantage of this process is that there are no benzene emissions. Other VOC emissions should not differ to a great extent from the benzene oxidation process.[6]

D. Phthalic Anhydride By-Product Process

Phthalic anhydride is manufactured from naphthalene and ortho-xylene. Maleic anhydride is recovered as a by-product from the plant effluent.[3] The emissions associated with MA recovery are believed to be insignificant and are not being investigated at this time. Relevant data will be obtained during the study of phthalic anhydride production.

E. Foreign Processes (Mixed Butenes)

1. Basic Process

The only significant foreign deviation from benzene oxidation for MA production is a process using feedstocks of 65-80% n-butenes with the remainder mostly butanes or isobutene. The general process description is very similar to that shown in Figure 2 for benzene oxidation and, therefore, is not repeated.[1,2]

The exhaust from the main process vent contains unreacted butane, butene, carbon monoxide, and various secondary products. Except for the absence of benzene, the emissions should be about the same as for the benzene oxidation process.[6]

2. Process Variations

The most significant process variation is the use of a fluidized catalyst bed rather than a fixed bed. This provides good temperature control within the bed and, thus, allows optimum ratios of butene/air. In contrast, optimum benzene/air ratios cannot be used with fixed bed systems because excess air is necessary to stay below the explosive range. The reduction of excess air in the fluidized bed feed will reduce emissions from the product recovery absorber; however, the product yields with the fluidized bed are not as good as with the fixed bed.[9]

SECTION III

REFERENCES

1. Lenz, D. and M. De Bouille, "The Bayer Process for the Production of Maleic Anhydride from Butenes", Rev. Assoc. Fr. Tech. Pet., Volume 236, No. 20-3, p. 17 (1976).

2. Shinji, Uemura and Kamimura Shiego, "Production of Anhydrous Maleic Acid from C_4 Distillate", Petroleum Academic Association Journal, Volume 16, No. 8 (1973).

3. Blackford, J. C. CEH Marketing Research Report on Maleic Acid. July 1976. Chemical Economics Handbook, Stanford Research Institute, Menlo Park, California.

4. J. F. Lawson, Trip Report for Visit to Denka Chemical Corporation, Houston, TX, Nov. 17, 1977.
~~Personal communication between J. F. Lawson, Hydroscience, Inc.~~, ~~and R. E. Hinkson, Denka USA, November 17, 197~~7.

5. ^Survey and Ranking Report^ in ~~preparation by~~ Hydroscience for ESED.

6. Lewis, W. A., Jr., G. M. Rinaldi, and T. W. Hughes, "Source Assessment: Maleic Anhydride Manufacture", Monsanto Research Corporation, received January 1978.

7. J. F. Lawson, Trip Report for visit to Reichhold Chemicals, Inc., Morris, Illinois, July 28, 1977.

8. "Maleic Makers Build on Hopes for Polyester", Chemical Week, February 2, 1977, pp. 37-38.

9. Technical Week, European Chemical News, April 5, 1974.

IV. EMISSIONS

A. Benzene Oxidation Process

1. Model Plant

The model plant for this study has a capacity of 22,700 Mg/yr (50 x 10^6 lb/yr), based on 8000 annual hours of operation. Though not an actual operating plant, it is typical of most plants. The model benzene oxidation process, shown in Figure 2, best fits today's maleic anhydride manufacturing and engineering technology. Single-process trains as shown are typical for the large plants except in the reaction area where multiple reactors are common. The model process uses partial condensation and azeotropic ~~drying~~ distillation with xylene.

Typical raw material, intermediate, and product storage tank capacities were estimated for a 22,700 Mg/yr plant. The storage tank requirements are covered under storage emissions. Estimates of potential fugitive sources were based on data from existing facilities and are also covered. Characteristics of the model plant important to air dispersion modeling are shown in Appendix B.

2. Sources and Emissions

All emission rates and sources for the benzene oxidation process are summarized in Table III.

a. Main Process Vent

The largest vent is the main process vent from the product recovery absorber (Vent A, Figure 2). All plants have this vent. The emissions are influenced by the excess air feed to the reactor to maintain the benzene concentration below the explosive limit. The composition of this stream for the model plant is shown in Table IV. The majority of the unreacted benzene is contained in this stream.

TABLE III

BENZENE AND TOTAL VOC EMISSIONS
MALEIC ANHYDRIDE FROM BENZENE
MODEL PLANT 22,700 Mg/Yr
UNCONTROLLED

Source	Stream Designation Figure 2	Benzene (kg/kg)x10^{-3} MA Produced	Total VOC (kg/kg)x10^{-3} MA Produced	Benzene kg/hr (B)	Total VOC kg/hr (B)
Product Recovery Absorber [1,2,3]	A	67.0	86.0	190 ~~193~~ (420)	~~243~~ 244
Refining Vacuum System [3]	B	--	0.1	--	0.28
Storage and Handling [4]	C	0.84 ~~0.69~~	0.98 ~~.004~~	2.2 ~~1.8~~ (5)	2.6 ~~2.2~~
Fugitive (A)	J	0.28 ~~0.31~~	1.02 ~~1.13~~	0.8 (2)	2.9
Secondary	K	--	~~0.057~~ 0.04	--	0.1
	Total	68 x 10^{-3}	88 x 10^{-3}	~~193.6~~ 193.0	~~248.5~~ 249.9

Note:

(A) Appendix C.

(B) Emission rates are annual averages at 8000 hrs/yr.

$\frac{kg}{hr} \times \frac{lb}{.454\ kg}$

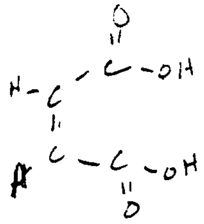


TABLE IV

WASTE GAS COMPOSITION - PRODUCT RECOVERY ABSORBER[1,2,3]

(Weighted Average)

Component	mw	Weight % 100 # basis	moles	Kg/hr	emission factor g/kg
Oxygen	32	16.67	.52	13,800	4863
Nitrogen	28	73.37	2.62	60,740	21,406
Carbon Dioxide	44	3.33	0,076	2,757	972
Carbon Monoxide	28	2.33	0.083	1,929	680
Water	18	4.00	0.222	3,312	1167
Benzene	78	0.23	0.0029	190	67
Maleic Acid	116	0.01	0.000086	8	2.8
Formaldehyde	30	0.05	0.0017	41	14
Formic Acid	46	0.01	0.0002	8	2.8
			[illegible]	82,785	based on 8000 hrs per year

Process upsets causing more benzene release will affect benzene emissions since the absorber can only remove benzene up to its solubility level in water. These upsets can cause short duration benzene and VOC emissions of three to five times normal. Process startup also increases benzene emissions three to five times normal because of incomplete benzene reaction. Shutdown will not affect emissions because benzene is shut off as the first step in the shutdown procedure. This immediately reduces the level of unreacted benzene emitted from the reactor.[5]

b. Refining Vacuum Vents

The refining vacuum system vent (Vent B, Figure 2) exhausts the noncondensibles from the three vacuum columns used to dehydrate and fractionate MA. The factors affecting this vent are the use of inerts to bleed the system for vacuum control and process leaks into the system. The VOC emissions will be maleic acid or xylene and are estimated to be relatively insignificant as indicated in Table III. Process upsets, startups, and shutdowns do not affect the VOC emissions from this vent.[3]

c. Fugitive Emissions

Didn't use

Process pumps and valves are potential sources of fugitive emissions. The model plant is estimated to have 15 pumps handling VOC, three of which handle benzene. The estimated number of valves is 500, with 175 handling benzene. The fugitive emission factors from Appendix C were applied to this valve and pump count to determine the fugitive emissions shown in Table III.

d. Storage and Handling Emissions

Didn't use

Emissions result from the storage and handling of benzene, maleic anhydride, and xylene. For the model plant, the sources are shown on the flow diagram in Figure 2 (Source C). The storage tank conditions for the model plant are given in Table V. The emissions in Table III were calculated based on fixed roof tanks, one-half full, and a 11°C diurnal temperature using the emission equations from AP-42.[4]

TABLE V

MODEL PLANT STORAGE

Content	Tank Size m^3	Turnovers Per Year	Bulk Liquid Temp. °C
Benzene	2460	~~8~~ 12.5	13
Benzene	75	~~250~~ 410	13
Maleic Anhydride	380	40	77
Maleic Anhydride	190	80	71
Maleic Anhydride	380	40	60
Xylene	150	12	13

Benzene freezes at 5.5°C; therefore, storage tanks are generally heated to maintain the temperature above freezing. Maleic anhydride freezes at 52.8°C; therefore, finished product is normally stored at 60°C, and in-process material is stored from 60-105°C.

Emissions from loading bulk maleic anhydride into tank cars and trucks were calculated using the equations from AP-42.[4] These emissions, included in Table III, are 0.02 kg/hr. Maleic anhydride dust is produced in the briquetting operation, but is not a significant atmospheric emission.

e. Secondary Emissions

Didn't use

R.C. Weber said these emissions are based on engineering estimates and not actual tests. He recomend not to use them. 4/22/80

Secondary emissions of VOC can result from the handling and disposal of process waste liquid and solid streams. For the model plant, four potential sources are indicated on the flow diagram, Figure 2 (Source K).

The spent catalyst removal and reclamation do not present significant emission potential because the catalyst is thoroughly purged and washed before removal. Reclamation is normally done off-site.[2]

For the model plant, ~~total~~ aqueous effluent from the vacuum system is estimated to be 417 kg per hour, of this 0.05 kg/hr is xylene. The fractionation column residue stream is 27 kg/hr of low vapor pressure organics. Even untreated, these streams do not represent significant emissions. In at least one operating facility, incineration is used to destroy the organics in these liquid streams.[2] Assuming a well-designed liquid incinerator with 99.5% destruction of organics, the secondary VOC emissions from this incinerator would be 0.13 kg/hr. Whether using liquid incineration or terminal wastewater treatment, the emission will be no greater than 0.13 kg/hr. This emission number is included in uncontrolled emissions, Table III.

The potential for secondary emissions from the vacuum system water discharge is minor. The stream should contain low levels of maleic acid and xylene. The amount has not been quantified.

B. n-Butane Process

As stated earlier, no benzene emissions are associated with the n-butane process. In all other respects, the VOC emissions are believed to be about the same as the benzene oxidation process, although there are no public data to support this.

C. Phthalic Anhydride By-Product Process

The emissions associated with MA recovery are believed to be insignificant and are not being investigated at this time. Relevant data will be obtained during the study of phthalic anhydride production.

SECTION IV

REFERENCES

1. Morse, P. L. "Maleic Anhydride". November 1973. Process Economic Program, Stanford Research Institute, Menlo Park, California.

2. J. F. Lawson, Trip Report for visit to Reichhold Chemicals, Inc., Morris, Illinois, July 28, 1977.

3. Pervier, J. W., et. al., Survey Reports on Atmospheric Emissions from the Petrochemical Industry, Volume III, Houdry Division of Air Products, Inc., EPA-450/3-73-005-C, EPA, Office of Air Quality Planning and Standards, Research Triangle Park, North Carolina, April 1974.

4. "Compilation of Air Pollutant Emission Factors", AP-42 Supplement No. 7, Second Edition, April 1977, pp. 4.3-6 to 4.311.

5. Personal communication between J. F. Lawson, Hydroscience, Inc., and G. R. Wood, Monsanto Chemical Company, October 20, 1977.

V. APPLICABLE CONTROL SYSTEMS

A. Benzene Oxidation Process

1. Main Process Vent

A carbon adsorption system or an incineration system can be used to effectively control process emissions. It is assumed that the vent from the refining vacuum system is combined with the main process vent. Therefore, treatment of the product recovery absorber vent stream will control all process emissions.

a. Carbon Adsorption

In order to use carbon adsorption, the exhaust gas stream must be scrubbed with caustic to remove organic acids and water-soluble organics. Benzene is likely the only VOC remaining. The stream is then conditioned by heating to reduce the relative humidity. Four carbon beds are specified for the model plant (see Carbon Adsorption Flowsheet, Appendix D). The exhaust stream passes through two parallel beds while the other two beds are being regenerated with steam. The steam condensate is decanted to separate the benzene for recycle to the process, and the benzene-saturated aqueous layer is ~~sent to waste disposal.~~ recycled to the product recovery absorber.

It was concluded based on engineering experience with similar applications that a carbon adsorption system can be designed and operated at a sustained removal efficiency greater than 99%. A removal efficiency of 99.5% is considered attainable and has been used to project the final emissions from the controlled model plant (Table VI). The cost of this system is shown in Table X of Section VI.

One carbon adsorption system is currently being used to recover benzene from a MA operation, but reportedly has only achieved 85% removal efficiency. For comparison purposes, the estimated costs for this installation are included in Table X.[1]

TABLE VI

BENZENE AND TOTAL VOC EMISSIONS
MALEIC ANHYDRIDE FROM BENZENE
MODEL PLANT 22,700 Mg/yr
CONTROLLED - CARBON ADSORPTION

Source	Stream Designation Figure 2	Control Device or Technique	Emission Reduction %	Benzene (kg/kg)x10^{-5} MA Produced	Total VOC (kg/kg)x10^{-5} MA Produced	Benzene Emission kg/hr	Total VOC Emission kg/hr
Product Recovery Absorber	A	Carbon Adsorption	99.5	34 35	34 ~~45~~	0.95	0.95
Refining Vacuum System	B	Vent thru Absorber	99.5				
Storage and Handling[(A)]	C	Floating Roof Tanks - Benzene	~~94~~ 90	--	--		
Storage and Handling	C	Scrubber - MA	99+	12 ~~4~~	13 ~~5~~	0.32 ~~0.1~~	0.34 ~~0.1~~
Storage and Handling	C	Xylene - None	0	--	--		
Fugitive[(B)]	J	Detect & Repair Major Leaks plus Mechanical Seals	93	3	9	0.06	0.2
Secondary	K	None	--	Trace	5		0.1
			Total	~~42~~ x 10^{-5} 49	64 x 10^{-5} 61 54	~~1.11~~ 1.33	~~1.35~~ 1.59

Notes:

(A) Approximately 65% of the storage emissions do not vary with production rate.

(B) Fugitive emissions do not vary with production rate.

b. Incineration

The direct fire incineration system for the model plant waste gas stream has: (1) a knockout demister tank to protect the incinerator from a liquid stream reaching the firing area, (2) an incinerator with a combustion chamber of sufficient volume to give a retention time of 0.5 sec, and (3) a stack designed for a velocity of 10 m/s.[2] Supplemental natural gas and makeup air are required to maintain the necessary combustion temperatures. Heat recovery can be used for steam generation or to preheat the feedstream and thereby reduce natural gas requirements. Stainless steel is specified for construction ahead of the combustion chamber because the waste gas contains corrosive organic acids. For more details of the incineration system specified for the model plant, see the incineration calculation section of Appendix D.

how much air? →

It was concluded based on engineering experience with similar incineration applications that a properly designed and operated incinerator will result in sustained benzene and VOC removal efficiencies greater than 99%. A removal efficiency of 99.5% is considered attainable and has been used to project final emissions from the controlled model plant (Table VII). The cost of incorporating this system is shown in Table X of Section VI.

A temperature of 871°C (1600°F) is specified to ensure complete combustion of the waste gas. While it is conceivable that greater than 99% VOC removal could be obtained at lower temperatures, it cannot be dependably predicted. This determination is consistent with government air pollution engineering manuals.[2,3] While the manuals contain no data on combustion temperatures above 800°C, extrapolation of the data presented combined with similar incineration experience justifies the projection of greater than 99% removal at 871°C. The high carbon monoxide content of the waste gas must be considered. Similar incineration experience indicates that the high temperature specified is necessary to

TABLE VII

BENZENE AND TOTAL VOC EMISSIONS
MALEIC ANHYDRIDE FROM BENZENE
MODEL PLANT 22,700 Mg/yr
CONTROLLED - INCINERATION

Source	Stream Designation Figure 2	Control Device or Technique	Emission Reduction %	Benzene (kg/kg)x10^{-5} MA Produced	Total VOC (kg/kg)x10^{-5} MA Produced	Benzene Emission kg/hr	Total VOC Emission kg/hr
Product Recovery Absorber	A	Incineration	99.5	~~35~~ 34	~~45~~ 43	0.95	1.23
Refinery Vacuum Vent	B	Vent thru Incinerator	99.5	--	--	--	--
Storage and Handling[(A)]	C	Floating Roof Tank - Benzene	~~94~~ 90				
Storage and Handling	C	Scrubber - MA	99+	~~4~~ 12	~~5~~ 13	0.32 ~~0.1~~	0.34 ~~0.1~~
Storage and Handling	C	Xylene - None	--				
Fugitive[(B)]	J	Detect & Repair Major Leaks plus Mechanical Seals	93	3	9	0.06	0.2
Secondary	K	None	--	--	5	--	0.1
			Total	~~42~~ x 10^{-5} 49	~~54~~ x 10^{-5} 70	~~1.11~~ 1.33	~~1.63~~ 1.87

Notes:

(A) Approximately 65% of the storage emissions do not vary with production rate.

(B) Fugitive emissions do not vary with production rate.

obtain acceptable oxidation of the carbon monoxide, as well as complete VOC oxidation. The cost of incorporating this system is shown in Table X of Section VI.

An incineration system currently being used for VOC removal from a MA process reportedly is operating with a combustion temperature of 760°C and achieving 93% removal efficiency.[4] For comparison purposes, a removal efficiency of 93% is used in Table X of Section VI to correspond with the 760°C combustion temperature. All costs are identical to the 871°C system except for the lower utilities (natural gas) requirement at 760°C. An incinerator designed for 871°C can be operated at lower temperatures if it is determined by operating experience that a lower temperature will still provide adequate benzene, VOC, and carbon monoxide removal.

c. Catalytic Incineration

One company is using a catalytic incinerator to control emissions from the product recovery absorber. Very little design data are available. The maximum practical achievable VOC removal efficiency reportedly is less than 95%. The high catalyst volume or high temperature requirement necessary to obtain greater removal efficiencies makes the unit uneconomical.[5]

d. n-Butane Process

Since the n-butane oxidation process has no potential benzene emissions, conversion to the n-butane process for MA production is an option for benzene emission control. No data are presently available on the control device options for the n-butane process. Also, no data are currently available comparing the economics of MA production by n-butane oxidation and benzene oxidation.

2. Refining Vacuum Vent

The refining vacuum vent is controlled by joining the waste stream ahead of the product recovery absorber control device or by joining the product recovery absorber feed stream. The incremental costs are relatively small since only piping additions are required, and no added utilities, manpower, or other operating costs are involved. Emissions from the refining vacuum system vent are included in all control system calculations.

3. Fugitive Sources

Didn't use

Controls for fugitive sources will be discussed in a future document which will cover fugitive emissions from the synthetic organic chemicals manufacturing industry. Control of emissions from pumps and valves can be attained by an appropriate leak detection system followed by repair maintenance. Controlled fugitive emissions have been calculated using the factors given in Appendix C and are included in Tables VI and VII. The factors are based on the assumption that major leaks are detected and repaired; therefore, the emission reduction is projected to be 93%.

4. Storage and Handling Sources

Didn't use

a. Benzene Storage

Control of benzene storage emissions will be covered in a future EPA document. Information on MA manufacturing locations indicates that benzene is stored in floating roof tanks at three locations and fixed roof tanks at the others.[5] A floating roof is commonly used to control storage tank emissions for VOC's in the vapor pressure range of benzene and is used in the model plant instead of fixed roof tanks. This results in a benzene emission reduction of ~~94%~~ 90%. The benzene emissions given in Tables VI and VII were calculated using the floating roof storage tank emission equations from AP-42.[6]

An alternate control method may be possible in MA plants: the fixed roof storage tank vents could be tied into the main process vent control. The resulting emission reduction should be >99%

which is better than that of a floating roof tank. However, this modification may add a safety hazard because benzene storage vapors may be in the flammable range.

b. Maleic Anhydride Storage and Handling

D. dr 4
usP

The vapor pressure of MA at storage conditions in the model plant is 550-1170 pascals, this results in calculated average emissions from fixed roof tanks of only 0.3 kg/hr. At two production plants, the tank vents (as well as air handling from the briquetting operation) are treated in aqueous scrubbers.[1,7] The effluent water is discharged to the sewer. The scrubbers are primarily for plant housekeeping to prevent solid MA buildup rather than for emission reduction. It is estimated that scrubber efficiency under these conditions should be at least 99%. The controlled VOC emissions given in Tables VI and VII were calculated based on this efficiency.

At one plant, the crude maleic anhydride is stored at 130°C and the vent is treated in a xylene scrubber with a stated 50% efficiency.[8] The xylene effluent can be recycled to the process for MA recovery. This control option may not be adequate without an aqueous scrubber to prevent the resulting MA emission from creating a condensed MA solid housekeeping problem.

c. Xylene Storage

D. dr 4
usP

The calculated emissions from xylene storage in fixed roof tanks for the model plant average less than 0.03 kg/hr, which is insignificant relative to other emissions from the process; therefore, control is not assumed to be needed.

5. Secondary Sources

Secondary emissions are also insignificant. In plants where a liquid incinerator is used to destroy organic effluents, the design must be adequate to ensure essentially complete combustion.

6. Current Emission Control

Table VIII shows the control devices currently being used by U.S. MA producers.

B. Other Processes

Data are not currently available to size emission control devices for other processes. The only other significant domestic process, n-butane oxidation, was discussed as a control option for benzene emissions.

TABLE VIII

CONTROL DEVICES CURRENTLY USED BY THE MALEIC ANHYDRIDE INDUSTRY IN THE U.S.[5]

	Emission Point			
Company	Flaking, Pelletizing and Packaging	Product Recovery Absorber	Vacuum System Vent	Storage Tank Vents
Amoco Chemicals Corp.	NR	NR	NR	NR
Ashland Oil, Inc.	A	Scrubber	NR	Floating Roof Tanks
Denka Chemical Corp.	Scrubber	Incineration	Scrubber	Floating Roof Tanks
Koppers Company, Inc.	Scrubber	Incineration	NR	Return Vents
Monsanto Co.	Scrubber	B	A	Scrubber
Reichhold Chemical, Inc.				
Elizabeth, NJ	Scrubber	Carbon Adsorber	NR	Conservation Vents
Morris, IL	A	Carbon Adsorber	B	Scrubber, Conservation Vents
Tenneco, Inc.	B	B	NR	Scrubber, Conservation Vents
U.S. Steel Corp.	Scrubber	Catalytic Incinerator	NR	Floating Roof Tanks

A - ~~Does not have this emission source~~. Vents to the product recovery absorber.

B - No control.

NR - Not reported.

SECTION V

REFERENCES

1. J. F. Lawson, Trip Report for visit to Reichhold Chemicals, Inc., Morris, Illinois, July 28, 1977.
2. Air Pollution Engineering Manual, U.S. Environmental Protection Agency, May 1973, pp. 176 and 181.
3. Air Pollution Engineering Manual, U.S. Department of Health, Education, and Welfare, 1967, p. 709.
4. Personal communication between J. F. Lawson, Hydroscience, Inc., and R. E. Hinkson, Denka USA, November 17, 1977.
5. Lewis, W. A., Jr., G. M. Rinaldi, and T. W. Hughes, "Source Assessment: Maleic Anhydride Manufacture", Monsanto Research Corporation, received January 1978.
6. "Compilation of Air Pollutant Emission Factors", AP-42 Supplement No. 7, Second Edition, April 1977, pp. 4.3-12 and 4.3-13.
7. Houdry Division of Air Products, Inc., "Engineering and Cost Study of Air Pollution Control for the Petrochemical Industry - Maleic Anhydride", prepared for EPA.
8. Permit Exemption Request from R. E. Pruessner, Petro-Tex Chemical to Mr. Charles Borden, Texas Air Control Board, February 23, 1976.

VI. IMPACT ANALYSIS

A. Control Cost Impact

The purpose of this section is to present estimated costs and cost effectiveness (CE) ratios for control of benzene and total VOC emissions resulting from the production of maleic anhydride. Details of the model plant (Figure 2) have been covered in Sections III and IV. Cost estimate calculations are included in Appendix D.

Basis for Capital Cost Estimates - Estimates represent the total investment required to purchase and install all equipment and material to provide a complete control system performing as defined for a new installation at a typical location. These estimates do not include the cost of: lost MA production during installation or start-up, research and development costs, and land purchases required. Costs for retrofitting these systems in existing plants usually will not be appreciably greater than for a new installation. The primary retrofit difficulty may be space to fit into the existing plant layout.

Bases for Annualized Cost Estimates - Estimates for the annualized costs for the control alternatives include: utilities, operating labor, maintenance supplies and labor, chemicals or raw materials, recovery credits, disposal costs, capital charges and miscellaneous recurring costs such as taxes, insurance, and administrative overhead. Cost factors used are itemized on Table IX. Capital charges are computed by a factor based on a depreciable life of the control system and an annual interest rate. The factor of 0.147 is based on an assumed 12-year life and 10% interest. Recovery credits are based on the market value for the material being recovered. Chemical or other raw material costs also are based on the market value for the material required. Annualized costs are for a one-year period beginning in mid-1978.

1. Benzene Oxidation Process

a. Main Process Vent

Estimated costs for two systems (carbon adsorption and incineration) used to control benzene and total VOC emissions from this point source are shown on Table X.

TABLE IX

COST FACTORS USED IN COMPUTING ANNUALIZED COSTS

Utilities	
Electricity	$0.03/kw-h
Natural Gas	$1.90/~~k~~GJ
Steam	$2.85/~~k~~GJ
Cooling Water	$0.15/m^3
Operating Time	8000 hrs/yr (91% of available time)
Operating Labor	$10/man hour
Maintenance Material & Labor	0.05 x capital cost
Capital Charges	
Capital Recovery	0.147* x capital cost
Misc (taxes, insurance, & administration)	0.04 x capital cost
Liquid Waste Disposal	$2.65/m^3
Recovery Credits	
Energy	$1.90/~~k~~GJ
Benzene	$0.22/kg
Chem. & Raw Material Costs	
50% Caustic Soda Soln.	$0~~.20~~.11/kg
Granular Activated Carbon	$1.90/kg

*Based on 12 year life and 10% interest rate.

TABLE X

CONTROL COST ESTIMATES - MALEIC ANHYDRIDE MODEL PLANT MAIN PROCESS VENT AND REFINING VACUUM VENT

Control System	760°C Incineration		871°C Incineration		Carbon Adsorption	
	Without Heat Recov.	With Heat Recov.	Without Heat Recov.	With Heat Recov.(B)	99.5% Benzene Removal	85%(A) Benzene Removal
Installed Capital Cost (1978) $1000	400	670	400	670	~~1400~~ 1620	1100
Direct Operating Costs (1978)						
Utilities $1000	544	544	798	798	313	235
Manpower $1000	16	16	16	16	24	24
Maintenance $1000	20	33	20	33	~~74~~ 81	55
Raw Material & Chemicals $1000	None	None	None	None	~~60~~ 46	~~60~~ 46
Disposal $1000	None	None	None	None	~~86~~ 7	~~64~~ 5
Capital Recov. Costs $1000	59	98	59	98	~~210~~ 238	162
Misc. Capital Related Costs $1000	16	27	16	27	~~59~~ 65	44
Credits(C) $1000	None	493	None	(618)	(330)	(272)
Net Annualized Costs (1978) $1000	655	225	909	354	~~504~~ 444	~~372~~ 299
Controlled Emissions						
Benzene Mg/yr	1420	1420	1520	1520	1520	1298
Total VOC Mg/yr	1809	1809	1936	1936	1938	1655
Emission Reduction						
Benzene %	93.0	93.0	99.5	99.5	99.5	85
Total VOC %	93.0	93.0	99.5	99.5	99.5	95
Cost Effectiveness						
Benzene $/Mg	461	158	598	233	~~332~~ 292	287 230
Total VOC $/Mg	362	124	469	183	~~260~~ 229	~~225~~ 181

(A)Based on capital cost and operation of existing system.[1]

(B)Figures apply to either steam generation or incineration feed preheat.

(C)Credit as natural gas (not steam).

Carbon Adsorption - Estimated cost bases for the carbon adsorption system described in Section V and Appendix D are as follows:

Carbon loading - 6 kg VOC/100 kg carbon
Steam for regeneration - 20 kg/kg of organic absorbed - 3810 kg/hr
Steam for gas conditioning - 825 kg/hr
Cooling water for the steam/VOC condenser - 30 m^3/hr
50% caustic solution for scrubber - ~~30~~ 37 kg/hr
Granular activated carbon replacement every 4 years at $1.90/kg

Corrosion resistant 316 stainless steel is required through the gas scrubber. Costs of the scrubber and gas heater are included. Costs for a carbon adsorption system operating at 85% VOC removal efficiency are also shown on Table X. This is for an existing system and shows that the cost effectiveness is not appreciably less for a system operating at a lower efficiency.[1] Costs of systems for removal efficiencies of more than 99.5% are not included because such higher efficiencies would be practically impossible to obtain.

Capital costs for carbon adsorption systems for MA production plants with capacities different from the model plant are estimated to vary approximately as the 0.7 power of the waste gas flow ratio rate, which is directly proportional to the MA production rate. This relationship is applicable over the size range of existing plants (50-165% of the model plant capacity).

Other annualized costs will vary with capacity as listed below.

(1) Varies directly with capacity:
    (a) Utilities
    (b) Raw materials and chemicals
    (c) Liquid disposal changes
    (d) Energy recovery

(2) Remains constant with capacity variance:
  (a) Manpower

(3) Varies as the ~~0.6~~ 0.7 power of the capacity ratio:
  (a) Capital recovery
  (b) Capital related rinse items
  (c) Maintenance

Figure 3 is a plot of capital and net annualized costs versus capacity for the 99.5% benzene removal carbon adsorption systems.

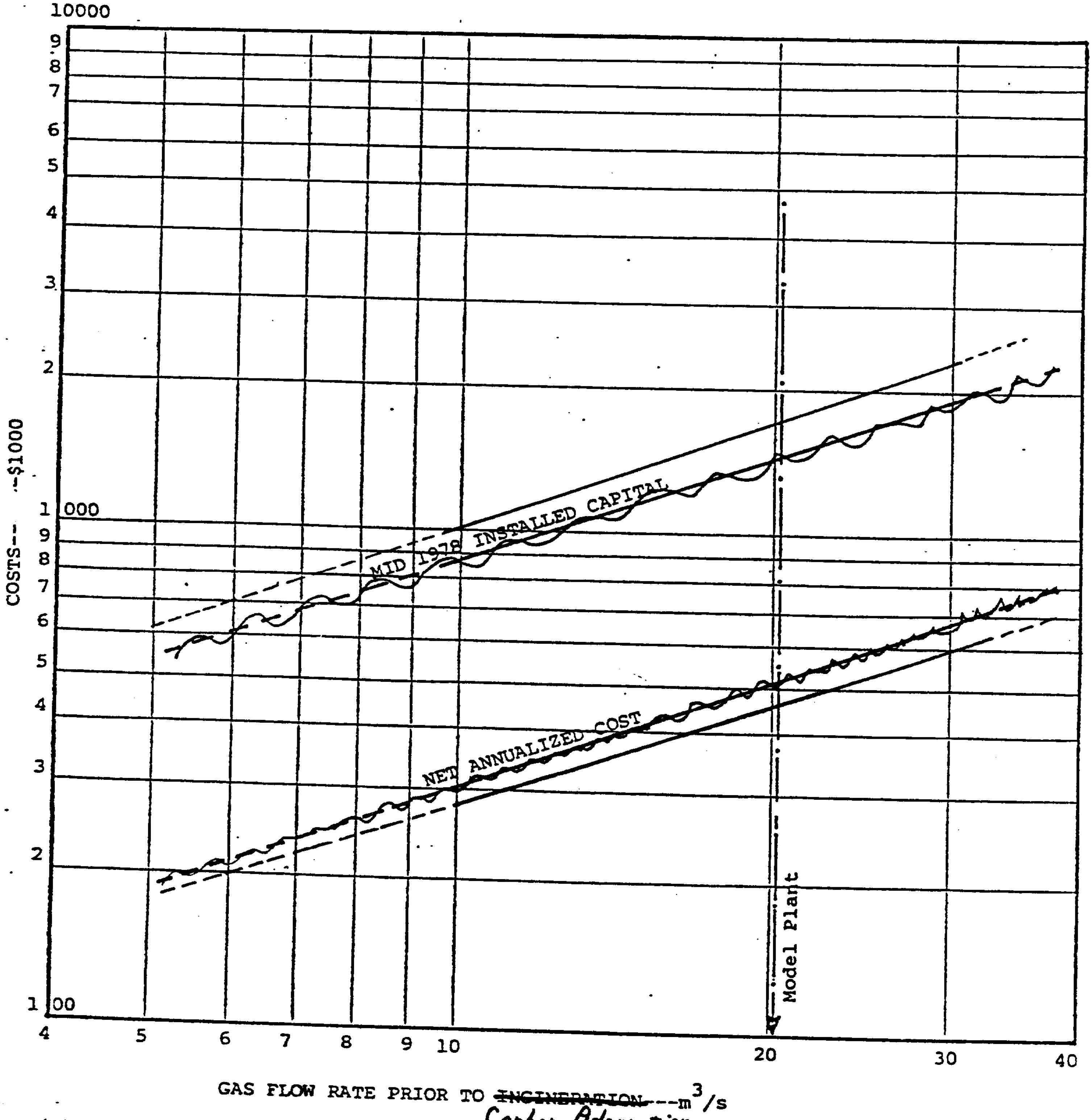
Incineration - Estimated costs are based on the incineration system described in Section V and Appendix D. A well-designed and operated incinerator will have a VOC removal efficiency greater than 99% (99.5% assumed). The cost savings of operating at a lower efficiency do not appear justified. To obtain greater than 99.5% removal efficiencies would increase the cost significantly, because of the disproportionately large increase in combustion temperature and residence time required. One incinerator is presently operating at 760°C with a VOC removal efficiency of 93%.[2] Costs are included in Table X to operate the specfied high temperature system at a lower temperature with an assumed efficiency of 93%.

Heat recovery from incineration can be utilized to generate steam or preheat the incinerator feed stream to reduce supplemental fuel requirements. For this comparative study, recovery of 50% of the energy in the exit stream is assumed. Credit for the value of this heat is computed as equivalent to natural gas at $1.9/GJ, not for the steam that may be generated. If the recovered heat is used to preheat the waste gas feed stream, the natural gas (utilities) requirement can be reduced by the full value of the heat recovered. For this preliminary estimate it is assumed that the capital costs for a preheat system or a steam generation system to augment an existing system are equivalent.

5/10/78

FIGURE 3

COST VS. GAS FLOW RATE
CARBON ADSORPTION



GAS FLOW RATE PRIOR TO ~~INCINERATION~~ Carbon Adsorption --m^3/s

(A) Includes Capital Cost Recovery.

Capital costs for incinerators at MA production plants with capacities different from the model plant are estimated to vary at approximately the 0.6 power of the waste gas flow ratio, which is directly proportional to the MA production rate. This relationship is applicable over the size range of existing plants (50-165% of the model plant capacity).

Other annualized costs will vary with capacity as listed below.

(1) Varies directly with capacity:
  (a) Utilities
  (b) Raw materials and chemicals
  (c) Liquid disposal charges
  (d) Energy recovery

(2) Remains constant with capacity variance:
  (a) Manpower

(3) Varies as the 0.6 power of the capacity ratio:
  (a) Capital recovery
  (b) Capital related miscellaneous items
  (c) Maintenance

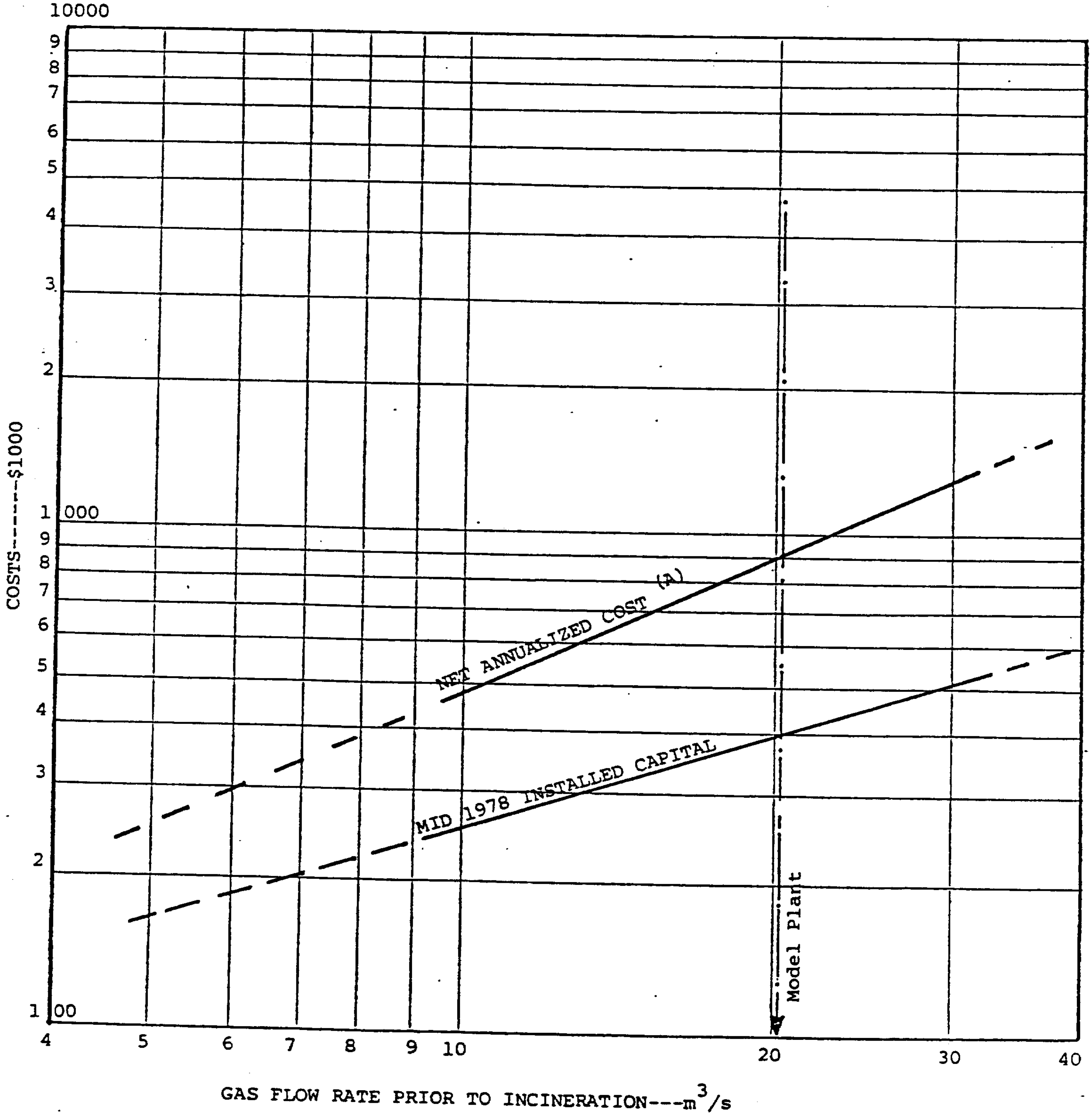
Figure 4 is a plot of capital and net annualized costs versus capacity for an incinerator without heat recovery, and Figure 5 is the same information for an incinerator with heat recovery. Figure 6 is a plot of the cost effectiveness versus capacity showing this relationship for each of the 99.5% benzene removal control systems considered.

Capital and operating costs for the incinerator installation will not be appreciably greater when retrofitting an existing plant, except in unusual circumstances. For example, if incineration is to be added to a crowded existing site, the cost of long distance piping to a remote location could be significant. Long distance piping could also be required to avoid flammable materials storage and handling areas.

FIGURE 4

COST VS. GAS FLOW RATE

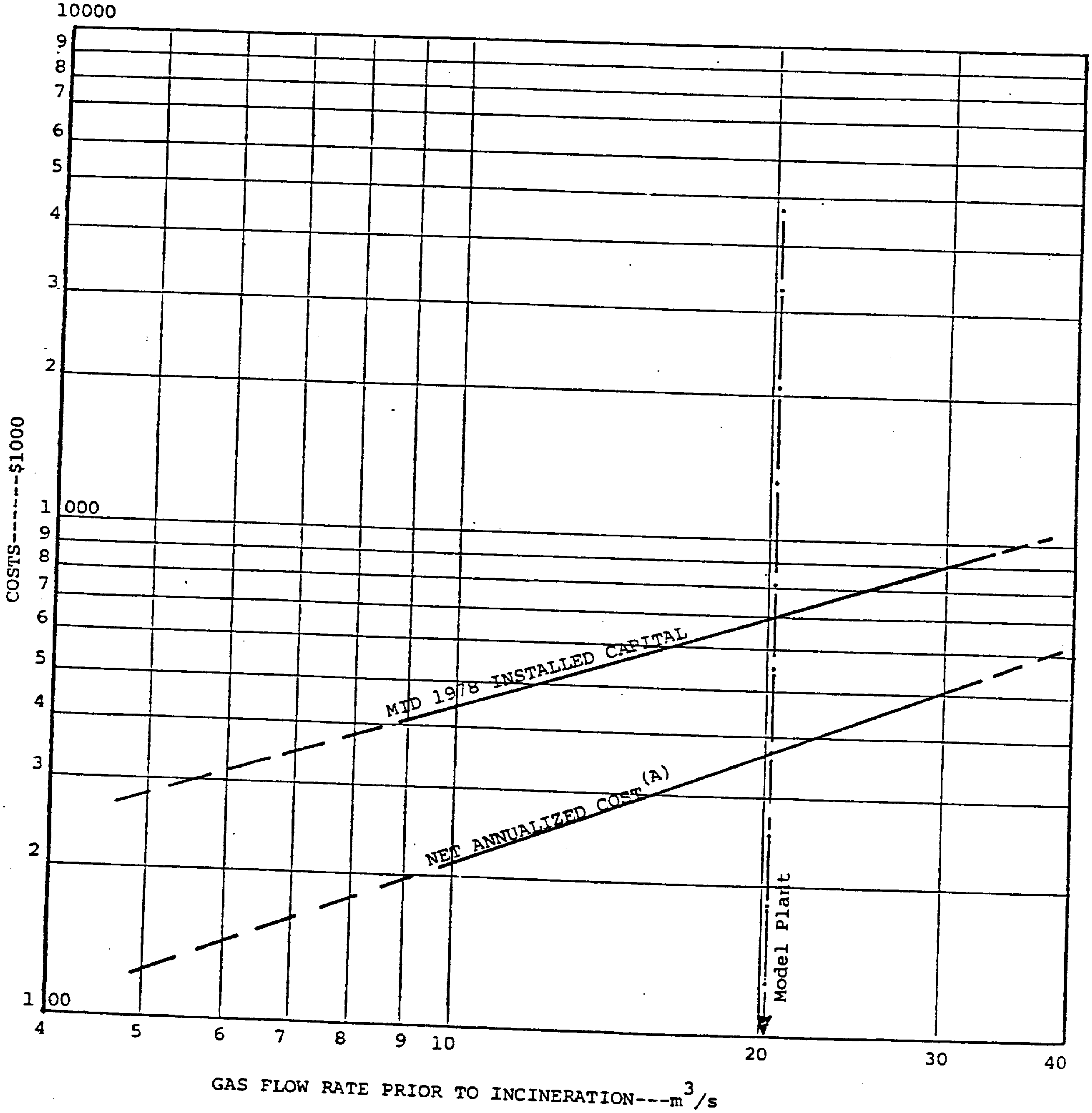
INCINERATION WITHOUT HEAT RECOVERY



(A) Includes Capital Cost Recovery.

FIGURE 5

COST VS. GAS FLOW RATE
INCINERATION WITH HEAT RECOVERY



(A) Includes Capital Cost Recovery.

5/10/78

COST EFFECTIVENESS VS. GAS FLOW RATE

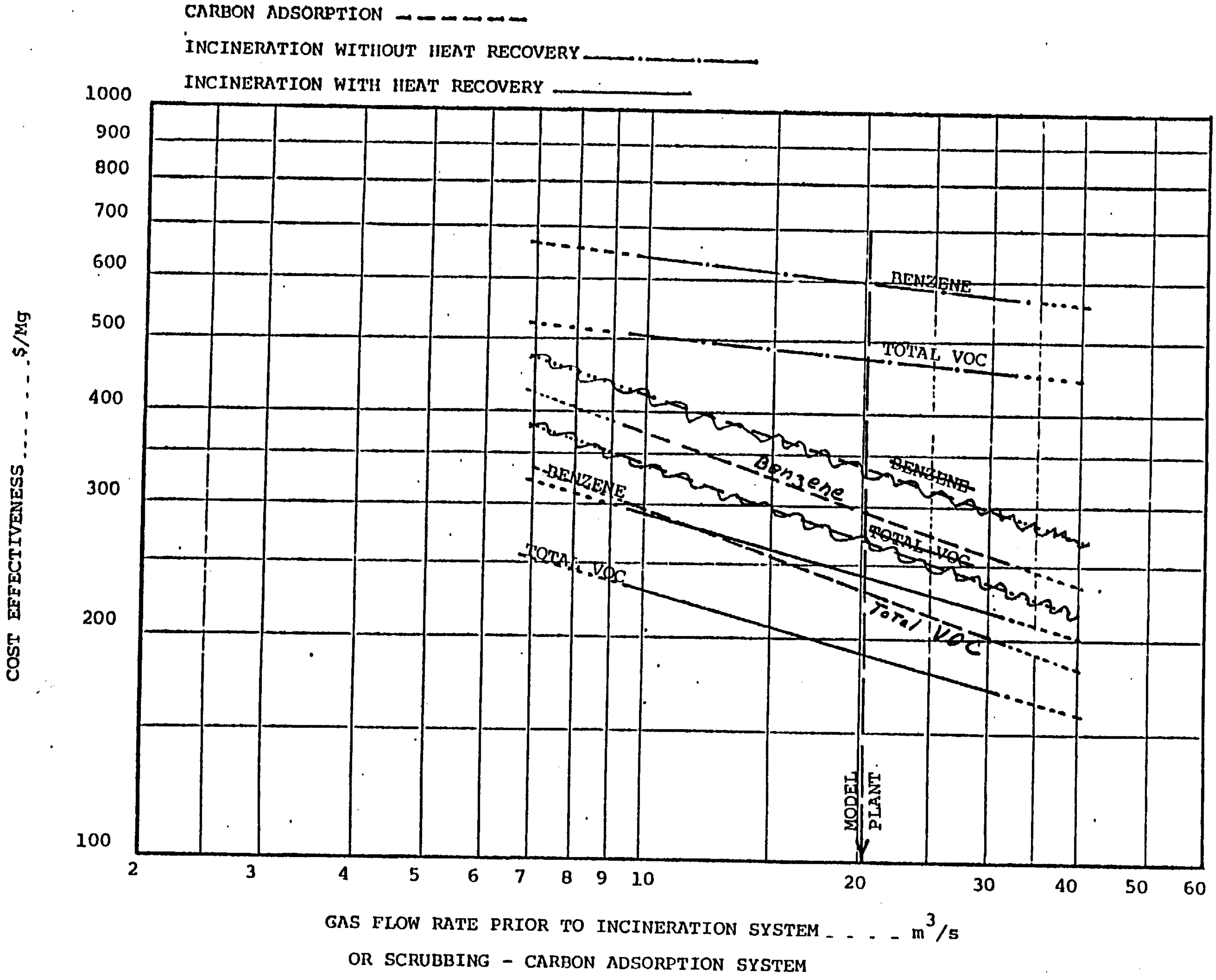


FIGURE 6

b. Refining Vacuum Vent

The vent stream from the vacuum system for the MA finishing section (Stream B, Figure 2) is controlled by connecting into product recovery absorber vent (Stream A) before the emission control devices previously described in Section V. The quantities and composition used for Stream A include Stream B, including any credits for recovered heat.

c. Fugitive Sources

A control system has been defined in Appendix C for fugitive sources; a future document will cover fugitive emissions and their applicable controls for all of the synthetic organic chemicals manufacturing industry.

d. Storage and Handling Sources

(1) Benzene Storage

Model plant benzene storage emissions are controlled by the use of floating roof tanks. The control cost and the cost effectiveness for retrofitting the large (2460 m^3) model plant fixed roof benzene storage tank are given in Table XI. ~~The total cost to build a cone roof tank and retrofit a floating roof is approximately the same as the cost of a floating roof tank.~~ For a new tank, the additional cost to include an internal floating roof is $42,000 including 19% allowance. (See installed capital graph added to Appendix D.)

It is normally not practical to incorporate a floating roof in small tanks such as the 75 m^3 benzene feed tank for the model plant; therefore, a cost impact is not presented.

(2) MA Storage and Handling

MA storage and handling emissions are controlled in the model plant by the use of an aqueous scrubber. The control cost and cost effectiveness for an assumed 0.6 meter diameter by 9.1 meter packed aqueous scrubber system to handle all model plant MA storage tanks, loading, and briquetting emissions are given in Table XI.

e. Secondary Sources

No significant secondary emission sources requiring a control system exist for the model plant.

2. Other Processes

No data are available for determining the cost impact of converting to the n-butane process as a benzene emission control option. Data also are unavailable to determine the cost of any control devices required to control emissions from a n-butane oxidation process.

B. Environmental and Energy Impacts

1. Benzene Oxidation Process

Table XII shows the environmental impact of reduced benzene and VOC emissions by application of described control systems to the model plant. From an energy standpoint, a typical uncontrolled MA process will produce a heat surplus of approximately 15 kJ/kg MA.[3] Individual impacts are discussed below.

a. Main Process Vent and Vacuum Refining Vent

(1) Carbon Adsorption

(a) The carbon adsorption system described to control emissions from these two sources (Vents A and B) will reduce benzene emissions by 1510 1520 Mg/yr and total VOC emissions by 1938 Mg/yr for the model plant. This is equivalent to an emission reduction of 15,780 Mg/yr benzene and 20,000 Mg/yr total VOC for the U.S. MA production assumed uncontrolled at capacity. Of this emission reduction, 1474 Mg/yr of benzene from the model plant is recovered for recycling to the process. The remainder of the benzene and the other VOC (~57 kg/hr) are picked up in the scrubber ~~or decanter water~~ and are removed as effluent wastewater which may cause a trace secondary emission.

TABLE XI

CONTROL COST ESTIMATES MA MODEL PLANT
STORAGE AND HANDLING

Plant Section Control System	Benzene Storage 2460 m^3 Tank Floating Roof Tank	MA Storage and Handling All Tanks and Briquetting Packed Tower Scrubber
Installed Capital Cost	60,000 ~~$18,000~~ (A)	$63,000
Direct Operating Costs		
Utilities		2,900
Manpower		1,000
Maintenance	3,000 ~~900~~	3,200
Waste Disposal		700
Capital Recovery Cost	8,800 ~~2,600~~	9,300
Misc. Capital Costs	2,400 ~~700~~	2,500
Recovery Credits	(3,000) (B)	
Net Annualized Cost	11,200 ~~1,200~~	19,600
Emission Reduction		
Benzene, Mg/yr	14.6 ~~13.8~~	
MA Mg/yr		2.8
Percent %	96 ~~97~~	99+
Cost Effectiveness, $/Mg	~~90~~ 767	7,000

Notes:

(A) Installed cost of retrofitting internal steel floating roof and closure seal on existing fixed roof tank; does not include cleaning tank, correcting tank defects and loss of use of tank during retrofit. Includes freight, taxes, engineering and 25% allowance.

(B) Benzene credit at $.22/kg

TABLE XII

ENVIRONMENTAL IMPACT
MODEL PLANT - 22,700 Mg/yr MA
CONTROLLED AS NOTED

Source	Stream No.	Control Device or Technique	Emission Reduction %	Emission Controlled Mg/yr Benzene	Emission Controlled Mg/yr Total VOC
Product Recovery Absorber	A	Carbon Adsorption	99.5	1520	1938 (A)
Refining Vacuum System	B	Incineration	99.5	1520	1936
Storage and Handling (B) Benzene	C	Floating Roof	90 ~~94~~		
Storage and Handling MA	C	Scrubber	99+	14.6 ~~13.8~~	17.4 ~~16.6~~
Storage and Handling Xylene	C	None	--	--	--
Fugitive (C)	J	Detect & Repair Major Leaks plus Mechanical Seals	93	6.4	23.2
Secondary	K	None	--	--	--
		Total		~~1540.2~~ 1541	~~1977.8~~ (A) 1978.6 ~~1975.8~~ 1976.6

Notes:

(A) VOC removal with carbon adsorption is slightly higher because of the VOC removed by the caustic scrubber.

(B) Approximately 65% of the storage emissions do not vary with production rate.

(C) Fugitive emissions do not vary with production rate.

(b) Energy (steam) is required to desorb the benzene off the carbon. For the model plant, this energy (as steam) is 57,000 MJ/Mg of benzene emission reduced and 45,000 MG J/Mg of total VOC removed. For the total U.S. MA production capacity, this is equivalent to ~~250~~ 900 x 10^6 MJ/yr. ~~This is a net usage, since the~~ Benzene recovered is recycled to the process rather than used as fuel.

The electrical energy required is 2500 MJ/Mg of benzene emission reduced and 2000 MJ/Mg of VOC removed. For the total U.S. MA production capacity this is equivalent to 40 x 10^6 MJ/yr.

(2) Incineration

(a) The incinerator system to control emissions from these two sources reduces benzene emissions 1520 Mg/yr and total VOC emissions 1936 Mg/yr. Extrapolated to total U.S. MA production, this is 15,780 Mg/yr reduction of benzene and 20,000 Mg/yr of total VOC emissions. No organics are recovered for recycle to the process with incineration. Stack gases from incineration can have a negative impact on the environment (NO_x, CO_2, CO) particularly if the carbon monoxide is not adequately oxidized.

(b) Supplemental fuel is required to maintain suitable operating conditions. The net amount of energy required for the model plant ranges from 278,000 MJ/Mg of benzene removed for an incinerator without heat recovery to 62,000 MJ/Mg for an incinerator with heat recovery. For total VOC removal the net energy requirement ranges from 218,000 MJ/Mg to 48,000 MJ/Mg of VOC removed. The net energy required for total U.S. MA production projects to 4~~4~~300 x 10^6 MJ/yr without heat recovery and 9~~60~~50 x 10^6 MJ/yr with heat recovery.

b. Other Emissions (Fugitive, Storage and Handling, and Secondary)

(1) Control methods described for these sources are floating roof storage tanks, scrubbing for product tank vents, and leak repair for fugitives.

(a) Application of these systems results in a VOC emission reduction of ~~40~~42 Mg/yr for the model plant and ~~415~~436 Mg/yr for the U.S. MA production capacity.

(b) The energy impact resulting from application of these systems is minimal because the only energy required is electricity for pumps.

2. Other Processes

Control systems for other processes have not been described.

SECTION VI

REFERENCES

1. J. F. Lawson, Trip Report for visit to Reichhold Chemicals, Inc., Morris, Illinois, July 28, 1977.
2. Personal Communication between J. F. Lawson, Hydroscience, Inc., and R. E. Hinkson, Denka USA, November 17, 1977.
3. Morse, P. L., "Maleic Anhydride". November 1973. Process Economic Program, Stanford Research Institute, Menlo Park, California.

VII. PRODUCT ASSESSMENT

A. Summary

Maleic anhydride is produced in the United States predominately by catalytic oxidation of benzene. In recent years producers have been looking very carefully at the feasibility of n-butane as a feedstock. Late in 1976 Amoco began operation of a 27,000 Mg/yr plant using butane feedstock. In 1974 Monsanto also dedicated 20% of its capacity toward development of butane technology. Denka USA has notified the regulatory agencies of its intent to perform butane process development. Approximately 83% of the 235,000 Mg/yr domestic capacity is based on benzene oxidation. Butane oxidation accounts for another 15% of capacity, and the remaining 2% is produced as a by-product of phthalic anhydride manufacture.

The maleic anhydride annual growth rate is estimated to be 11% through 1982. Based on 1977 U.S. MA production capacity, there is sufficient present capacity to meet the growth rate through 1982. No shortage of either benzene or n-butane is expected during this period.[1]

The emission sources and control levels for the model plant are summarized in Table XIII. Projecting these values for the U.S. industry at full capacity operation, the benzene emission would be ~~2010~~ 2014 1680 kg/hr with all plants uncontrolled and ~~11.5~~ 13.8 11.5 kg/hr controlled (carbon adsorption or incineration option). The total VOC emissions for the industry at capacity would be ~~2596~~ 2600 2160 kg/hr uncontrolled versus ~~16.9~~ 19.4 16.2 kg/hr controlled with the incineration option, and ~~14.0~~ 15.9 13.8 kg/hr controlled with the carbon adsorption option. Considering the degree of control currently being employed, the present total capacity emission projection for the U.S. MA industry is 830 kg/hr benzene and ~~1060~~ 1070 kg/hr total VOC.[2,3] Based on these figures the present removal efficiency is approximately 59%, and the fully controlled emission reduction would be approximately 99.4%.

The predominant emission points are the main process vent and benzene storage facilities. Control of these sources alone can reduce emissions by 99%.

TABLE XIII

EMISSION SUMMARY
MODEL PLANT - 22,700 Mg/yr
UNCONTROLLED AND CONTROLLED

	Kg/hr				
	Uncontrolled		Controlled		
Emission Source	Benzene	VOC	Benzene[(A)]	VOC	
Process	191	243	0.95	1.23[(B)]	0.95[(C)]
Storage and Handling	~~1.8~~ 2.2	~~2.2~~ 2.6	~~0.1~~ 0.32	~~0.1~~ 0.34	~~0.1~~ 0.34
Fugitive	0.8	2.9	0.06	0.2	0.2
Secondary	-	0.1	-	0.1	0.1
Total	~~193.6~~ 194	~~248.2~~ 248.6	~~1.1~~ 1.33	~~1.63~~ 1.87	~~1.35~~ 1.59

(A) Benzene controlled emissions are identical for carbon adsorption and incineration.

(B) Based on incineration option.

(C) Based on carbon adsorption. VOC emissions are lower because of the VOC removed by caustic scrubbing prior to carbon adsorption.

The model plant main process vent, which contains 241.2 kg/hr total VOC, of which 190.8 kg/hr is benzene, can be controlled by either carbon adsorption or incineration. The carbon adsorption efficiency is 99.5% for benzene and VOC. ~~Cost~~ The cost for emission control for a model plant would be $1.480 million, with cost effectiveness of $332/Mg for benzene and $260/Mg for VOC. The energy as steam required is 57,000 MJ/Mg of benzene controlled, and the energy as electricity required is 2500 MJ/Mg of benzene controlled. Incineration is 99.5% efficient for benzene and VOC. Cost for a model plant with heat recovery would be $0.670 million, with cost effectiveness of $233/Mg for benzene and $183/Mg for VOC. The net supplemental fuel energy requirement is 278,000 MJ/Mg benzene without heat recovery and 62,000 MJ/Mg benzene with heat recovery. With either carbon adsorption or incineration, the vacuum vent system is tied to the control device feed; and all calculations include these emissions.

Benzene storage can be controlled by retrofitting a floating roof to a fixed roof tank. The emission removal efficiency is ~~94~~ 90%. Cost for the model plant would be $~~10,000~~ 60,000, with a cost effectiveness of $~~90~~ 82/Mg benzene controlled.

Maleic anhydride storage and handling can be controlled by fume scrubbing, which is greater than 99% efficient based on MA emissions. Cost for the model plant would be $63,000, with a cost effectiveness of $7,000/Mg MA controlled.

One carbon adsorption system is currently being used to recover benzene from a MA operation, but reportedly has only achieved 85% efficiency.[4] One MA producer using incineration for VOC emission control reports a removal efficiency of 93%.[5]

Since the n-butane oxidation process has no inherent benzene emissions, it is an option for benzene emission control. However, data regarding the potential economic incentive for switching to n-butane oxidation are not presently available. Its increased popularity is evidenced by the fact that a growth rate of 24.3% is projected for n-butane oxidation through 1982 compared to 9.1% for benzene oxidation.[1]

B. Supplemental Information

The process emissions from the benzene oxidation model plant are based on the emissions which were reported in the Houdry study, Reichhold trip report, SRI, and an understanding of the process chemistry and yields. The main process vent (the largest source of benzene and VOC emissions) will be sampled in at least two plants (Reichhold Chemicals, Inc., Morris, Illinois; and Denka USA, Houston, Texas) with additional data released concerning the emissions from a third plant (Monsanto Company, St. Louis, Missouri).

In addition, the two plants being sampled have emission control devices on the main process vent, which will also extend the data base for control technology. Denka USA has an incineration system controlling the main process vent, and samples will be taken of the emissions before and after the incinerator. The incinerator operates at a temperature of 760°C; but a temperature of 871°C will be attempted to compare the removal efficiency of an actual operating system at the combustion temperature specified for the model plant emission control.

With respect to fugitive emissions, the data used for this study were factors based on the petroleum refinery data referenced in Appendix C. The adequacy of these data will be determined by the programs presently underway. Radian, under EPA contract, is performing measurements in petroleum refineries; and Monsanto Research Corporation, also under EPA contract, is performing similar measurements in three organic chemical plants. In addition to the above on-going tests, the EPA has plans for testing various leak detection monitoring methods in both organic chemical plants and petroleum refineries. Completion of these programs should provide a sound basis for quantification of fugitive emissions and assessment of emission control approaches.

The storage and handling emissions were calculated using AP-42 data. AP-42 equations used for calculating fixed roof breathing do not take storage temperature control into consideration; therefore, the benzene and maleic anhydride storage emissions projected may be higher than actual.

SECTION VII

REFERENCES

1. Survey and Ranking Report in preparation by Hydroscience for ESED.
2. Lewis, W. A., Jr., G. M. Rinaldi, and T. W. Hughes, "Source Assessment: Maleic Anhydride Manufacture", Monsanto Research Corporation, received January 1978.
3. Pruessner, W. D. and L. D. Broz, "Hydrocarbon Emission Reduction Systems Utilized by Petro-Tex", presented at AIChE Meeting, March 1977.
4. J. F. Lawson, Trip Report for visit to Reichhold Chemicals, Inc., Morris, Illinois, July 28, 1977.
5. Personal communication between J. F. Lawson, Hydroscience, Inc., and R. E. Hinkson, Denka USA, November 17, 1977.

APPENDIX A

PHYSICAL PROPERTIES OF MALEIC ANHYDRIDE

TABLE A-1

PHYSICAL PROPERTIES OF MALEIC ANHYDRIDE(A)

Synonyms	Toxilic Anhydride Cis Butene Dioicanhydride 2,5-Furandione (B)
Molecular Formula	$C_4H_2O_3$
Molecular Weight	98.06
Physical State	Solid
Vapor Pressure	1 mm at 44°C
Vapor Density	3.4
Boiling Point	197-199°C
Melting Point	60°C
Density	1.48 at 20°C/4°C
Water Solubility	Reacts with Water

(A) Mitre Toxicity Study on file at offices of ESED, EPA, Research Triangle Park, North Carolina.

(B) The Merck Index, 8th Edition, Merck & Co., Rahway, New Jersey, 1968.

APPENDIX B

AIR DISPERSION PARAMETERS

TABLE B-1

ATMOSPHERIC DISPERSION PARAMETERS

MODEL PLANT 22,700 Mg/yr

Source	Emission Rate (g/sec)	Height (m)	Diameter (m)	Discharge Temp. (°K)	Flow Rate (m^3/sec)	Discharge Velocity (m/sec)
Product Recovery Absorber (Uncontrolled)	53.1 (Benzene) 67.5 (Total VOC)	27.4	1	311	20.3	25.8
Refining Vacuum System (Uncontrolled)	0.078 (Total VOC)	25.9	0.05	300	0.0075	3.8
Incinerator (With heat recovery)	0.26 (Benzene) 0.34 (Total VOC)	14.0	2	533	31.4	10.0
Carbon Adsorber (Two vents)	0.13 (Benzene) 0.13 (Total VOC)	15.2	1.5	333	18.0	10.2
Storage and Handling (Uncontrolled)						
A) Benzene	0.530 0.450	12	16	Amb		
B) Benzene	0.075 0.050	6	4	Amb		
C) Maleic Anhydride	.615 0.045	9	7	77 350		
D) Maleic Anhydride	0.021	6	6	71 344		
E) Maleic Anhydride	0.024	9	7	60 333		
F) Xylene	0.008	6	5	Amb		
Storage and Handling (controlled)						
A) Benzene	0.012 .052	12	16	Amb		
B) Benzene	0.018 .034	6	4	Amb		
C) Maleic Anhydride D) Maleic Anhydride E) Maleic Anhydride *	Trace	9.1	0.6	298	3.34	11.8
F) Xylene	0.008	6	5	Amb		
Fugitive Emissions** (Uncontrolled)	0.22 (Benzene) 0.81 (Total VOC)					
Fugitive Emissions** (Controlled)	0.017 (Benzene) 0.056 (Total VOC)					
Secondary Emissions (Uncontrolled)	0.028 (Total VOC)			Amb		

*Storage scrubber.

**Fugitive emissions are evently distributed over a rectangular area 150m by 30m.

APPENDIX C

FUGITIVE EMISSION FACTORS

Fugitive emission factors have been established for petroleum refinery operation and published in AP-42, but are based on emission quantities per unit throughput and are unsatisfactory for use here.[1] The emission factors per equipment component used in this report are based on the original emission studies[2,3] used to establish the AP-42 factors with assumptions as follows:

A. Pump Seals

1. "Uncontrolled" is the average loss measured for packed seals.
2. "Controlled" is the average loss measured for mechanicals seals assuming the major leaks are fixed.

	Uncontrolled	Controlled
Pump seals kg/day/seal	2.4	0.16

B. Valves

1. "Uncontrolled" is the average loss measured for all valves.
2. "Controlled" is the average loss measured assuming the large leaks are fixed.

	Uncontrolled	Controlled
Pipeline valves, kg/day/valve	0.068	0.006

[1] "Compilation of Air Pollutant Emission Factors". Publication No. AP-42, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, March 1975. pp. 9.1-1 to 9.1-8.

[2] Palmer, R. K. "Hydrocarbon Losses from Valves and Flanges". Report No. 2, (PB-216-682), Joint District, Federal and State Project for the Evaluation of Refinery Emissions. Air Pollution Control District, County of Los Angeles, California, March 1957. 17 pp.

[3] Steigerwald, B. J. "Emissions of Hydrocarbons to the Atmosphere from Seals on Pumps and Compressors". Report No. 6, (PB-216-582), Joint District, Federal and State Project for the Evaluation of Refinery Emissions. Air Pollution Control District, County of Los Angeles, California, April 1958. 37 pp.

APPENDIX D

COST ESTIMATE DETAILS AND CALCULATIONS

INCINERATION-CARBON ADSORPTION

COST ESTIMATES DETAILS

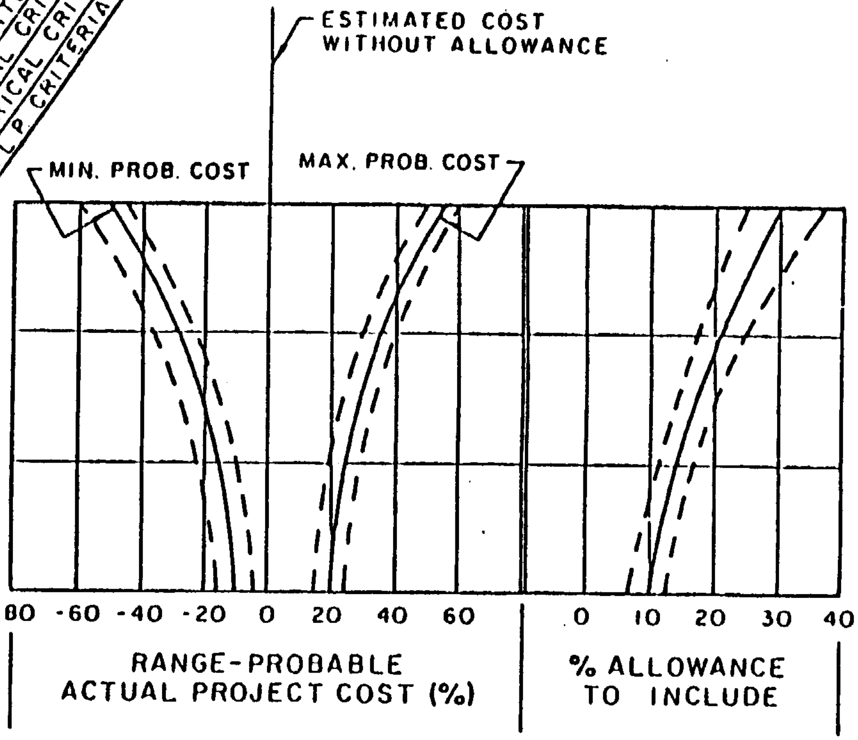
This appendix contains the details of the estimated costs used in this report.

Capital costs shown are based on an accuracy range of +30% to -30%. This range is a function of the degree of detailed data available when the estimate is made. The evaluation made in this report is a screening study based on general design criteria, block flowsheets, approximate material balances, and general equipment requirements. The attached figure illustrates the relationship between the degree of accuracy of an estimated cost and the amount of data available. The allowance indicated on this chart to cover the undefined scope of the project has been included in the estimated costs.

This type estimate is an acceptable basis to provide a screening estimate to indicate the most cost-effective alternate, within the limits of accuracy indicated.

INFORMATION USED BY ESTIMATOR

ESTIMATE TYPE	GENERAL DESIGN BASIS	BLOCK FLOWSHEET	M & E BALANCE	EQUIPMENT GEN'L DESCRIP	DETAIL PROCESS FLOWSHEET	EQ. PERF SPEC. SHEETS	GENERAL PLANT LAYOUT	PLAN & ELEV OF FACILITIES	DESIGN SK. (SPEC. EQ.)	COMPLETE EQ. SPEC. SHEETS	COMPLETE INSTR SPEC. SHEETS	P & I D'S	ELECTRICAL ONE LINE DIA	PIPELINE SCHEDULE	ARCHITECT. CRITERIA	MECHANICAL CRITERIA	ELECTRICAL CRITERIA	S B L P CRITERIA
SCREENING (PRELIM. ENG. STUDY)	●	●	●	●														
PHASE II (PRELIM. PROC. ENG.)	●	●	●	●	●	●	●	●										
PHASE III (COMPLETE PROCESS ENG. DESIGN)	●	●	●	●	●	●	●	●	●	●	●	●	●	●	●	●	●	●

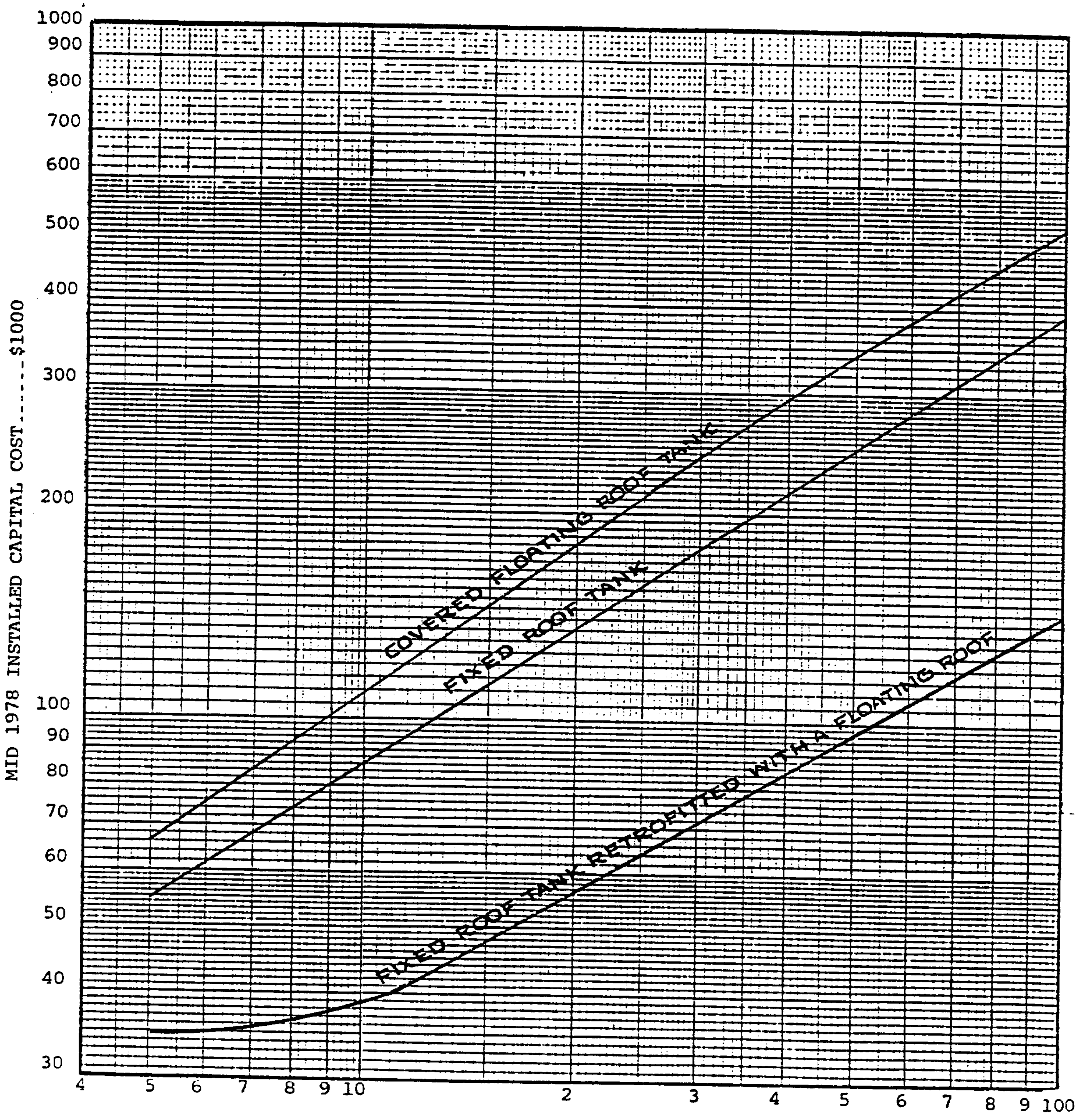


FIGURE

PRECISION OF CAPITAL COST ESTIMATES

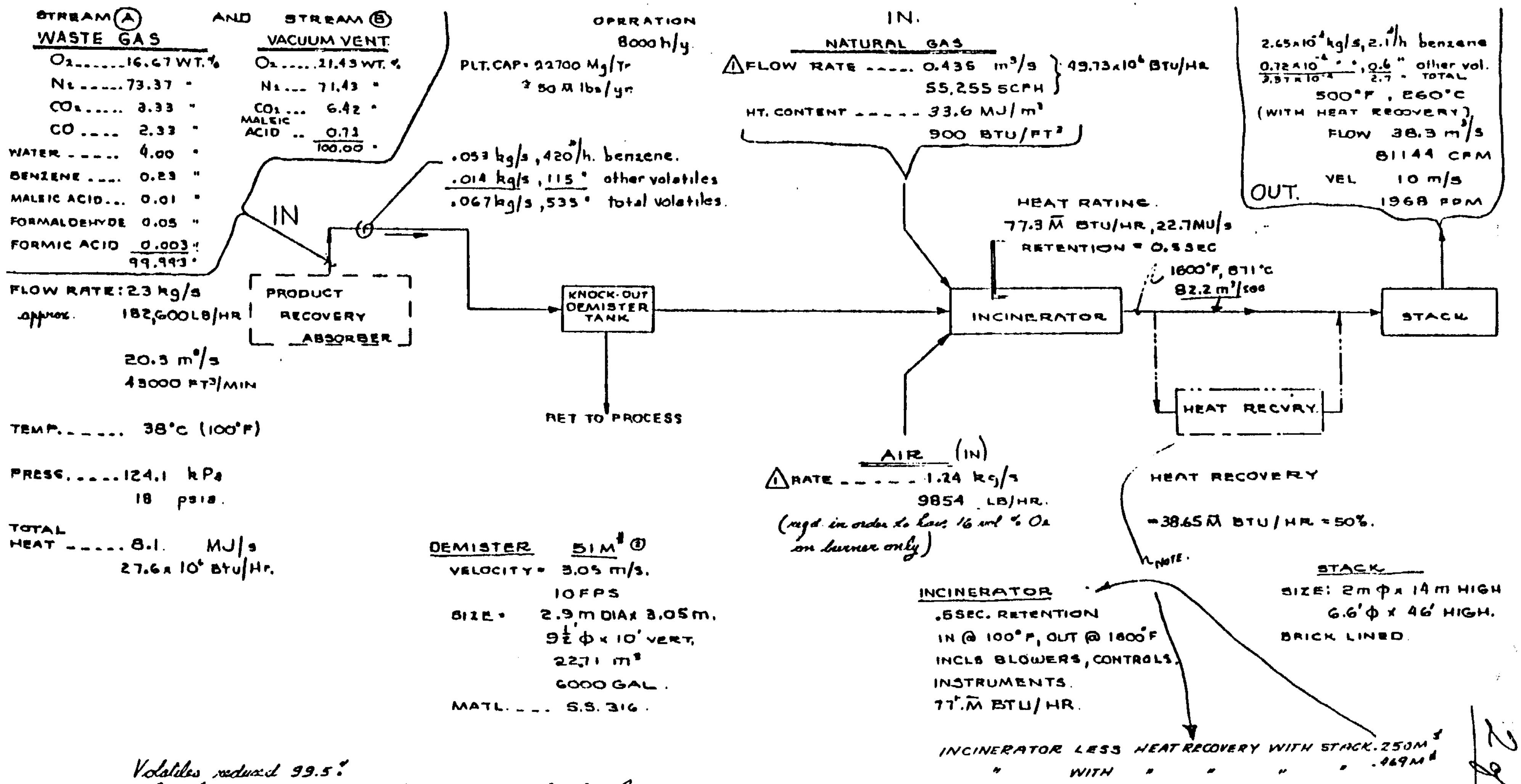
D-2(b)

INSTALLED CAPITAL COSTS FOR TANKS



TANK CAPACITY........1000 BBLS.

"INSTALLED" includes tank installed on a
foundation, painted, plus indirect capital.



Volatiles reduced 99.5%
This sheet serves direct fired incineration with & less heat recovery.
.2 Manpower

COSTS MID 1978 INSTALLED CAPITAL ... { LESS HEAT RECOVERY + 30% ALLOWANCE 430M$
WITH " " " " " 730M$

capital cost references.

① Booz-Allen Hamilton
② Hydroscience
△ Rev. 1 3.1.78 JRF

Hydroscience Environmental Systems

MALEIC ANHYDRIDE
INCINERATION
J.R.F. 9107 Rev. 1 2.17.78

PROJ. Maleic Anhydride (Incineration CHG. 9107 REV. 1 SHEET 3 OF 13

△1 Rev. 1.. 3/1/78 JRF.

Annualized Cost Details.

WASTE & RATE OF HDLG.	GASEOUS WASTE CONTAINING BENZENE 20.3m³/s, 43000 CFM	
MID 1978 INSTALLED CAPITAL COST ~~[illegible]~~		$ 400,000
△1 UTILITIES:	NATL. GAS .435 m³/s (55255 CFH x 900 BTU/FT³ x 8000 HR/YR x $2 / 10⁶ BTU) =	$ 796,000
	ELEC assume (10 KW x 8000 HR x $.03/KWH) =	2,000
	COOL H2O - =	NIL.
RAW MATLS:	DEMISTER REPLACEMENT =	NIL.
MANPOWER	.2 MAN (.2 x 10$/hr x 8000 hrs) =	$ 16,000 D-4
MAINTENANCE	5% CAP. (400M$ x .05)	$ 20,000
CAPITAL RECOVERY	14.7% CAP. (✓ x .147)	$ 59,000
MISC. CAP. COSTS	4% CAP. (✓ x .04)	16,000
WASTE DISPOSAL	NONE	none
RECOVERY CREDITS	NONE	none
△1 NET ANNUALIZED COST		$ 909,000
△1 COST EFFECTIVENESS	Benzene = $909,000/yr ÷ 1520 Mg/yr =	$ 598 /Mg
	Total VOC: 909,000/ ÷ 1936 Mg/yr =	$ 469 /Mg

PROJ. Maleic Anhydride (Reichhold Info) D-5 CHG. 9107 SHEET 4 OF 13

△1 Rev. 1 3-9-78 JRF

Reichhold System @
85% Removal.

Annualized Cost Details

	WASTE & RATE OF HDLG.		
	MID 1978 INSTALLED CAPITAL COST 1000 $	BASED ON REICHHOLD DATA & ESCLATE (J. Lawson)	$ 1,100,000
△1	UTILITIES: *	STEAM $ 245,000 from sheet 9 x 75%	184,000
△1		ELEC. $ 32,000 " " " x 75%	235 { 24,000
△1	*	COOL H₂O $ 36000 " " " x 75%	27,000
		NaOH DIL. H₂O	NIL.
	RAW MATLS:	50% NAOH (~~64~~ 81 #/h x 8000 h/y x $ .09/#)	~~60~~ 46 { ~~46,000~~ 32,000
△1		CARBON (16800#/yr x $ .85/lb) =	14,000
	MANPOWER	.3 MAN x $ 10/hr/man x 8000 h/y	24,000
	MAINTENANCE	.05 x $ 1,100,000	55,000
	CAPITAL RECOVERY	.147 x $ 1,100,000	162,000
	MISC. CAP. COSTS	.04 x $ 1,100,000	44,000
△1	WASTE DISPOSAL *	17.9 GPM x $ .01/GAL x 480,000 min x 75%	~~64,000~~ 5,000
△1	RECOVERY CREDITS	BENZENE 340#/h. x 8000 hr/y x $ .10/#	— (272,000)
	NET ANNUALIZED COST		$ ~~372,000~~ $ 299,000
△1	COST EFFECTIVENESS	BENZENE $ ~~372,000~~ 299,000 ÷ 1298 Mg/Yr = $ ~~287~~ → 236	
		TOT VOC " ÷ 1655 " " = $ ~~225~~ → 181	

* Use 75% on steam & water rqmts. because carbon loading difference. (& elec).

INSTALLED CAPITAL DETAILS.

CAPITAL REFERENCE	DESCRIPTION OF FACILITY OR ITEM	CALCS OR NOTES	CAPITAL MID 1978 $ 1000
BOOZ ALLEN FIG. IV-1	43000 CFM THERMAL INCINERATOR (.5 SEC. RETENTION) INCLS; BLOWER, INSTRUMENTATION, CONTROLS, STACK, DUCTS. -- NO HEAT RECOVERY.	$80M^{\$} \times \overset{INST.}{2.6} \times \overset{ESC}{1.117}$ =	232
H.I.	DEMISTER $9\frac{1}{2}'\phi \times 10'$ VERT. ≅ 6000GAL, SS 316,	$15M^{\$} \times \overset{ESC}{1.82} \times (\overset{MATL.}{1.66} + \overset{INT}{.20})$ =	51
		SUBTOTAL =	283
	INSTRUMENTS @ DEMISTER & W/ PIPING.	+	27
		SUBTOTAL =	310
H.I	Allowance consideration for ; magnitude @ $300^{+}M^{\$}$ and hazardous with pre-phase I information (approx. 30%)	=	90
		MID 1978 INST. CAP =	400
	RANGE -	HIGH +30% = 520 M$	
		LOW -23% = 308 M$	

PROJ. MALEIC ANHYDRIDE (INCINERATION) [DIRECT FIRE LESS HE]

CHG. 91(SHEE 1 OF 1

PROJ. Maleic Anhydride Incineration CHG. D-7 9107 SHEET 6 OF 13

△1 Rev. 1 .1.78 J

ANNUALIZED COST DETAILS.

WASTE & RATE OF HDLG.	GASEOUS WASTE CONTAINING BENZENE 20.3m^3/s, 43000CFM	
MID 1978 INSTALLED CAPITAL COST ~~[illegible]~~		$ 670,000
△1 UTILITIES:	NATL. GAS .435 m^3/s (55255 CFH x 900 BTU/FT^3 x 8000 HRS/YR x $2 / 10^6 BTU) ~~49.73 M Btu/Hr~~	$ 796,000
	ELEC. assume (10KW x 8000 HR x $.03/KWH)	$ 2,000
		798
RAW MATLS:		
MANPOWER	.2 MAN (.2 x 10$/hr x 8000 hr/yr)	$ 16,000
MAINTENANCE	5% CAP (670,000 x .05)	$ 33,000
CAPITAL RECOVERY	14.7% CAP (" x .147)	$ 98,000
MISC. CAP. COSTS	4% CAP (" x .04)	$ 27,000
WASTE DISPOSAL	NONE	
RECOVERY CREDITS	38.65 M BTU/HR x 8000 Hr x $2 / 10^6 BTU	($ 618,000)
△ NET ANNUALIZED COST		$ 354,000
△1 COST EFFECTIVENESS	Benzene = $354000/YR ÷ 1520 Mg/YR = $ 233	
	VOC = " ÷ 1936 " = $ 183	

~~2.1#/hr x 8000 hr/yr = 16800 #/yr = 7.62 Mg/yr.~~
~~2.7 " " = 21600 # = 9.80~~

INSTALLED CAPITAL DETAILS.

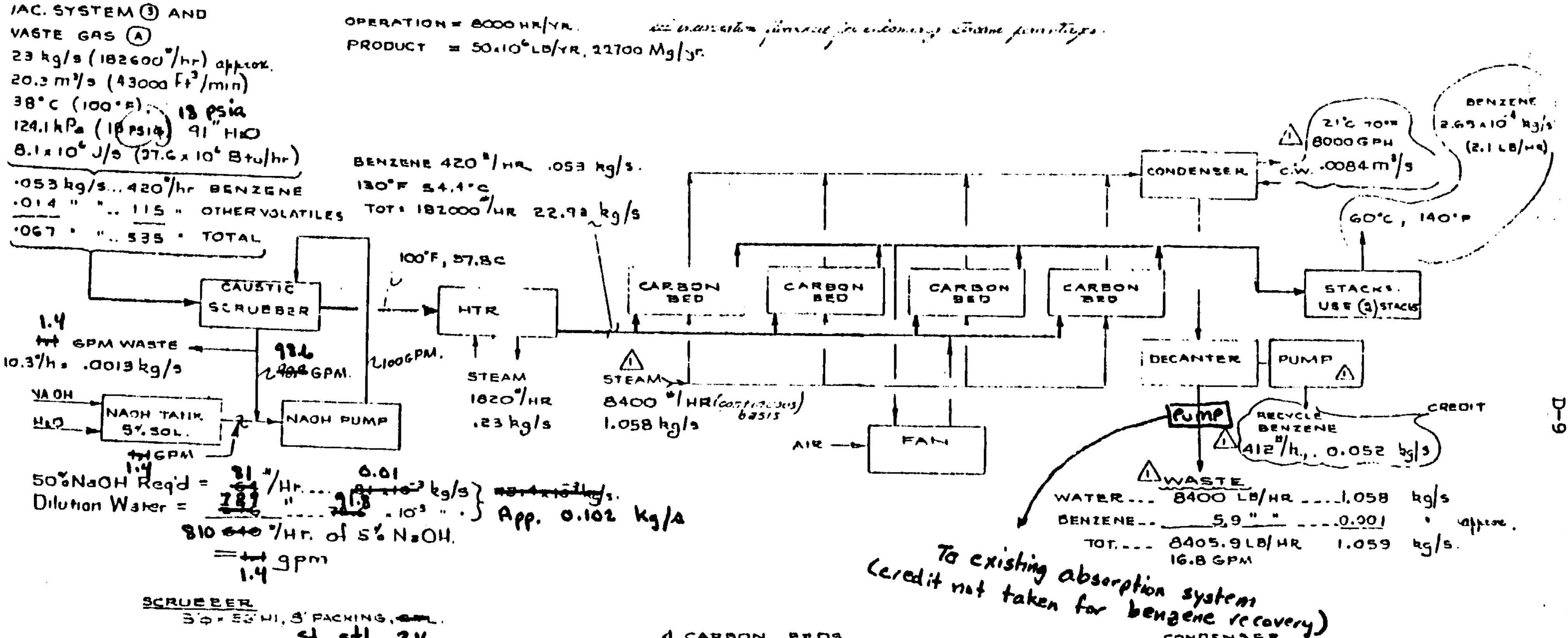
CAPITAL REFERENCE	DESCRIPTION OF FACILITY OR ITEM	CALCS OR NOTES	CAPITAL MID 1978 $ 1000
Booz Allen FIG. IV-1	43000 CFM THERMAL INCINERATOR (0.5 SECOND RETENTION TIME) INCLS; BLOWER, CONTROLS, STACK, DUCTS, INSTRUMENTATION --- WITH HEAT RECOVERY @ 50% EFF.	150 M$ x 2.6 (INST) x 1.117 (ESC) =	436
			—
H.I.	DEMISTER (assume same as direct fired less H.E.)		51
		SUBTOTAL =	487
	INSTS. @ DEMISTER & w/ PIPING + H.R.	INSTS + PIPE =	33
		SUBTOTAL =	520
H.I.	ALLOWANCE (assume 30% same as less H.E.)	approx 30% =	150
	MID 1978 INSTALLED CAPITAL	=	670
	RANGE HIGH + 30%	871 M$	
	LOW - 23%	516 M$	

PROJ: MALEIC ANHYDRIDE (INCINERATION)

CHG. 915 SHEET 1 OF 1

Direct Fired
With H.E.

79/w



SCRUBBER
3'φ x 52' HI, 8' PACKING, ~~STL.~~ St. stl. 316

NAOH PUMP
100 GPM, 120 HD, 5 HP, ~~STL.~~ St. Stl. 316

NAOH TANK
500 GAL. STL.

AIR HTR.
200 FT², STL.

4 CARBON BEDS
10'φ x 30' LG. HORIZ.
16800 LBS OF CARBON / BED 7620 kg
2 BEDS ON STREAM 5 HR LOADING CYCLE
CARBON USE: 16800 #/YR 7620 kg/YR
Δ
BED DEPTH = 2 FT,
BED VELOCITY = 75 FPM

FAN
50000 CFM, 15", 150 HP

CONDENSER
1040 FT², STL., S & T., HORIZONTAL
96.6 m²

DECANTER
50 GAL, STL.,

PUMP Δ
5 GPM, 1/4 HP, STL.

(2) STACKS
2000 FPM VELOCITY, 10.16 m/s STL
φ x 25' HI, 1.5 m DIA x 15.2 m HIGH

Δ Rev. No. 1 d. 3/8/78

MALEIC ANHYDRIDE
CARBON ADSORPTION (MODEL PLT.)
2107
J.R FORDYCE 3-8-78
REV. 1

D-10 REV. 1

PROJ. Maleic Anhydride Absorption CHG. 2-10-78 9107 SHEET 10 OF 13.

△ Rev. 1 - 3/8/78 JRF.

Annualized Cost Details.

WASTE & RATE OF HDLG.	GASEOUS WASTE CONTAINING BENZENE 20.3 m^3/s, 43000 CFM	
△ MID 1978 INSTALLED CAPITAL COST ~~[illegible]~~		\$ 1,620,000 ~~1,480,000~~
△ UTILITIES:		
△	STEAM 1.29 kg/s (10220 #/h x 8000 h x 1000 Btu x \$3/10^6 Btu)	\$ 245,000 (313
	ELEC (133 KWH x 8000 h x \$.03/kwh)	\$ 32,000
	COOL H2O .0084 m^3/s (8000 GPH x 8000 x \$.57/1000 GAL)	36,000
	NAOH DILUTION WATER 576 #/HR	NIL
△ RAW MATLS:	50% NAOH 0.01 ~~.0081~~ kg/s (81 ~~64~~ #/h x 8000 h x \$.05 ~~.09~~/#)	\$ 32,000 ~~46,000~~ 60
	CARBON 7620 kg/yr (16800 #/yr x \$.85/#)	\$ 14,000 46
MANPOWER	.3 MAN (.3 x \$10/hr x 8000 hr)	\$ 24,000
MAINTENANCE	5% CAP	\$ 81,000 ~~74,000~~
CAPITAL RECOVERY	14.7% CAP	\$ 238,000 ~~218,000~~
MISC. CAP. COSTS	4% CAP	\$ 65,000 ~~59,000~~
△ WASTE DISPOSAL	1.13 kg/s (17.9 GPM x \$.01 x 48 x 10^4 MIN)	\$ 7,000 ~~86,000~~
△ RECOVERY CREDITS	BENZENE .052 kg/s (412 #/h x 8000 x \$.10/#)	(\$ 330,000)
△ NET ANNUALIZED COST		\$ ~~504,000~~ 444,000
△ COST EFFECTIVENESS	Benzene \$ 444,000 ~~604,000~~/yr ÷ 1520 Mg/yr = \$ ~~332~~ △ — 292	
	" ÷ 1938 = \$ ~~260~~ △ — 229	

elec: (150 H. + 5 HP) 155 x .746 = 116 KW + 17 KW (insts & lights) = 133 KW.
& fractional mtrs.

△1 Rev. of Carbon Adsorption Sys. 3/10/78 J.R.F D-11 Sheet 10a of 13

PRELIMINARY CAPITAL - (MODEL PLT.) 50×10^6 lb/yr = 22700 Mg/yr

43000 CFM = 20.3 m^3/s

PHASE: MALEIC ANHYDRIDE — CARBON ADSORPTION — 100 % SIZE

CASE NUMBER: PRE-J-
SECTOR NUMBER & NAME: J.R. FORDYCE
DATE: MAR. 8, 78
JOB NUMBER: 9107

SUMMARY SHEET

SECTIONS - ALLOWANCES - ETC.	MID 1978 \$1000
CARBON ADSORPTION BED SYSTEM	~~890~~ 902
CAUSTIC SCRUB SYSTEM	386
SUBTOTAL	~~1,176~~ 1,288
ALLOWANCES (8 % HAZARD EXISTING + 18 % LITER) = 26	304
MID 1978 INSTALLED CAPITAL =	~~1,480~~ 1,620

PROBABILITY + 30%, - 23%

HIGH = 1.3 X = ~~1820~~ 2,110 M\$

LOW = 0.77 X = ~~1,140~~ 1,250 M\$

△ Rev. of Carbon Adsorption Sys. 3/10/78 J.R.F.

Sheet 10b. of 13

PRELIMINARY CAPITAL

D-12

50 x 10^6 lb/yr = 22700 Mg/yr

43000 SCFM = 20.3 m³/s

MALEIC ANHYDRIDE — CARBON ADSORPTION — 100 % SIZE

PRE-1 — J. R. FORDYCE — MARCH 8, 78 — 910

CARBON ADSORPTION BED SYSTEM

NAME OF FACILITY DESCRIPTION QUANT.		MID 197 $ 00
(4) Carbon Adsorption Beds: 10' ϕ x 30' long, horizontal, 19 M gal, stl, w/internals, ~~partial~~ st. stl 316 bed ~~panel coils~~, 67.2 M# carbon, insulate, 25 psi design		387
Ductwork & Piping (extra) 6' ϕ x 200' stl main 4½' ϕ x 160' " branches 4½' ϕ – (8) " motorized dampers 6" ϕ x 100' " stm line 4" ϕ x 160' " " " 4" ϕ – (4) " motorized valves 8" ϕ – (2 @ 80') vent 4 - 8" Valves & 100' - 10" vent.		311
(1) Condenser 1040 ft², horiz, stl, insul.		77
(1) Decanter 50 gal, vert, stl, t & c.		3
(1) Product Pump 5 gpm, cent. ¼ HP, t & c		11

△1 Rev. 1 of Carbon Adsorption Sys. 3/10/78 JRF. D-13

PRELIMINARY CAPITAL

50 x 10^6 lbs/yr = 22700 Mg/yr.

MALEIC ANHYDRIDE

43000 CFM = 20.3 m^3/s

CARBON ADSORPTION 100 % SIZE

CASE NUMBER PRE-1-

SECTION NUMBER

BY J.R. FORDYCE

DATE MAR. 8, 78

9107

SECTOR NUMBER & NAME CARBON ADSORPTION BED SYSTEM

NAME OF FACILITY DESCRIPTION QUANT.		MID 19 $ 1000
(1) Fan		67
50,000 CFM, 15", 150 HP		
Piping (extra)		15
Condenser to; decanter, to pump, to sewer, to product stg.		
Instrumentation; 15%		119
Exit vent gas pressure control Inlet-exit analysis, bed temp. control, stm controls, condenser, decanter controls.		
△3 Cond. Return Pump		
20 GPM, cent. 1/2 HP, ttc		12
Subtotal This Sector		902 ~~890~~

Δ Rev. 1 of Carbon Adsorption Sys. 3/10/78 JRF. Sheet 10 a. of 13

PRELIMINARY CAPITAL

50 x 10^6 lb/yr = 22700 Mg/yr

Sheet 4

43000 CFM = 20.3 m^3/s

MALEIC ANHYDRIDE CARBON ADSORPTION 100 % SIZE

PRE-1-

MARCH 8, 78

91

CAUSTIC SCRUB SYSTEM

NAME OF FACILITY DESCRIPTION QUANT.		MID 197 $1 00
(1) Caustic Scrubber		252 ~~102~~
8 ϕ' x 36' high, 8' ϕ x 8' pkd.		
@ 402 ft^3 of saddles, insulate		
st. stl. 316		
(1) Circulation Pump cent - or - sump		23 ~~13~~
100 gpm, 120' hd, 5 HP stl	st. stl. 316, t + c	
(1) Sump Tank stl vessel - or - ingrd. sump		6
500 gal, stl, vert, 25 psi		
(1) Heater		79
500 ft^2, shell & tube, stl, insu		
Instrumentation		26
Controls for ph, level, temp.		
Subtotal This Sector		386 ~~226~~

PROJ. Maleic Anhydride D-15 CHG. 9107 SHEET 11 OF 13

⚠1 Rev. 1 1/78 JF WASTE & RATE OF HDLG.	20.3 m³/s 43000 CFM $ COST 1,000	⊗ FACTOR =	MID 1978 8600 CFM COST 4 m³/s 1000 $	FACTOR =	MID 197_ 6 m³/s 86000 CFM COST 1000 $
MID 1978 INSTALLED CAPITAL COST 1000 $ ③	670	× 0.381 =	255	1.516 =	1016
⚠1 UTILITIES: ①	798	× 0.2 =	160	2 =	1596
RAW MATLS: ①	NONE	× 0.2 =	0	2 =	0
MANPOWER ②	16	× 1 =	16	1 =	16
MAINTENANCE ③	33	× 0.381 =	13	1.516 =	50
CAPITAL RECOVERY ③	98	× 0.381 =	37	1.516 =	149
MISC. CAP. COSTS ③	27	× 0.381 =	10	1.516 =	41
WASTE DISPOSAL ①	NONE	× 0.2 =	0	2 =	0
chemicals, or utilities, RECOVERY CREDITS ①	(618)	× 0.2 =	− (124)	2 =	− (1236)
⚠1 NET ANNUALIZED COST	354		112		61.6
⚠1 COST EFFECTIVENESS Bz	.234		.369		.203
Tot. VOC	.183		.290		.159

INCINERATION WITH HEAT RECOVERY

÷ 303.3
÷ 386.7

3032.7
3865.1

8600CFM

86000CFM

△1 Rev. 1 3.1.78 JRF WASTE & RATE OF HDLG.		20.3 m³/s 43000 CFM $ COST 1000	ⓧ FACTOR =	MID 1978 8600CFM COST 4 m³/s 1000 $	FACTOR =	MID 1978 40.6 m³/s 86000CFM COST 1000 $
MID 1978 INSTALLED CAPITAL COST 1000 $	③	400	× 0.381 =	152	1.516 =	606
△1 UTILITIES:	①	798	× 0.2 =	160	2 =	1596
RAW MATLS:	①	none	× 0.2 =	0	2 =	0
MANPOWER	②	16	× 1 =	16	1 =	16
MAINTENANCE	③	20	× 0.381 =	8	1.516 =	30
CAPITAL RECOVERY	③	59	× 0.381 =	22	1.516 =	89
MISC. CAP. COSTS	③	16	× 0.381 =	6	1.516 =	24
WASTE DISPOSAL	①	none	× 0.2 =	0	2 =	0
chemicals, or utilities, RECOVERY CREDITS	①	none	× 0.2 =	0	2 =	0
△1 NET ANNUALIZED COST		909		212		1755
△1 COST EFFECTIVENESS Bz		.600		.699		.579
Tot. VOC		.470		.548		.[illegible]4

INCINERATION LESS HEAT RECOVERY

△1 Rev. 3-78 = (all values)	50% x MODEL PLANT	165% x MODEL PLANT
WASTE & RATE OF HDLG.	21500 CFM = 10.15 m^3/s	71000 CFM = 33.5 m^3/s
MID 1978 INSTALLED CAPITAL COST 1000 $	1010 ~~900~~ M$	2,340 ~~2,090~~
UTILITIES:	41×10^6 lbs steam. 123 6×10^5 KWH elec. 18 32×10^6 Gal. Cool H_2O 18 159 M$	135×10^6 lbs steam. 405 14×10^5 KWH 42 106×10^6 Gal Cool H_2O 60 507 M$
RAW MATLS:	NaOH & Carbon @ 50% of model plt. 23 ~~30~~ M$	model plt x 165% 76 ~~99~~ M$
MANPOWER	same as model plt. 24 M$	same as model plt. 24 M$
MAINTENANCE	5% x 1010 ~~900~~ M$ 50 ~~45~~ M$	5% x 2340 ~~2090~~ 117 ~~104~~ M$
CAPITAL RECOVERY	14.7% x 1010 ~~900~~ M$ 148 ~~132~~ M$	14.7% x 2340 ~~2090~~ 344 ~~307~~ M$
MISC. CAP. COSTS	4% x 1010 ~~900~~ M$ 40 ~~36~~ M$	4% x 2340 ~~2090~~ 94 ~~84~~ M$
WASTE DISPOSAL	50% of model plt. 4 ~~49~~ M$	165% x model plt. 12 ~~147~~ M$
RECOVERY CREDITS	50% of model plt. (165) ~~(150)~~ M$	165% x model plt (544) M$
NET ANNUALIZED COST	283 ~~319~~ M$	630 ~~723~~ M$
COST EFFECTIVENESS BENZENE	283 ~~319~~ M$ ÷ 760 Mg/Yr = 372 ~~420~~ $/(Mg/yr)	630 ~~723~~ M$ ÷ 2508 Mg/yr = 251 ~~260~~ $/(Mg/yr)
TOTAL VOC	" ÷ 968 " " = 292 ~~330~~ $ "	" ÷ 3198 " " = 197 ~~226~~ $/(" ")

CARBON ADSORPTION.

Total Heat: 27.6 Btu content of waste gas
34.0 natural gas
61.6

For the 1400°F inc. system
w/& less heat recovery.
Per. J. Lawson 3/1/78.

1. The only change considered here is the utility of natural gas.

$$37{,}800 \text{ CFH} \times 900 \text{ Btu/ft}^3 \times 8{,}000 \text{ hr/yr} \times \frac{\$2}{10^6 \text{ Btu}} = \$544{,}000$$

Use \$544,000 + \$2,000 elec. = \$546,000

2. For less heat recovery, other costs are same.

3. For with heat recovery, credit for 50% of heat $\frac{61.6\ \overline{M}}{2} = 30.8\ \overline{M}\ \frac{\text{Btu}}{\text{hr}}$

$$30.8\ \overline{M} \text{ Btu/hr} \times 8000 \text{ hr/yr} \times \frac{\$2}{10^6 \text{ Btu}} = \$493{,}000\text{/yr savings.}$$