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Air

&EPA **Polymer Manufacturing Industry - Background Information for Proposed Standards**

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Preliminary Draft

NSPS

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Emission Standards and Engineering Division

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1.0 SUMMARY

1.1 REGULATORY ALTERNATIVES

Standards of performance for stationary sources of volatile organic compounds (VOC) from process and fugitive emission sources in the polymers and resins industry are being developed under the authority of Section 111 of the Clean Air Act. These standards would, in general, affect new and modified/reconstructed existing facilities that produce the following basic polymers: polypropylene, polyethylene, polystyrene, and poly(ethylene terephthalate). The fugitive emission standards would not apply to poly(ethylene terephthalate) facilities.

Because of production and emission differences, nine model plants and regulatory alternatives specific to each model plant were developed. The model plants and their regulatory alternatives are presented in detail in Chapter 6. Regulatory Alternative 1 for each model plant represents the levels of control within each industry segment in the absence of new regulations. It provides the basis for comparison of the impacts of the other regulatory alternatives.

Regulatory Alternative 2 for each model plant except poly(ethylene terephthalate) plants examines the control of fugitive emissions. These requirements are as follows:

- Monthly monitoring for leaks from valves in gas and light liquid service, and pump seals in light liquid services;

- Weekly visual inspection for liquid leakage from pump seals in light liquid service;

- Installation of controlled degassing vents on compressors, rupture disks on relief valves, and caps on open-ended lines; and

- Closed-purge sampling on sampling connections.

This particular set of requirements was adopted based upon the results of alternative levels of fugitive emission control already analyzed for the proposed fugitive emission standards for the synthetic organic chemical manufacturing industry and the petroleum refining industry. Similarities between the polymers and resins industry and these other two industries enable the results to be transferred. As this particular set of fugitive emission controls was already found to be reasonable and representative of the best system for reducing VOC fugitive emissions, no other fugitive emission control alternatives were considered.

For the poly(ethylene terephthalate) model plants, Regulatory Alternative 2 examines the reduction of ethylene glycol emissions from the process by using an alternative ethylene glycol recovery system that is more efficient than the system under Regulatory Alternative 1. No additional regulatory alternatives were developed for the poly(ethylene terephthalate) model plant using the terephthalic acid process.

Regulatory Alternative 3 for the eight remaining model plants further reduces emissions through the additional control of process emissions. For the polypropylene and polyethylene model plants, additional control was achieved by applying combustion devices to groups of emission streams that were combined on the basis of their emanating from equipment performing a particular task, such as polymerization or material recovery, within a production line. For the polystyrene model plant, Regulatory Alternative 3 attains additional control by applying additional recovery to the process emissions. Regulatory Alternative 3 for poly(ethylene terephthalate) plants using the dimethyl terephthalate process achieves additional control by combustion of the methanol stream from the methanol recovery system.

A fourth regulatory alternative (Regulatory Alternative 4) was developed for three of the nine model plants. This alternative applied combustion control to additional process emissions.

1.2 ENVIRONMENTAL IMPACT

The environmental and energy impacts of each regulatory alternative for each model plant are presented in Chapter 7. Table 1-1 presents a summary of the aggregate environmental and energy impacts; that is, each model plant's Regulatory Alternative 1 is combined and the resulting impact is reported under Regulatory Alternative I. Similarly, Regulatory

Table 1-1. ASSESSMENT OF ENVIRONMENTAL AND ECONOMIC IMPACTS
FOR EACH REGULATORY ALTERNATIVE CONSIDERED

Administrative Action	Air impact	Water impact	Solid waste impact	Energy impact	Noise impact	Economic impact
Regulatory Alternative I	-2**	0	0	0	-1**	0
Regulatory Alternative II	+2**	+1**	0	+2*	0	-1*
Regulatory Alternative III	+4**	+1**	-1*	+3*	-1**	-1*
Regulatory Alternative IV	+4**	+1**	-1*	+3*	-2**	-1*

KEY: + Beneficial impact   0 No impact   * Short term impact

      - Adverse impact   1 Negligible impact   ** Long-term impact

                                      2 Small impact   *** Irreversible impact

                                      3 Moderate impact

                                      4 Large impact

Alternative 2 impacts for each model plant are totaled as Regulatory Alternative II. Regulatory Alternative III represents Regulatory Alternative 3 impacts for eight model plants plus Regulatory Alternative 2 impacts for the poly(ethylene terephthalate) model plant using the terephthalatic acid process as no Regulatory Alternative 3 was developed for it. Finally, Regulatory Alternative IV corresponds to Regulatory Alternative 4 for three model plants plus Regulatory Alternative 3 impacts for five model plants plus Regulatory Alternative 2 impacts for one model plant.

1.2.1 Air Emissions Impacts

Total VOC emissions from new plants in these industry segments in 1988 are projected to be approximately 17.9 gigagrams (Gg) under Regulatory Alternative I, compared to 13.6, 3.4, and 3.3 Gg under Regulatory Alternatives II, III, and IV, respectively. The average percent emission reductions from the Regulatory Alternative I level achieved by Regulatory Alternatives II, III, and IV are 24, 81, and 82 percent, respectively.

1.2.2 Water, Solid Waste, and Noise Impacts

Little adverse affect on water quality is expected under any of the regulatory alternatives. Implementation of fugitive controls under Regulatory Alternative II would result in a small positive effect on water by curtailment of potential liquid leaks.

Minor adverse solid waste impacts could occur under Regulatory Alternatives III and IV due to the use of catalytic incinerators. Spent catalyst from catalytic incinerator use may be generated at an annual rate of 2.9 m³ (102 ft³) and 3.0 m³ (135 ft³), respectively, under these two alternatives.

Some noise impact could arise from increased use of flares under the regulatory alternatives. By employing noise mitigation techniques, additional noise impact on surrounding communities should be minimal.

1.2.3 Energy Impacts

Under Regulatory Alternative II, implementation of fugitive controls in seven of the nine model plants and of a more efficient ethylene glycol recovery system in poly(ethylene terephthalate) plants result in a net decrease in energy usage from what otherwise would

occur under Regulatory Alternative 1. This occurs because the energy credit obtained through recovered VOC is greater than the energy expended to implement the controls.

Under Regulatory Alternative III, a net decrease in energy usage is also obtained. Six of the nine industry segments realize a net decrease, while the other three segments realize a net increase in energy usage.

Under Regulatory Alternative IV, a net decrease in energy usage across the industry is again obtained, with six segments realizing a net decrease and three segments realizing a net increase in energy usage.

1.3 ECONOMIC IMPACT

As was done for the environmental and energy impacts, the aggregate economic impacts that result from the costs for each of the regulatory alternatives are summarized in Table 1-1. A more detailed economic analysis is presented in Chapter 9 and a more detailed analysis of costs for each industry segment is presented in Chapter 8.

Under Regulatory Alternative II, the industry as a whole would realize a net annual credit of around $0.2 million in the fifth year (1988) compared to what the industry as a whole would otherwise spend under Regulatory Alternative I. Fifth year annual costs compared to Regulatory Alternative I for individual industry segments range from an annual credit of $0.6 million in the poly(ethylene terephthalate) plants using the terephthalatic acid process up to $1.6 million in polyethylene plants using the gas phase technology.

Under Regulatory Alternative III, the industry as a whole would spend an net annual amount of $2.8 million in the fifth year (1988) over and above what they would otherwise spend under Regulatory Alternative I. Fifth year annual costs for individual industry segments range from annual credits in poly(ethylene terephthalate) and polystyrene plants up to $3.2 million for high density polyethylene plants using a solution process.

Under Regulatory Alternative IV, a fifth year annual cost of $4.3 million over and above what would be otherwise spent under Regulatory Alternative I would be realized by the industry as a whole. Again,

the range of annual costs for individual industry segments range from net annual credits in the poly(ethylene terephthalate) and polystyrene segments to a cost of $3.2 million in high density polyethylene, solution process segment.

Under the most costly combination of individual regulatory alternatives for each model plant, total additional annualized costs of controls in 1988 are estimated to be $4.9 million. The potential adverse economic impacts of these regulatory alternatives are expected to be very minor in view of the small price increases anticipated as a result of control costs.

2.0 INTRODUCTION

2.1 BACKGROUND AND AUTHORITY FOR STANDARDS

Before standards of performance are proposed as a Federal regulation, air pollution control methods available to the affected industry and the associated costs of installing and maintaining the control equipment are examined in detail. Various levels of control based on different technologies and degrees of efficiency are expressed as regulatory alternatives. Each of these alternatives is studied by EPA as a prospective basis for a standard. The alternatives are investigated in terms of their impacts on the economics and well-being of the industry, the impacts on the national economy, and the impacts on the environment. This document summarizes the information obtained through these studies so that interested persons will be able to see the information considered by EPA in the development of the proposed standard.

Standards of performance for new stationary sources are established under Section 111 of the Clean Air Act (42 U.S.C. 7411) as amended, hereinafter referred to as the Act. Section 111 directs the Administrator to establish standards of performance for any category of new stationary source of air pollution which "... causes, or contributes significantly to air pollution which may reasonably be anticipated to endanger public health or welfare."

The Act requires that standards of performance for stationary sources reflect,"... the degree of emission reduction achievable which (taking into consideration the cost of achieving such emission reduction, and any nonair quality health and environmental impact and energy requirements) the Administrator determines has been adequately demonstrated for that category of sources." The standards apply only to stationary sources, the construction or modification of which commences after regulations are proposed by publication in the Federal Register.

The 1977 amendments to the Act altered or added numerous provisions that apply to the process of establishing standards of performance. Examples of the effects of the 1977 amendments are:

1. EPA is required to review the standards of performance every 4 years and, if appropriate, revise them.

2. EPA is authorized to promulgate a standard based on design, equipment, work practice, or operational procedures when a standard based on emission levels is not feasible.

3. The term "standards of performance" is redefined, and a new term "technological system of continuous emission reduction" is defined. The new definitions clarify that the control system must be continuous and may include a low- or non-polluting process or operation.

4. The time between the proposal and promulgation of a standard under section 111 of the Act may be extended to 6 months.

Standards of performance, by themselves, do not guarantee protection of health or welfare because they are not designed to achieve any specific air quality levels. Rather, they are designed to reflect the degree of emission limitation achievable through application of the best adequately demonstrated technological system of continuous emission reduction, taking into consideration the cost of achieving such emission reduction, any nonair quality health and environmental impacts, and energy requirements.

Congress had several reasons for including these requirements. First, standards with a degree of uniformity are needed to avoid situations where some States may attract industries by relaxing standards relative to other States. Second, stringent standards enhance the potential for long-term growth. Third, stringent standards may help achieve long-term cost savings by avoiding the need for more retrofitting when pollution ceilings may be reduced in the future. Fourth, certain types of standards for coalburning sources can adversely affect the coal market by driving up the price of low-sulfur coal or effectively excluding certain coals from the reserve base because their untreated pollution potentials are high. Congress does not intend that new source performance standards contribute to these problems. Fifth, the standard-setting process should create incentives for improved technology.

Promulgation of standards of performance does not prevent State or local agencies from adopting more stringent emission limitations for the same sources. States are free under Section 116 of the Act to establish even more stringent emission limits than those established under Section 111 or those necessary to attain or maintain the National Ambient Air Quality Standards (NAAQS) under Section 110. Thus, new sources may in some cases be subject to limitations more stringent than standards of performance under Section 111, and prospective owners and operators of new sources should be aware of this possibility in planning for such facilities.

A similar situation may arise when a major emitting facility is to be constructed in a geographic area that falls under the prevention of significant deterioration of air quality provisions of Part C of the Act. These provisions require, among other things, that major emitting facilities to be constructed in such areas are to be subject to best available control technology. The term Best Available Control Technology (BACT), as defined in the Act, means

> ... an emission limitation based on the maximum degree of reduction of each pollutant subject to regulation under this Act emitted from, or which results from, any major emitting facility, which the permitting authority, on a case-by-case basis, taking into account energy, environmental, and economic impacts and other costs, determines is achievable for such facility through application of production processes and available methods, systems, and techniques, including fuel cleaning or treatment or innovative fuel combustion techniques for control of each such pollutant. In no event shall application of "best available control technology" result in emissions of any pollutants which will exceed the emissions allowed by any applicable standard established pursuant to Sections 111 or 112 of this Act. (Section 169(3))

Although standards of performance are normally structured in terms of numerical emission limits where feasible, alternative approaches are sometimes necessary. In some cases physical measurement of emissions from a new source may be impractical or exorbitantly expensive. Section 111(h) provides that the Administrator may promulgate a design or equipment standard in those cases where it is not feasible to prescribe or enforce a standard of performance. For example, emissions of hydrocarbons from storage vessels for petroleum liquids are greatest during tank filling. The nature of the emissions, high concentrations

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for short periods during filling and low concentrations for longer periods during storage, and the configuration of storage tanks make direct emission measurement impractical. Therefore, a more practical approach to standards of performance for storage vessels has been equipment specification.

In addition, Section 111(j) authorizes the Administrator to grant waivers of compliance to permit a source to use innovative continuous emission control technology. In order to grant the waiver, the Administrator must find: (1) a substantial likelihood that the technology will produce greater emission reductions than the standards require or an equivalent reduction at lower economic energy or environmental cost; (2) the proposed system has not been adequately demonstrated; (3) the technology will not cause or contribute to an unreasonable risk to the public health, welfare, or safety; (4) the governor of the State where the source is located consents; and (5) the waiver will not prevent the attainment or maintenance of any ambient standard. A waiver may have conditions attached to assure the source will not prevent attainment of any NAAQS. Any such condition will have the force of a performance standard. Finally, waivers have definite end dates and may be terminated earlier if the conditions are not met or if the system fails to perform as expected. In such a case, the source may be given up to three years to meet the standards with a mandatory progress schedule.

2.2 SELECTION OF CATEGORIES OF STATIONARY SOURCES

Section 111 of the Act directs the Adminstrator to list categories of stationary sources. The Administrator "... shall include a category of sources in such list if in his judgment it causes, or contributes significantly to, air pollution which may reasonably be anticipated to endanger public health or welfare." Proposal and promulgation of standards of performance are to follow.

Since passage of the Clean Air Act of 1970, considerable attention has been given to the development of a system for assigning priorities to various source categories. The approach specifies areas of interest by considering the broad strategy of the Agency for implementing the Clean Air Act. Often, these "areas" are actually pollutants emitted by stationary sources. Source categories that emit these pollutants

are evaluated and ranked by a process involving such factors as:
(1) the level of emission control (if any) already required by State
regulations, (2) estimated levels of control that might be required
from standards of performance for the source category, (3) projections
of growth and replacement of existing facilities for the source category,
and (4) the estimated incremental amount of air pollution that could
be prevented in a preselected future year by standards of performance
for the source category. Sources for which new source performance
standards were promulgated or under development during 1977, or earlier,
were selected on these criteria.

The Act amendments of August 1977 establish specific criteria to
be used in determining priorities for all major source categories not
yet listed by EPA. These are: (1) the quantity of air pollutant
emissions that each such category will emit, or will be designed to
emit; (2) the extent to which each such pollutant may reasonably be
anticipated to endanger public health or welfare; and (3) the mobility
and competitive nature of each such category of sources and the consequent
need for nationally applicable new source standards of performance.

In some cases it may not be feasible immediately to develop a
standard for a source category with a high priority. This might
happen when a program of research is needed to develop control techniques
or because techniques for sampling and measuring emissions may require
refinement. In the developing of standards, differences in the time
required to complete the necessary investigation for different source
categories must also be considered. For example, substantially more
time may be necessary if numerous pollutants must be investigated from
a single source category. Further, even late in the development
process the schedule for completion of a standard may change. For
example, inability to obtain emission data from well-controlled sources
in time to pursue the development process in a systematic fashion may
force a change in scheduling. Nevertheless, priority ranking is, and
will continue to be, used to establish the order in which projects are
initiated and resources assigned.

After the source category has been chosen, the types of facilities
within the source category to which the standard will apply must be

determined. A source category may have several facilities that cause
air pollution, and emissions from some of these facilities may vary
from insignificant to very expensive to control. Economic studies of
the source category and of applicable control technology may show that
air pollution control is better served by applying standards to the
more severe pollution sources. For this reason, and because there is
no adequately demonstrated system for controlling emissions from
certain facilities, standards often do not apply to all facilities at
a source. For the same reasons, the standards may not apply to all
air pollutants emitted. Thus, although a source category may be
selected to be covered by a standard of performance, not all pollutants
or facilities within that source category may be covered by the standards.

2.3 PROCEDURE FOR DEVELOPMENT OF STANDARDS OF PERFORMANCE

Standards of performance must (1) realistically reflect best
demonstrated control practice; (2) adequately consider the cost, the
nonair quality, health and environmental impacts, and the energy requirements
of such control; (3) be applicable to existing sources that are modified
or reconstructed as well as new installations; and (4) meet these
conditions for all variations of operating conditions being considered
anywhere in the country.

The objective of a program for developing standards is to identify
the best technological system of continuous emission reduction that
has been adequately demonstrated. The standard-setting process involves
three principal phases of activity: (1) information gathering, (2)
analysis of the information, and (3) development of the standard of
performance.

During the information-gathering phase, industries are queried
through a telephone survey, letters of inquiry, and plant visits by EPA
representatives. Information is also gathered from many other sources,
and a literature search is conducted. From the knowledge acquired about
the industry, EPA selects certain plants at which emission tests are con-
ducted to provide reliable data that characterize the pollutant emissions
from well-controlled existing facilities.

In the second phase of a project, the information about the industry
and the pollutants emitted is used in analytical studies. Hypothetical

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"model plants" are defined to provide a common basis for analysis. The model plant definitions, national pollutant emission data, and existing State regulations governing emissions from the source category are then used in establishing "regulatory alternatives." These regulatory alternatives are essentially different levels of emission control.

EPA conducts studies to determine the impact of each regulatory alternative on the economics of the industry and on the national economy, on the environment, and on energy consumption. From several possibly applicable alternatives, EPA selects the single most plausible regulatory alternative as the basis for a standard of performance for the source category under study.

In the third phase of a project, the selected regulatory alternative is translated into a standard of performance, which, in turn, is written in the form of a Federal regulation. The Federal regulation, when applied to newly constructed plants, will limit emissions to the levels indicated in the selected regulatory alternative.

As early as is practical in each standard-setting project, EPA representatives discuss the possibilities of a standard and the form it might take with members of the National Air Pollution Control Techniques Advisory Committee. Industry representatives and other interested parties also participate in these meetings.

The information acquired in the project is summarized in the Background Information Document (BID). The BID, the standard, and a preamble explaining the standard are widely circulated to the industry being considered for control, environmental groups, other government agencies, and offices within EPA. Through this extensive review process, the points of view of expert reviewers are taken into consideration as changes are made to the documentation.

A "proposal package" is assembled and sent through the offices of EPA Assistant Administrators for concurrence before the proposed standard is officially endorsed by the EPA Administrator. After being approved by the EPA Administrator, the preamble and the proposed regulation are published in the Federal Register.

As a part of the Federal Register announcement of the proposed regulation, the public is invited to participate in the standard-setting process. EPA invites written comments on the proposal and also holds a

public hearing to discuss the proposed standard with interested parties. All public comments are summarized and incorporated into a second volume of the BID. All information reviewed and generated in studies in support of the standard of performance is available to the public in a "docket" on file in Washington, D. C.

Comments from the public are evaluated, and the standard of performance may be altered in response to the comments.

The significant comments and EPA's position on the issues raised are included in the "preamble" of a "promulgation package," which also contains the draft of the final regulation. The regulation is then subjected to another round of review and refinement until it is approved by the EPA Administrator. After the Administrator signs the regulation, it is published as a "final rule" in the Federal Register.

2.4 CONSIDERATION OF COSTS

Section 317 of the Act requires an economic impact assessment with respect to any standard of performance established under Section 111 of the Act. The assessment is required to contain an analysis of: (1) the costs of compliance with the regulation, including the extent to which the cost of compliance varies depending on the effective date of the regulation and the development of less expensive or more efficient methods of compliance; (2) the potential inflationary or recessionary effects of the regulation; (3) the effects the regulation might have on small business with respect to competition; (4) the effects of the regulation on consumer costs; and (5) the effects of the regulation on energy use. Section 317 also requires that the economic impact assessment be as extensive as practicable.

The economic impact of a proposed standard upon an industry is usually addressed both in absolute terms and in terms of the control costs that would be incurred as a result of compliance with typical, existing State control regulations. An incremental approach is necessary because both new and existing plants would be required to comply with State regulations in the absence of a Federal standard of performance. This approach requires a detailed analysis of the economic impact from the cost differential that would exist between a proposed standard of performance and the typical State standard.

Air pollutant emissions may cause water pollution problems, and captured potential air pollutants may pose a solid waste disposal problem. The total environmental impact of an emission source must, therefore, be analyzed and the costs determined whenever possible.

A thorough study of the profitability and price-setting mechanisms of the industry is essential to the analysis so that an accurate estimate of potential adverse economic impacts can be made for proposed standards. It is also essential to know the capital requirements for pollution control systems already placed on plants so that the additional capital requirements necessitated by these Federal standards can be placed in proper perspective. Finally, it is necessary to assess the availability of capital to provide the additional control equipment needed to meet the standards of performance.

2.5 CONSIDERATION OF ENVIRONMENTAL IMPACTS

Section 102(2)(C) of the National Environmental Policy Act (NEPA) of 1969 requires Federal agencies to prepare detailed environmental impact statements on proposals for legislation and other major Federal actions significantly affecting the quality of the human environment. The objective of NEPA is to build into the decisionmaking process of Federal agencies a careful consideration of all environmental aspects of proposed actions.

In a number of legal challenges to standards of performance for various industries, the United States Court of Appeals for the District of Columbia Circuit has held that environmental impact statements need not be prepared by the Agency for proposed actions under Section 111 of the Clean Air Act. Essentially, the Court of Appeals has determined that the best system of emission reduction requires the Administrator to take into account counter-productive environmental effects of a proposed standard, as well as economic costs to the industry. On this basis, therefore, the Court established a narrow exemption from NEPA for EPA determination under Section 111.

In addition to these judicial determinations, the Energy Supply and Environmental Coordination Act (ESECA) of 1974 (PL-93-319) specifically exempted proposed actions under the Clean Air Act from NEPA requirements. According to Section 7(c)(1), "No action taken under the Clean Air Act

shall be deemed a major Federal action significantly affecting the
quality of the human environment within the meaning of the National
Environmental Policy Act of 1969." (15 U.S.C. 793(c)(1))

Nevertheless, the Agency has concluded that the preparation of
environmental impact statements could have beneficial effects on
certain regulatory actions. Consequently, although not legally required
to do so by Section 102(2)(C) of NEPA, EPA has adopted a policy requiring
that environmental impact statements be prepared for various regulatory
actions, including standards of performance developed under Section 111
of the Act. This voluntary preparation of environmental impact statements,
however, in no way legally subjects the Agency to NEPA requirements.

To implement this policy, a separate section in this document is
devoted solely to an analysis of the potential environmental impacts
associated with the proposed standards. Both adverse and beneficial
impacts in such areas as air and water pollution, increased solid
waste disposal, and increased energy consumption are discussed.

2.6 IMPACT ON EXISTING SOURCES

Section 111 of the Act defines a new source as "... any stationary
source, the construction or modification of which is commenced ..."
after the proposed standards are published. An existing source is
redefined as a new source if "modified" or "reconstructed" as defined
in amendments to the general provisions of Subpart A of 40 CFR Part
60, which were promulgated in the Federal Register on December 16,
1975 (40 FR 58416).

Promulgation of a standard of performance requires States to
establish standards of performance for existing sources in the same
industry under Section 111(d) of the Act if the standard for new
sources limits emissions of a designated pollutant (i.e., a pollutant
for which air quality criteria have not been issued under Section 108
or which has not been listed as a hazardous pollutant under Section 112).
If a State does not act, EPA must establish such standards. General
provisions outlining procedures for control of existing sources under
Section 111(d) were promulgated on November 17, 1975, as Subpart B of
40 CFR Part 60 (40 FR 53340).

2.7 REVISION OF STANDARDS OF PERFORMANCE

Congress was aware that the level of air pollution control achievable by any industry may improve with technological advances. Accordingly, Section 111 of the Act provides that the Administrator "... shall, at least every 4 years, review and, if appropriate, revise ..." the standards. Revisions are made to assure that the standards continue to reflect the best systems that become available in the future. Such revisions will not be retroactive, but will apply to stationary sources constructed or modified after the proposal of the revised standards.

3.0 THE POLYMERS AND RESINS INDUSTRY

The polymers and resins industry consists of operations that combine monomer or chemical intermediate materials obtained from the basic petrochemical industry and the synthetic organic chemical manufacturing industry (SOCMI) into polymeric or copolymeric products. (A copolymer is formed when two different monomers are polymerized together so that both occur in the same polymer chain. The copolymer will generally combine to some extent the properties of the individual polymers and will often have lower strength and a lower melting point than either of the polymers.) Such products include plastic materials, synthetic resins, synthetic rubbers, and synthetic fibers. This chapter describes the polymers and resins industry, its production processes and associated emissions of volatile organic compounds (VOC), and industry practices and State and Federal regulations affecting VOC emissions.

3.1 INDUSTRY DESCRIPTION

A large number of polymers are produced domestically by a variety of processes. Polymers can be grouped into two categories: thermoplastic, those which melt upon reheating and, thus, can be reshaped after initial fabrication, and thermosetting, those which do not. Thermoplastic polymers are linear chain polymers with little or no crosslinking between the individual chains. Common end-uses include safety shields, clothing, appliance parts, boiling bags, sutures, textiles and woven goods, bottles for a variety of fluids, toys, and hot/cold insulated drink cups. Thermosetting polymers are extensively crosslinked, making them far more rigid and often insoluble. These resins are often used in applications where rigidity or heat-resistant characteristics are required. End-uses include molding compounds, adhesives and bonding resins, laminating resins, paper and surface coatings, and as a binder for fiber glass and other reinforced plastics for construction and transportation applications.

The selection of a suitable polymer for a particular end-use application depends on the specific properties of the polymer.

The U.S. Environmental Protection Agency (EPA) study by the Pullman-Kellogg Company ranked segments of the polymers and resins industry by magnitude of emissions for the purpose of new source performance standard (NSPS) development priority setting.[1] This study examined 16 polymers and resins categories with potentially large VOC emissions:

Acrylics	Polyester Fibers
Alkyds	Polypropylene
High Density Polyethylene	Polystyrene
Low Density Polyethylene	Polyvinyl Acetate
Melamine Formaldehyde	Polyvinyl Alcohol
Nylon 6	Styrene-Butadiene Latex
Nylon 66	Unsaturated Polyester Resins
Phenol Formaldehyde	Urea Formaldehyde

The majority of these 16 polymers are of the thermoplastic type (acrylics, polyethylene, nylon 6 and 66, (saturated) polyester resin, polypropylene, polystyrene, polyvinyl acetate, polyvinyl alcohol, and styrene-butadiene latex); the remainder are thermosetting resins.

The survey of these 16 polymers and resins categories showed that emissions from five of these categories amount to approximately 75 percent of the current total estimated VOC emissions from these 16 polymers and resins manufacturing operations. These five source categories, all of which were found to be experiencing growth, were chosen for NSPS development. They are:

1. Polypropylene
2. Low density polyethylene
3. High density polyethylene
4. Polystyrene, and
5. Polyester resin, poly(ethylene terephthalate), [PET].

Tables 3-1 to 3-5 list existing production locations and capacities for plants that produce these five polymers and resins.

3.1.1 End-Uses of the Five Polymers Chosen for NSPS Development

The 16 commercial polymers and resins covered by the Pullman-Kellogg report have an extremely wide variety of end-uses and are found in

Table 3-1. POLYPROPYLENE (PP) PLANT LIST[a]

Company	Location	Capacity Gg/yr
ARCO Polymers, Inc.	La Porte, TX	181
Amoco Chemical Corp.	Chocolate Bayou, TX	234
El Paso Polyolefins Co.	Odessa, TX	68
	Pasadena, TX	68
Exxon Chemical Co.	Baytown, TX	181
Gulf Oil Chemical Co.	Cedar Bayou, TX	181
Hercules, Inc.	Bayport, TX	204
	Lake Charles, LA	395
Northern Petrochemical Co.	Morris, IL	91
Phillips Chemical Co.	Pasadena, TX	91
Shell Chemical Co.	Norco, LA	136
	Woodbury, NJ	136
Soltex Polymer Corp.	Deer Park, TX	91
Texas Eastman Co.	Longview, TX	64
USS Chemicals	La Porte, TX	159
	Kenova, WV	75

[a]Source: SRI International, 1982 Directory of Chemical Producers, United States.

Table 3-2. LOW DENSITY POLYETHYLENE (LDPE) PLANT LIST[a]

Company	Location	Average capacity, Gg/yr
Allied Chemical Co.[b]	Orange, TX	-
ARCO Polymers, Inc.	Port Arthur, TX	181
Chemplex Co.	Clinton, IA	188
Cities Service Co.[c]	Lake Charles, LA	329
Dow Chemical U.S.A.	Freeport, TX	447
	Plaquemine, LA	254
E.I. du Pont de Nemours & Co., Inc.	Orange, TX	211
	Victoria, TX	109
El Paso Polyolefins Co.	Odessa, TX	181
	Pasadena, TX	68
Exxon Chemical Co.	Baton Rouge, LA	299[d]
	Mt. Belview, TX[d]	154[d]
Gulf Oil Chemical Co.	Cedar Bayou, TX	259
	Orange, TX	129
	Baytown, TX[b]	-
Mobil Chemical Co.	Beaumont, TX	136
National Distillers & Chemical Corp.	Deer Park, TX	249
	Tuscola, IL	75
Northern Petrochemical Co.	Morris, IL	273[e]
Phillips Chemical Co.	Pasadena, TX	-[f]
Texas Eastman Co.	Longview, TX	166
Union Carbide Corp.	Seadrift, TX	544[g]
	Taft, LA	272
	Penuelas, P.R.	141
United Foam Corp.[h]	Louisville, KY	73

[a]Source: SRI International, 1982 Directory of Chemical Producers, United States, unless otherwise indicated.

[b]Source: Texas Air Control Board communication.

[c]In a letter dated March 9, 1982, Cities Services Co. indicated that they were closing all their polymer and resins manufacturing plants.

[d]Source: Chemical Engineering. October 18, 1982. p. 26.

[e]Capacity obtained from April 14, 1982, letter from Northern Petrochemical Company.

[f]Primarily HDPE produced; small portion of 680 Gg total capacity used for LDPE.

[g]Capacity is about 245 Gg liquid phase and 299 Gg gas phase.

[h]Source: Organic Chemical Producers Data Base. Product Data Report-Nationwide. February 17, 1981, p. 926.

Table 3-3. HIGH DENSITY POLYETHYLENE (HDPE) PLANT LIST[a]

Company	Location	Capacity, Gg/yr
Allied Chemical Corp.	Baton Rouge, LA	299
American Hoechst Corp.	Bayport, TX	136
Amoco Chemical Corp.	Chocolate Bayou, TX	172
ARCO Polymers, Inc.	Port Arthur, TX	159
Chemplex Co.	Clinton, IA	122
Dow Chemical U.S.A.	Freeport, TX	54
	Plaquemine, LA	150
E.I. du Pont de Nemours & Co., Inc.	Orange, TX	104
	Victoria, TX	102
Gulf Oil Corp.	Orange, TX	261[b]
Hercules, Inc.	Lake Charles, LA	7
Nat'l. Petrochemical Corp.	La Porte, TX	283
Phillips Chemical Co.	Pasadena, TX	680
Soltex Polymer Corp.	Deer Park, TX	340
Union Carbide Corp.	Seadrift, TX	91

[a]Source: SRI International, 1982 Directory of Chemical Producers, United States.

[b]Expansion from 191 Gg/yr to 261 Gg/yr completed in 1982. Chemical Engineering. October 4, 1982. p. 25.

Table 3-4. POLYSTYRENE (PS) PLANT LIST[a]

Company	Location	Capacity, Gg/yr	Process[b]
A&E Plastics, Inc.	City of Industry, CA	16	-
American Hoechst Corp.	Chesapeake, VA	118	-
	Leominster, MA	54	-
	Peru, IL	113	-
Amoco Chemical Corp.	Joliet, IL	136	Continuous
	Torrance, CA	16	Batch
	Willow Springs, IL	41	Batch
ARCO Polymers, Inc.	Beaver Valley, PA	213	-
BASF Wyandotte Corp.	South Brunswick, NJ	79	Batch
Carl Gordon, Ind., Inc.	Owensboro, KY	23	Batch
	Oxford, MA	45	Batch, Continuous
Cosden Oil & Chemical Co.	Windsor, NJ	54	-
	Calumet City, IL	122	Continuous
	Big Spring, TX	20	-
	Orange, CA	27	-
Crest Container Corp.[c]	Fort Worth, TX	3.5	-
Dow Chemical Corp.	Gales Ferry, CT	86	-
	Midland, MI	100	Continuous
	Torrance, CA	91	Continuous
	Ironton, OH	86	-
	Joliet, IL	64	-
Gulf Oil Chemical Co.	Marietta, OH	141	Continuous
	Channelview, TX[d]	18	-
Huntsman-Goodson Chem. Corp.	Troy, OH	9	-
Kama Corp.	Hazelton, PA	11	-
Mobil Chemical Co.	Holyoke, MA	41	-
	Joliet, IL	18	-
	Santa Ana, CA	29	Continuous
Monsanto Co.[e]	Addyston, OH	136	Continuous
	Decatur, AL	45	Continuous
	Springfield, MA	136	Continuous
Polysar Resins, Inc.	Copley, OH	54	Continuous
	Leominster, MA	54	-
	Forest City, NC	18	-
Richardson Company	West Haven, CT	-[f]	-
Shell Chemical Co.	Belpre, OH	136	Continuous
Texstyrene Plastics, Inc.	Fort Worth, TX	23	-
U.S.S. Chemicals[g]	Haverhill, OH	9	-
Vititek Inc.	Delano, CA	2	-

[a]Source: SRI International, 1982 Directory of Chemical Producers, United States, unless otherwise indicated.

[b]Source: Industry communications.

[c]Source: Organic Chemical Producers Data Base. Product Data Report - Nationwide. February 17, 1981. p. 943.

[d]This plant is not currently in production. Letter from Gulf Oil to Texas Air Control Board. July 28, 1982.

[e]Monsanto's Long Beach plant has been closed.

[f]In mid-1977, this company switched its 18 Gg PS plant to production of other styrene copolymers. Small quantities of specialty grade PS are still being produced.

[g]Telephone conversation on October 12, 1982, with U.S.S. Chemicals indicated that this plant has been closed.

Table 3-5. POLYETHYLENE TEREPHTHALATE (PET) POLYESTER PLANT LIST[a]

Plant	Location	Process[b]	Average capacity, Gg/yr	Product
Akzona Inc.	Central, SC	DMT	25	Fiber
	Lowland, TN	DMT	43	Fiber
Allied Corp.	Moncure, NC	TPA	39	Fiber
American Hoechst Corp.	Spartanburg, SC	DMT/TPA	261	Fiber, Bottle Resins
	Greer, SC	-	32	Bottle Resins, Film
Avtex Fibers, Inc.	Lewistown, PA	DMT/TPA	18	Fiber
E.I. du Pont de Nemours	Camden, SC			
	Charleston, SC			
	Chattanooga, TN	DMT/TPA	736	Fiber
	Kinston, NC			
	Old Hickory, TN			
	Wilmington, NC			
	Old Hickory, TN	-	249	-
	Wilmington, NC	-	566	-
	Brevard, NC	-	14	Film
	Parlin, NJ	-	7	Film
	Circleville, OH	-	29	Film
	Florence, SC	-	34	Film
Eastman Kodak Co.	Columbia, SC	DMT	204	Fiber
	Kingsport, TN	DMT	238	Fiber, Bottle Resins
	Rochester, NY	-	25	Film
	Windsor, CO	-	11	Film
Fiber Industries, Inc.	Fayetteville, NC	DMT/TPA[c]		
	Florence, SC		(36)	Fiber, Bottle Resins
	Greenville, SC	DMT/TPA[c]	680	Fiber
	Salisbury, NC	TPA[c]		
	Shelby, NC	DMT[c]		
Firestone Tire & Rubber Co.	Hopewell, VA	TPA	23	Fiber
Goodyear Tire & Rubber Co.	Scottsboro, AL	DMT/TPA	11	Fiber, Bottle Resins
	Point Pleasant, WV	-	122	Bottle Resins
ICI Americas Inc.	Hopewell, VA	-	36	Film
Minnesota Mining & Manufacturing Co.	Decatur, AL	-	25	Film
	Greenville, SC	-	10	Film
Monsanto Co.	Decatur, AL	DMT/TPA	91	Fiber
Ronm and Haas Co.	Fayetteville, NC	-	91	Bottle Resins

[a]Source: SRI International, 1982 Directory of Chemical Producers, United States. Does not include manufacturers of unsaturated resins or processors using resins as a raw material (generally to produce fibers). Saturated resins listing is also not included as it is comprised of manufacturers of polyesters other than PET.

[b]DMT - Dimethyl terephthalate process.
TPA - Terephthalic acid process.

[c]Industry correspondence.

numerous sectors of the economy. Forms include shapes for structural housings or parts, films, sheets, surface coatings, adhesive liquids, foams, fibers, and filaments. Many types of manufacturing processes are used to shape resin into these forms. The various kinds of shaping techniques used include blow molding, tubular film blowing, calendaring, injection molding, rotational molding, casting, coating, extrusion, foaming, and elongation to orient fibers. These shaping operations are noted, but not elaborated on in this document. (This document discusses the manufacturing of the above selected polymers but does not include the fabrication of polymer products.)

These products are used in every sector of the economy with particularly large applications in the construction, transportation, clothing, consumer goods, and electrical industries. Generally, end-use functions include structural components in equipment or appliances, insulation, film for packaging wrap, and fiber for lines or clothing.

Each major polymer or resin product has its own properties, forms, and end-use sectors. The important end-uses of each polymer chosen for NSPS development are summarized below.

3.1.1.1 <u>End-Uses of Polypropylene</u>. Polypropylenes, which are made by many different processes, are lightweight, water- and chemical-resistant plastics, somewhat rigid, but easy to process. They are thermoplastic and belong to the olefins family. Polypropylene products can be formed in many ways, including molding, extrusion, rotational molding, powder coating, thermoforming, foam molding, and fiber orientation.

Molded applications include bottles for syrups and foods, caps, auto parts, appliance parts, toys, housewares, and furniture components. Polypropylene fibers and filaments are used in carpets, rugs, carpet backing, woven bags, and cordage. Film uses include packaging for cigarettes, records, toys, and housewares. Extrusions include pipes, profiles, wire and cable coatings, and corrugated packing sheets.[2]

Products formed by injection molding consume about 41 percent of the polypropylene produced domestically. The second most utilized form, fibers and filaments, accounts for 31 percent of the total production. Other forms account for the remaining 28 percent.[3] The major sectors using polypropylenes are consumer/institutional (19 percent), furniture/furnishings (18 percent), packaging (16 percent),

transportation (12 percent), and electrical/electronics (7 percent). Other uses account for the remaining 28 percent.[3]

3.1.1.2 **End-Uses of Polyethylenes: Low Density and High Density.** Polyethylenes are the largest volume plastics produced, both domestically and internationally. These thermoplastic polymers are valued for their structural strength, water and chemical resistance, and easy processing characteristics.

There is nearly an infinite variety of polyethylenes that differ in melting point, clarity, and density. They are generally divided into two broad categories, low and high density, both of which are flexible, although high density polyethylene (HDPE) is more rigid. Within the last few years, a new class of LDPE has appeared - linear low density polyethylene (LLDPE). LLDPE combines the linear molecular structure of HDPE with the physical and optical properties of conventional LDPE, and its overall properties are superior to those of conventional LDPE.[4] Polyethylenes are often extruded into film, sheets, pipe, or profiles, injection molded, blow molded, rotationally molded, foamed, or formed in other ways.[5]

3.1.1.2.1 **Low density polyethylene (LDPE).** Conventional LDPE is used primarily in packaging. Specific applications include packaging film and wrap, trash bags, garment bags, and molded forms (toys, housewares, containers, and others).[6] End-uses are found in many segments of the economy, including the packaging industry (62 percent), consumer/institutional industries (11 percent), electrical/electronics industries (7 percent), and other sectors (21 percent).[7]

Like conventional LDPE, LLDPE is suitable for many end-uses. Specific applications may include housewares, lids, closures, blow molded parts (such as toys, bottles, and drums), wire and cable insulation, extruded pipe and tubing, and industrial and consumer films such as food packaging, trash bags, and garment bags.[8]

3.1.1.2.2 **High density polyethylene (HDPE).** The primary application for HDPE is the manufacture of blow molded bottles for bleaches, liquid detergents, milk, and other fluids. Other blow molded forms for which HDPE is used include automotive gas tanks, drums, and carboys. HDPE is

also used for injection molded forms including material handling pallets, stadium seats, trash cans, and auto parts. The film is used in shopping bags.[6] HDPE is of special value where high impact resistance is required.

Products formed by blow molding represent 40 percent of the total domestic HDPE production. Another 22 percent is injection molded, while 6 percent is attributed to film and sheet applications. Other uses account for 32 percent.[9] End-use sectors for HDPE include packaging (45 percent), consumer/institutional (11 percent), building and construction (9 percent), and other sectors (35 percent).[9]

3.1.1.3 **End-Uses of Polystyrene.** Polystyrene plastics are durable, provide good electrical insulation, and are easy to process. This thermoplastic is used in molded forms, extrusions, liquid solutions, adhesives, coatings, and foams.[10]

Molded uses include toys, auto parts, housewares, kitchen items, appliances, wall tiles, refrigerated food containers, radio and television housings, small appliance housings, furniture, packages, and building components such as shutters. Extruded sheets also are used in packaging, appliances, boats, luggage, and disposable plates. Foamed styrene is a good insulator and is used in construction, packaging, boats, housewares, toys, and hot/cold insulated drink cups.[10]

Fifty percent of the domestic polystyrene production is molded into its consumer form. An additional 33 percent of the domestic production is formed by extrusion, while other forming operations are used for the remaining 17 percent of polystyrene produced.[11] Segments of the economy using products from the polystyrene industry include the packaging industry (35 percent), the consumer/institutional industries (22 percent), and the building/construction and electrical/electronics industries (10 percent each). End-uses in all other sectors account for the remaining 23 percent.[11]

3.1.1.4 **End-Uses of Polyester Resin, Poly(Ethylene Terephthalate), [PET].** Poly(ethylene terephthalate) [PET] polyester resins are spun into fiber, blown into film, molded into bottles and other forms, or blended into adhesive products. Most of the PET produced in the U.S. is used for fiber production.

Polyester fibers are used widely in clothes, textiles, and woven goods. They are thermoplastic polymers which retain their original

shape, enabling clothing to have permanent press characteristics. A
specialty PET fiber with high density and high tensile strength is used
for tires, seat belts, and other industrial applications. Some specialty
PET molding and extruding materials are engineering thermoplastics with
high gloss, hard scratch resistance, and high rigidity.

3.2 POLYMERIZATION PROCESSES AND PROCESS EMISSIONS

All processes for manufacturing the five polymers and resins chosen
for NSPS development follow a general series of steps and procedures.
Figure 3-1 illustrates a simplified stepwise process for polymer production.
The manufacture of a polymer may be considered as a five step operation:

1. Raw materials storage and preparation
2. Polymerization reaction
3. Materials recovery
4. Product finishing
5. Product storage

Raw materials storage and preparation includes methods of storing
monomers and other raw materials to be used in the polymerization reaction.
Raw material drying and other purification steps may be taken. Raw
materials are then routed to the polymerization reactor.

In the reactor, raw materials and catalyst are combined with other
processing materials to produce the polymer. Reactor conditions, such
as temperature and pressure, are specific to the product being made.
After polymerization, unreacted materials are recovered and returned to
raw material storage, and the polymer is routed to product finishing.

The product finishing stage of the polymerization process may
include extruding and pelletizing, cooling and drying, introduction of
additives, shaping operations, and curing operations. The polymer is
then ready for product storage and shipping.

Pollutant emissions from any chemical process, including polymerization,
may be considered in two categories; those that can be anticipated based
on the process flow diagram and those that can be identified only by
sampling procedures, such as leakage at valves, pumps, compressors, and
flanges. The first type of emissions will be referred to as "process"
emissions and the second type as "fugitive" emissions. This NSPS would
limit VOC emissions from raw material preparation, polymer production,

```
┌──────────────────┐
│  RAW MATERIALS   │◄──────────────┐
│   STORAGE AND    │               │
│   PREPARATION    │               │
└────────┬─────────┘               │
         │                         │
         ▼                         │
┌──────────────────┐    ┌──────────────────┐
│  POLYMERIZATION  │    │  REACTANTS AND   │
│                  │───►│     REACTION     │
│     REACTION     │──┐ │      MEDIA       │
│                  │  └►│     RECYCLE      │
└────────┬─────────┘    └──────────────────┘
         │
         │              ┌──────────────────┐
         │              │   ORGANICS TO    │
         ▼          ┌──►│                  │
┌──────────────────┐│   │   BY-PRODUCTS    │
│                  ││   └──────────────────┘
│     MATERIAL     │┤
│                  ││   ┌──────────────────┐
│     RECOVERY     │└──►│    WATER TO       │
└────────┬─────────┘    │     WASTE         │
         │              │   TREATMENT       │
         ▼              └──────────────────┘
┌──────────────────┐
│     PRODUCT      │
│                  │
│    FINISHING     │
└────────┬─────────┘
         │
         ▼
┌──────────────────┐
│     PRODUCT      │
│                  │
│     STORAGE      │
└──────────────────┘
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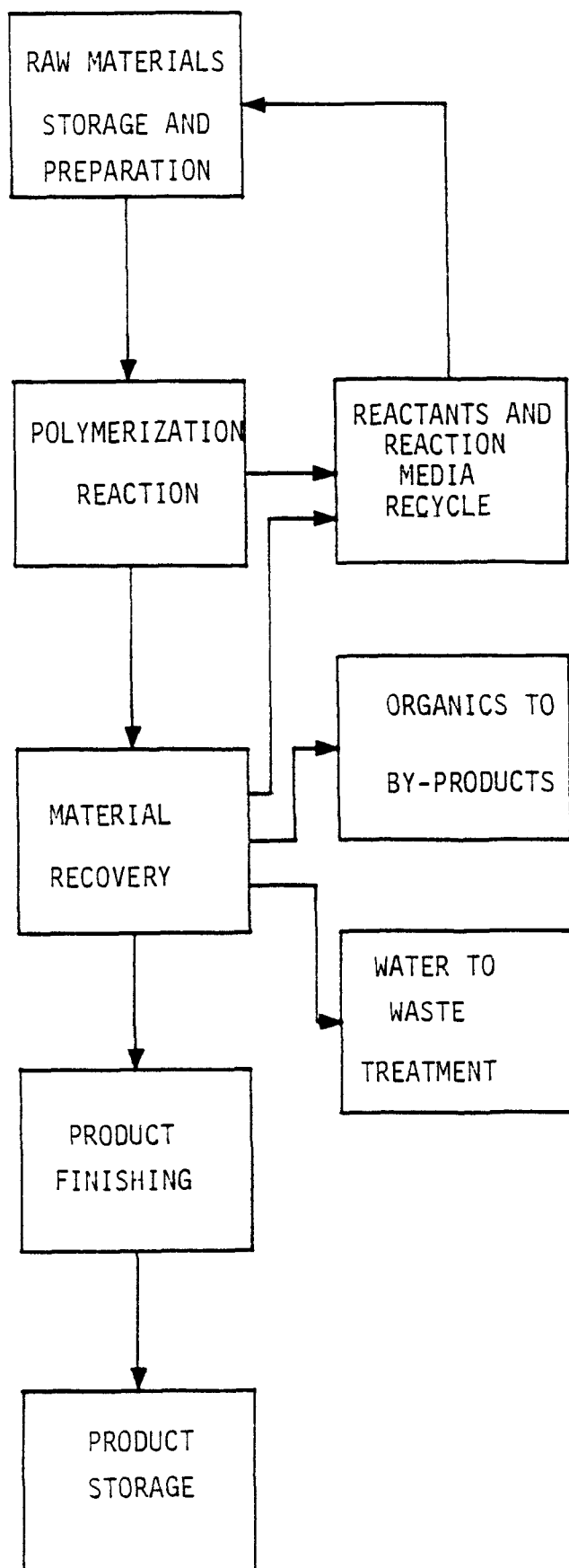
Step 1: Raw Materials Storage and Preparation may include

(a) Raw materials storage,

(b) Raw materials purification,

(c) Recovered raw materials recycle return,

(d) Raw materials drying, and

(e) Catalyst activation.

Step 2: Polymerization Reaction may be

(a) Batch or continuous·homogeneous polymerization,

(b) High or low pressure polymerization, and

(c) Liquid or gas phase polymerization.

Step 3: Material Recovery may include

(a) Product/raw materials separation,

(b) Catalyst deactivation,

(c) Product recovery and devolatilization,

(d) Reactants and reaction media recycle,

(e) Organic by-product separation and recovery.

Step 4: Product Finishing may include

(a) Extruding and pelletizing,

(b) Product cooling and drying,

(c) Additives introduction,

(d) Product shaping (e.g., fiber spinning, molding, fabricating), and

(e) Product curing, annealing or modification (e.g., fiber stretching and crimping).

Step 5: Product Storage consists of

(a) Product storage, and

(b) Product shipping.

Figure 3-1. General Polymerization Process

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material recovery, polymer extrusion and pelletizing, and product cooling and drying. The process descriptions in this section are representative of most of the polymerization processes used to manufacture the products of the source categories chosen for NSPS development.

The remainder of this section presents information on the polymers and resins listed below, including their production processes and the process emissions associated with each process.

1. Polypropylene - continuous, liquid phase slurry process.
2. Polypropylene - gas phase process.
3. LDPE - high-pressure, liquid phase process.
4. LDPE - low-pressure, gas phase process.
5. HDPE - low-pressure, liquid phase slurry process.
6. HDPE - low-pressure, liquid phase solution process.
7. HDPE - low-pressure, gas phase process.
8. Polystyrene - batch process.
9. Polystyrene - continuous process.
10. PET - Dimethyl terephthalate (DMT) process.
11. PET - Terephthalic acid (TPA) process.

Fugitive emissions, which originate from equipment components common to all of the processes, are treated in the same manner for the entire industry and are discussed separately in Section 3.3.

3.2.1 Polypropylene

Polypropylene is a high molecular weight, crystalline polymer of propylene. With a density of 0.902 to 0.904 g/cm^3, it is one of the lightest of the commercial thermoplastics. The general formula for polypropylene is:

$$\cdots - CH_2 - \underset{\underset{CH_3}{|}}{CH} - CH_2 - \underset{\underset{CH_3}{|}}{CH} - CH_2 - \underset{\underset{CH_3}{|}}{CH} - CH_2 - \cdots$$
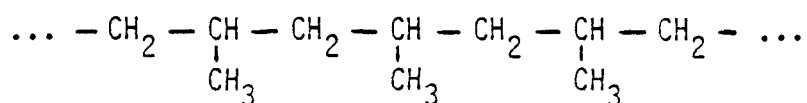
Polypropylene can be stereospecific, which means that each repeating methyl group of the polymer chain can be attached to its neighboring groups in two different geometrical arrangements. Depending on the geometrical arrangement of the methyl groups, the polymer exists in the following three forms: (1) isotactic - with all methyl groups aligned

on the same side of the chain as shown above, (2) syndiotactic - with
methyl groups alternating, and (3) atactic - all other forms in which
the methyl groups are randomly aligned on either side of the chain.
Both isotactic and syndiotactic forms, because of their regular structure,
are highly crystalline, whereas the atactic form has little crystallinity.
Only the isotactic polypropylene is of commercial interest. Atactic
resin, an undesirable byproduct of all polypropylene processes, represents
about 7 percent of the total product.[12]

Polypropylene is produced by either a liquid phase or a gas phase
process. Two basic types of liquid phase process are employed - slurry
and solution. The slurry process is the predominant liquid phase process,[13]
and can be a batch or continuous process. Batch polymerization is
particularly applicable when low volume specialty resins are to be
produced. The continuous, liquid phase slurry process and the gas phase
process are described in this section.

3.2.1.1 Polypropylene Continuous, Liquid Phase Slurry Process.
Polypropylene resins are produced through coordination polymerization
using a heterogeneous Ziegler-Natta type catalyst system, which is
typically a combination of titanium chlorides and aluminum alkyls.[13] In
conventional liquid phase processes, the catalyst provides a relatively
low polymer yield on the order of 500 to 1,000 units/unit of catalyst.[14]
In addition, a significant amount of catalyst or its residue remains in
the reaction product and must be removed. More recently, some slurry
processes have used recently developed high yield catalysts with improved
activity. These catalysts are known to provide a relatively high polymer
yield on the order of 5,000 to 7,000 units/unit of catalyst.[14] In these
processes, the catalyst is present in such small quantities that it can
remain in the product, eliminating the need for the additional process
equipment previously required for catalyst removal and recovery. The
elimination of a process operation results in lower VOC emissions.

3.2.1.1.1 Process description. Both conventional and high yield
catalyst continuous slurry processes are represented by Figure 3-2.
(The identification numbers used for process equipment in this section
and the identification letters used for VOC emission streams in the
following section refer to this figure.) Liquid phase slurry processes

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Figure 3-2. Simplified Process Block Diagram for the Polypropylene Continuous, Liquid Phase Slurry Process
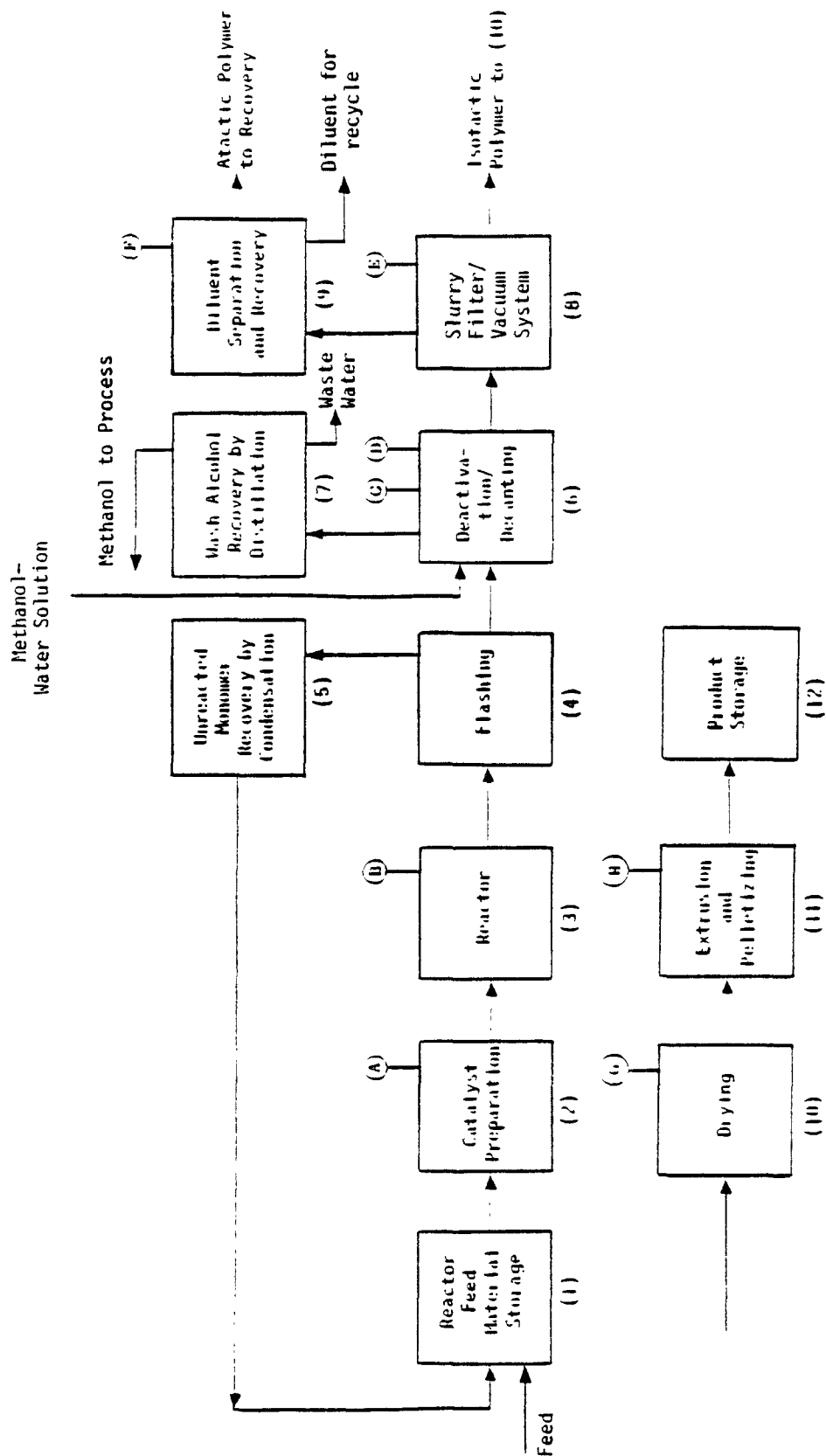
Note. Process Steps 6 and 7 are unnecessary in the high yield catalyst process.

may use either a different organic or simply liquid propylene as a diluent in which the polymer forms a slurry. The only difference between the two processes is that process steps 6 and 7 are unnecessary in the high yield process.

Reactor feed material (1) consists mainly of liquid propylene, the monomer, comonomer ethylene (if a copolymer product is desired), process diluent (which acts as a heat transfer agent), a polymer suspension medium, and a heterogeneous Ziegler-Natta type catalyst. Hexane is often used as a process diluent, although some processes use mixtures of other aliphatic hydrocarbons. The catalyst (2) is sometimes manufactured on-site. The catalyst solution is prepared by mixing the catalyst with the process diluent. Propylene is charged to the polymerization reactor while the catalyst solution and process diluent are metered in separately. Hydrogen is also introduced into the reactor for molecular weight control. Spent diluent from the catalyst preparation operation is sent to the diluent recovery section for reuse.

Polymerization is carried out in either of two types of reactors (3), a continuously stirred, jacketed vessel or a loop reactor. Most conventional slurry processes employ jacketed, continuous, stirred-tank reactors. The pipe or loop reactor is more prevalent in high yield catalyst plants. Operating pressures of 2,070 to 2,760 kPa (300 to 400 psig) are common, but they can be as high as 4,140 kPa (600 psig) when higher operating temperatures are used.[15] Reaction is carried out at temperatures of about 60°C (140°F) for approximately 8 hours.[16] A portion of the reaction effluent, which consists of polymer, monomer, and diluent, is continuously drawn from the reactor to a flash tank (4) in which the unreacted propylene and propane (an impurity in the monomer) are vaporized, and subsequently condensed by compression and cooling (5).

If the catalyst residues must be removed from the product polymer, the residual slurry from the flash tank is fed to the deactivation/decanting section (6) for washing with a methanol-water solution. Some processes use isopropyl alcohol instead of methanol to deactivate the catalyst. The washing with alcohol decomposes the catalyst, dissolves the residues, and results in two phases - a lighter diluent/crude product phase and a heavier methanol-water phase. The crude methanol from this latter phase is refined in a distillation column (7) and leaves the column in the

overhead for recycle to the process. The column bottoms containing catalyst metals are sent to the plant wastewater treatment facility.

The crude product slurry, containing isotactic polymer and atactic polymer-diluent (hexane) solution, is decanted from the methanol-water phase and fed to a slurry vacuum/filter system (8) where isotactic polymer solids are separated from the atactic polymer which is dissolved in the diluent by vacuum filtration. (Alternatively, a centrifuge may be used in place of the slurry vacuum/filter system.) The atactic-diluent solution is then introduced into a diluent purification unit (9) containing a stripping column in which the diluent is evaporated, condensed, and purified, after which it is dried for recycle. The atactic solids may then be recovered or burned in an incinerator. Liquid and gaseous waste streams from the diluent separation and purification unit may be burned in the same incinerator with the atactic waste.

The isotactic product from the slurry filter goes through a product dryer (10), extruder and pelletizer (11), and then is sent to storage (12). The type of polymer dryer used varies with the facility, but the fluidized bed dryer with a hot nitrogen or air purge is the most common.

Except for the high yield catalyst process, the variations in the various processes are minor and have little effect on VOC emissions. The high yield slurry process, however, does not require catalyst removal. The absence of deactivation/decanting and alcohol recovery processes eliminates several major VOC emission sources. The units that would not be required by the high yield process are identified in Figure 3-2.

3.2.1.1.2 <u>Emissions from the polypropylene continuous, liquid phase slurry process</u>. The characteristics of vent streams from this process are shown in Table 3-6. Each is identified in Figure 3-2. The total process VOC emission rate for a conventional slurry process is almost 37 kg VOC/Mg product. The high yield process requires neither a decanter nor a neutralizer. Therefore, the high yield process would not have emissions from these sources and the emission rate would be about 22.8 kg VOC/Mg product. The emission streams are continuous, or nearly so, and consist mainly of propylene, ethylene, propane, and a small amount of the diluent used by the process, usually hexane. The temperatures of the vent streams vary from ambient to 104°C (220°F), and the pressure is about atmospheric.

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Table 3-6. CHARACTERISTICS OF VENT STREAMS FROM THE POLYPROPYLENE CONTINUOUS
LIQUID PHASE SLURRY PROCESS[a]

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition[d,e]
RMP	A	Catalyst preparation	Continuous	0.07	29	0	$C_{10}HC$, IPA
PR	B	Reactor vents	Continuous	4.07	54	0	C_3HC, $C_{10}HC$
MR	C	Decanter vents	Continuous	11.49	38	0	C_3HC, $C_{10}HC$, IPA
MR	D	Neutralizer vents	Continuous	1.82	32-71	0	C_3HC, C_4HC, $C_{10}HC$, IPA
MR	E	Slurry vacuum/filter system vent	Continuous	7.93	32	0	$C_{10}HC$, IPA
MR	F	Diluent separation and recovery	Continuous	8.72	104	0	$C_{10}HC$, IPA
PF	G	Dryer vents	Continuous	0 to 0.6[f]	85-104	0	Air and small amount of VOC
PF	H	Extrusion/pelletizing vent	Continuous	2.0			2½% HC
		Total Emission Rate		36.7[g]			

[a]Source of information: Industry correspondence.

[b]RMP = raw material preparation; PR = polymerization reaction; MR = material recovery; PF = product finishing.

[c]See Figure 3.2 for stream identification.

[d]Streams are diluted in 10-30 percent nitrogen.

[e]C_3HC - Propylene or any other hydrocarbon compound with three carbon atoms such as propane.
   C_4HC - Butylene or any other hydrocarbon compound with four carbon atoms.
   $C_{10}HC$ - A mixture of aliphatic hydrocarbons with 10-12 carbon atoms.
   IPA - Isopropyl alcohol.

[f]The range reflects the fluidized bed dryer emission at different operating pressures. Other types of dryers may have even higher emissions.

[g]Including upper end to the range of emissions from dryer vents.

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The vent from the catalyst preparation section, Stream A, releases diluent continuously. The reactor vent, Stream B, is continuous and releases monomer and diluent.

The combined total of the decanter and neutralizer vents, Streams C and D, is usually the largest VOC emission source in the process. The constituents are methanol or isopropyl alcohol (if used), C_3 hydrocarbons, and diluent. Newer, high yield catalyst processes do not require these process steps. The absence of these vents significantly reduces the overall VOC emission rates for the high yield process.

The slurry vacuum/filter system vent, Stream E, is one of the largest VOC emission streams, venting process diluent and alcohol that has remained in the polymer. It is common to both the conventional and high yield slurry processes and releases at atmospheric pressure.

The vacuum jet exhaust, Stream F, from the by-product and diluent recovery section, can be the second largest VOC emission stream in the entire process. The diluent recovery section, which consists of an evaporator, an extractor, and distillation units, is common to both the conventional and high yield processes. It emits process diluents and traces of alcohol.

The vents from the product drying section, Stream G, emit diluent, methanol, and propane diluted in large quantities of air at a relatively high temperature, 104°C (220°F), and near atmospheric pressure.

Emissions are also released from the extrusion/pelletizing section vent, Stream H. Significant quantities of hydrocarbon still remain in the polypropylene powder as it exits the dryer and enters the extruder feed chute. At this point, the powder is in equilibrium with a vapor that can contain up to 25 percent hydrocarbon by weight. As a result of heating and compression in the extruder, there is some VOC loss through the extruder/pelletizer section and further losses from the powder/pellet transfer system downstream from the product dryer as the transfer medium acts as a stripping gas.

The stream properties and VOC concentrations of these process vents can vary depending on process conditions. The variation generally depends on the grade or type of product being manufactured, process variables such as temperature, pressure, catalyst concentration, or catalyst activity, and the amount of hydrogen used for molecular weight control.

3.2.1.2 <u>Polypropylene Gas Phase Process</u>. The gas phase process
for producing polypropylene is relatively new and is currently being
used in only one plant. Processing steps are less complicated than in
the traditional continuous liquid phase slurry process. The gas phase
process is similar to the high yield liquid phase slurry process in that
there is no need for catalyst removal. The product processing techniques,
however, are different due to the gas phase reaction. The uncontrolled
continuous VOC emission rate from the gas phase and the high yield
liquid phase slurry processes are comparable. These two newer processes
are expected to predominate in future polypropylene capacity because of
their simplicity, lower capital cost, and high yield.

3.2.1.2.1 <u>Process description</u>. Figure 3-3 shows a simplified flow
diagram of the gas phase process. (The identification numbers used for
process equipment in this section and the identification letters used
for VOC emission streams in the following section refer to this figure.)
In the gas phase process, catalysts and hexane are premixed in the
catalyst mix drum (1) from which they are fed to the gas phase reactor (2).
Propylene monomer (3) is introduced into the gas phase reactor by a
separate line.

The product, which is in a powder form containing contaminants of
propylene and finely divided catalyst, is transferred from the gas phase
reactor to a fluidized bed catalyst deactivator (4). The gaseous components
containing unreacted propylene from the reactor are recovered and purified
in a material recovery operation (5) and then recycled back to the
reactor. The recycle system contains bag filters, an entrained gas
scrubber, a recycle scrubber, and a recycle gas compressor. The gas
stream from the fluidized bed catalyst deactivator passes through a bag
filter system (6) to recover product. A nitrogen purge gas stream from
the bag filters is then fed to a scrubber (7). The purge gas from the
HCl scrubber is sent to a flare for combustion.

Polymer product from the fluidized bed deactivator and the bag
filter system (6) is sent through the product finishing steps of extrusion,
pellet blending, and storage (8).

3.2.1.2.2 <u>Emissions from the polypropylene gas phase process</u>. Two
VOC offgas streams from the process, the scrubber vent which is a con-
tinuous stream (A), and the reactor blowdown vent which is an intermittent
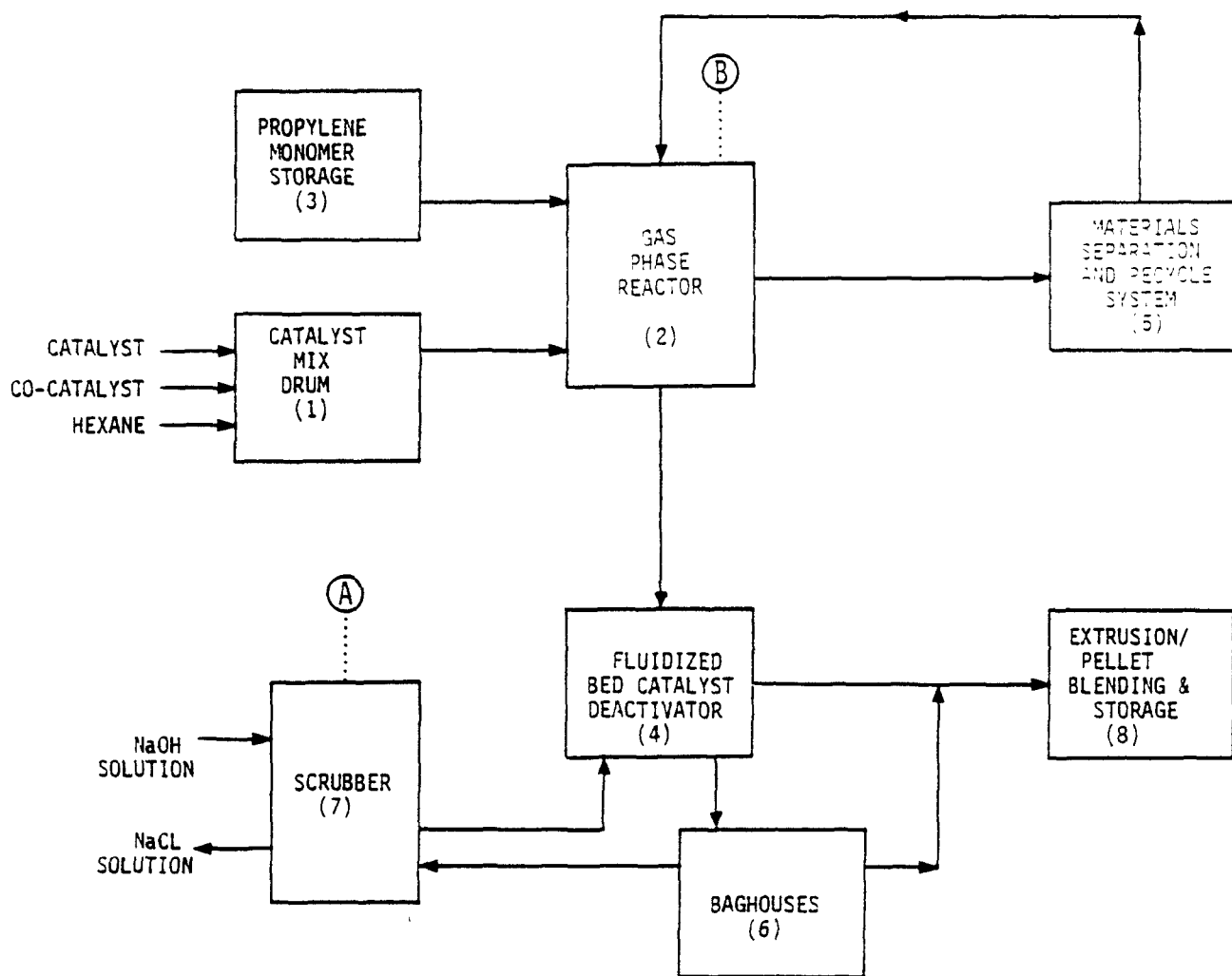
Figure 3-3. Simplified Process Block Diagram for the Polypropylene
Gas Phase Process

stream (B), are currently controlled by a flare system which handles both streams, according to one industry source. The uncontrolled VOC emission rate for the entire process, based on the average expected flowrate from these two streams, is 36.5 kg VOC/Mg product. The VOC emissions consist primarily of propylene, propane, and hexane. Table 3-7 summarizes the emission characteristics of this process.

3.2.2 Low Density Polyethylene (LDPE)

Ethylene polymerizes in the presence of a suitable catalyst in the following manner.

$$n \left(\begin{array}{cc} H & H \\ | & | \\ C & = C \\ | & | \\ H & H \end{array} \right) \xrightarrow{\text{Catalyst}} \cdots - \overset{\displaystyle H}{\underset{\displaystyle H}{C}} - \overset{\displaystyle H}{\underset{\displaystyle H}{C}} - (C_2H_4)_{n-2} - \overset{\displaystyle H}{\underset{\displaystyle H}{C}} - \overset{\displaystyle H}{\underset{\displaystyle H}{C}} - \cdots$$
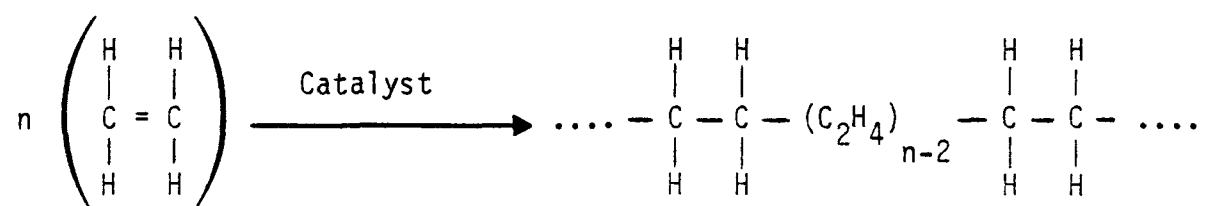
Conventional LDPE resins have a high degree of branching with densities that range from 0.910 to 0.935 g/cm^3 and are produced in either an autoclave or tubular reactor.

LDPE resins may also be produced commercially at low pressures in both gas phase fluidized bed reactors and liquid phase solution process reactors. The low pressure process yields a much more linear structure. Linear LDPE can be made in the gas phase with melt indices and densities over the full commercial range.

Both processes are described below.

3.2.2.1 LDPE High-Pressure, Liquid Phase Process. The high-pressure process is currently used more widely than the low-pressure process. However, due to the more favorable economics of the low-pressure process, few, if any, high-pressure plants are likely to be built in the future.[17] The high-pressure process is a free radical process utilizing pressures of several thousand atmospheres to polymerize ethylene and copolymers of ethylene, and nonolefinic comonomers such as vinyl acetate.[18] High-pressure processes, however, are not capable of polymerizing propylene or higher olefins to their corresponding polyolefin.[19] Free radical catalysts, or initiators, are usually used and are predominantly oxygen and peroxides. The product grade can be changed by varying the throughput rate of the polymer in the reactor, the reactor configuration,[20] reaction temperature, or type and concentration of the catalyst.[21]

Table 3-7. CHARACTERISTICS OF VENT STREAMS FROM THE POLYPROPYLENE
GAS PHASE PROCESS[a]

Process[b] Section	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition
MR	A	Scrubber vent	Continuous	25	38	75	Propylene, Propane, Hexane
PR	B	Reactor blow-down vent	Intermittent	11.5	38	75	Hydrogen, Propylene, Propane, Hexane
		Total Emission Rate		36.5			

[a]Source of information: Industry correspondence.

[b]MR = material recovery; PR = polymerization reaction.

[c]See Figure 3-3 for stream identification.

3.2.2.1.1 Process description. Figure 3-4 is a schematic representation of the liquid phase, high-pressure process. In this process, ethylene is dried (1) and fed to the suction of a first stage compression system (2) which raises its pressure to about 28 MPa (4,000 psig). The exact pressure depends on the pressure employed in the product separation phase of the process. The primary compressor discharge is combined with recycle ethylene and introduced to the suction of a second stage compression system (3) which raises the pressure of the ethylene stream to 275 to 345 MPa (40,000 to 50,000 psig). A comonomer is also added if a copolymer product is desired. The compressor discharge effluents are fed to the reactor (4). An initiator solution of organic peroxide in isopropyl alcohol is injected directly into the reactor to initiate polymerization. The amount of peroxide catalysts required varies from about 10 to 100 ppm.[22]

Either tubular or autoclave reactors are used. Temperatures in the reactors may vary from about 150 to 300°C (300 to 570°F), although ranges from 200 to 250°C (390 to 480°F) are more common. Pressures within the reactors vary from about 100 to 345 MPa (15,000 to 50,000 psig). For many types of polyethylene, the common pressure range is 130 to 240 MPa (20,000 to 35,000 psig).[22] The total residence time of the reactants varies from 45 to 60 seconds in a tube reactor and from 25 to 40 seconds in an autoclave reactor.[23] Polyethylene product from the reactor is continuously throttled into the high-pressure separator (5) which operates at a pressure of 6 to 28 MPa (900 to 4,000 psig). Most of the unreacted ethylene is flashed and withdrawn overhead from this separator. It is cooled, separated from low molecular weight polymers (wax) in the high-pressure wax knock-out drum (6), and recycled to the second stage compressor (3) suction. The separator bottoms pass through a throttle valve to the low-pressure separator (7) which operates at 35 to 70 kPa (5 to 10 psig). The remaining ethylene is flashed from the product and withdrawn overhead from the separator. The withdrawn gas is also cooled to condense waxes and routed through the low-pressure wax knock-out drum (8) to the olefins recovery unit (9). From this unit the recovered ethylene is returned to the monomer storage area. The degassed homopolymer or copolymer product is extracted as the low-pressure separator bottoms which still contain residual ethylene monomer. These product
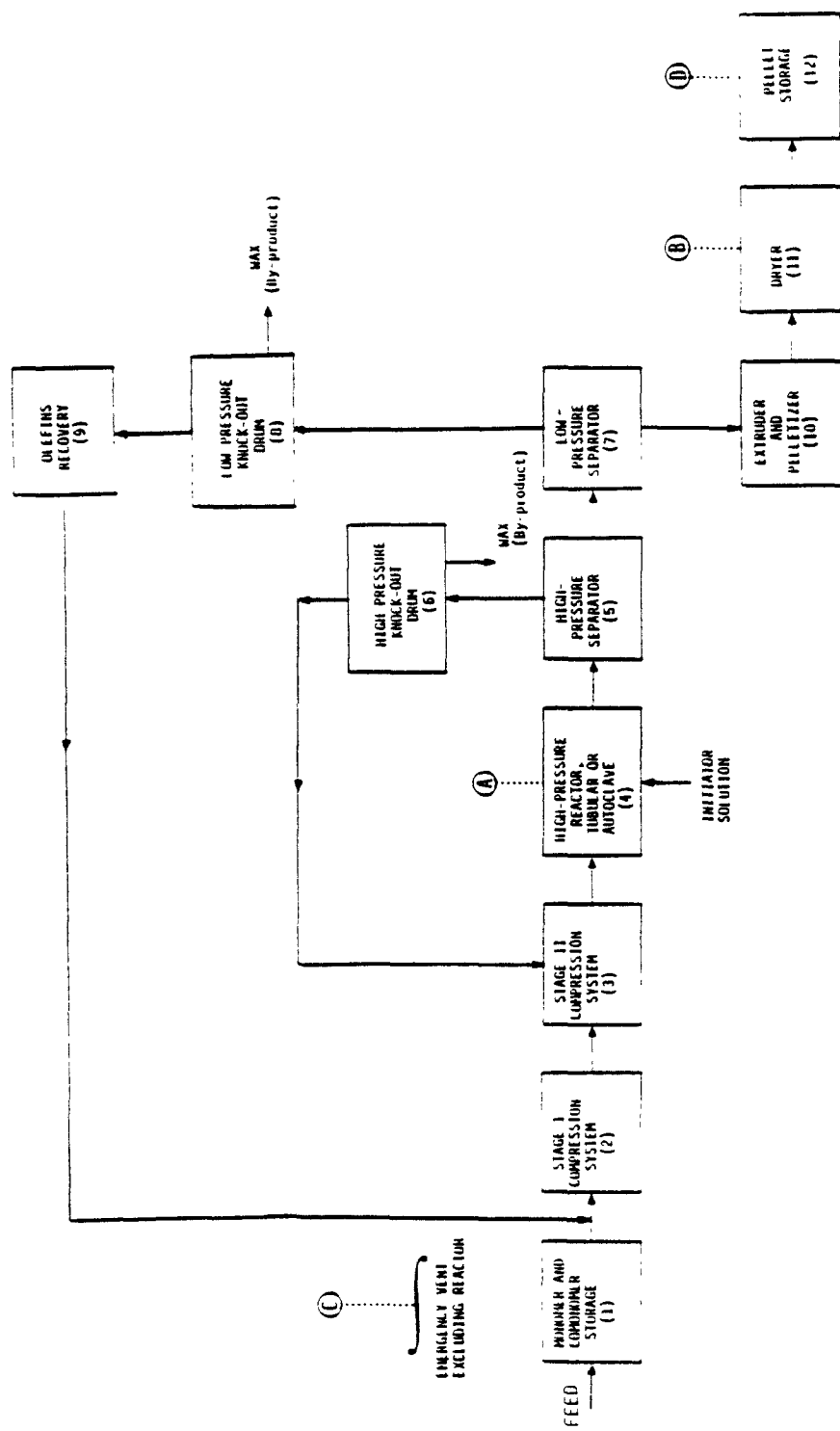
Figure 3-4. Simplified Process Block Diagram for the Low Density
Polyethylene High-Pressure, Liquid Phase Process

resins are routed to the finishing line where antioxidants are added to resin which is then melted and pelletized in an extruder-pelletizer (10). The hot pellets are water cooled and conveyed to a hot air dryer (11). The residual ethylene within the polymer is very low after extrusion and drying. Dried pellets are conveyed to storage (12).

3.2.2.1.2 <u>Emissions from the LDPE high-pressure, liquid phase process.</u> The offgas stream characteristics for the liquid phase process are shown in Table 3-8. The uncontrolled VOC emission rate for the entire process is 2.9 kg VOC/Mg product. In general, the composition of the streams varies based on the type and amount of product made, separation pressures involved, and ambient temperature. The emission streams consist mainly of ethylene. Other compounds, such as ethane, methane, propane, propylene, and isopropanol, are also present. The temperature of the streams varies from 50 to 260°C (106 to 500°F), and the pressure varies from atmospheric to 140 kPa (20 psig).

The emergency reactor vent, Stream A, is intermittent. It is activated either manually or automatically as an emergency blowdown during a process upset. The total mass and instantaneous flowrate can vary dramatically depending on the nature of the upset. This, the major VOC emission source in the process, generally emits directly to the atmosphere because it is an extremely large quantity emitted at very high pressure at infrequent intervals. It consists of monomer and polymers, with trace amounts of catalyst, in a three-phase flow. In order for this stream to be safely destroyed, such as in a flare system, the solids would have to be separated from the bulk flow before it reached the combustion zone.

The dryer and storage bin vents, Streams B and D, are continuous and consist of mostly air with small amounts of ethylene. These streams are usually emitted to the atmosphere.

The emergency vents, Stream C, are intermittent and consist of VOC gases from a number of sources from the plant other than the reactors. The wax blowdown system can be a source of significant ethylene losses. The emission rate is highly dependent upon the design of the wax blowdown and discharge system.

A considerable number of LDPE plants are located near or integrated with a plant which manufactures ethylene. As a result, the unreacted

Table 3-8. CHARACTERISTICS OF VENT STREAMS FROM THE LOW DENSITY POLYETHYLENE HIGH-PRESSURE, LIQUID PHASE PROCESS[a]

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition, Wt.%
PR	A	Emergency reactor vent	Intermittent	1.7	Varies	0-20	95.3 Ethylene 1.0 Ethane 1.5 Propylene 2.2 Isopropanol
PF	B	Dryer vent	Continuous	0.2	54	0	0.2 Ethylene 99.8 Air
OTHER	C	Emergency vent excluding reactor	Intermittent	0.2	Varies	0-20	95.3 Ethylene 1.0 Ethane 1.5 Propylene 2.2 Isopropanol
PS	D	Storage bin vent	Continuous	0.8	54	0	0.2 Ethylene 99.8 Air
		Total Emission Rate		2.9			

[a]Source of information: Industry correspondence.

[b]PR = polymerization reaction; PF = product finishing, PS = product storage; other = vents from all process sections except the emergency reactor vent in the polymerization reaction section.

[c]See Figure 3-4 for stream identification.

3-27

ethylene from the polymer plant is purified by recycling through the ethylene manufacturing unit. Overall process VOC emissions from these units can be expected to be 8 to 10 percent lower than those where the LDPE plant and the ethylene plant are not located at the same site.

3.2.2.2 LDPE Low-Pressure, Gas Phase Process.[24] Most, and probably all, new low density polyethylene plants will use the relatively new gas phase process because of its much more favorable economics. The most significant cost aspect of the new process is the drastic reduction in reaction pressure. The low-pressure, gas phase process uses reaction pressures of only 0.69 MPa to 2.1 MPa (100 to 300 psig) in comparison to pressures as high as 345 MPa (50,000 psig) in the high-pressure process. Changes in product grade are accomplished primarily by changing the catalyst composition; reactor operating conditions remain the same.[25]

The new process has reduced capital investment requirements by 50 percent, energy consumption by 75 percent, and the operating cost of making low density polyethylene by 50 percent. One significant technical aspect of the gas phase process is its capability of producing high or low density polyethylenes in the same process equipment. The advantages of the new process are so overwhelming that it is likely to be the preeminent process for future expansion of the polyethylene industry.

3.2.2.2.1 Process description. The process flow diagram presented in Figure 3-5 is based on Union Carbide's Unipol process. Ethylene is polymerized in the presence of a chain transfer agent and an alpha-olefin comonomer to produce polymers having desired melt indices, densities, and molecular weight distributions. The alpha-olefin comonomer is usually 1-butene, or the more costly and higher-boiling 1-hexene or 1-pentene. Before entering the reactor, the monomers (depending on their sources) are subjected to varying degrees of pretreatment (1) to remove impurities that could poison the catalyst. Monomer is then fed continuously into the fluidized bed reactor (2). Catalyst is added separately (3).

The process uses a fluidized bed reactor technology and a new family of catalysts that trigger the desired chemical reaction at pressures between 690 and 2,070 kPa (100 and 300 psig) and temperatures of about 100°C (212°F). The fluid bed in the reactor is granular polyethylene, the product of the polymerization reaction. Circulated up through the
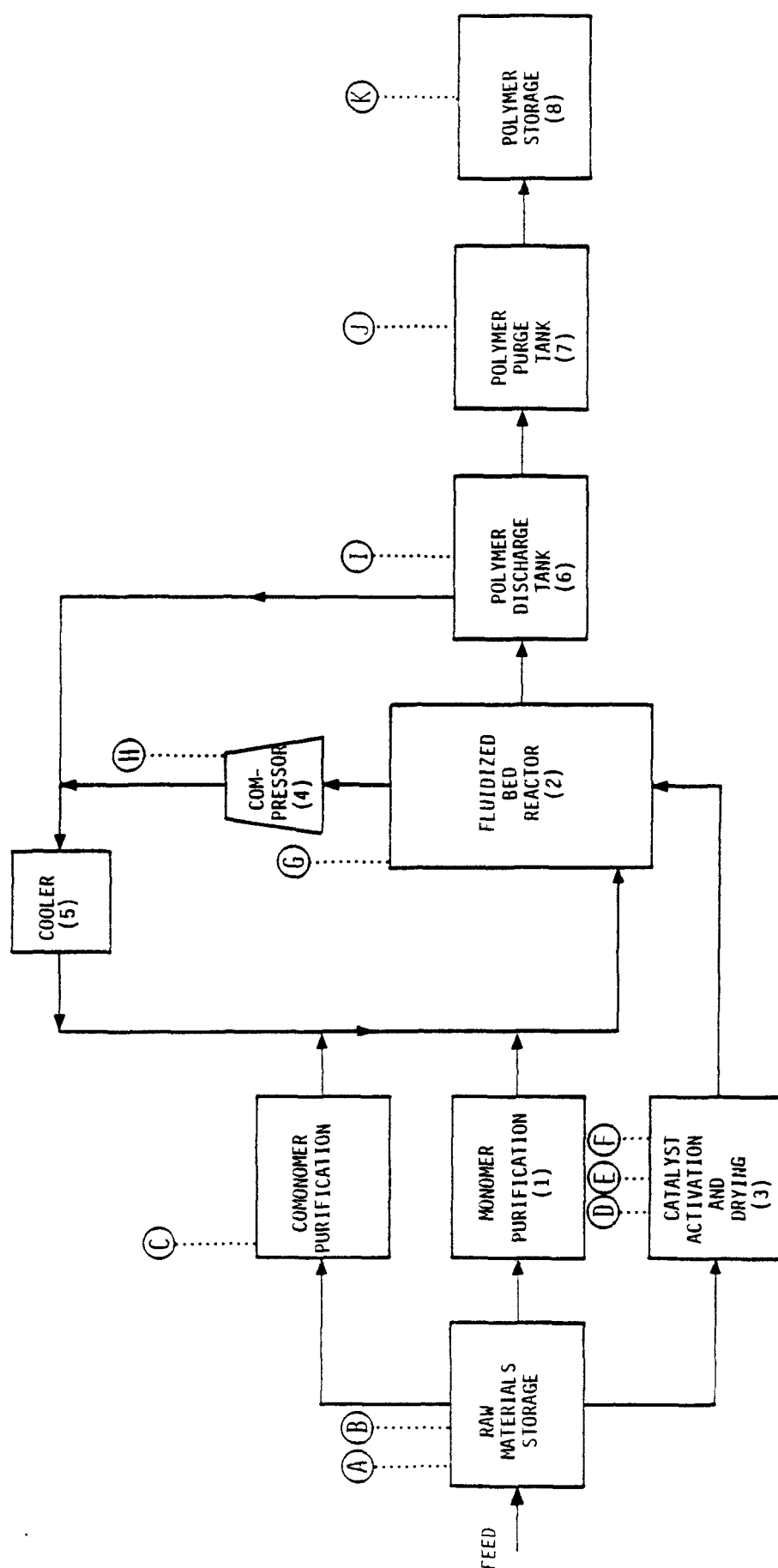
Figure 3-5. Simplified Process Block Diagram for the Low Density and High Density Polyethylene Low-Pressure, Gas Phase Process

bed, the gas stream containing the unreacted ethylene and comonomer passes out of the reactor through an enlarged top section designed to reduce velocity, thereby disengaging most of the fine particles. It then goes to a cycle compressor (4) and through an external cooler (5) before returning to the reactor. Dry, free-flowing solid product is removed intermittently from the continuously growing fluid bed through a discharge system (6) to keep the volume of the bed approximately constant.

Although most of the unreacted monomer is recycled, some residual VOC must be purged (7) from the granular product before it can be safely conveyed in air. As an optional final step, one or more conventional additives (e.g., antiblocking, antislipping, antioxidizing, ultraviolet-light-stabilizing) may be added to the granular product before it is stored or shipped (8).

The overall combined conversion rate of ethylene and comonomer is approximately 97 percent. The average residence time of the polymer in the reactor bed is 3 to 5 hours, during which the particles grow to an average size of about 1,000 microns. Polymer density is regulated by the type and concentration of alpha-olefin comonomer, which controls the frequency of short chain branches. Molecular weight is influenced by the reaction temperature and the concentration of chain-transfer agent in the circulating gas. Molecular weight distribution is manipulated primarily by catalyst type and composition but also, to a much lesser extent, by reactor operating conditions. [21]

3.2.2.2.2 <u>Emissions from the LDPE low-pressure, gas phase process.</u> The waste gas stream characteristics for the LDPE gas phase process are shown in Table 3-9. The combined process VOC emission rate for this process is about 23 kg VOC/Mg product. One process vent, the product discharge vent, contributes almost 96 percent (22.3 kg VOC/Mg product) of the total VOC discharged annually from the model plant for this process.

Process emissions consist of VOC with two to six carbon atoms along with nitrogen or air. The temperature of the streams varies from 38 to 85°C (100 to 185°F), and the pressure varies from 0.7 to 138 kPa (0.1 to 20 psig).

Table 3-9. CHARACTERISTICS OF VENT STREAMS FROM THE LOW DENSITY POLYETHYLENE LOW-PRESSURE, GAS PHASE PROCESS[a]

Process Section[b]	Stream[c]	Name[d]	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition, Wt.%
RMP	A	Raw materials purification vent	Intermittent	0.04	-	-	0.5 VOC 99.5 N_2
RMP	B	Catalyst additive vent	Intermittent	0.01	40	1.0	35 VOC 65 N_2
RMP	C	Comonomer purification vent	Continuous	0.01	-	-	92 VOC 8 N_2
RMP	D	Catalyst dehydrator handling vent	Intermittent	0.01	827	0.1	2 VOC 98 Air
RMP	E	Vacuum pump vent	Intermittent	0.03	39	1.0	41 VOC 59 N_2
RMP	F	Recovery vessel vent	Intermittent	0.06	39	1.0	46 VOC 54 Air
PR	G	Emergency reactor blowdown vent	Intermittent	0.05	85	60	83 VOC 17 N_2
PR	H	Compressor seal vent	Continuous	0.01	38	1.0	50 VOC 50 N_2
PF	I	Product discharge vent	Continuous	22.3	85	1 – 20	35 VOC 65 N_2
PF	J	Purge tank vent	Intermittent	0.71	-	-	1-83 VOC 99-17 N_2
PS	K	Bin vent	Continuous	0.05	50	0.2	0.02 VOC 99.98 Air
		Analyzer vents[e]	Continuous	0.01	49	0.1	100 VOC
		Total Emission Rate		23.3			

[a] Source of information: Industry correspondence.

[b] RMP = raw material preparation; PR = polymerization reaction; PF = product finishing; PS = product storage.

[c] See Figure 3-5 for stream identification.

[d] All streams except D, H, K, and the analyzer vents are currently controlled by the use of safety flares.

[e] Not shown in Figure 3-5. Analyzer vents come from all areas and are one type of sampling connection system covered as a fugitive, rather than process, emission source.

3.2.3 High Density Polyethylene (HDPE)

HDPE resins are linear thermoplastic polymers of ethylene with densities higher than 0.96 g/cm^3 and copolymers of ethylene with densities as low as 0.94 g/cm^3. HDPE resins are typically produced at low pressures by either liquid phase or gas phase processes. There are two liquid phase processes, slurry and solution. In both, the solvent dissolves the ethylene monomer and comonomer and suspends the solid catalyst. The basic difference between them is that in the solution process the solvent also dissolves the polyethylene product.

The gas phase process is virtually identical to the LDPE gas phase process. All three HDPE processes are described in this section.

3.2.3.1 HDPE Low-Pressure, Liquid Phase Slurry Process. Of the two liquid phase processes, the slurry process is predominant and is the only one capable of producing the whole range of HDPE polymers.[26] The slurry or particle form process of Phillips Petroleum Company serves as the basis for this description, but it is intended to illustrate all liquid phase slurry processes.

3.2.3.1.1 Process description. As illustrated by the schematic for this process, Figure 3-6, the feed section (1) consists of the reactor feed storage and a catalyst purification and activation system. The chromium oxide catalyst is suspended in a solvent (pentane or isobutane) and is continuously fed to the reactor (2). Other slurry processes use Ziegler catalysts or molybdenum oxide catalysts. Purified ethylene monomer and a comonomer (1-butene or hexane) are fed to the reactor where suspension polymerization takes place. The reactor is usually a closed-loop pipe reactor.

Temperatures of 20 to 100°C (68 to 212°F) in the reactor have been reported, but the polymerization rates at 20°C (68°F) are probably very low. Pressures are from about 690 to 3,450 kPa (100 to 500 psig). Higher pressures are normally required at the higher temperatures to dissolve sufficient ethylene in the liquid phase. The slurry in the reactor contains 18 to 25 percent solid polyethylene. Settling legs and screw conveyors are used to withdraw concentrated slurries containing 50 to 80 percent solids.[15] Unreacted monomer and diluent (3) are separated from the product by flashing. Final stripping of the gases from the
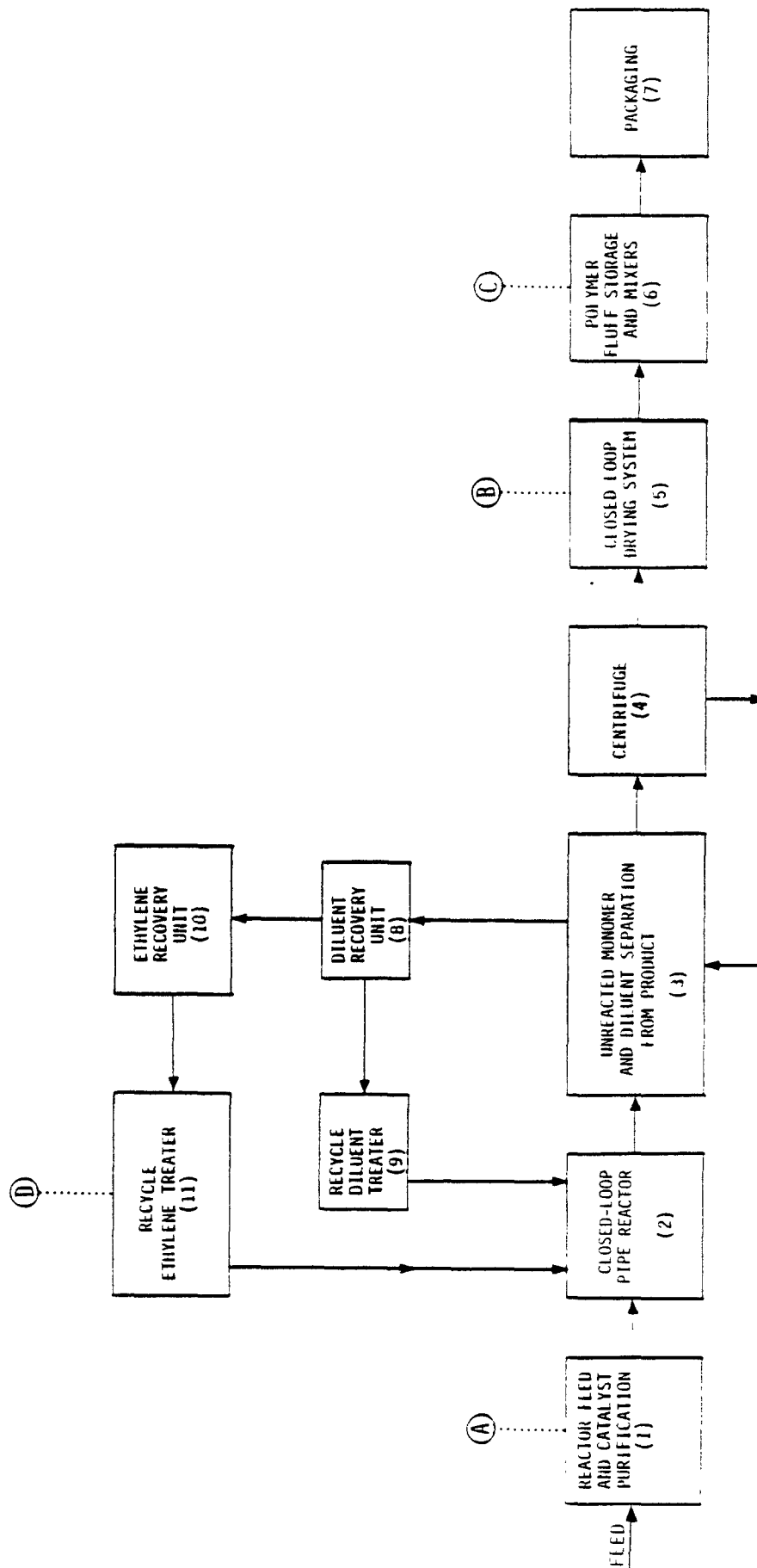
Figure 3-6. Simplified Process Block Diagram for the High Density Polyethylene
Low-Pressure, Liquid Phase Slurry Process

polymer is performed using steam. The wet polymer solids are then centrifuged (4) to remove water, and dried in a closed-loop nitrogen or fluidized air drying system (5). The resulting polymer fluff is mixed with various finishing agents (6), and packaged (7).

Vapors from the flashing vessels are sent through a diluent recovery unit (8) which condenses the diluent and recycles it through diluent treaters (9) back to the reactor. The ethylene-rich stream is then sent to the ethylene recovery unit (10) for purification and sent to recycle ethylene treaters (11) and back to the reactor.

3.2.3.1.2 Emissions from the HDPE low-pressure, liquid phase slurry process. This process has one intermittent and three continuous process emission sources. The major emission source for this process is the recycle treater vent with an emission rate of about 13 kg VOC/Mg product. The total emissions from the four streams is about 14 kg VOC/Mg product. Table 3-10 shows the composition of these streams.

The emissions from the closed-loop drying system (Stream B) are a dilute mixture of process solvent in nitrogen. The continuous mixer in which antioxidants are added to the polymer (Stream C) vents a low VOC emission stream. Some of the process solvent that is still in the polymer is emitted along with a large quantity of nitrogen, usually to the atmosphere. The recycle treater system has continuous emissions (Stream D) which are about 80 weight percent VOC. Treaters are vessels containing materials such as adsorbers, dessicants, or molecular sieves which remove water and other impurities in the recycle ethylene stream. Emissions occur when the vessels are purged prior to regeneration.

Most plants use a separate recycle treater for each individual VOC component since components are usually recovered by fractional distillation. Therefore, an HDPE plant recycling ethylene, isobutane, and butene would generally have three treaters.

The intermittent offgas stream is the feed preparation stream (Stream A) which consists mostly of ethylene. Sources for this stream include drying/dehydrating and other feed purification operations. Its emission rate is 0.2 kg VOC/Mg product. Emissions occur when the treating vessels are purged prior to regeneration, usually once a month.

The HDPE liquid phase slurry process described above has an ethylene recycle and closed-loop nitrogen drying system. These greatly reduce

Table 3-10. CHARACTERISTICS OF VENT STREAMS FROM THE HIGH DENSITY
POLYETHYLENE LOW-PRESSURE, LIQUID PHASE SLURRY PROCESS[a,b]

Process Section[c]	Stream[d]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Composition, Wt.%
RMP	A	Feed preparation	Intermittent[e]	0.2	21	100.0 Ethylene
PF	B	Dryer nitrogen blower	Continuous	0.2-0.4	21	0.3 Isobutane 99.7 Nitrogen
PF	C	Continuous mixer	Continuous	0.006	21	0.6 Isobutane 99.4 Nitrogen
MR	D	Recycle treaters	Continuous	12.7	21	61.0 Ethylene 18.0 Isobutane 20.0 Ethane 1.0 Hydrogen
		Total Emission Rate		13.1-13.3		

[a] Source of information: Industry correspondence.
[b] Gas stream pressure data unknown.
[c] RMP = raw material preparation; PF = product finishing; MR = material recovery.
[d] See Figure 3-6 for stream identification.
[e] One occurrence per month.

the emissions. Some plants, however, vent unreacted monomer and use simple single pass dryers. These plants have substantially higher emissions. The major VOC source for these plants is the flash tank where the unreacted monomer stream is about 50 percent VOC. This stream is often burned in a boiler because of its high heat value.

A considerable number of HDPE plants are located near or integrated with olefins distillation trains. Some of these units do not need recycle treaters since the ethylene is purified by recycling it through the olefins manufacturing unit. In these cases, overall process VOC emission from these units can be expected to be only 2 percent of the emissions from separate HDPE plants.

3.2.3.2 HDPE Low-Pressure, Liquid Phase Solution Process. The solution process differs from the slurry process primarily in that the polymerization process is carried out at a temperature higher than the solution temperature of the polymer in the selected solvent medium so that the polymer is in solution (completely dissolved) rather than in particle form suspended in the reaction mixture. The solvent medium may be cyclohexane, pentane, hexane, or heptane. Operating temperatures for the solution process may range from 100 to 200°C (212 to 392°F), depending upon the solution temperature of the polymer in the solvent.[27] Emphasis in HDPE manufacturing, however, is shifting from the solution process to the slurry and gas phase processes.

3.2.3.2.1 Process description.[28,29] As illustrated in Figure 3-7, ethylene monomer goes through preparation step (1) before being fed via a compressor (2) to a stirred reactor (3). The preparation step assures removal of acid gases and moisture. Catalyst is introduced as a slurry into the reactor from the catalyst preparation section (8). The reactor may be operated in a continuous or batch mode. Operating conditions may be around 260°C (500°F) and 7 MPa (1,000 psig). The exact operating conditions depend in part on the particular polymer being produced.

From the reactor, the reaction mixture proceeds to the separation section (4) where the catalyst, solvent mixture, and polymer are separated. If desired, the catalyst may be recovered and recycled. The solvent mixture is sent to solvent recovery (9). Vapors of unreacted monomer and solvent pass through a condenser (10). The unreacted monomer is

Figure 3-7. Simplified Process Block Diagram for the High Density Polyethylene, High-Pressure, Liquid Phase Solution Process

NOTE: Stream A, from raw material storage tanks, is not shown.

recycled to the monomer preparation area through recycle ethylene treaters (11), while the condensed solvent is returned to the flash drum. Alternatively, a reflux condenser (10a) may be placed on the reactor to recycle unreacted monomer directly back into the reactor. If an ethylene plant is located on-site, the ethylene vapors from the reactor may be sent directly to the ethylene plant, thereby eliminating the recycle ethylene treaters in the HDPE plant.

As it leaves the separator, the polymer is extruded and melt cut (5). The melt cut polymer still contains solvent that is subsequently stripped from the polymer in a steam still (6), which may be operated under pressure or vacuum. For copolymer processing, the stripping is carried out under vacuum (1.0 to 700 mm Hg) in order to lower the vaporization point of the solvent below 71 to 77°C (160 to 170°F) and prevent agglomeration and melting of the copolymer charge. The vaporized mixture of steam and reaction medium from the steam still is sent to a distillation train (9). The recovered solvent is recycled to the catalyst preparation section and to the reactor.

The steam stripped extrudate from the extruder is then dewatered, re-extruded in a semi-dry state, and dried to reduce the water content to about 1 to 2 weight percent or less. The final product may be blended with additives at this time. The extrudate is then cooled, chopped, and transferred to packaging facilities (7).

3.2.3.2.2 Emissions from the HDPE low-pressure, liquid phase solution process. This process has numerous offgas streams. The major one is the recycle treater vent which has an emission rate of about 13 kg VOC/Mg product, with other streams contributing about 19 kg VOC/Mg product. If the HDPE plant is integrated with an olefin manufacturing operation, the plant may not have recycle treaters since the ethylene can be purified in the olefin manufacturing plant. In this case, overall process VOC emissions can be expected to be about 60 percent of the emissions from those HDPE plants which do not have an adjacent ethylene plant. Table 3-11 shows the basic characteristics of the process VOC emissions from a solution process at an HDPE facility which has ethylene treaters.

3.2.3.3 HDPE Low-Pressure, Gas Phase Process. The HDPE gas phase process is similar to the LDPE gas phase process described earlier

Table 3-11. CHARACTERISTICS OF VENT STREAMS FROM THE HIGH DENSITY POLYETHYLENE LOW-PRESSURE LIQUID PHASE SOLUTION PROCESS[a]

Process Section[b]	Stream	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Composition, % by volume
RMP	A	Tank vents	continuous; cyclic	0.024	–	–
RMP	B	Treater vents	intermittent	(see stream Z)		
RMP	C	Catalyst preparation vent	generally continuous	(see stream Z)	–	–
RMP	D	Compressor vents	continuous and intermittent	(see stream Z)		–
PR	E	Reactor vents	generally intermittent	(see stream Z)	–	–
MR	F	Separation vents	some continuous and intermittent	(see stream Z)	–	–
MR	G[c]	Recycle ethylene treaters	continuous	12.7	21[c]	61.0 ethylene[d] 1.0 hydrogen 18.0 isobutal 20.0 ethane
MR	H	Recovery distillation	continuous and intermittent	(see stream Z)	–	–
PF	I	Extruder vents	continuous	0.63	105	VOC[e] – 0.05 stream air – 99.95[f]
PF	J	Stripper vents	continuous[g]	0.85	40	VOC[e] – 0.01 air – 99.99
	Z	(All indicated streams to flare)		17.4	–	VOC[h] air
		Total Emission Rate		31.6		

[a]Source of Information: Industry correspondence.

[b]RMP – raw material preparation; PR – polymerization reaction; MR = material recovery; PF = product finishing.

[c]Emission characteristics are assumed to be the same as for HDPE, slurry process.

[d]Percent by weight.

[e]Cyclohexane.

[f]Mostly steam.

[g]Upsets occur 6 to 7 times per year.

[h]Includes ethylene, cyclohexane, and other VOC contaminers. Ethylene is emitted at about 5.2 kg VOC/Mg product; cyclohexane at about 3.6 kg VOC/Mg product, and other VOC at about 3.6 kg VOC/Mg product.

(Section 3.2.2.2) in terms of processing steps (Section 3.2.2.2.1), potential VOC emissions (Section 3.2.2.2.2), and potential growth. Further, as indicated in Section 3.2.2.2 for the LDPE low-pressure, gas phase process, some process equipment can produce both HDPE and LDPE. Thus, for the purposes of this project, the process and emission descriptions for the LDPE low-pressure, gas phase process serve also for production of HDPE when the same process is used.

3.2.4 <u>Polystyrene</u>

Styrene readily polymerizes to polystyrene by a relatively conventional free radical chain mechanism. Either heat or a catalyst, typically benzoyl peroxide or di-tert-butyl per-benzoate, will initiate the polymerization. Styrene will homopolymerize in the presence of inert materials and copolymerize with a variety of monomers. Pure polystyrene has the following structure:

$$\left(\begin{array}{c} H \\ | \\ -C-CH_2- \\ | \\ \bigcirc \end{array}\right)_n$$
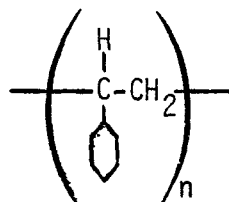
The homopolymers of styrene are also referred to as general purpose or crystal polymers. The copolymers of styrene are generally produced in the presence of particular elastomers to improve the strength of the polymer. They are called impact or rubber-modified polystyrenes. For them, the styrene content varies for rubber modified polystyrenes from about 88 to 97 percent, for styrene-acrylonitrile copolymers (SAN) from about 70 to 75 percent, and for styrene-butadiene copolymers from 50 percent and above. (Styrene-acrylonitrile copolymers and styrene-butadiene rubbers are not part of this source category.)

Homopolymers and copolymers can be produced by bulk (or mass), solution (a modified bulk), suspension, or emulsion polymerization techniques. In solution, or modified bulk, polymerization the reaction takes place as the monomer is dissolved in a small amount of solvent, such as ethylbenzene. Suspension polymerization takes place with the monomer suspended in a water phase. The bulk and solution polymerization processes are homogeneous (i.e., take place in one phase), whereas the suspension and emulsion polymerization processes are heterogeneous (i.e., take place in more than one phase). The bulk (mass) process is

the most widely used process for polystyrene today.[30] The suspension process is also commonly used, especially in the production of expandable beads.[30] The use of the emulsion process for producing homopolymer of styrene has decreased significantly since the mid-1940's.[31] This section describes both the bulk (mass) batch and continuous processes.

3.2.4.1 <u>Polystyrene Batch Process.</u> Various grades of polystyrene can be produced by a variety of batch processes. Batch processes generally have a high conversion efficiency, leaving only small amounts of unreacted styrene to be emitted if the reactor is purged or opened between batches. A typical plant will have multiple process trains, each usually capable of producing a variety of grades of polystyrene.

3.2.4.1.1 <u>Process description.</u> Figure 3-8 is a schematic representation of the polystyrene batch bulk polymerization process. Pure styrene monomer and comonomer (if a copolymer product is desired) are pumped from storage (1) to the mix feed tank (2), then usually to an agitated tank, often a prepolymerization reactor, for mixing the reactants. Small amounts of mineral oil (as a lubricant and plasticizer), the dimer of alpha-methylstyrene (as a polymerization regulator), and an antioxidant are added. Polybutadiene may be added in the case of production of an impact grade polystyrene. The blended or partially polymerized feed is then pumped into a batch reactor (3). During the reactor filling process, some styrene vaporizes and is vented through an overflow drum (4). When the reactor is charged, the vent is closed and polymerization is thermally initiated. The reaction may also be initiated by introducing a free radical initiator into the feed tank along with other reactants. After polymerization is complete, the polymer melt, which contains some unreacted styrene monomer, ethylbenzene (an impurity from the styrene feed), and low molecular weight polymers (dimers, trimers, and other oligomers) is pumped to a vacuum devolatilizer (5). In the devolatilizer the residual monomer, ethylbenzene and low polymers are separated, condensed (6), and sent to the by-product recovery unit (7). Overhead vapors from the condenser are usually exhausted through a vacuum pump (8). Molten polystyrene from the bottom of the devolatilizer is pumped through a stranding dieplate into a cold water bath. The cooled strands are pelletized (9) and sent to product storage (10).
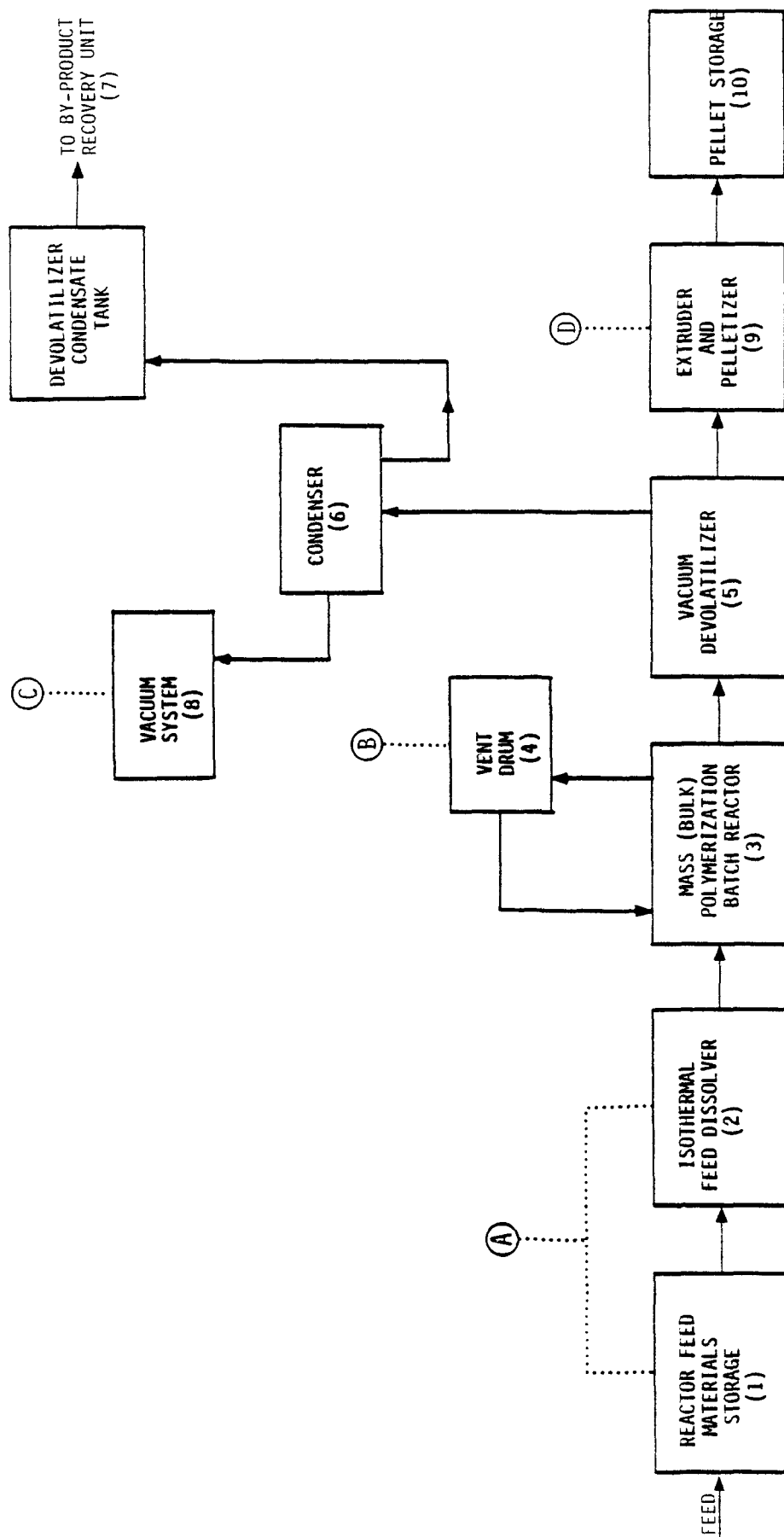
Figure 3-8. Simplified Process Block Diagram for the Polystyrene Batch Process

3.2.4.1.2 <u>Emissions from the polystyrene batch process.</u> The process has four major emission sources, which are: the monomer storage and feed dissolver vent, the reactor drum vent, the styrene condenser vent, and the extruder quench vent. These are labeled A, B, C, and D, respectively, in Figure 3-8.

The total emission rate is estimated to range from 0.64 to 2.5 kg VOC/Mg product. The major vent is the devolatilizer condenser vent (Stream B). This continuous offgas vent has an emission rate of 0.25 to 0.75 kg VOC/Mg product. The emissions consist of unreacted styrene, which is flashed from the product polymer in the vacuum devolatilizer, extremely diluted in air due to leakage. The stream is exhausted through a vacuum system (i.e., vacuum pump and oil demister), to the atmosphere.

The VOC emissions associated with other continuous offgas streams (A and C) are 0.09 kg VOC/Mg product and 0.15 to 0.3 kg VOC/Mg product, respectively. Pure styrene is emitted directly to the atmosphere from the monomer storage and feed dissolver vent (Stream A), whereas steam and styrene vapor are usually vented through a forced-draft hood (Stream C) and passed through a mist separator pad or electrostatic precipitator before venting to the atmosphere from the extruder quench vent.

The only intermittent offgas stream from a batch process is the reactor drum vent (Stream D). Its VOC emissions range from 0.12 to 1.35 kg VOC/Mg product. Emissions occur from the reactor drum vent only during reactor filling periods. Their filling frequency is once per day, and the associated offgases are vented to the atmosphere. Table 3-12 summarizes these four sources of VOC emissions.

3.2.4.2 <u>Polystyrene Continuous Process.</u> As with the batch process, various continuous processes are used to make a variety of grades of polystyrene or copolymers of styrene. The chemical reaction in continuous processes does not approach completion as efficiently as the reaction in batch processes. As a result, a lower percentage of styrene is converted to polystyrene, and larger amounts of unreacted styrene may be emitted from continuous process sources. A typical plant may contain more than one process train, each producing either the same or different grades of polymer or copolymer.

3.2.4.2.1 <u>Process description.</u> The bulk (mass) continuous process is represented in Figure 3-9. The feed dissolver tank (1) is charged

Table 3-12. CHARACTERISTICS OF VENT STREAMS FROM THE POLYSTYRENE BATCH PROCESS[a]

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition, Wt.%
d	A	Monomer storage & feed dissolver[c]	Continuous	0.09	-	-	100 Styrene
PR	B	Reactor vent drum vent	Intermittent[e]	0.12-1.35[f]	43	-	6.9 Styrene 35.0 Water 58.1 Air
MR	C	Devolatilizer condenser vent	Continuous	0.25-0.75	27	0	3 Styrene 97 Air
PF	D	Extruder quench vent	Continuous	0.15-0.3	21	-	Styrene & steam
		Total Emission Rate		0.61-2.5			

[a]Source of information: Industry correspondence.
[b]MR = material recovery; PF = product fininshing; PR = polymerization reaction.
[c]See Figure 3-8 for stream identification.
[d]Raw material storage emissions are not covered in this project.
[e]One occurrence per day.
[f]Average over a year.
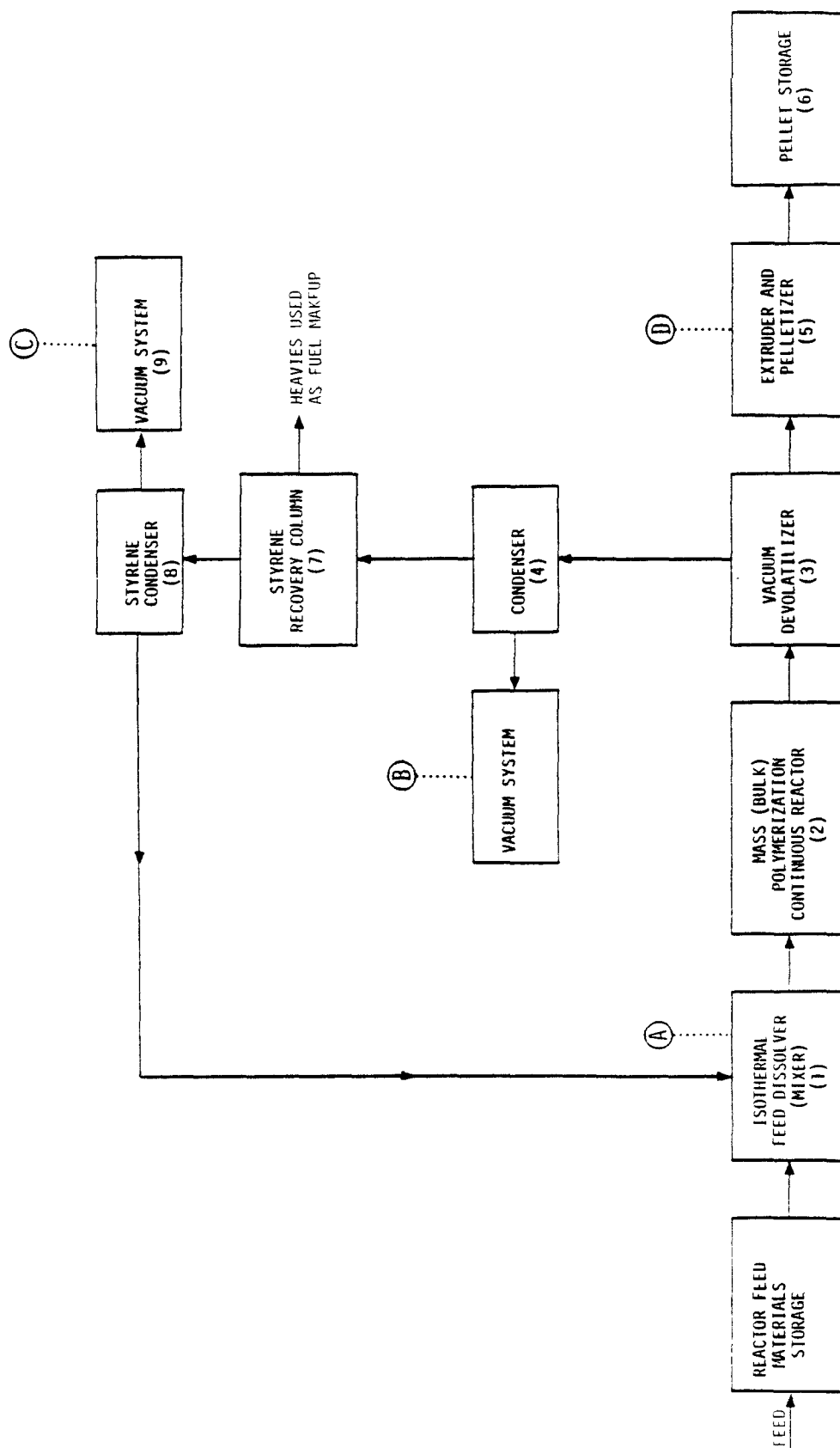
Figure 3-9. Simplified Process Block Diagram for the Polystyrene Continuous Process

with styrene, polybutadiene (if an impact grade product is desired), mineral oil (lubricant and plasticizer), and small amounts of recycled polystyrene, antioxidants and other additives, in proportions that vary according to the grade of resin to be produced. Blended feed is pumped continuously to the reactor system (2) where it is thermally polymerized to polystyrene. A process train usually employs more than one reactor in series. Some polymerization occurs in the initial reactor, often referred to as the prepolymerizer. Polymerization to successively higher levels occurs in subsequent reactors in the series. Either stirred autoclaves or tower reactors are employed, depending on the variation in the process. The polymer melt, which contains unreacted styrene monomer, ethylbenzene (an impurity from the styrene feed) and low polymers, is pumped to a vacuum devolatilizer (3). In the devolatilizer, most of the monomer, ethylbenzene, and low molecular weight polymers are separated, condensed (4), and sent to the styrene recovery unit. Noncondensables (i.e., overhead vapors) from the condenser are typically exhausted through a vacuum pump. Molten polystyrene from the bottom of the devolatilizer is pumped by an extruder through a stranding dieplate into a cold water bath. The solidified strands are then pelletized (5) and sent to storage (6).

In the styrene recovery unit, the crude styrene monomer recovered from the condenser (4) is purified in a distillation column (7). The styrene overhead from the tower is condensed (8) and returned to the feed dissolver tank. Noncondensables are vented through a vacuum system (9). Column bottoms containing low molecular weight polymers are sometimes used as a fuel supplement.

3.2.4.2.2 <u>Emissions from the polystyrene continuous process</u>. The process has four types of vent streams, all of which are continuous. These are: the feed dissolver vent, the devolatilizer condenser vent, the styrene recovery unit condenser vent, and the extruder quench vent. These are Streams A, B, C, and D, respectively, in Figure 3-9. Industry's experience with continuous polystyrene plants indicates a wide range of emission rates from plant to plant. It is estimated that the typical total VOC emission rate is about 3.25 kg VOC/Mg product. Table 3-13 presents the VOC vent stream characteristics.

Table 3-13. CHARACTERISTICS OF VENT STREAMS FROM THE POLYSTYRENE CONTINUOUS PROCESS[a]

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition, Wt. %
RMP	A	Feed Dissolver	Continuous	0.009	-	-	92 Styrene 7.5 Polybutadiene 1.5 Other
MR	B	Devolatilizer condenser vent	Continuous	0.05-2.96	100	0	21.8 Styrene 78.2 Steam
MR	C	Styrene recovery unit condenser vent	Continuous	0.05-0.13	100	0	2.1 Styrene 97.9 Steam
PF	D	Extruder quench vent	Continuous	0.15	21	-	99.99 Steam Trace Styrene
		Total Emission Rate		0.26-3.25			

[a] Source of Information: Industry correspondence.

[b] RMP = raw material preparation; MR = material recovery; PF = product finishing.

[c] See figure 3-9 for stream identification.

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These vent streams differ from those of the batch process. Emissions from the devolatilizer condenser vent tend to be higher from the continuous process than from the batch process. In addition, there is no reactor vent drum vent in the continuous process. The other offgas streams (from storage and the extruder) have emissions comparable to those of the batch processes.

Two types of vacuum systems are now used in the industry. One relies on steam ejectors; the other, on vacuum pumps. Where steam ejectors are used, the overheads from the devolatilizer condenser vent and the styrene recovery unit condenser vent are composed mainly of steam. Some companies have recently replaced these steam ejectors with vacuum pumps. When vacuum pumps are used, emissions as well as energy consumption are lower than with steam ejectors.

3.2.5 Polyester Resin

Polyester resins used for fiber production may be classified chemically as either poly(ethylene terephlate [PET] resins or non-PET resins. Of these two types, PET resins are the most important polyester as a thermoplastic synthetic fiber.

PET resins are produced commercially from ethylene glycol (EG) by either the dimethyl terephthalate (DMT) process or the terephthalic acid (TPA) process. Both processes first produce the intermediate bis-(2-hydroxyethyl)-terephthalate (BHET) monomer and then polymerize it to PET under reduced pressure with heat and catalyst. Emissions from the two processes differ in that the DMT process produces methanol as a by-product during esterification, whereas water is a by-product in the TPA process. The production of methanol vapor in the DMT process creates the need for methanol recovery and purification operations and their attendant VOC emissions. Both the DMT and TPA processes are described in this section.

3.2.5.1 PET/DMT Process. The DMT process is the older of the two processes for making PET. Currently, most PET is produced using the DMT process, which may be either batch or continuous. The basic differences in going from the batch to the continuous process are (1) the replacement of the kettle-reactor with a column-type reactor for esterification, (2) "no-back-mix" (i.e., no stirred tank) reactor designs are required

in the continuous process at the polymerizer, and (3) different additives and catalysts are used to assure proper product characteristics.[32]

3.2.5.1.1 <u>Process description</u>. A schematic diagram of the continuous DMT process is presented in Figure 3-10. The primary reaction is:

$$CH_3OOC \text{—}\bigcirc\text{—} COOCH_3 + HOCH_2CH_2OH \rightarrow HO - (OC \text{—}\bigcirc\text{—} COOCH_2CH_2O)_n H + 2nCH_3OH$$

DMT   EG   PET
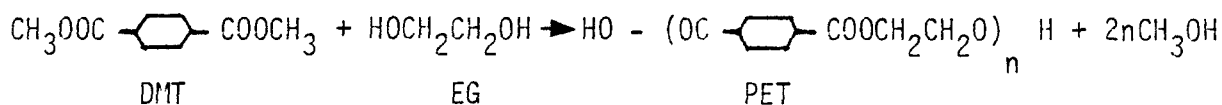
Dimethyl terephthalate and ethylene glycol are purified (1) before being fed to the esterification reactor (2). There, in the presence of a catalyst, they react to form bis-hydroxyethyl terephthalate (BHET) and methanol. The continuous removal of methanol is necessary in order to shift the reaction equilibrium in favor of increased production of BHET. Therefore, a vent stream is withdrawn from the esterifier. This stream is fed to a methanol recovery process (3) where methanol is condensed and purified by distillation, before being forwarded to the methanol storage tank (4). The BHET monomer is polymerized (5) to PET in a second reaction step under reduced pressure with heat and a catalyst. The polymerization reaction may be carried out in two or more reactors in series operated at increasing temperatures and successively lower pressures. Unreacted ethylene glycol is flashed from the polymer product, condensed, and, if desired, can be purified and recycled.

3.2.5.1.1 <u>Emissions from the PET/DMT process</u>. The methanol recovery section and the reactor are the major sources of VOC emissions from the DMT process. Both are continuous. These are shown as Streams A and B in Figure 3-10. A summary of the vent stream characteristics is provided in Table 3-14. Emissions from the methanol recovery section are recovered by condensers, which results in relatively low emission rates. This stream is typically composed of methanol and nitrogen.

The emission stream from the reactor is composed primarily of ethylene glycol with small amounts of methanol vapors and volatile feed impurities. The amount of ethylene glycol that is emitted to the atmosphere depends upon where the ethylene glycol is recovered. A plant may recover the ethylene glycol by using a spent ethylene glycol spray condenser directly off of the reactors and before the stream passes through the vacuum system. The condensed ethylene glycol may then be recovered through distillation. This type of recovery system results in low emission
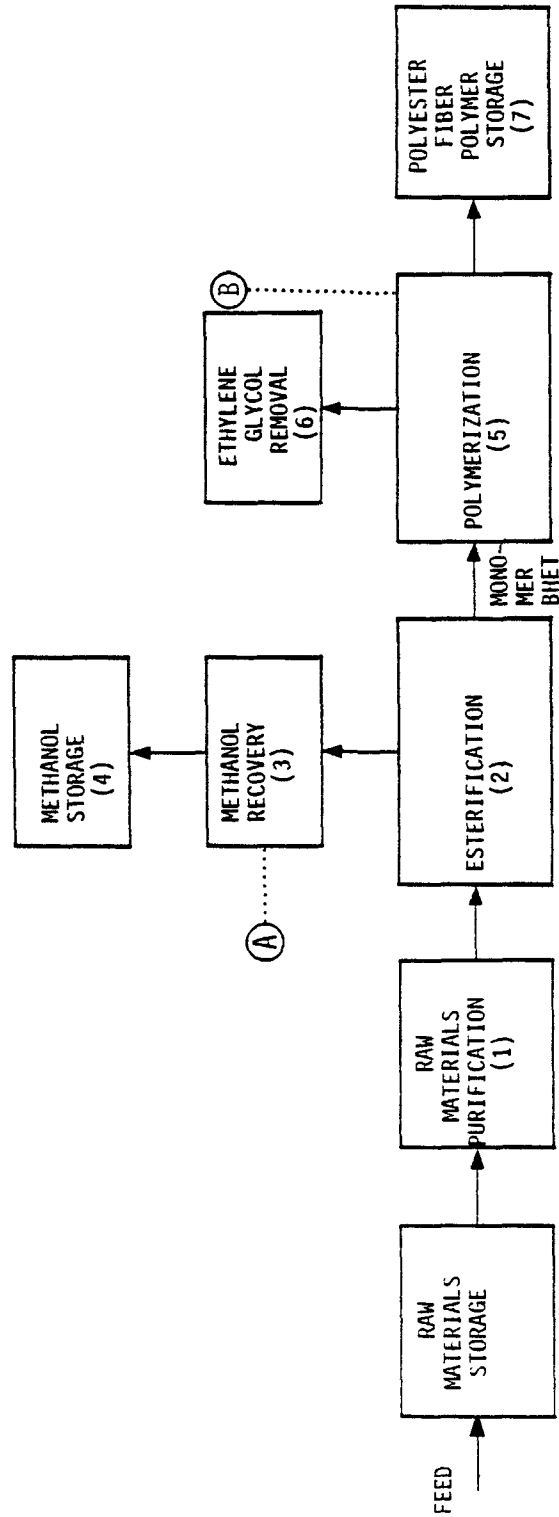
Figure 3-10. Simplified Process Block Diagram for the PET/DMT Process

Table 3-14. CHARACTERISTICS OF VENT STREAMS FROM THE POLY(ETHYLENE TEREPHTHALATE) DMT PROCESS

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature °C	Pressure, psig	Composition, wt.%
MR	A	Esterifier Vent (Methanol recovery vent)	Continuous	0.18	38	0	49 Methanol 51 Nitrogen
PR	B	Polymerizer Reactor Cooling Water Tower	Continuous		25	0.2	VOC Air H2O
		– with spent ethylene glycol spray condenser		0.21[d]			
		– without spent EG spray condenser		9.5[e]			
		Total Emission Rate		—			
		– with EG spray condenser		0.39			
		– without EG spray condenser		9.68			

[a] Source of information: Industry correspondence.

[b] MR = material recovery; PR = polymerization reaction.

[c] See Figure 3-10 for stream identification.

[d] Based upon 0.02 kg VOC/Mg product from recovery equipment and 0.19 kg VOC/Mg product from cooling water tower.

[e] Estimated to be the same as for the DMA process. All emissions from cooling water tower.

rates. Alternatively, a plant may send the emission stream directly through the vacuum system (typically composed of steam ejectors) without using a spent ethylene glycol spray condenser. The steam ejectors used to produce a vacuum will result in contaminated water, which is then cooled for reuse. Ethylene glycol in the cooling tower is recovered from the bottom of the cooling water with distillation columns. This system of recovering ethylene glycol results in much higher emission levels of ethylene glycol, because of the cooling water's contact with the atmosphere.

3.2.5.2 **PET/TPA Process.** The TPA process, which has been available only since 1963, is now generally preferred over the DMT process because it avoids the recovery and purification of the methanol by-product generated in the DMT process. The TPA process can also be a batch or continuous process.

3.2.5.2.1 **Process description.** A schematic diagram of the continuous TPA process is presented in Figure 3-11. The primary reaction is:

$$HOOC \underset{}{-\!\!\!\bigcirc\!\!\!-} COOH + HOCH_2CH_2OH \longrightarrow HO - (OC \underset{}{-\!\!\!\bigcirc\!\!\!-} COOCH_2CH_2O)_n H + 2nH_2O$$

| TPA | EG | PET |
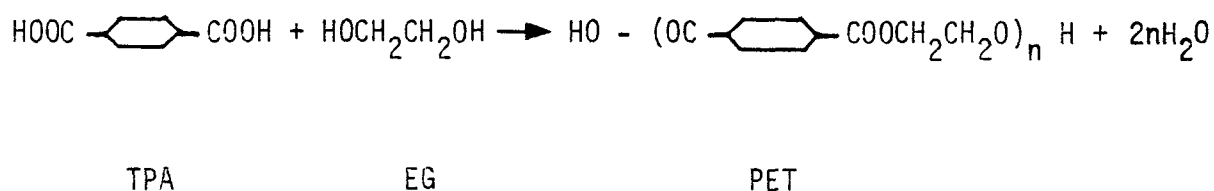
The processing steps for producing PET by this method are very similar to those of the DMT process except that water rather than methanol is the by-product from the esterifiers. Thus, the TPA process does not have methanol recovery and purification.

3.2.5.4 **Emissions from the PET/TPA process.** There are two sources of VOC emissions from this process as illustrated by Figure 3-11. These continuous streams are emitted from the esterifier (Stream A), and from the polymerizer (Stream B). Offgas characteristics for these streams are summarized in Table 3-15.

As with the DMT process, the overall emission rate from this process depends on the type of system used to recover the ethylene glycol. If the ethylene glycol is recovered from the cooling water, emissions occur from the cooling tower. In this system, offgases from the esterifiers are sent through a distillation column to recover the ethylene glycol.
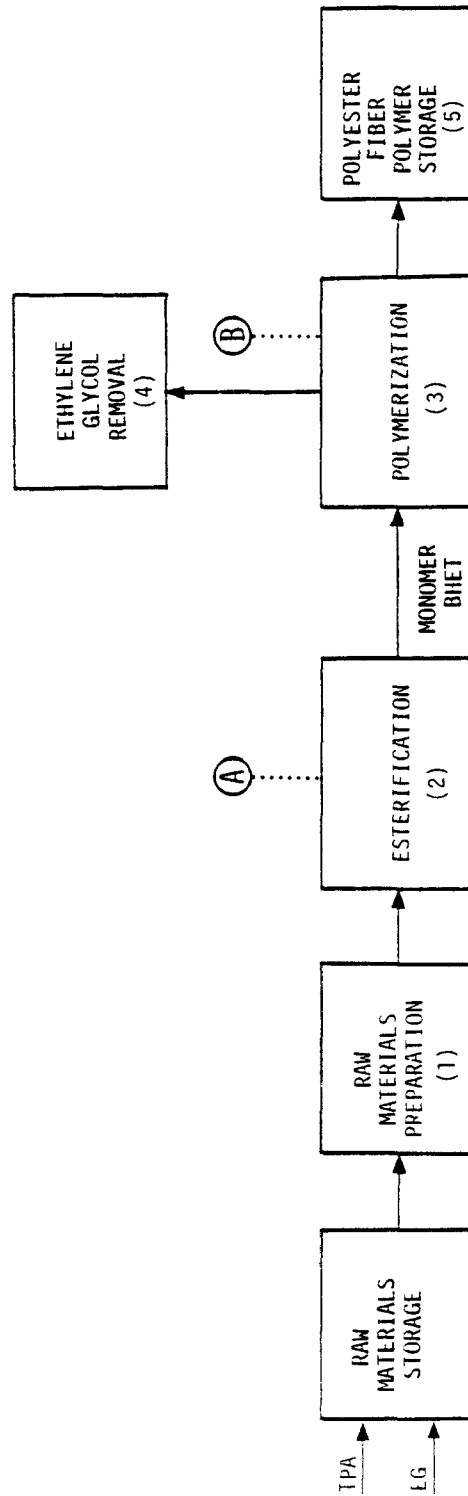
Figure 3-11. Simplified Process Block Diagram for the PET/TPA Process

Table 3-15. CHARACTERISTICS OF VENT STREAMS FROM THE POLY(ETHYLENE TEREPHTHALATE) TPA PROCESS

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature °C	Pressure, psig	Composition, Wt. %
RMP	A	Esterifier Vent[d]	Continous	0.04	38	Atm.	VOC, N$_2$, H$_2$O
PR	B	Polymerizer Reactor Cooling Water Tower	Continous		25	0.2	VOC, Air, H$_2$O
		- with spent ethylene glycol spray condenser		0.21[e]			
		- without spent ethylene glycol spray condenser		9.5[f]			
		Total Emission Rate		--			
		- with EG spray condenser		0.25			
		- without EG spray condenser		9.5			

[a]Source of information: Industry correspondence.
[b]RMP = raw material preparation; PR = polymerization reaction.
[c]See Figure 3-11 for stream identification.
[d]The potential emissions from this stream are sent to reactor cooling water tower when the spray condenser is not used.
[e]Based upon 0.02 kg VOC/Mg product from recovery equipment and 0.19 kg VOC/Mg product from cooling water tower.
[f]Includes potential emissions from Stream A.

The overheads from this distillation column are fed to the cooling tower. Under this setup, there is only one major emission source--the cooling tower. The ethylene glycol emission rate has been calculated to be about 9.5 kg VOC/Mg of product. Alternatively, where a spent ethylene glycol spray condenser is used to recover the ethylene glycol directly from the reactors, reflux condensers may be used on the esterifiers. This set-up results in emissions of about 0.04 kg VOC/Mg of product from the esterifiers and about 0.21 kg VOC/Mg of product from the reactors.

3.3 FUGITIVE VOC SOURCES AND EMISSIONS

Fugitive VOC emissions result when process fluids leak from the plant equipment. The potential fugitive VOC sources in the polymers and resins processes are similar to those in synthetic organic chemicals manufacturing and petroleum refining. Sources include valves, pump seals, compressor seals, safety or relief valves, flanges, sampling connections, and open-ended lines. Fugitive emission sources are extensively described in References 33 through 37.

Table 3-16 lists the vapor pressures of the various organic compounds used in the polymer and resin processes. Those compounds with vapor pressures greater than 0.3 kPa (2.25 mm Hg) at 20°C (68°F) are considered to be "light liquids," and those with vapor pressures equal to or less than 0.3 kPa at 20°C are considered to be "heavy liquids." For the purposes of this project, only those organic compounds considered as "light liquids" under this classification scheme are of concern with regard to fugitive emissions.

Data characterizing the uncontrolled levels of fugitive emissions in the polymers and resins industry are generally unavailable at the present time. However, data of this type have been obtained for the synthetic organic chemical manufacturing industry (SOCMI) and the petroleum refining industry. Because the operation of the various process equipment in the polymers and resins industry is not expected to differ greatly from the operation of the same equipment in the SOCMI industry, the SOCMI fugitive emission data can be used to approximate the levels of fugitive emissions in the polymers and resins industry. The final SOCMI data, which are used for the purposes of this project, are presented in

Table 3-16. VAPOR PRESSURES OF MAJOR ORGANIC COMPOUNDS USED
IN THE POLYMERS AND RESINS SEGMENTS CHOSEN FOR NSPS DEVELOPMENT

Organic Compound	Vapor Pressure	Process
Ethylene	>38,000 mm Hg @ 8.9°C	PP, PE
Ethane	>30,000 mm Hg @ 23.6°C	PE
Propylene	7,600 mm Hg @ 19.8°C	PP, PE
Propane	7,600 mm Hg @ 26.9°C	PP
Iso-butane	1,520 mm Hg @ 7.5°C	PE
Butene	1,410 mm Hg @ 21°C	PE
Hexane	100 mm Hg @ 15.8°C	PP
Methanol	60 mm Hg @ 12.1°C	PET (DMT)
Isopropyl Alcohol	20 mm Hg @ 12.7°C	PP, PE
Butylene	5 mm Hg @ 11.6°C	PP
Ethylbenzene	5 mm Hg @ 13.9°C	PS
Styrene	5 mm Hg @ 18°C	PS
Light/Heavy Liquid Break	2.25 mm Hg @ 20°C	--
Ethylene glycol	0.05 mm Hg @ 20°C	PET

Key: PP = Polypropylene
      PE = Polyethylene
      PS = Polystyrene
      PET = Polyethylene terephthalate
      DMT = Dimethyl terephthalate

Source: Perry, Robert H. and Cecil H. Clinton, Chemical Engineers'
         Handbook. 5th Edition. McGraw Hill. New York, 1973.

Table 3-17. (See References 36 and 37 for a detailed discussion of the derivation of these emission rates.)

3.4 BASELINE EMISSIONS

The baseline emission level is that level of emissions achieved in the absence of additional EPA standards; in this instance, in the absence of a polymers and resins NSPS. This section describes briefly the various industrial practices and existing regulations that affect baseline emissions.

3.4.1 Process Emissions

Process emissions in the polymers and resins industry are controlled through both industrial practices and government regulations. Industrial practices for polyolefin production are distinctly different than for either polystyrene or polyester production. The following sections describe briefly the relevant industrial practices and regulations.

3.4.1.1 Industrial Practices. In general, most polyolefin plants control the large volume, intermittent streams with flares to avoid buildup of explosive concentrations within the plant. Continuous streams may also, on occasion, be controlled by a flare.

Liquid phase polypropylene plants do not routinely install VOC control equipment on the smaller continuous emission streams. The polymerization reactors and the diluent separation and purification units are generally provided with emergency relief valves leading to a flare (for safety purposes) in case of an upset. These emergency vents usually pass through "knock-out" drums, which disentrain liquids and polymer particles before the vapors are released to the flare. As polypropylene units are subject to plugging, most are provided with emergency relief valves throughout the entire process and, in the great majority of cases, these relief valves discharge to the flare header.

Both vent streams from the gas phase polypropylene plant are flared.

In the liquid phase, low density polyethylene plant, flares are generally used to control emissions from safety relief valves, except from the reactor. The emergency reactor vent stream typically is not flared, because safety considerations dictate the need for a particulate removal system. Based on available information, only one company has a particulate polymer removal technology that can handle high-pressure

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Table 3-17. UNCONTROLLED FUGITIVE EMISSION RATES

Fugitive emission source	Uncontrolled emission rate,[a] kg/hr/source
Valves	
Gas	0.0056
Light liquid[b]	0.0070
Heavy liquid[c]	0.00023
Pump seals	
Light liquid[b]	0.0494
Heavy liquid[c]	0.0214
Compressor seals	0.228
Safety or relief valves	
Gas	0.1040
Flanges	0.00083
Sampling connections	0.015
Open-ended lines	0.0017

[a]These uncontrolled emission levels are based on the data presented in Reference 35.

[b]Light liquid is defined as a petroleum liquid with a vapor pressure greater than that of kerosene.

[c]Heavy liquid is defined as a petroleum liquid with a vapor pressure equal to or less than that of kerosene.

emergency vent gas, and this system is used on tubular reactors. Controls are not routinely applied to the dryer and bin storage vents.

Gas phase polyethylene plants (based upon Union Carbide's Unipol process), currently flare both the continuous and intermittent emission streams. The flare system has a liquid seal and is air assisted.[38]

In liquid phase, high density polyethylene plants, a flare is generally installed as part of the safety system. Safety relief devices leading to the flare are utilized to avoid accidents resulting from equipment overpressurization or other malfunction. While recycle treater vent streams are usually flared, dryer vent streams are usually vented directly to the atmosphere. This is done because traditional dryers dilute the organic with large quantities of air, making the cost of burning the organic prohibitive.

In the polystyrene industry, no routine control is applied to the batch or continuous processes other than condensation operations in which styrene is recovered, due to its value and ease of recovery. Offgas from the styrene condenser and other vents are usually vented directly to the atmosphere. Flares are not usually installed in these plants.

In the polyester industry, control devices (incinerators or flares) are not used. Rather, as in the polystyrene industry, condensers and distillation columns are installed to recover methanol and/or unreacted ethylene glycol because of their value.

3.4.1.2 <u>Current State VOC Regulations</u>. Many of the these types of polymer and resin plants are located in five States (California, Illinois, Louisiana, New Jersey, and Texas) which have regulations that limit VOC emissions. Almost all of the polyolefin plants are located in Texas or Louisiana. Only one of the polyester plants, however, is located in any of these five States; it is in New Jersey.

The five States have different VOC regulations. These are summarized below:

1. In California, the South Coast Air Quality Management District (SCAQMD), which contains five of the six polystyrene plants in the State, imposes an emission limitation of 34 kg/day (75 lbs/day).

2. Illinois allows new sources to comply with either of two standards:

   a. No waste gas stream discharged into the atmosphere in excess of 100 ppm equivalent methane (molecular weight of 16.0) [Rule 205(g)(1)(A)(iii)], or

   b. A maximum of 8 pounds per hour of organic material [Rule 205(g)(1)(C)(i)].

   Emissions of organic material in excess of 8 pounds per hour is allowed, provided such emissions are controlled by State agency-approved air pollution control methods or equipment capable of reducing 85 percent or more of the uncontrolled organic material that otherwise would be emitted to the atmosphere [Rule 205(g)(1)(C)(ii)].

   In addition, in the case of emissions from vapor blowdown systems or any safety relief valve, except such valves not capable of causing an excessive release (a discharge of more than 0.65 pound of mercaptans and/or hydrogen sulfide into the atmosphere in any 5-minute period), such emissions must be controlled to 10 ppm equivalent methane or less, through combustion in a smokeless flare, or by some other State agency-approved control device [Rule 205(g)(2)(A-C].

3. Under Louisiana law, new sources must burn VOC waste gas streams at a minimum temperature of 704°C (1,300°F) for 0.3 second or longer in a direct flame afterburner or any equally effective device [Section 22.8(a)]. This law allows the following exemptions [Section 22.8(c)(1-3)]:

   a. Waste gas streams with less than 100 tons/yr VOC emissions.

   b. Waste gas streams that will not support combustion without the addition of auxiliary fuel.

   c. Waste gas streams where disposal cannot be practically or safely accomplished by other means without causing an economic hardship.

   Also significant is that this section (Section 22.8) does not apply to safety relief and vapor blowdown systems where control cannot be accomplished because of safety or economic considerations.

In addition, Section 22.9 limits emissions of organic solvents from any source which uses organic solvents to 1.3 kilograms (3 pounds) per hour or 6.8 kilograms (15 pounds) per day. Where emissions exceed these amounts, they must be reduced where feasible by one or more of the following methods:

a. Incineration, provided 90 percent of the carbon in the organic compounds being incinerated is oxidized to carbon dioxide [except as provided in 22.9.3(a)].

b. Carbon adsorption of the organic material.

c. Any other equivalent means as may be approved by the Technical Secretary.

Where a waste gas stream may be subject to both Sections 22.8 and 22.9, then that stream must show compliance with both.

4. New Jersey law regulating VOC emissions uses a "sliding scale" to determine allowable emissions. Vapor pressure and concentration of VOC in the vent stream are used to determine applicable exclusion rates and maximum allowable emissions. The exclusion rates range from 0 to 3.2 kilograms (7 pounds) per hour and the control efficiencies required to comply with the maximum allowable emissions range from 85 to 99.7 percent.

5. The Texas Air Control Board (TACB) regulates new polymer and resin plants on a case-by-case basis, requiring the application of best available control technology (BACT). The TACB follows several general "rules of thumb" in making BACT determinations:

a. All waste gas streams, including analyzer vents, cannot be vented directly to the atmosphere, unless provided for in a special provision in the operating permit;

b. Reactors are to have pressure and temperature controls to minimize pressure excursions that require emergency releases;

c. Inlet and outlet valves to the reactors are to be placed as close to the reactor as possible to help minimize emissions in cases of emergency releases; and

d. VOC emissions from the extruder, pelletizier, and end-product storage should be able to meet a limit of 350 lbs of VOC

per million lbs of product. The exact emission limit may
be higher or lower, depending on the individual case.

3.4.2 Fugitive Emissions

There are presently no Federal regulations that specifically reduce
emissions from polymers and resins manufacturing plants. However, some
fugitive emission reduction is achieved by operating practices currently
followed by industry and applicable State or local regulations.

3.4.2.1 Industrial Practices. The industrial practices used by
the SOCMI industry are typically used in the polymers and resins industry.
Their primary reason for controlling fugitive emissions is the economic
loss from leaks. Such leaks are usually large enough to be physically
evident (that is, can be seen, heard, or smelled) and are termed "easily
detectable leaks." These are normally repaired to minimize the loss of
product. Fugitive emissions, as considered in this report, are considerably
smaller and less readily identified than "easily detectable leaks." For
a detailed description of fugitive emission control practices that may
be used, see References 34 and 35.

3.4.2.2 Existing Regulations. There are two types of regulations
that affect fugitive VOC emissions from polymer and resin plants. The
first regulates industrial operating practices on the basis of worker
health and safety. Because some aspects of these regulations deal with
worker exposure to process emissions, they may have some impact on
fugitive VOC emissions. The second type is regulations that were specifically
developed to limit fugitive emissions.

3.4.2.2.1 Health and safety regulations. Several regulations have
been developed under the direction of the Occupational Safety and Health
Administration (OSHA) and the National Institute for Occupational Safety
and Health (NIOSH) to limit worker exposure to chemical substances.
Protecting the workers may be accomplished by limiting emissions or by
providing workers with individual safety protection from the emissions.
Thus, although present health and safety regulations do not mandate a
reduction in fugitive VOC emissions, and any reduction in fugitive
emissions resulting from these regulations is "incidental," OSHA regulations
may affect a company's attitude about leaks and fugitive VOC emissions.

3.4.2.2.2 <u>Fugitive emissions regulations</u>. Currently, only California and Texas have fugitive regulations and they apply only to new polymer and resin plants.

California presently prohibits open-ended process lines to minimize fugitive VOC emissions. This State also requires relief valves to discharge to a flare system, and be monitored and maintained, or a rupture disk to be used. In addition to these regulations, the SCAQMD also requires these plants to vent fugitive emissions from compressor seals to a fired-heater or flare system.

Texas requires new polyolefin plants to use enclosed compressors to vent fugitive emissions to a fired-heater or flare system.

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4.0 EMISSION CONTROL TECHNIQUES

Volatile organic compounds (VOC), used as solvents and key raw
materials in the manufacture of polymers and resins, are emitted to the
atmosphere from a variety of process equipment. The emissions may be
considered as two large groups: process emissions, which result from
fundamental operations of the process, and fugitive emissions, those
that escape directly to the atmosphere rather than through a flare or
exhaust system. Process VOC emissions can be reduced either by installing
emission control devices or by reducing the VOC in the vent streams by a
process modification such as recovery of monomer or solvent. Fugitive
VOC emissions can be reduced or essentially eliminated by increased
surveillance and maintenance or by installation of specified controls or
leakless equipment. This chapter describes emission control techniques
that may be used to reduce these emissions from the polymers and resins
industry. Control techniques for process emissions are discussed in
Section 4.1 and for fugitive emissions in Section 4.2.

4.1 CONTROL TECHNIQUES FOR PROCESS EMISSIONS

Process emissions from the manufacture of polymers and resins are
diverse in both composition and flow. Streams contain a wide range of
VOC concentrations, i.e., less than 1 percent to essentially 100 percent,
but most are of high concentration. Some streams are continuous, while
others are intermittent. Process emissions also differ in temperature,
pressure, heating value, and miscibility. These factors are extremely
important in the selection and design of VOC emission control equipment.

Due to this diversity, different control techniques may be appropriate
for different vent streams. The control techniques may be characterized
by two broad categories: combustion techniques and recovery techniques.
Combustion techniques such as flares and incinerators are applicable to a
variety of VOC streams. Recovery techniques such as condensation,

absorption, and adsorption, are effective for some select vent streams. Economic incentives may encourage the use of either type of VOC control, since certain combustion configurations may permit heat recovery, and recovery techniques permit the conservation and reuse of valuable materials. The selection of a control system for a particular application is based primarily on considerations of technical feasibility and process economics.

The most common control techniques form the basis for this chapter. Basic design considerations for flares, thermal and catalytic incinerators, industrial boilers, condensers, absorbers, and adsorbers, are briefly described. The conditions affecting the VOC removal efficiency of each type of device and its applicability for use in the polymers and resins industry are examined. Emphasis has been given to flares, thermal incinerators, and condensers because of their wide applicability to a variety of VOC streams. Combustion techniques are discussed in Subsection 4.1.1 and recovery techniques in Subsection 4.1.2.

4.1.1 Control by Combustion Techniques

The four major combustion devices that are or can be used to control VOC emissions from the polymers and resins industry are: flares, thermal or catalytic incinerators, and boilers. Flares are the most widely used control devices at polyethylene and polypropylene manufacturing plants. Incinerators and boilers are also used, to a lesser extent, to control continuous vent streams. Although these control devices are founded upon basic combustion principles, their operating characteristics are very different. While flares can handle both continuous and intermittent streams, neither boilers nor incinerators can effectively handle large volume intermittent streams. Subsection 4.1.1 discusses the general principles of combustion, and then the design and operation, VOC destruction efficiency, and applicability of these four combustion devices at polymers and resins manufacturing plants.

Combustion is a rapid oxidation process, exothermic in nature, which results in the destruction of VOC by converting it to carbon dioxide and water. Poor or incomplete combustion results in the production of other organic compounds including carbon monoxide. The chemical reaction sequence which takes place in the destruction of VOC by combustion is a complicated process. It involves a series of reactions that produce

free radicals, partial oxidation products, and final combustion products. Several intermediate products may be created before the oxidation process is completed. However, most of the intermediate products have a very short life and, for engineering purposes, complete destruction of the VOC is the principal concern.[1]

Destruction efficiency is a function of temperature, turbulence, and residence time. Chemicals vary in the magnitudes of these parameters that they require for complete combustion. An effective combustion technique must provide:[2]

1. Intimate mixing of combustible material (VOC) and the oxidizer (air),

2. Sufficient temperature to ignite the VOC/air mixture and complete its combustion,

3. Required residence time for combustion to be completed, and

4. Admission of sufficient air (more than the stoichiometric amount) to oxidize the VOC completely.

4.1.1.1 Flares. Flaring is an open combustion process in which the oxygen required for combustion is provided by the air around the flame. Good combustion in a flare is governed by flame temperature, residence time of components in the combustion zone, turbulent mixing of components to complete the oxidation reaction, and oxygen for free radical formation.

There are two types of flares: ground level flares and elevated flares. Kalcevic (1980) presents a detailed discussion of different types of flares, flare design and operating considerations, and a method for estimating capital and operating costs for flares.[3] Elevated flares are most common in the polymers and resins industry. The basic elements of an elevated flare system are shown in Figures 4-1 and 4-2. Process offgases are sent to the flare through the collection header. The offgases entering the header can vary widely in volumetric flowrate, moisture content, VOC concentration, and heat value. The knock-out drum removes water or hydrocarbon droplets that could create problems in the flare combustion zone. Offgases are usually passed through a water seal before going to the flare. This prevents a possible flame flashback, caused when the offgas flow to the flare is too low and the flame front pulls down into the stack.

4-3
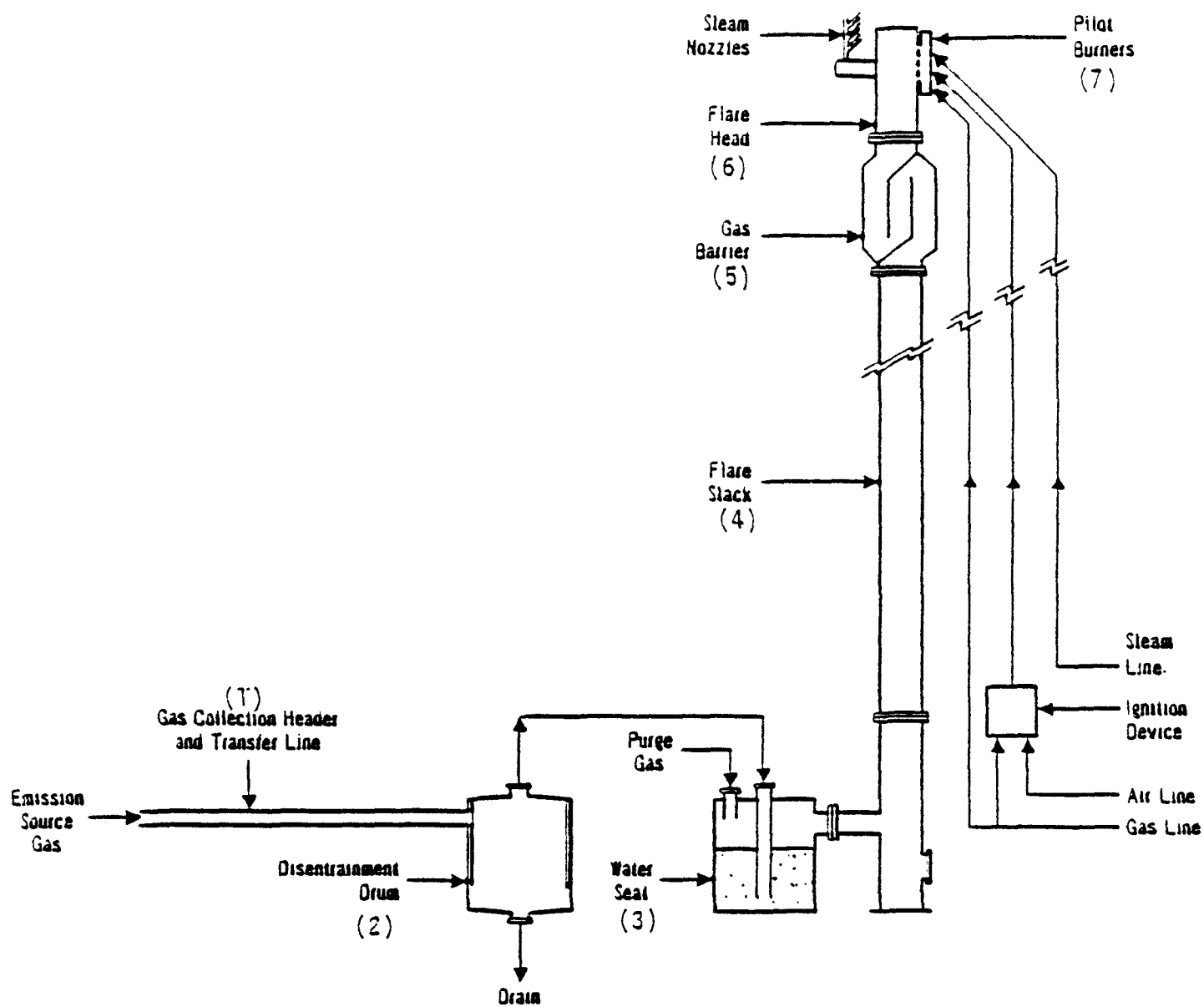
Figure 4-1. Steam Assisted Elevated Flare System
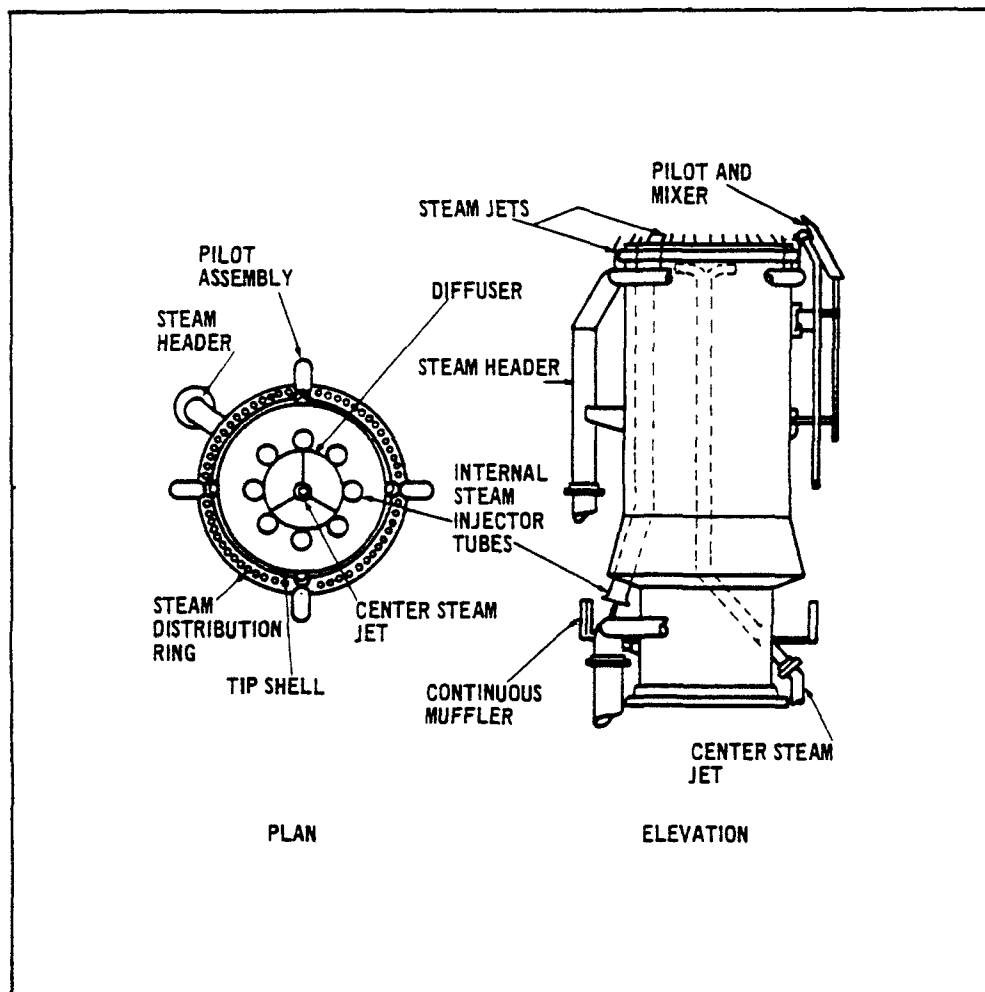
Figure 4-2. Steam Injection Flare Tip

Purge gas (N_2, CO_2, or natural gas) also helps to prevent flashback in the flare stack caused by low offgas flow. The total volumetric flow to the flame must be carefully controlled to prevent low flow flashback problems and to avoid a detached flame (a space between the stack and flame with incomplete combustion) caused by an excessively high flowrate. A gas barrier or a stack seal is sometimes used just below the flare head to impede the flow of air into the flare gas network.

The VOC stream enters at the base of the flame where it is heated by already burning fuel and pilot burners at the flare tip. Fuel flows into the combustion zone where the exterior of the microscopic gas pockets is oxidized. The rate of reaction is limited by the mixing of the fuel and oxygen from the air. If the gas pocket has sufficient oxygen and residence time in the flame zone it can be completely burned. A diffusion flame receives its combustion oxygen by diffusion of air into the flame from the surrounding atmosphere. The high volume of fuel flow in a flare requires more combustion air at a faster rate than simple gas diffusion can supply, so flare designers add steam injection nozzles to increase gas turbulence in the flame boundary zones, drawing in more combustion air and improving combustion efficiency. The steam injection promotes smokeless flare operation by minimizing the cracking reactions that form carbon. Significant disadvantages of steam usage are the increased noise and cost. The steam requirement depends on the composition of the gas flared, the steam velocity from the injection nozzle, and the tip diameter. Although some gases can be flared smokelessly without any steam, typically 0.15 to 0.5 kg of steam per kg of flare gas is required.

Steam injection is usually controlled manually with the operator observing the flare (either directly or on a television monitor) and adding steam as required to maintain smokeless operation. Several flare manufacturers offer devices which sense flare flame characteristics and adjust the steam flowrate automatically to maintain smokeless operation.

Some elevated flares use forced air instead of steam to provide the combustion air and mixing required for smokeless operation. These flares consist of two coaxial flow channels. The combustible gases flow in the center channel and the combustion air (provided by a fan in the

bottom of the flare stack) flows in the annulus. The principal advantage of air assisted flares is that expensive steam is not required. Air assist is rarely used on large flares because air flow is difficult to control when the gas flow is intermittent. About 600 J/sec (0.8 hp) of blower capacity is required for each 45 kg/hr (100 lb/hr) of gas flared (Klett and Galeski, 1976).[4]

Ground flares are usually enclosed and have multiple burner heads that are staged to operate based on the quantity of gas released to the flare. The energy of the gas itself (because of the high nozzle pressure drop) is usually adequate to provide the mixing necessary for smokeless operation and air or steam assist is not required. The fence or other enclosure reduces noise and light from the flare and provides some wind protection.

Ground flares are less numerous and have less capacity than elevated flares. Typically they are used to burn gas "continuously" while steam assisted elevated flares are used to dispose of large amounts of gas released in emergencies (Payne, 1982).[5]

4.1.1.1.1 Flare combustion efficiency. The flammability limits of the gases flared influence ignition stability and flame extinction (gases must be within their flammability limits to burn). When flammability limits are narrow, the interior of the flame may have insufficient air for the mixture to burn. Outside the flame, so much air may be induced that the flame is extinguished. Fuels with wide limits of flammability are therefore usually easier to burn (for instance, H_2 and acetylene). However, in spite of wide flammability limits, CO is difficult to burn because it has a low heating value and slow combustion kinetics.

The auto-ignition temperature of a fuel affects combustion because gas mixtures must be at high enough temperature and at the proper mixture strength to burn. A gas with a low auto-ignition temperature will ignite and burn more easily than a gas with a high auto-ignition temperature. Hydrogen and acetylene have low auto-ignition temperatures while CO has a high one.

The heating value of the fuel also affects the flame stability, emissions, and structure. A lower heating value fuel produces a cooler flame which does not favor combustion kinetics and also is more easily

extinguished. The lower flame temperature will also reduce buoyant forces, which reduces mixing (especially for large flares on the verge of smoking). For these reasons, VOC emissions from flares burning gases with low heat content may be higher than those from flares which burn high heat content gases.

Some fuels, also, have chemical differences (slow combustion kinetics) sufficient to affect the VOC emissions from flares. For instance, CO is difficult to ignite and burn, and so flares burning fuels with large amounts of CO may have greater VOC emissions than flares burning pure VOC.

The density of the gas flared also affects the structure and stability of the flame through the effect on buoyancy and mixing. The velocity in many flares is very low, and, therefore, most of the flame structure is developed through buoyant forces on the burning gas. Lighter gases thus tend to burn better, all else being equal. The density of the fuel also affects the minimum purge gas required to prevent flashback and the design of the burner tip.

Poor mixing at the flare tip or poor flare maintenance can cause smoking (particulate). Fuels with high carbon-to-hydrogen ratios (greater than 0.35) have a greater tendency to smoke and require better mixing if they are to be burned smokelessly.

The following review of flares and operating conditions summarizes five studies of flare combustion efficiency. Each study can be found in complete form in the docket.

Palmer (1972) experimented with a 1/2-inch ID flare head, the tip of which was located 4 feet from the ground.[6] Ethylene was flared at 15 to 76 m/sec (50 to 250 ft/sec) and $0.12-0.62 \times 10^6$ J/sec ($0.4-2.1 \times 10^6$ Btu/hr) at the exit. Helium was added to the ethylene as a tracer at 1 to 3 volume percent and the effect of steam injection was investigated in some experiments. Four sets of operating conditions were investigated; destruction efficiency was measured as greater than 99.9 percent for three sets and 97.8 percent for the fourth. The author questioned the validity of the 97.8 percent result due to possible sampling and analytical errors. He recommended further sampling and analytical techniques development before conducting further flare evaluations.

Siegel (1980) made the first comprehensive study of a commercial flare system.[7] He studied burning of refinery gas on a commercial flare head manufactured by Flaregas Company. The flare gases used consisted primarily of hydrogen (45.4 to 69.3 percent by volume) and light paraffins (methane to butane). Traces of H_2S were also present in some runs. The flare was operated with from 130 to 2,900 kilograms of fuel/hr (287 to 6,393 lb/hr), and the maximum heat release rate was approximately 68.9×10^6 J/sec (235×10^6 Btu/hr). Combustion efficiency and local burnout was determined for a total of 1,298 measurement points. Combustion efficiency was greater than 99 percent for 1,294 points and greater than 98 percent for all points except one, which had a 97 percent efficiency. The author attributed the 97 percent result to excessive steam addition.

Lee and Whipple (1981) studied a bench-scale propane flare.[8] The flare head was 2 inches in diameter with one 13/16-inch center hole surrounded by two rings of 16 1/8-inch holes, and two rings of 16 3/16-inch holes. This configuration had an open area of 57.1 percent. The velocity through the head was approximately 1 m/sec (3 ft/sec) and the heating rate was 0.09×10^6 J/sec (0.3×10^6 Btu/hr). The effects of steam and crosswind were not investigated in this study. Destruction efficiencies were greater than 99 percent for three of four tests. A 97.8 percent result was obtained in the only test where the probe was located off the centerline of the flame. The author did not believe that this probe location provided a valid gas sample for analysis.

Howes, et al. (1981) studied two commercial flare heads at John Zink's flare test facility.[9] The primary purpose of this test (which was sponsored by the EPA) was to develop a flare testing procedure. The commercial flare heads were an LH air assisted head and an LRGO (Linear Relief Gas Oxidizer) head manufactured by John Zink Company. The LH flare burned 1,045 kg/hr (2,300 lb/hr) of commercial propane. The exit gas velocity based on the pipe diameter was 8.2 m/sec (27 ft/sec) and the firing rate was 12.9×10^6 J/sec (44×10^6 Btu/hr). The LRGO flare consisted of three burner heads 1 meter (3 feet) apart. The three-burner combination fired 1,909 kg/hr (4,200 lbs/hr) of natural gas. This corresponds to a firing rate of 24.5×10^6 J/sec (83.7×10^6 Btu/hr). Steam was not used for either flare, but the LH flare head was in some

trials assisted by a forced draft fan. In four of five tests, combustion efficiency was determined to be greater than 99 percent when sampling height was sufficient to ensure that the combustion process was complete. One test resulted in combustion efficiency as low as 92.6 percent when the flare was operated under smoking conditions.

An excellent detailed review of the above four studies was done by Payne, et al. in January 1982,[10] and a fifth study [McDaniel, et al. (1982)] determined the influence on flare performance of mixing, heat content, and gas flow velocity.[11] A summary of these studies is given in Table 4-1. Steam assisted and air assisted flares were tested at the John Zink facility using the procedures developed by Howes. The test was sponsored by the Chemical Manufacturers Association (CMA) with the cooperation and support of EPA. All of the tests were with an 80 percent propylene, 20 percent propane mixture diluted as required with nitrogen to give different Btu/scf values. This was the first work which determined flare efficiencies at a variety of "nonideal" conditions where lower efficiencies had been predicted. All previous tests were of flares which burned gases that were very easily combustible and did not tend to soot. This was also the first test which used the sampling and chemical analysis methods developed for the EPA by Howes.

The steam assisted flare was tested with exit flow velocities up to 19 m/sec (62.5 ft/sec), with heat contents of 11-81 x 10^6 J/scm (294 to 2,183 Btu/scf) and with steam-to-gas (weight) ratios varying from zero (no steam) to 6.86:1. Flares without assist were tested down to 7.2 x 10^6 J/scm (192 Btu/scf). All of these tests, except for those with very high steam-to-gas ratios, showed combustion efficiencies of over 98 percent. Flares with high steam-to-gas ratios (about 10 times more steam than required for smokeless operation) had lower efficiencies (69 to 82 percent) when combusting 81 x 10^6 J/scm (2,183 Btu/scf) gas.

The air assisted flare was tested with flow velocities up to 66 m/sec (218 ft/sec) and with Btu contents of 3.1-81 x 10^6 J/scm (83 to 2,183 Btu/scf). Tests at 10.5 x 10^6 J/scm (282 Btu/scf) and above gave over 98 percent efficiency. Tests at 6.3 x 10^6 J/scm (168 Btu/scf) gave 55 percent efficiency.

Table 4-1. FLARE EMISSION STUDIES

Investigator	Sponsor	Flare Tip Design	Flared Gas	Throughput (10^6 Btu/hr)	Flare Efficiency (percent)	Reference
Palmer (1972)	E.I. du Pont	0.5" diameter	Ethylene	0.4 - 2.1	97.8	6
Lee & Whipple (1981)	Union Carbide	Discrete Holes in 2" diameter cap	Propane	0.3	96 - 100	8
Siegel (1980)	Ph.D. Dissertation University of Karlsruhe	Commercial Design (27.6" dia. steam)	50% H_2 plus light hydrocarbons	49 - 178	97 - 99	7
Howes et al. (1981)	EPA	Commercial Design (6" dia. air assist)	Propane	44	92.6 - 100	9
		Commercial Design II.P. (3 tips of 4" dia.)	Natural Gas	28 (per tip)	99	
McDaniel et al. (1982)	CMA-EPA	Commercial Design (6"dia. air assist)	Propylene	0.003 - 58	59.6 - 99.9	11
		Commercial Design (4" dia. steam assist)		0.009 - 57	83 - 99.9	

After consideration of the results of these five tests, EPA has concluded that 98 percent combustion efficiency can be achieved by steam assisted flares if these flares are operated with combustion gas heat contents and exit flow velocities within ranges determined by the tests. Steam flares obtain 98 percent combustion efficiency combusting gases with heat contents over 11.2×10^6 J/scm (300 Btu/scf) at velocities of less than 18.3 m/sec (60 ft/sec). Steam flares are not normally operated at the very high steam-to-gas ratios that resulted in low efficiency in some tests because steam is expensive and operators make every effort to keep steam consumption low. Flares with high steam rates are also noisy and may be a neighborhood nuisance. Nonassisted pipe flares obtain 98 percent efficiency with heat contents over 200 Btu/scf at velocities of less than 18.3 m/sec. Air assisted flares obtain 98 percent efficiency with heat contents over 11.2×10^6 J/scm and at velocities not exceeding that determined by the following formula:

$$v \text{ (ft/sec)} = 28.75 + 0.867 \text{ HC, where}$$

$$v = \text{maximum gas velocity in ft/sec, standard conditions,}$$

$$HC = \text{heat content of the combusted gas in Btu/scf.}$$

The EPA has a program underway to determine more exactly the efficiencies of flares used in the petroleum/SOCMI industries and a flare test facility has been constructed. The combustion efficiency of four flares (3.8 to 30.5 cm dia.) will be determined and the effect on efficiency of flare operating parameters, weather factors, and fuel composition will be established. The efficiency of larger flares will be estimated by scaling. A final report of this work should be available in the summer of 1983.

4.1.1.1.2 Applicability. A typical polymer plant produces several hundred million pounds of product per year. Because of this huge throughput, the VOC emissions that result from frequent process upsets are also large. Flares are used mainly to minimize the safety risk caused by emergency blowdowns where large volumes of gases with variable composition must be released from the plant almost instantaneously. Flares are ideal for this service and their reliability, as measured by absence of explosions and plant fires, has been demonstrated repeatedly. Flares

also effectively eliminate the hazard of process streams which, during startup or shutdown, would otherwise vent to the atmosphere and could also create an explosion or toxic hazard. Finally, flares are also used to burn co-products or by-products of a process that has too little value to reclaim, and thus would otherwise be a continuous VOC emission during normal operation of the unit. This practice, once the norm, has abated considerably during the past decade as the value of VOC stream components has dramatically increased.

4.1.1.2 Thermal Incinerators. The design and operation of thermal incinerators are influenced by operating temperature, residence time, desired VOC destruction efficiency, offgas characteristics, and combustion air. Operating temperatures may typically be between 650°C (1,200°F) and 980°C (1,800°F) with a residence time of 0.3 to 1.0 second.[12] The temperature theoretically required to achieve complete oxidation depends on the nature of the chemical involved and can be determined from kinetic rate studies.[13] The design of the combustion chamber should maximize the mixing of the VOC stream, combustion air, and hot combustion products from the burner. This helps ensure that the VOC contacts sufficient oxygen while at combustion temperature, for maximum combustion efficiency.

The heating value and water content of the waste gas feed and the excess combustion air delivered to the incinerator also affect incinerator design and operation. Heating value is a measure of the heat produced by the combustion of the VOC in the waste gas. Gases with a heating value less than 1,860 kJ/scm (50 Btu/scf) will not burn and require auxiliary fuel to maintain combustion. Auxiliary fuel requirements can be reduced and sometimes even eliminated by transferring heat from the exhaust gas to the inlet gas. Offgases with a heating value between 1,860 kJ/scm and 3,720 kJ/scm (100 Btu/scf) can support combustion but require some auxiliary fuel to ensure flame stability, i.e., avoid a flameout. Theoretically, offgases with a heating value above 3,720 kJ/scm possess enough heat content to not require auxiliary fuel (although practical experience has shown that 5,580 kJ/scm (150 Btu/scf) and above may be necessary,[14] and these gases may be used as a fuel gas or boiler feed gas.[15] A thermal incinerator handling offgas streams with varying heating values and moisture content requires periodic adjustment to

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maintain the proper chamber temperatures and operating efficiency. Increases in heat content reduce auxiliary fuel requirements, whereas increases in water content can substantially increase fuel requirements.

Incinerators are always operated with excess air to ensure a sufficient supply of oxygen. The amount of excess air used varies with the fuel and burner type but should be kept as low as possible. Using too much excess air wastes fuel because this air must be raised to the combustion temperature but does not contribute any heat by participating in the oxidation reaction. Large amounts of excess air also increase the flue gas volume and may cause an operator to invest in a larger system than required.

A thermal incinerator usually contains a refractory-lined chamber (which may vary in cross-sectional size along its length) containing a burner at one end. Because of the risk to the refractory, incinerators are neither brought quickly up to nor cooled down quickly from operating temperatures. They require a fairly constant fuel input to maintain combustion temperature. A diagram of a thermal incinerator using discrete burners is shown in Figure 4-2. (Numbers in parentheses following the mention of equipment parts or streams denote the numbered items on the referenced figures.) Discrete dual fuel burners (1) and inlets for the offgas (2) and combustion air (3) are arranged in a premixing chamber (4) to thoroughly mix the hot products from the burners with the offgas air streams. The mixture of hot reacting gases then passes into the main combustion chamber (5). This section is sized to allow the mixture enough time at the elevated temperature for the oxidation reaction to be completed (residence times of 0.3 to 1 second are common). Energy can then be recovered from the hot flue gases with the installation of a heat recovery section (6). Preheating of combustion air or the process waste offgas fed to the incinerator by the incinerator exhaust gases will reduce auxiliary fuel usage. In some instances, the incinerator exhaust gas may be used in a waste heat boiler to generate steam. Insurance regulations require that if the process waste offgas is preheated, the VOC concentration must be maintained below 25 percent of the lower explosive limit (LEL) to minimize explosive hazards.[16]

Thermal incinerators designed specifically for VOC incineration with natural gas as the auxiliary fuel may use a grid-type (distributed) gas burner similar to that shown in Figure 4-3. The tiny gas flame jets (1) on the grid surface (2) ignite the vapors as they pass through the grid. The grid acts as a baffle for mixing the gases entering the chamber (3). This arrangement ensures burning of all vapors using less fuel and a shorter burning length in the duct than conventional forward flame burners. Overall, this system makes possible a shorter reaction chamber while maintaining high efficiency.[17]

Thermal incinerators used to burn halogenated VOC's often use additional equipment to remove the corrosive combustion products. The flue gases are quenched to lower their temperature and routed through absorption equipment such as spray towers or liquid jet scrubbers to remove the corrosive gases from the exhaust.[18]

Packaged, single unit thermal incinerators are available in many sizes to control streams with flowrates from a few hundred scfm up to about 50,000 scfm. A typical thermal incinerator built to handle a VOC waste stream of 850 scm/min (30,000 scfm) at a temperature of 870°C (1,600°F) with 0.75 second residence time would probably be a refractory-lined cylinder. With the typical ratio of flue gas to waste gas of about 2.2, the chamber volume necessary to provide for 0.75 second residence time at 870°C (1,600°F) would be about 100 m^3 (3,500 ft^3). If the ratio of the chamber length to the diameter is 2, and if a 30.5 cm (1 ft) wall thickness is allowed, the thermal incinerator would measure 8.3 m (27 ft) long by 4.6 m (15 ft) wide, exclusive of heat exchangers and exhaust equipment.

4.1.1.2.1 VOC destruction efficiency. The destruction efficiency of an incinerator can be affected by variations in chamber temperature, residence time, inlet concentration, compound type, and flow regime (mixing). Of these, chamber temperature, residence time, and flow regime are the most important.

When the temperature exceeds 700°C (1,290°F), the oxidation reaction rate is much faster than the rate at which mixing can take place, so VOC destruction becomes more dependent upon the fluid mechanics within the combustion chamber.[19] Variations in inlet concentration also affect the
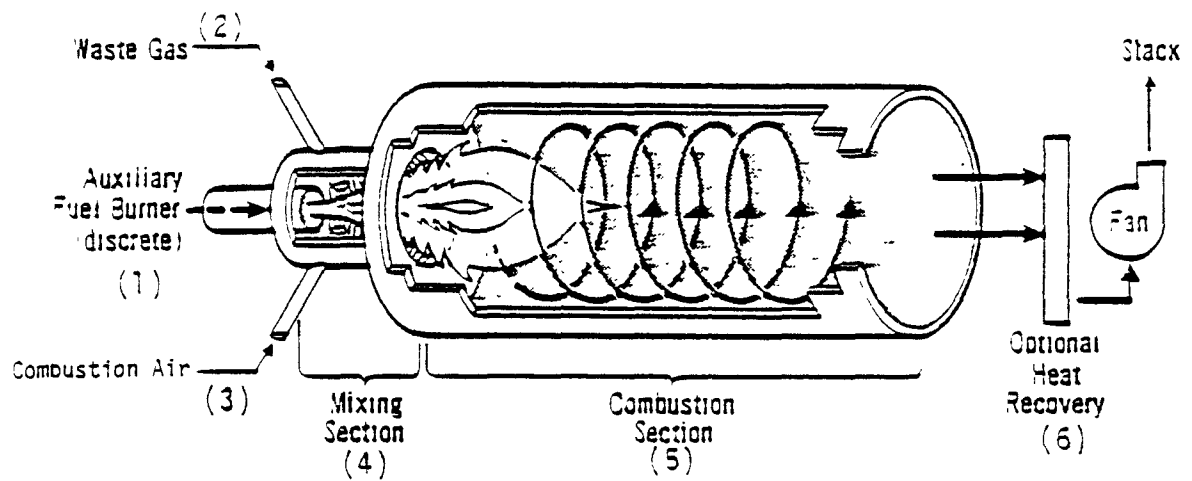
Figure 4-2. Discrete Burner Thermal Incinerator
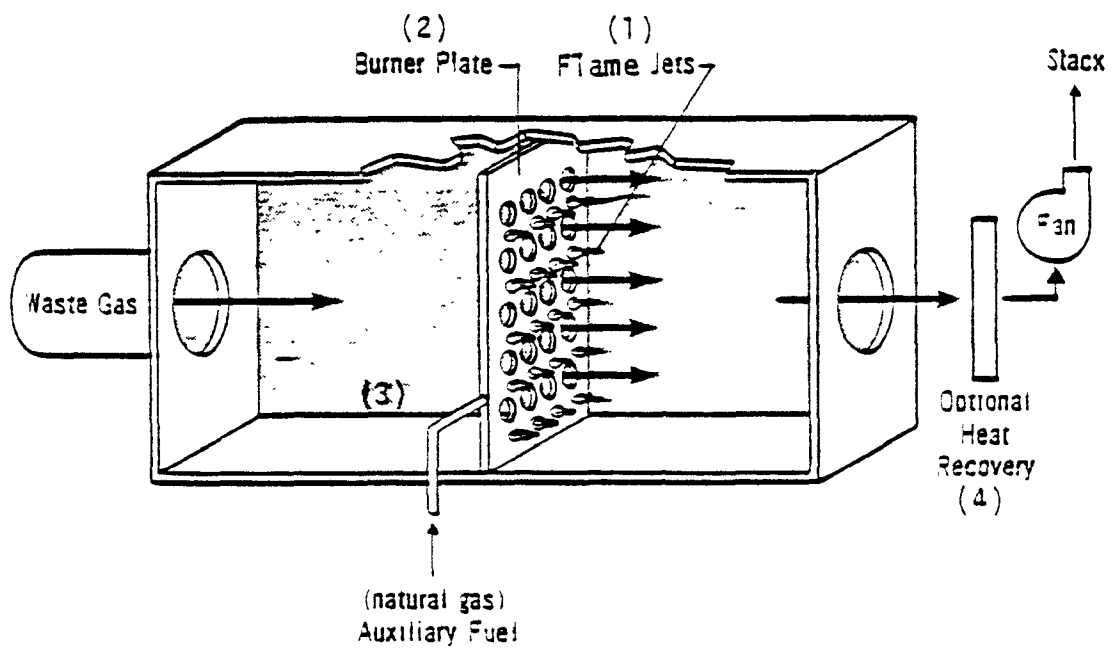
Figure 4-3. Distributed Burner Thermal Incinerator

VOC destruction efficiency achievable; kinetics calculations describing the combustion reaction mechanisms indicate much slower reaction rates at very low compound concentrations. Therefore, at low VOC concentration, a greater residence time is required to achieve a high combustion efficiency.

Test results show that a VOC control efficiency of 98 percent can be achieved consistently for many VOC compounds by well-designed units and can be met under a variety of operating conditions:[20,21] combustion chamber temperatures ranging from 700 to 1,300°C (1,300 to 2,370°F) and residence times of 0.5 to 1.5 seconds. The test results covered the following VOC compounds: C_1 to C_5 alkanes and olefins, aromatics (benzene, toluene, and xylene), oxygenated compounds (methyl ethyl ketone and isopropanol), chlorinated organics (vinyl chloride), and nitrogen-containing species (acrylonitrile and ethylamines). At chamber temperatures below 760°C (1,400°F), a wide range of efficiencies were reported for several VOC compounds. This information, used in conjunction with kinetics calculations, indicates overall that the minimum combustion chamber parameters for ensuring at least a 98 percent VOC destruction efficiency are a combustion temperature of 870°C (1,600°F), and a residence time at combustion temperature of 0.75 second. A thermal incinerator designed to produce these conditions in the combustion chamber should be capable of high destruction efficiency for almost any VOC even at low inlet concentrations.

Based on the studies of thermal incinerator efficiency, auxiliary fuel use, and costs, EPA has concluded that 98 percent VOC destruction, or a 20 parts per million by volume (ppmv) compound exit concentration (whichever is less stringent), is the highest reasonable control level achievable by all new incinerators considering current technology.[22] This estimate is predicated on thermal incinerators operated at 870°C (1,600°F) with a 0.75 second residence time.

4.1.1.2.2 <u>Applicability</u>. Thermal incinerators can be used to control a wide variety of continuous waste gas streams (one has been observed in a polypropylene plant[23]). They can be used to destroy VOC in streams with any concentration and type of VOC. Although they accommodate minor fluctuations in flow, incinerators are not well suited to streams with intermittent flow because of the large auxiliary fuel

requirements during periods when there is no fuel contribution from the waste gas, yet the chamber temperature must be maintained to protect the incinerator lining.

For extremely dilute streams, a catalytic incinerator might be a favorable choice over a thermal incinerator if supplemental fuel requirements are of principal concern. However, most waste gas streams in this industry contain enough heating value to support a flame by itself on a properly designed flame burner. Such streams can be considered for use as fuel gas or boiler feed gas, from which the recovery of energy may more than compensate for a thermal incinerator's capital costs.

4.1.1.3 <u>Catalytic Incinerators</u>. The control principles and equipment used in catalytic incineration are similar to those employed in conventional thermal incineration. The VOC-containing waste gas stream is heated to an appropriate reaction temperature and then oxidation is carried out at active sites on the surface of a solid catalyst. The catalyst increases the rate of oxidation, allowing the reaction to occur at a lower temperature than in thermal incineration. This technique may offer advantages over thermal incineration in auxiliary fuel savings where low VOC content makes large fuel usage necessary. Catalytic incinerators also may produce less NO_x because of lower combustion temperatures and smaller excess air requirements.

Combustion catalysts are made by depositing platinum or platinum alloys, copper oxide, chromium, or cobalt on an inert substrate, which is suitably shaped to fit the mechanical design of the incinerator. The operating temperature of the catalyst is usually from 315°C (600°F) to 650°C (1,200°F). Combustion may not occur below 315°C and temperatures higher than 650°C may shorten the catalyst life or even evaporate catalyst from the support substrate.[24] Accumulation of particulate matter, condensed VOC's, or polymerized hydrocarbons on the catalyst can block the active sites and reduce its effectiveness. Catalysts can also be contaminated and deactivated by compounds containing sulphur, bismuth, phosphorous, arsenic, antimony, mercury, lead, zinc, tin, or halogens. If the catalyst is so "poisoned," VOC's will pass through unreacted or only partially oxidized. Catalytic incinerators can operate efficiently treating offgas streams with VOC concentrations below the lower explosive

limit. This is a distinct advantage over thermal incinerators which would in this situation require auxiliary fuel.

A schematic of a catalytic incinerator unit is shown in Figure 4-4. During operation, the waste gases (1) first enter the mixing chamber (also called the preheat zone) (3) where they are heated by contact with the hot combustion products of a burner (2). The mixing chamber temperature may vary as a function of the composition and type of contaminants to be oxidized, but will generally operate in the range of 343°C (650°F) to 593°C (1,100°F).[25] The heated mixture then passes through the catalyst bed (4) where oxygen and VOC's diffuse to the catalyst and are adsorbed on its surface. The oxidation reaction takes place at these "active sites." Reaction products desorb from the active sites and diffuse back into the waste gas. As with the exhaust gases from thermal incinerators, the products of combustion leaving the bed may be used in a waste heat recovery device (5) before being exhausted to the atmosphere.

4.1.1.3.1 VOC destruction efficiency. The destruction efficiency of catalytic incinerators is a function of many variables, including type of catalyst, its surface area, volume, and pore size distribution, gas composition, uniformity of flow through the catalyst bed, oxygen concentration, and temperature in the unit.[26,27]

The efficiency of a catalytic incinerator will deteriorate over time, necessitating periodic replacement of the catalyst. The replacement time varies widely, depending on the service of the unit, from less than 1 year up to 10 years,[12] with an average life between 3 and 5 years.[28]

A 1980 study by Engelhard Industries for the EPA involved testing of both pilot and full-scale catalytic incineration systems. The full-scale unit installed on a formaldehyde plant achieved control efficiencies ranging from 97.9 to 98.5 percent. These efficiencies represent overall control levels for carbon monoxide, methanol, dimethyl ether, and formaldehyde. Measurements indicated the ability of the system to control at this level consistently over a 1-year period. No trend in the data points gave indication of a maximum catalyst life.[29]

4.1.1.3.2 Applicability. A catalytic incinerator is best applied to a continuous stream that is (1) low in VOC (higher VOC concentrations

Stack

Fan

Optional
Heat Recovery
(5)

Catalyst Bed (4)

Mixing Chamber (3)

Auxiliary
Fuel Burners
(2)

Waste Gas
(1)
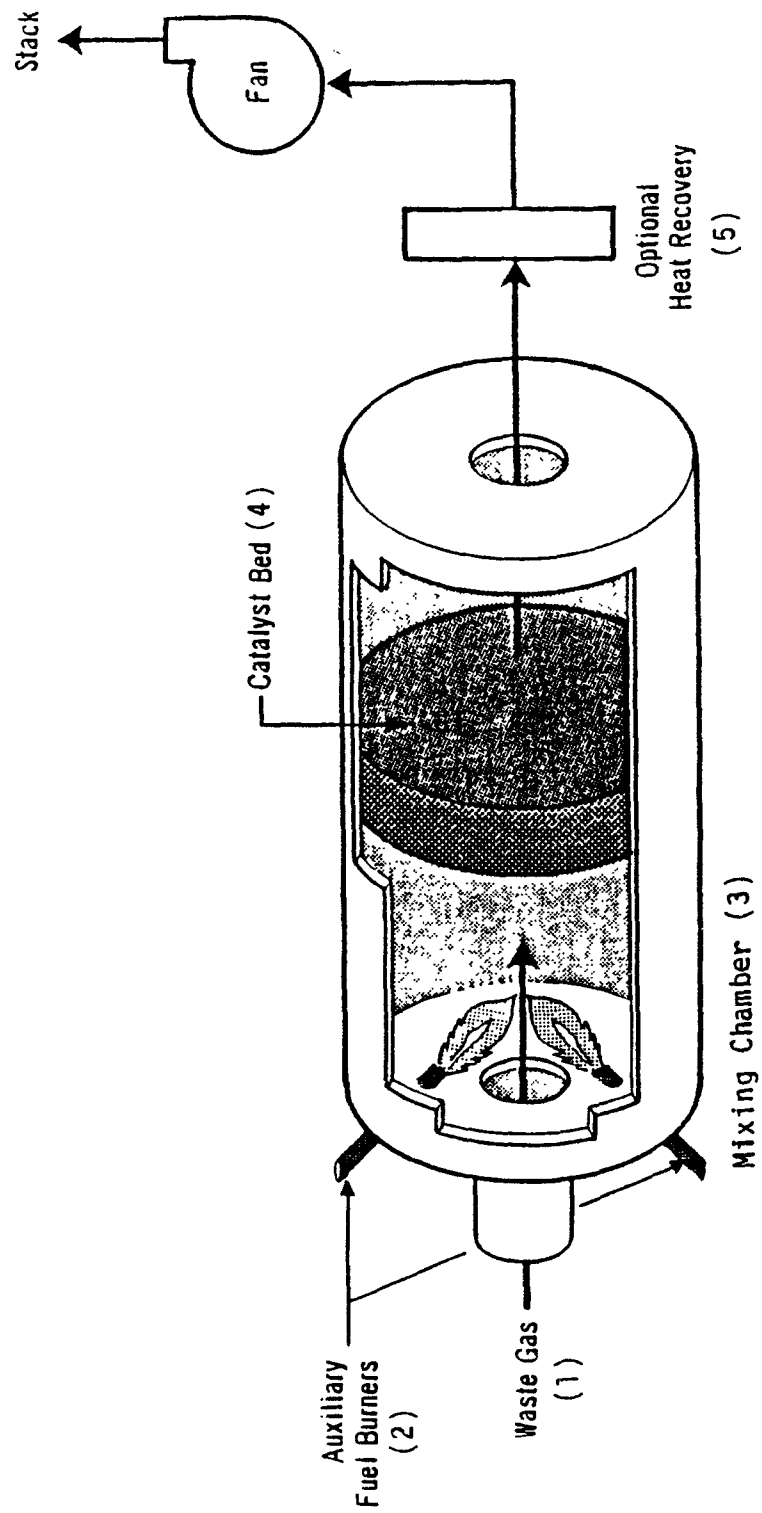
Figure 4-4. Catalytic Incinerator

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lead to higher catalyst temperatures, which can seriously damage the catalyst activity and possibly create fire hazards) and (2) free from solid particles and catalyst "poisons." A catalytic incinerator in many situations may be favored over a thermal incinerator because it can destroy the VOC at a lower temperature and, therefore, use less fuel. However, since most of the streams involved in the polymers and resins industry are high enough in heating value to self-combust without using auxiliary fuel, virtually no advantage is achieved by using a catalytic unit and their applicability in this industry is very limited.

4.1.1.4 <u>Industrial Boilers.</u> Fireboxes of boilers and fired heaters can be used, under proper conditions, to incinerate waste streams that contain VOC's. Combustible contaminants, including smoke, organic vapors, and gases can be converted essentially to carbon dioxide and water in boiler fireboxes. As the primary purpose of the boiler is to generate steam, all aspects of operation must be thoroughly evaluated before this method of air pollution control can be used. Any breakdown in the boiler can result in expensive process downtime. Consequently, the risk of shutdown should be kept small and only streams that do not threaten boiler performance should be introduced.

For the satisfactory use of boilers as a control device, there are several prerequisites. Generally, the burner must be modified, the boiler must operate continuously and concurrently with the pollution source, the contaminants must be completely combustible, and the products of combustion must not corrode the materials used to construct the boiler. Corrosive VOC compounds can be combusted in a boiler, but special attention must be given to operate above the dew point of the flue gases. If these gases are allowed to condense, severe corrosion problems will occur. Further, the volumetric flowrate of low VOC concentration emission streams must be taken into consideration because they can reduce thermal efficiencies in the same way as excess combustion air does. The pressure drop caused by additional products of combustion should not exceed the draft provided by boiler auxiliaries. Boiler life, efficiency, and capacity can be affected by the presence of contaminants in the VOC emission streams. Halogens, for example, would be devastating to the life of boiler tubes. Finally, a personnel safety

hazard may occur if coal-fired boilers that are not pulverized coal-fired are used to destroy organic waste. Any interruption in the air supply to these types of boilers would release into the boiler house combustion vapors and any hazardous or toxic substances that may have been injected.[30] Great care, therefore, must be exercised in selecting this mode of pollution control.

The large majority of industrial boilers are of water tube design. Water, circulated through the tubes, absorbs the heat of combustion. Drums store the superheated water from which steam is directed to external heat exchangers for use as process steam. Boilers typically operate at combustion chamber temperatures above 1,650°C (3,000°F) with a residence time of about 1 second.[31]

Both forced and natural draft burners, designed to thoroughly mix the incoming fuel and combustion air, may be used. After ignition, the mixture of hot reacting gases passes through the furnace section that is sized to allow the oxidation reaction to reach completion and to minimize abrasion on the banks of the water tubes. Energy transfer from the hot flue gases to form steam can attain greater than 85 percent efficiency. Additional energy can be recovered from the hot exhaust gases by installation of a gas-gas heat exchanger to preheat combustion air.

Boilers designed specifically for use as a VOC control device typically use discrete or vortex burners, depending on the heating value of the vent stream. For vent streams with heating values between 1,100 kJ/scm (300 Btu/scf) and 1,850 kJ/scm (500 Btu/scf), a discrete burner would be best suited.[31] Streams with lower heating values would probably require vortex burners to ensure the desired VOC destruction.

4.1.1.4.1 VOC destruction efficiency. VOC destruction efficiency achievable by boilers depends on the same factors that affect any combustion technique. Since boiler furnaces typically operate at higher peak temperatures and with longer combustion residence times than thermal incinerators, the VOC destruction efficiency usually would be expected to match or exceed the 98 percent efficiency demonstrated in incinerators.

4.1.1.4.2 Applicability. Use of a boiler for VOC emission control in the polymers and resins industry is uncommon. Despite the potential problems, boilers are being used in at least two polypropylene plants[32] and a high-density polyethylene plant.[33] The polypropylene plants

supplement boiler fuel with waste gas that otherwise would be flared. The high density polyethylene plant sends the dehydrator regeneration gas (a mixture of natural gas and nitrogen) and a degassing stream from the recycle diluent step (mostly ethylene) to steam-generating boilers as a fuel.

A boiler would be used as a control device only if the process generated its own steam or the fuel value of the waste gas was sufficient to make the process a net exporter of steam. Whenever either condition exists, installation of a boiler is an excellent control measure that provides greater than 98 percent VOC destruction and very efficient recovery of the heat of combustion of the waste gas.

4.1.2 Control by Recovery Techniques

The three major recovery devices are condensers, adsorbers, and absorbers. These devices permit many organic materials to be recovered and, in some cases, reused in the process.

Condensers are widely used for recovering organics from both continuous and intermittent rich by-product streams in polystyrene manufacturing processes. The VOC is mainly styrene which is easily condensed because of its relatively high condensation temperature. The ease of styrene recovery and the ability of a condenser to handle an intermittent stream makes it a desirable control technology for all process VOC emissions in the polystyrene industry. Another application of condensers occurs in polyethylene manufacturing processes. In the DMT process, emissions from the methanol recovery section are minimized through the use of condensers. In both the DMT and TPA polyethylene processes, ethylene glycol (EG) is often recovered using an EG spray condenser, or with a distillation column incorporating a reflux condenser. Condensers may also be used in series with other air pollution control systems. A condenser located upstream of an incinerator, adsorber, or absorber will reduce the VOC load entering the downstream control device. The downstream device will abate most of the VOC that passes through the condenser.

Adsorbers are used on gas streams which contain relatively low VOC concentrations. Concentrations are usually well below the lower explosive limit in order to guard against overheating of the adsorbent bed.

Adsorbers are often neither suitable nor the most efficient means of control for the higher VOC concentration streams characteristic of the polymers and resins industry.

Absorbers, which use low volatility liquids as absorbents, are another control option. Their use is generally limited to applications in which the spent absorbent can be used directly in a process, since desorption of the VOC from the absorbent is often prohibitively expensive.

Recovery techniques either condense the organic or contact the VOC-containing gas stream with an appropriate liquid or solid. Gases containing only one or two organic gases are easier to process by recovery techniques than multi-component mixtures. The presence of inert or immiscible components in the waste gas mixture complicates recovery techniques.

4.1.2.1 <u>Condensers</u>. Condensation devices transfer thermal energy from a hot vapor to a cooling medium, causing the vapor to condense. Condenser design thus typically requires knowledge of both heat and mass transfer processes. Heat may be transferred by any combination of three modes: conduction, convection, or radiation.

The design of a condenser is significantly affected by the number and nature of components present in the vapor stream. The entering gases may consist of a single condensable component or any number of gaseous components which may or may not all be condensable or miscible with one another. Example gas streams found in the polystyrene industry may consist of a single condensable component (styrene); a mixture of condensable and noncondensable components (styrene and air); a mixture of condensable, but immiscible, components (styrene and steam); or a mixture of condensable, but immiscible, components with a noncondensable component (styrene, steam, and air).

Condensers are designed and sized using the principles or thermodynamics. At a fixed pressure, a pure component will condense isothermally at the saturation or equilibrium temperature, yielding a pure liquid condensate. A vapor mixture, however, does not have a single condensate temperature. As the temperature drops, condensation progresses, and the composition, temperature, enthalpy, and flowrate of both the remaining vapor and the condensate will change. These changes

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can be calculated from thermodynamics data, if it is assumed that the vapor and liquid condensate are in equilibrium. Variations in composition and temperature will affect most of the physical and transport properties which must be used in condenser design calculations. When these properties change, the calculations governing the heat transfer process are adjusted to accommodate these changes.

In a two-component vapor stream with one noncondensable component, condensation occurs when the partial pressure of the condensable component is equal to the component's vapor pressure. To separate the condensate from the gas at fixed pressure, the temperature of the vapor mixture must be reduced. The liquid will begin to appear when the vapor pressure of the condensable component becomes equal to its partial pressure, the "dew point." Condensation continues as the temperature is further reduced. The presence of a noncondensable component interferes with the condensation process, because a layer of noncondensable on the condensate acts as a heat transfer barrier.

Two types of condensers are employed: contact and surface. Contact, or direct, condensers cause the hot gas to mingle intimately with the cooling medium. Contact condensers usually operate by spraying a cool liquid directly into the gas stream. Contact condensers also may behave as scrubbers since they sometimes collect noncondensable vapors which are immiscible with the coolant. The direct contact between the vapor and the coolant limits the application of contact condensers since the spent coolant can present a secondary emission source or a wastewater treatment problem,[34] unless it is economically feasible to separate the two in a subsequent process.

Surface, or indirect, condensers are usually common shell-and-tube heat exchangers. The coolant usually flows through the tubes and the vapor condenses on the outside of the tubes. In some cases, however, it may be preferable to condense the vapor inside the tubes. The condensate forms a film on the cool tube and drains to storage.[35] The shell-and-tube condenser is the optimum configuration from the standpoint of mechanical integrity, range of allowable design pressures and temperatures, and versatility in type of service. Shell-and-tube condensers may be designed to safely handle pressures ranging from full vacuum to approximately

41.5 MPa (6,000 psig), and for temperatures in the cryogenic range up to approximately 1,100°C (2,000°F).[36] Surface condensers usually require more auxiliary equipment for operation (such as a cooling tower or a refrigeration system) but offer the advantage of recovering valuable VOC without contaminating the coolant, thereby minimizing waste disposal problems. The successively more volatile material returned from the condenser to the distillation column is termed "reflux," or overhead product. The heavier compounds removed at the bottom are often called column "bottoms."[37]

The major pieces of equipment used in a typical refrigerated surface condenser system are shown in Figure 4-5.[38] Refrigeration is often required to reduce the gas phase temperature sufficiently to achieve low outlet VOC concentrations. This type of system includes dehumidification equipment (1), a shell-and-tube heat exchanger (2), a refrigeration unit (3), recovery tank (4), and operating pumps (5). Heat transfer within a shell-and-tube condenser occurs through several material layers, including the condensate film, combined dirt and scale, the tube wall, and the coolant film. The choice of coolant used depends on the saturation temperature of the VOC stream. Chilled water can be used to cool down to 4°C (40°F), brines to -34°C (-30°F), and chlorofluorocarbons below -34°C (-30°F).[39] Temperatures as low as -62°C (-80°F) may be necessary to condense some VOC streams.[34]

4.1.2.1.1 <u>Condenser control efficiency</u>. VOC removal efficiency of a condenser is dependent upon the composition of the stream. Single component streams with a relatively high boiling point will easily condense, resulting in essentially 100 percent control efficiency. Thus, very high efficiencies would be expected for condensers controlling such streams in the polystyrene industry. Ethylene glycol spray condensers in PET polyester production reduce EG emissions to the atmosphere from 9.5 to 0.21 kg/Mg of product, or 97.8 percent (Tables 3-14 and 3-15). A less condensable component in the stream, however, will reduce the control efficiency because of the lower temperatures required for higher percentage removal. Water-cooled condensers sometimes cannot achieve a sufficiently low temperature to ensure high control efficiency. Better control, of course, is possible by use of a chilled coolant or even a
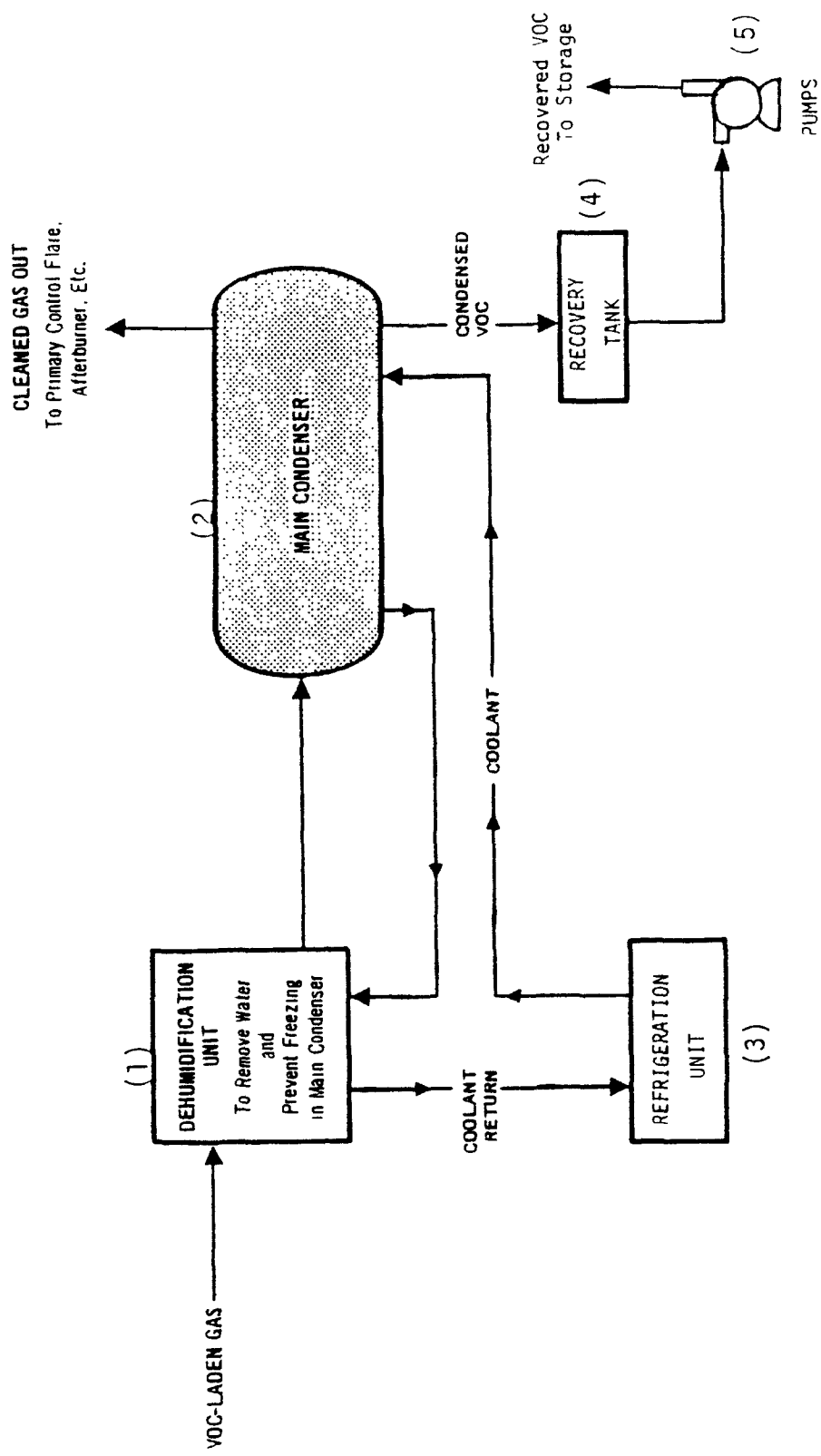
Figure 4-5. Condensation System

refrigerated condenser at an increased cost. Outlet concentrations for low boiling organics may be above 10,000 ppmv to 20,000 ppmv.[40]

4.1.2.1.2 Applicability of condensers. Water-cooled condensers are effective in reducing potential emissions of high boiling, easily condensable organics, and find broad application in the polystyrene manufacturing segment. Surface condensers are used to recover styrene from polystyrene vents and methanol and ethylene glycol from PET polyester esterifier vents. Spray condensers are highly efficient at recovering EG from PET polyester esterifier and reactor vents. Condensers cannot be used to condense low boiling organics such as ethylene or propylene in streams containing large quantities of inert gases such as nitrogen. Refrigerated condensers may be a viable option unless the stream contains water or heavy organics which would freeze and foul the condenser.

4.1.2.2 Adsorbers. Vapor-phase adsorption utilizes the ability of certain solids to preferentially adsorb and thereby concentrate certain components from a gaseous mixture onto their surfaces. The gas phase (adsorbate) is pumped through a packed bed of the solid phase (adsorbent) where selective components are captured on its surface by physical adsorption. The organic molecules are retained at the surface of the adsorbent by means of intermolecular or Van-der-Waals forces. The adsorbed organics can be readily removed and the adsorbent regenerated.

The most common industrial vapor-phase adsorption systems use beds of activated carbon. Carbons made from a variety of natural materials (wood, coal, nut shells, etc.) are marketed for their special adsorbent properties. The multiple bed system maintains at least one bed online while another is being regenerated. Most systems direct the vapor stream downward through a fixed carbon bed. Granular carbon is usually favored because it is not easily entrained in the exhaust stream.

Figure 4-6 is a schematic of a typical fixed bed, regenerative carbon adsorption system. The process offgases are filtered and cooled (1) to minimize bed contamination and maximize adsorption efficiency. The offgas is directed through the porous activated carbon bed (2) where adsorption of the organics progresses until the activated carbon bed is "saturated". When the bed is completely saturated, the organic will "breakthrough" the bed with the exhaust gas and the inlet gases must

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then be routed to an alternate bed. The saturated bed is then regenerated
to remove the adsorbed material.

Low-pressure steam (3) is usually used to heat the carbon bed
during the regeneration cycle, driving off the adsorbed organics, which
are usually recovered by condensing the vapors (4) and separating them
from the steam condensate by decanting or distillation (5). The adsorption/
regeneration cycle can be repeated numerous times, but eventually the
carbon loses its adsorption activity and must be replaced. The carbon
can sometimes be reactivated by recharring.

4.1.2.2.1 Adsorber control efficiency. The efficiency of an
adsorption unit depends on the properties of the carbon and the adsorbate,
and on the conditions under which they contact. Lower temperatures aid
the adsorption process, while higher temperatures reduce the adsorbent's
capacity.[41] Removal efficiencies of 95 to 99 percent are achieved by
well-designed and well-operated units.[42]

4.1.2.2.2 Applicability. Adsorbers effectively control streams
with dilute concentrations of organics. In fact, to prevent excessive
temperatures within the bed due to the heat of adsorption, inlet concen-
trations of organics are usually limited to about 0.5 to 1 percent.[40]
The maximum practical inlet concentration is about 1 percent, or
10,000 ppmv.[43] Higher concentrations are frequently handled by allowing
some condensate to remain from the regeneration process to remove the
heat generated during adsorption. Also, the inlet stream can be diluted
by use of a condenser or addition of air or nitrogen upstream of the
adsorber. If the organic is reactive or oxygen is present in the vent
stream, then additional precautions may be necessary to safeguard the
adsorption system.

Adsorbers can foul and hence are not very suitable for streams
containing fine particles or polymerizeable monomers. Both can contaminate
the beds and result in poor performance, or even introduce safety problems.
Because of their limitations in certain gas streams, carbon adsorbers
are not ideally suited for most of the emission streams encountered in
the polymers and resins industry.

4.1.2.3 Absorbers. Absorption is a gas-liquid mass transfer
operation in which a gas mixture is contacted with a liquid (solvent)
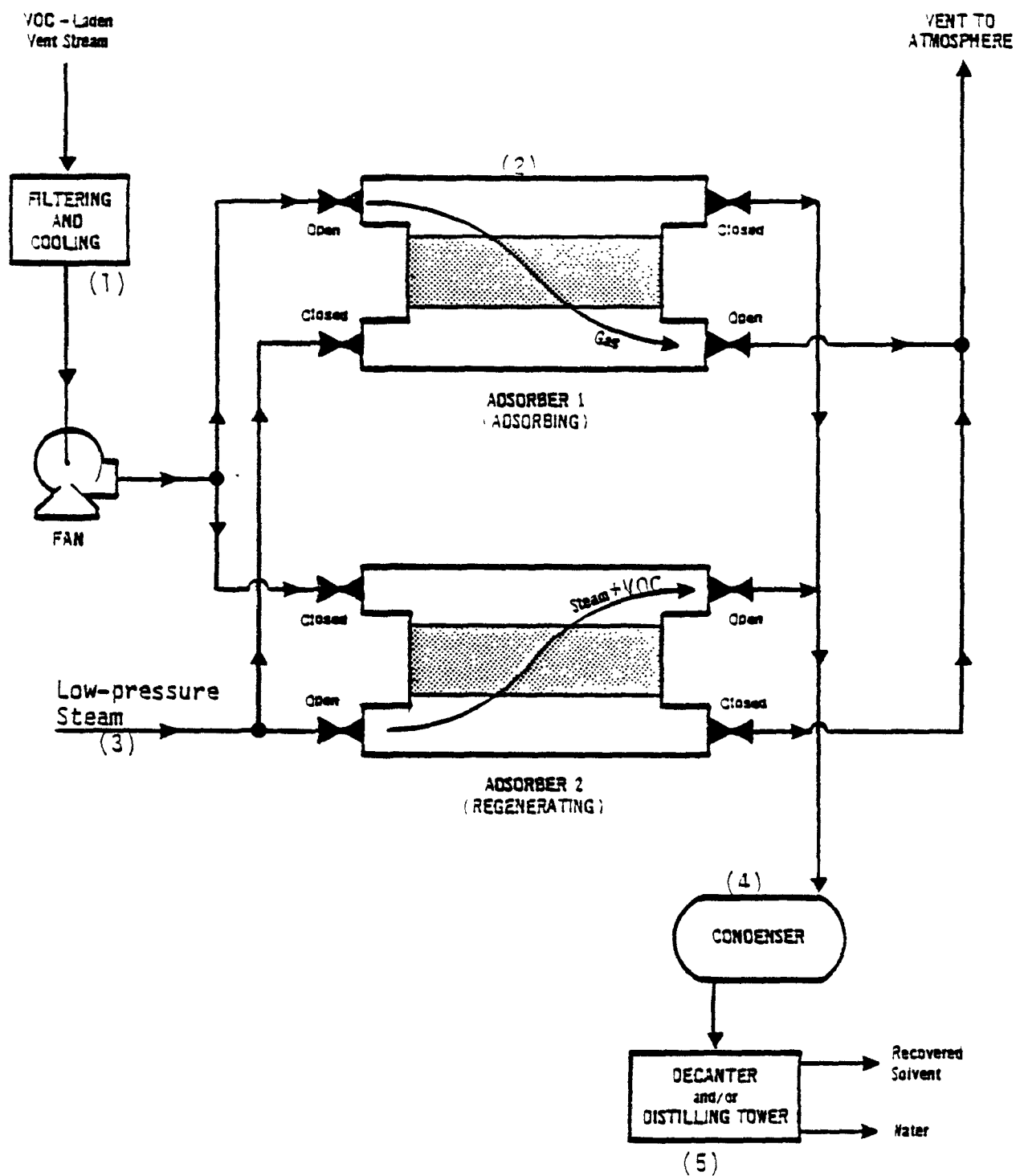
Figure 4-6. Two Stage Regenerative Adsorption System

for the purpose of preferentially dissolving one or more components (solutes) of the gas. Absorption may entail only the physical phenomenon of solution or may also involve chemical reaction of the solute with constituents of the solvent.[44]

For any given solvent, solute, and set of operating conditions, there exists a theoretical equilibrium ratio of solute concentration in the gas mixture to solute concentration in the solvent. The driving force for mass transfer in an operating absorption tower is related to the difference between the actual concentration ratio and this equilibrium ratio.[45] The solvents used are chosen for high solute (VOC) solubility and include liquids such as water, mineral oil, nonvolatile hydrocarbon oils, and aqueous solutions of oxidizing agents like sodium carbonate and sodium hydroxide.[46]

Devices based on absorption principles include spray towers, venturi scrubbers, packed columns, and plate columns. Spray towers and venturi scrubbers are generally restricted to particulate removal and control of high-solubility gases.[47] Most VOC control by gas absorption is by packed or plate columns. Packed columns are used mostly for handling corrosive materials, liquids with foaming or plugging tendencies, or where excessive pressure drops would result from the use of plate columns. They are less expensive than plate columns for small-scale or pilot plant operations where the column diameter is less than 0.6 m (2 ft). Plate columns are preferred for large-scale operations, where internal cooling is desired, or where low liquid flowrates would inadequately wet the packing.[48]

A schematic of a packed tower is shown in Figure 4-7. The gas is introduced at the bottom (1) and rises through the packing material (2). Solvent flows by gravity from the top of the column (3), countercurrent to the vapors, absorbing the solute from the gas phase and carrying the dissolved solute out of the tower (4). Cleaned gas exiting at the top is ready for release or final treatment such as incineration.

The major tower design parameters, column diameter and height, pressure drop, and liquid flowrate, are based on the specific surface area of the tower packing, the solubility and concentration of the components, and the quantity of gases to be treated.
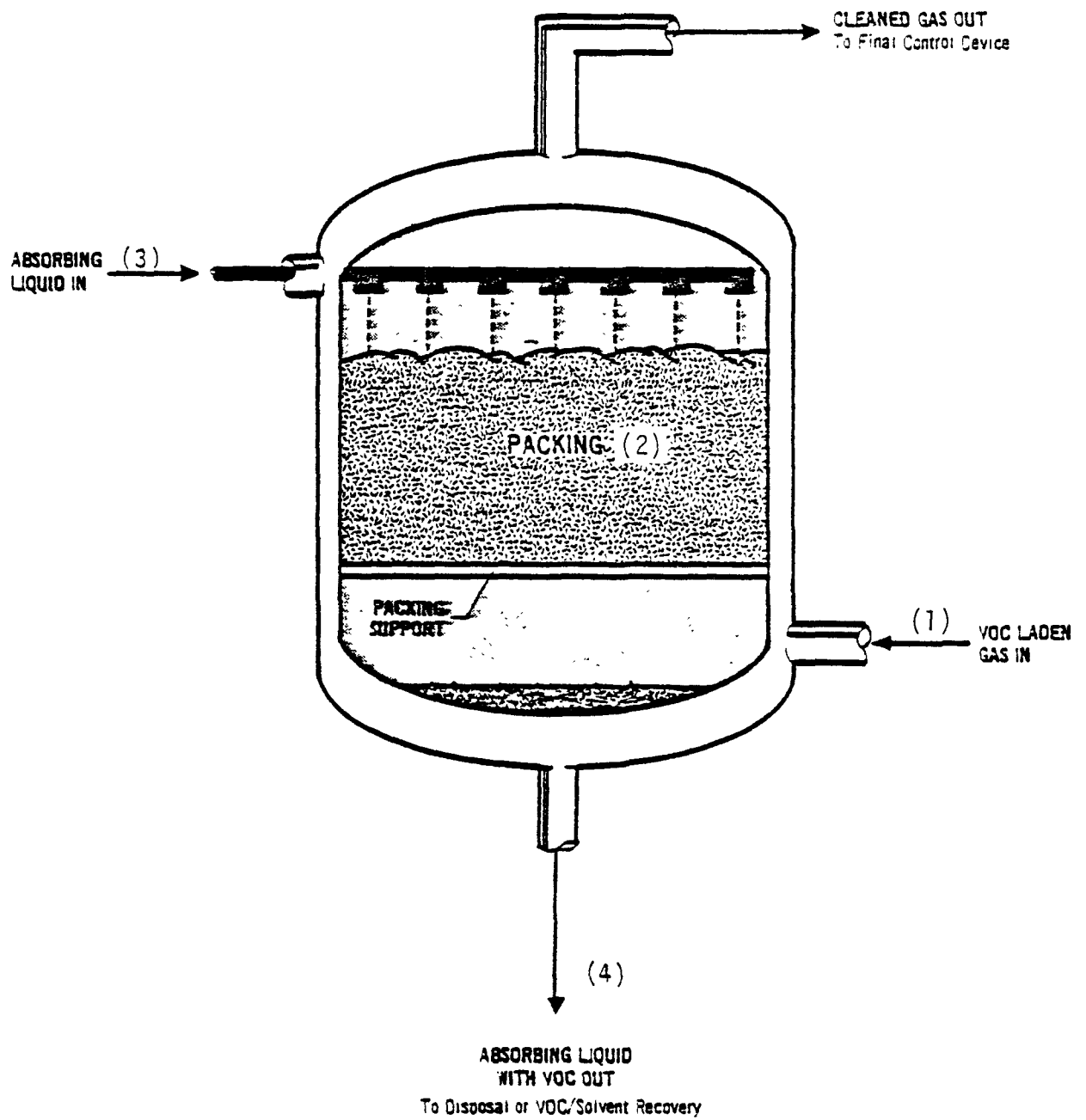
Figure 4-7. Packed Tower for Gas Absorption

4.1.2.3.1 <u>Absorber control efficiency</u>. The VOC removal efficiency of an absorption device is very dependent on the characteristics of the solvent and the design and operation of the tower. Generally, for a given solvent and solute, an increase in absorber size or a decrease in the operating temperature can increase the VOC control efficiency of the system.

Systems that utilize organic liquids as the solvent usually include a separate item of equipment to strip the adsorbed gas so that the solvent can be recycled to the absorber. The efficiency of the absorber is affected by the efficiency of the stripper. For example, a theoretical absorber calculated to achieve a removal efficiency of 99.9 percent with once-through solvent usage (equivalent to 100 percent stripping efficiency), would achieve only 98.5 percent VOC removal if the solvent were recycled through a stripper which was 98 percent efficient.[49]

4.1.2.3.2 <u>Applicability</u>. The selection of absorption for VOC control depends on the availability of an appropriate solvent for the specific VOC. Absorption is usually not considered when the VOC concentration is below 200-300 ppmv.[50]

The use of absorbers is generally limited to applications in which the stripped absorbent can be reused directly or with minimum treatment. Absorption may not be practical if the waste gas stream contains a mixture of organics, since all will likely not be highly soluble in the same absorbent. Absorbers have found limited use as a VOC emission control device in the polymers and resins industry.

4.2 CONTROL TECHNIQUES FOR FUGITIVE EMISSIONS

This section discusses control techniques that can be applied to reduce fugitive VOC emissions in the polymers and resins industry. Two approaches are available. The first involves a leak detection and repair program in which fugitive emission sources are located and repaired at specified intervals. The second is a preventive approach whereby fugitive emissions never materialize because of the installation of specified controls or leakless equipment. The technical application of these methods is briefly explained in the following subsections.

4.2.1 <u>Leak Detection and Repair Program</u>

4.2.1.1 <u>Leak Detection</u>. The most common types of equipment that have the potential to release fugitive emissions are valves, pump and

compressor seals, pressure relief devices, flanges, open ended lines, and sampling connections. When a leak develops, it can be detected in several ways - instrument monitoring of the individual component, a unit area survey, or by means of a fixed point monitoring system. These methods would generally yield only a qualitative indication of the size of a leak.[51]

4.2.1.2 Repair Program. When a leak is located, the leaking component can be scheduled for repair or replacement. Many components can be serviced on-line without disturbing the plant operations. An example would be the tightening of a valve seal around the packing material. If tightening the packing gland does not stop the leak, the valve must be isolated from the process. Control valves can often be isolated, but block valves generally cannot be.

Repairs of rotary equipment seals usually require isolation of the leaking device. Almost all process pumps are "spared" (two installed in parallel, but only one required for operation) so that either one can be isolated for repair. However, most compressors are not spared, and so compressor seal replacements often necessitate a partial or complete shutdown.

Most leaking flanges can be resealed simply be retightening the flange bolts. A flange leak that requires a gasket seal replacement would likely require a total or partial shutdown of the entire unit. When the leaks can be corrected only by a total or partial shutdown, the temporary emissions resulting from a shutdown and startup could be larger than the continuous fugitive emissions that would take place before a scheduled shutdown.[52] For this reason, the repair of certain leaks is best delayed until the next scheduled shutdown.

4.2.1.3 Effectiveness of Leak Detection and Repair Programs. The emission reduction achieved by a leak detection and repair program is dependent on several factors, including the leak definition, the inspection interval, the allowable delay until repair, and the effectiveness of the repair.

In order to implement a monitoring program, an instrument meter reading (VOC concentration) which will be presumed to indicate an equipment leak must be defined. The meter reading selected may vary from 1,000 ppmv

to 100,000 ppmv.[53] The theoretical efficiency can then be estimated by relating the leak definition to the percentage of total mass emissions that can be expected from sources with concentrations at the source greater than the leak definition. In general, defining a leak by a low meter reading results in larger potential emission reductions. However, the difficulty of repairing relatively minor leaks may dictate that these leaks be excluded from the leak definition.

The inspection interval depends on the expected occurrence and recurrence of leaks after a piece of equipment has been checked and/or repaired. This interval can also vary with the type of equipment and service conditions. Monitoring may be scheduled on an annual, quarterly, monthly, or weekly basis.

If a leak is detected, the equipment should be repaired as soon as practicable. The longer the delay, the less effective will be the overall reduction program.

4.2.2 Preventive Programs

An alternative approach to the leak repair program is to replace the potentially leaky equipment with components based on "leakless" technology, or with equipment that captures the emissions for control. This approach is referred to as a preventive program. For example, in many cases, leakage from a pump seal can be reduced to a negligible level through the installation of an improved shaft sealing mechanism, such as dual mechanical seals, or it can be eliminated entirely by installing sealless pumps, which do not have a shaft/casing junction and thus do not leak during normal operation.[54] A barrier fluid can be circulated between the mechanical seals. Degassing vents in the barrier fluid system allow the transport of emissions in a closed system to a control device. These solutions are sometimes not feasible. For example, the maximum service temperature of a dual mechanical seal is usually about 260°C (500°F). Mechanical seals also cannot be used on pumps with reciprocating shaft motion or those handling extremely corrosive or abrasive fluids.

As in the case of pumps, compressor emissions occur at the junction of the moving shaft and the stationary casing. Emissions from both centrifugal and rotary compressors can be controlled either with mechanical

seals with barrier fluid systems or with liquid film seals. As with
pumps, the degassing vents for the seal fluid must discharge into a
closed system to prevent process gas from escaping. Leakage from recip-
rocating compressors can be controlled by creating a void in the packing
and inserting one or more spacer rings into the packing gland and venting
the void to a collection system.

Leakage from safety/relief valves can be minimized by installing a
rupture disk upstream of each valve. Such combinations can be spared by
installation of a two-way valve which assures that one emergency relief
system is always operational. The other could then be repaired while
process operations continue. An alternative method for controlling
relief valve emissions in some types of service is to use a soft elastomer
seat in the valve.

Caps, plugs, and double block and bleed valves (that vent to closed
systems) are devices that can reduce fugitive emissions from open-ended
lines. VOC emissions from the purging of sampling lines can be minimized
by a closed-purge sampling system that enables purge organics to be
recycled to the process or contained for subsequent disposal.

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5.0 MODIFICATIONS AND RECONSTRUCTIONS

The provisions of Title 40 Code of Federal Regulations, Sections 60.14 and 60.15 (40 CFR 60.14 and 60.15) state that an existing facility can become an affected facility and consequently, subject to the new source performance standard (NSPS) if there is a modification or reconstruction to that operation. An "existing facility," defined in 40 CFR 60.2, is a facility of the type for which a standard of performance is promulgated and the construction or modification of which was commenced prior to the proposal date of the applicable standards. This chapter gives the definition of modification and reconstruction as found in 40 CFR 60.14 and 60.15. A discussion of possible process changes at a polymers and resins facility that would constitute a modification or reconstruction, making it subject to the proposed NSPS, is also included in the chapter.

5.1 DEFINITIONS

5.1.1 Modification

Modification is defined in Section 60.14 as any physical or operational change to an existing facility which results in an increase in the emission rate of the pollutant(s) to which the standard applies. Paragraph (e) of Section 60.14 lists exceptions to this definition which will not be considered modifications, irrespective of any changes in the emission rate. These changes include:

1. Routine maintenance, repair, and replacement at the facility,

2. An increase in the production rate not requiring a capital expenditure as defined in Section 60.2 (bb),

3. An increase in the hours of operation,

4. Use of an alternative fuel or raw material if the existing facility was designed to accommodate the alternate fuel or raw material prior to the date of any applicable standard,

5. The addition or use of any system or device whose primary function is to reduce air pollutants, except when a system is removed or replaced by a system considered to be less efficient, and

6. Relocation or change in ownership.

As stated in paragraph (b), emission factors, material balances, continuous monitoring systems, and manual emission tests are to be used to determine emission rates expressed as kg/hr of pollutant. Paragraph (c) affirms that the addition of an affected facility to a stationary source through any mechanism -- new construction, modification, or reconstruction -- does not make any other facility within the stationary source subject to standards of performance. Paragraph (f) provides for superseding any conflicting provisions and (g) stipulates that compliance be achieved within 130 days of the completion of any modification.

5.1.2 Reconstruction

A "reconstruction" occurs when replacement of components at an existing facility takes place to such extent that: (1) the fixed capital cost of the new components exceeds 50 percent of the fixed capital cost that would be required to construct a comparable new facility, and (2) it is economically and technologically feasible for the facility to comply with the applicable standards set forth. Any existing facility undergoing "reconstruction" becomes an affected facility subject to requirements of NSPS, irrespective of any change in pollutant emission rate.

The owner or operator of any existing facility who proposes to replace components such that the fixed capital cost exceeds 50 percent of the fixed capital of a comparable new facility, must notify the Administrator in writing 60 days prior to commencement of the replacement. The notice must include identification of the owner and plant, a description of the replacement being made including present and proposed air pollution control equipment, an estimate of the proposed replacement costs and the cost of a comparable new facility, an approximation of the operating life of the existing facility after replacement, and a discussion of any economic or technical limitations the facility may have in complying with applicable standards.

The Administrator will determine whether the proposed replacement constitutes a "reconstruction" within 30 days of receipt of the operators notice and any additional information he or she may reasonably require for making a decision. The final decision is based upon the costs involved, the estimated operating life of the facility after replacement compared to a new facility, the extent to which the components being replaced cause or contribute to air pollutant emissions from the facility, and any economic or technical limitations for compliance with applicable standards.

The "reconstruction" provision requires a facility to comply with the NSPS if it undergoes changes that make it essentially a new source. The purpose of the provision is to assure that during reconstruction the facility will install the appropriate emission control equipment.

5.2 MODIFICATIONS AND RECONSTRUCTIONS AT POLYMERS AND RESINS FACILITIES

The polymers and resins industry is expected to experience some growth in the coming years. Within the industry, process technology and operational procedures are undergoing continual change. For example, in the polypropylene industry, catalysts and other technology continue to improve.[1] In addition, some companies are diversifying their product lines by shifting to copolymers or to new combinations of comonomers. These factors may lead to process changes at existing facilities. Changes in operating conditions would mean that an existing facility would be subject to new source standards of performance if the changes cause increased emissions. Under these conditions, the facility becomes a modified facility. However, it is difficult to determine what kinds of physical or operational changes made at polymers and resins facilities will constitute a modification, owing to wide variations in VOC flow and concentration in process vent streams. These variations make it difficult to assess emission increases under paragraph (a) of Section 60.14. Several changes that could be encountered in polymer and resin plants and their possible effects on emissions are presented below.

5.2.1 Process Emissions

In general, a number of modifications may be made to polyolefin plants to increase production. Changes may include upgrading, adding, or improving such equipment as electric drivers, reactor compressors,

catalyst addition systems, refrigeration equipment, heat exchange equipment and piping, and the reactors themselves. In general, these changes will result in increases in baseline emissions, especially those which result from the increase in pellet production.

Existing high pressure, LDPE facilities and liquid phase, HDPE facilities are being, and others will be, converted to the production of linear LDPE. This conversion may result from (1) replacing existing equipment with the new gas phase technology for producing LLDPE, or (2) modifying existing equipment in high pressure, LDPE facilities to allow LLDPE production with either the gas phase process[2] or the liquid phase solution process.[3] These changes, on the whole, are likely to result in a decrease in baseline level of emissions and emission rates as given in Chapter 6. The cost of modifying existing facilities to accept the gas phase technology has been reported to be relatively low.[4]

Recent advances in LLDPE gas phase technology have resulted in capacity increases in existing facilities by 35 percent with little or no capital investment and by 65 percent with "some" capital investment.[2] One company is increasing its capacity by approximately 33 percent through changes in the gear train drive on the cycle gas compressor, installation of an additional transfer line to the fluff storage vessels, and installation of a new C_3 unloading system.[5] No increase in process emissions is expected from these changes. However, fugitive emissions from these specific changes are expected to increase.[5]

The conversion to copolymer production or to new copolymers may result in increased baseline emissions as the comonomer is more likely to be less volatile than the olefins, hence retained longer in the process and not released until product finishing or storage steps, which are not typically under baseline control. However, steam stripping of the product before or immediately after extrusion, followed by condensation, can control these emissions so that overall emissions from the process line will not be increased. At least one company has sought to recover a comonomer (vinyl acetate) from ethylene recycle streams by adding a distillation column and auxiliary equipment.[6] The addition of this equipment does not create any new process emission sources[6] as the ethylene will be the overhead product and the vinyl acetate the bottoms product.

Other potential changes in polymer and resin processes include changes in type or operation of the product dryer, condenser, or distillation columns. In polypropylene and high pressure, LDPE processes, most newly installed dryers are fluidized bed, closed loop systems, which recycle the nitrogen used for drying polymer products. However, some existing plants use rotary dryers, which vent emissions directly to the atmosphere. These dryers may be replaced with fluidized dryers because the latter are more efficient. This conversion is likely to result in a decrease in emissions. Another operational change may involve increasing the operating pressure of product dryers, which may increase emission rates.

There are many instances where a polymers and resins facility may need to replace parts that have failed or not performed well with a different, improved part. This is often the case with distillation columns and their associated condensers. The trays or packing materials in a distillation column in conjunction with the operation of the condenser can have an important impact on pollutant emissions from the process. Replacing column or condenser parts with improved equipment can reduce pollutant emissions. If the replacement results in an increase in emissions, it is not exempt from being considered a modification. An example where such replacement may occur is in the gas phase process for producing LDPE and HDPE. In this process, the butene (comonomer) used for producing these polymers must be purified by distillation before use in the reaction processes. The distillation column and condenser used in the purification may be replaced. If emissions could be offset elsewhere so that there was no net increase in emissions, the existing facility would not become subject to the standards.

5.2.2 Fugitive Emissions[7]

Routine equipment changes and additions at polymers and resins facilities for increased ease of maintenance, plant productivity or plant safety can cause an increase in the fugitive emission rate. However, fugitive emissions from other sources could be reduced to compensate for this increase.

Potential fugitive emission sources, such as pumps or valves, may be replaced. If such a source is replaced with an equivalent source (such as is done during routine repair and replacement), the fugitive

emissions from the facility should not increase because the number of potential sources in the same vapor pressure service (i.e., handling the same monomer or comonomer) remains unchanged.

As noted above, process equipment pieces may be modified or added to existing facilities to increase the capacity of or to optimize a process. The addition of new equipment would normally increase fugitive emissions from a facility due to the increased number of potential emission sources (pumps, valves, sampling connections, etc.) that are associated with the process equipment.

In some cases a facility can be converted, as noted above, from the production of one polymer or copolymer to another. In such a case, whenever either the number of fugitive emission sources or the vapor pressure of the new comonomer is higher than the original, the emissions could be expected to increase. As shown in Table 3-17, emission rates for equipment in vapor service are higher than the rates for the same equipment in light liquid service which, in turn, are higher than those rates for equipment in heavy liquid service. So that, if the vapor pressure of the new monomer or comonomer is higher than the vapor pressure of the original, the fugitive emissions could be expected to increase. However, the most common change is to a comonomer of lower vapor pressure.

The process can also be changed without changing the polymer. One such case would be a change in catalyst. In this case, fugitive emissions would not be expected to change because neither the number of fugitive sources nor the vapor pressure of the monomer or comonomer(s) would change.

5.2.3 Summary

In general, some alterations are likely to be made in existing polypropylene and polyethylene plants that would be considered modifications or reconstructions. However, most changes likely to be made in existing polymer plants will result in reduced emissions and hence will not be a modification as defined by Section 60.14. For those changes where there is a potential for increased emissions, relatively inexpensive equipment can be installed to control these emissions, so that there will be no increase in emissions from the production line, and hence no modification.

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6.0 MODEL PLANTS AND REGULATORY ALTERNATIVES

This chapter presents model plants and their parameters and regulatory
alternatives for the reduction of process and fugitive VOC emissions
from the five polymers and resins categories chosen for NSPS development.
Section 6.1 presents the model plants chosen to represent the five
polymers and resins categories, and Section 6.2 describes the regulatory
baseline (i.e., the level of control that is likely to be employed in
new plants in the absence of a new source performance standard) and the
individual regulatory alternatives for each model plant.

6.1 MODEL PLANTS

A plant is modeled to describe process parameters and offgas stream
characteristics that are representative of a typical new source being
regulated by the standard. The purpose of developing a model plant is
to study the effect of regulatory alternatives on each type of facility
regulated by the NSPS and to estimate the energy, environmental, and
economic impacts associated with each regulatory alternative.

As described in Chapter 3, the five polymer and resin categories,
especially the polyolefins, share certain fundamental similarities in
terms of their processes. At the same time, however, there are among
the various processes many differences that affect process parameters
and process emission characteristics. Thus, no single model plant can
adequately characterize the process emissions of all five polymer and
resin categories. Therefore, nine model plants were developed:

    1.   Polypropylene - continuous, liquid phase slurry process,

    2.   Polypropylene - gas phase process,

    3.   Low density polyethylene - high-pressure, liquid phase process,

    4.   Low density and high density polyethylene - low-pressure, gas
phase process,

5. High density polyethylene - low-pressure, liquid phase slurry process,

6. High density polyethylene - low-pressure, liquid phase solution process,

7. Polystyrene - continuous process,

8. Poly(ethylene terephthalate), (PET)-DMT process, and

9. Poly(ethylene terephthalate), (PET) - TPA process.

The model plants were selected to be representative of basic manu-facturing processes used in plants making these polymers and resins and not of any individual processes used by a specific plant. However, in order to provide a basis for economic analysis, data from a specific plant and its process were used. No model plant was developed based upon the polystyrene batch process because available information indicates that new plants are unlikely to use this process.

Tables 6-1 through 6-9 summarize the parameters for process emissions for each model plant used in the study. The model plants are based upon the respective plant descriptions in Chapter 3. Each model plant is presented on the basis of its process sections; thus, the parameters and offgas stream characteristics presented in Tables 6-1 through 6-9 are the combined characteristics of the individual streams in each process section as identified in Chapter 3.

As noted in Chapter 6, "Model Process Units and Regulatory Alternatives," of the report "VOC Fugitive Emissions in Synthetic Organic Chemicals Manufacturing Industry - Background Information for Proposed Standards" (EPA-450/3-80-033a), fugitive emissions are proportional to the number of potential sources, but are not related to capacity, throughput, or age. Based on a qualitative assessment of both the emission source counts and emission estimates (see Appendix C), a single model unit, as presented in Table 6-10, was chosen to represent the fugitive emission characteristics of the polymers and resins industry. This model plant, however, does not apply to the poly(ethylene terephthalate) model plants because only a "heavy" liquids (i.e., ethylene glycol) and solids are used in these plants.

Table 6-1. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE LIQUID PHASE POLYPROPYLENE PROCESS

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate (kg VOC/Mg Product)	Uncontrolled VOC Emissions		Estimated Flow Total Plant (scfm)	Temp. (°C)	Heating Value (Btu/scf)	Composition (wt. %)	Average VOC Molecular Weight
			Total (Mg/yr)	Process Train (Mg/yr)					
1. Raw Material Preparation	Cont.	0.07	11	4	0.7	29	4,320	100 VOC[a]	86
2. Polymerization Reaction	Cont.	4.1	615	205	25	54	2,260	100 VOC[b]	44
3. Material Recovery	Cont.	30	4,500	1,500	140	56[e]	2,210	100 VOC[c]	4
4. Product Finishing	Cont.	2.6	390	130	230	85[e]	115	9.3 VOC[d] 90.7 Air	45
5. Product Storage	--	--	--	--	--			--	--
TOTALS		36.8	5,516	1,840					

Model Plant Capacity: 150 Gg/yr
No. of Process Trains: 3
Capacity per Process Train: 50 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

[a] Assumes VOC is diluent, which, in turn, is assumed to be hexane.

[b] Assumes 17.5 percent is C_3H_6, 47.5 percent is C_3H_8, and 5 percent is hexane.

[c] Assume 2.07 percent is C_4H_6, 2.07, C_4H_8; 1.25 percent, C_4H_{10}; 62.04 percent hexane; and 33.52 percent, methanol.

[d] Assumes 90.7 percent is air; 1.0 percent, hexane; 3.0 percent propane; and 3.3 percent, methanol.

[e] Assumed.

Table 6-2. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE GAS PHASE POLYPROPYLENE PROCESS

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate (kg VOC/Mg Product)	Uncontrolled VOC Emissions Total (Mg/yr)	Uncontrolled VOC Emissions Process Train (Mg/yr)	Estimated Flow Total Plant (scfm)	Temp. (°C)	Heating Value (Btu/scf)	Composition (wt. %)	Average VOC Molecular Weight
1. Raw Material Preparation	--	--	--	--	--	--	--	--	--
2. Polymerization Reaction	Inter.	11.5	1,210	400	48,670[a]	38	7,100	99.8 VOC[b] / 0.2 H_2	43
3. Material Recovery	Cont.	25	2,625	875	92	38	2,400	100 VOC[c]	47
4. Product Finishing	--	--	--	--	--	--	--	--	--
5. Product Storage	--	--	--	--	--	--	--	--	--
TOTALS		46.5	3,835	1,275					

Model Plant Capacity: 105 Gg/yr
No. of Process Trains: 3
Capacity per Process Train: 35 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

[a] Total flow time assumed to be 8.6 hours per year.

[b] Assumes 36.8 percent propylene; 10.8 percent, propane; and 2.2 percent, hexane.

[c] Assumes 71.6 percent is propylene; 7.9 percent, propane, and 20.5 percent, hexane.

Table 6-3. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE HIGH PRESSURE, LIQUID PHASE LDPE PROCESS

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate (kg VOC/Mg Product)	Uncontrolled VOC Emissions		Estimated Flow Total Plant (scfm)	Temp. (°C)	Heating Value (Btu/scf)	Composition (wt. %)	Average VOC Molecular Weight
			Total (Mg/yr)	Process Train (Mg/yr)					
1. Raw Material Preparation	--	--	--	--	--	--	--	--	--
2. Polymerization Reaction	Inter.	1.7	476	120	240,000[a]	--	1,490	99 VOC[b] 1 Ethane	28
3. Material Recovery	--	--	--	--	--	--	--	--	--
4. Product Finishing	Cont.	0.2	56	14	1,600	54	3.0	0.2 VOC[c] 99.8 Air	28
5. Product Storage	Cont.	0.8	224	56	6,390	54	3.0	0.2 VOC[c] 99.8 Air	28
6. Other	Inter.	0.2	56	14	28,200[a]	--	1,480	99 VOC[b] 1 Ethane	28
TOTALS		2.9	812	204					

Model Plant Capacity: 280 Gg/yr
No. of Process Trains: 4
Capacity per Process Train: 70 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

[a] Total flow time assumed to be 1 hour per year.
[b] Assumes 95.3 percent is ethylene; 1.5 percent propylene; and 2.2 percent isopropanol.
[c] Assumes all VOC is ethylene.

Table 6-4. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE LOW PRESSURE, GAS PHASE LDPE/HDPE PROCESSES

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate (kg VOC/Mg Product)	Uncontrolled VOC Emissions		Estimated Flow Total Plant (scfm)	Temp. (°C)	Heating Value (Btu/scf)	Composition (wt. %)	Average VOC Molecular Weight
			Total (Mg/yr)	Process Train (Mg/yr)					
1. Raw Material Preparation	Cont.	0.01	1.5	.75	0.05	40[a]	2,431	92 VOC / 8 N_2	56.1[b]
	Inter.	0.15	22.5	11.3	704,110[c]	76	23	1.6 VOC / 85.9 N_2 / 12.5 Air	66.5[b]
2. Polymerization Reaction	Cont.	0.01	1.5	0.75	0.16	38	804	50 VOC / 50 N_2	35[b]
	Inter.	0.05	7.5	3.8	3,824[c]	85	1,441	83 VOC / 17 N_2	35[b]
3. Material Recovery	--	--	--	--	--	--	--	--	--
4. Product Finishing	Cont.	22.3	3,345	1,673	508	85	560	35 VOC / 65 N_2	39[b]
	Inter.	0.7	105	53	102,146[c]	40[a]	755	42 VOC / 58 N_2	55[d]
5. Product Storage	Cont.	0.05	7.5	3.8	2,138	50	0.3	0.02 VOC / 99.93 Air	55[b]
TOTALS		23.3	3,490	1,745					

Model Plant Capacity: 150 Gg/yr
No. of Process Trains: 2
Capacity per Process Train: 75 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

[a] Assumed.
[b] Average molecular weight from industry data.
[c] Total flow time equals 1 hour per year.

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Table 6-5. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE LOW PRESSURE, LIQUID PHASE HDPE SLURRY PROCESSES

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate (kg VOC/Mg Product)	Uncontrolled VOC Emissions Total (Mg/yr)	Uncontrolled VOC Emissions Process Train (Mg/yr)	Estimated Flow Total Plant (scfm)	Temp. (°C)	Heating Value (Btu/scf)	Composition (wt. %)	Average VOC Molecular Weight
1. Raw Material Preparation	Inter.	0.7	42	14	1,710[a]	21	1,470	100 VOC[b]	28
2. Polymerization Reaction	--	--	--	--	--	--	--	--	--
3. Material Recovery	Cont.	12.7	2067	889	205[d]	21	1,450	79 VOC[c] 20 Ethane 1 H_2	32
4. Product Finishing	Cont.	0.406	85	28	1,670[d]	21	4.3	0.3 VOC[e] 99.7 N_2	58
5. Product Storage	--	--	--	--	--	--	--	--	--
TOTALS		13.3	2194	931					

Model Plant Capacity: 210 Gg/yr
No. of Process Trains: 3
Capacity per Process Train: 70 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

a. One occurrence per month, assumed 1 hour per occurrence.
b. Assumes 100 percent ethylene.
c. Assumes 61 percent is ethylene and 18 percent, isobutane.
d. Assumes that the emission rate of 12.7 kg/Mg includes methane also.
e. Assumes all VOC is isobutane.

Table 6-6. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE LOW PRESSURE, LIQUID PHASE HDPE SOLUTION PROCESS

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate (kg VOC/Mg Product)	Uncontrolled VOC Emissions Total (Mg/yr)	Uncontrolled VOC Emissions Process Train (Mg/yr)	Estimated Flow Total Plant (scfm)	Temp. (°C)	Heating Value (Btu/scf)	Composition (wt. %)	Average VOC Molecular Weight
1. Raw Material Preparation	Cont. Inter.	--	--	--	--	--	--	--	--
2. Polymerization	Inter.	30.1	2,709	903	158	30	1,600	75 VOC[a] 8 Ethane 16 Air	38
Reaction									
3. Material Recovery	Cont. Inter.	--	--	--	--	--	--	--	--
4. Product Finishing	Cont.	1.5	135	45	48,800[b]	50	0.2	0.02 VOC[b] 7.4 Steam 92.2 Air	84
5. Product Storage	--	--	--	--	--	--	--	--	--
TOTALS		31.6	2,844	948					

Model Plant Capacity: 90 Gg/yr
No. of Process Trains: 3
Capacity per Process Train: 30 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

[a] Assumes 37.4 percent is ethylene, 21.5 percent is cyclohexane, 7.3 percent is isobutane, and 9.0 percent is other VOC.
[b] Flow after condensation.
[c] Cyclohexane.

TABLE 6-7. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE CONTINUOUS POLYSTYRENE PROCESS

| Process Section | Nature of Stream(s) | Uncontrolled Emission Rate (kg VOC/ Mg Product) | Uncontrolled VOC Emissions | | Estimated Flow | | | | |
			Total (Mg/yr)	Process Train (Mg/yr)	Total Plant (scfm)	Temp. (°C)	Pressure	Composition (wt %)	VOC Molecular Weight
1. Material Recovery	Continuous	3.09	231.8	46	119	100	Atmospheric	15 Styrene 85 Steam	104
2. Product Finishing	Continuous	0.15	11.3	2	10,280	21	-	Trace Styrene	104
TOTALS		3.24	243	48					

Model Plant Capacity: 75 Gg/yr.
Number of Process Trains: 5
Capacity per Process Train: 15 Gg/yr.
Total Operating Hours per Year: 8,600.
Standard Conditions: 21°C (70°F) and 0 psig.

a Stream A has been deleted from the analysis. Stream A is considered not to be within the scope of this project.

TABLE 6-8. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE DMT POLY(ETHYLENE TEREPHTHALATE) PROCESS

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate (kg VOC/ Mg Product)	Uncontrolled VOC Emissions		Estimated Flow Total Plant (scfm)	Temp. (°C)	Pressure (psig)	Composition (wt %)	VOC Molecular Weight
			Total (Mg/yr)	Process Train (Mg/yr)					
1. Material Recovery	Continuous	0.18	19	2.7	2.1	38	-	49 VOC[b] 51 Nitrogen	32
2. Polymerization Reaction[a]	Continuous	9.5	997.5	142.5	-	25	0.2	-	62
TOTALS		9.68	1,016.5	145.2					

Model Plant Capacity: 105 Gg/yr.
Number of Process Trains: 7
Capacity per Process Train: 15 Gg/yr.
Total Operating Hours per Year: 8,600.
Standard Conditions: 21°C (70°F) and 0 psig.

[a] Based on recovery of ethylene glycol from the cooling water tower so that all emissions occur from the cooling water tower.
[b] Methanol.

TABLE 6-9. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE TPA POLY(ETHYLENE TEREPHTHALATE) PROCESS

| Process Section | Nature | Uncontrolled Emission Rate (kg VOC/ Mg Product) | Uncontrolled VOC Emissions | | Estimated Total Plant Flow (scfm) | Temp. (°C) | Pressure (psig) | Composition (wt %) | VOC Molecular Weight |
			Total (Mg/yr)	Process Train (Mg/yr)					
1. Raw material preparation and polymerization reaction	Continuous	9.5	997.5	142.5	-	25	0.2	-	62
TOTALS		9.5	997.5	142.5					

Model Plant Capacity: 105 Gg/yr
Number of Process Trains: 7
Capacity per Process Train: 15 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

^{a}Based on recovery of ethylene glycol from the cooling water tower so that all emissions occur from the cooling water tower.

Table 6-10. FUGITIVE VOC EMISSION MODEL PLANT PARAMETERS

Equipment Component	Equipment Counts	Uncontrolled Emission Rate (kg/hr/Source)
Valves		
Vapor service	402	0.0056
Light liquid service	524	0.0070
Heavy liquid service	524	0.00023
Pumps seals		
Light liquid service	29	0.0494
Heavy liquid service	30	0.0214
Compressor seals	2	0.228
Safety relief valves		
Vapor service	42	0.104
Flanges	2400	0.00083
Sampling connections	104	0.015
Open-ended lines	415	0.0017

TOTAL UNCONTROLLED FUGITIVE EMISSIONS: 151 Mg/yr

6.2 REGULATORY ALTERNATIVES

This section defines various regulatory alternatives or possible courses of action EPA could take to reduce VOC emissions from the polymers and resins industry. These alternatives provide a basis for determining the air quality and nonair quality environmental impacts, energy requirements, and costs associated with varying degrees of VOC emissions reduction and represent comprehensive programs for reduction of emissions.

The regulatory alternatives were developed in two basic steps: (1) determination of baseline control and (2) determination of more stringent levels of VOC control based upon applicable VOC control techniques.

6.2.1 Baseline Control

Baseline control reflects the level of VOC control that is likely to be employed in a new plant in the absence of the new source performance standard. Its determination is difficult. As discussed in Chapter 3, not all States regulate VOC emissions and the States that do regulate these emissions have regulations of different stringency. In addition, the level of control employed by two plants producing the same product with the same basic process in the same State may vary due to differences in detail of the process. Finally, the actual capacity of the new plant may determine the level of emission control employed. Given these difficulties, the following methods were used for determining baseline control.

6.2.1.1 Process Emissions. For polyolefin plants, individual process VOC emission streams from the plant descriptions in Chapter 3 were identified as the streams most likely to be controlled in the absence of a standard by the following conditions:

- Intermittent streams;
- Continuous streams of mass flow rates greater than 91 Mg/yr (100 tons/yr); and
- Exceptions to the above based on specific process information pertaining to the plant process.

The streams identified as most likely to be controlled were assumed to be presently flared. Intermittent streams are generally controlled by a flare for safety reasons and large continuous streams are likely to be controlled, typically by a flare since flares are acceptable control devices in States with VOC regulations.

For the polystyrene and the PET model plants, a different basis was used to identify the regulatory baseline. Both industries currently use various recovery technologies that achieve varying levels of VOC emission reductions. For the polystyrene segment, baseline control reflects the use of condensers on the material recovery emission streams (i.e., devolatilizer vent stream and styrene condenser vent stream). For the poly(ethylene terephthalate) segment, baseline control reflects the use of distillation columns to recover the ethylene glycol from the cooling tower water.

6.2.1.2 Fugitive Emissions. Fugitive emission baseline control, which applies to all model plants except the PET plants, assumes that 75 percent of all gas safety/relief valves and sampling connections, most of the open-ended lines, and 0 percent of all other fugitive emission sources are controlled. These assumptions are consistent with those made in the analysis of fugitive emission baseline chosen for the SOCMI and petroleum refinery industries.

6.2.2 Control Techniques

Process emissions may be controlled by flares, thermal incinerators, catalytic incinerators, boilers, condensers, absorbers, and adsorbers. Fugitive emissions may be controlled through leak detection and repair programs and equipment, design, and operational requirements. These control techniques and their associated emission reductions, which are discussed in detail in Chapter 4, "Emission Control Techniques," were examined to determine technical feasibility when applied to each model plant and the possible level of VOC emission reduction. This information was then used to develop the most effective control options.

6.2.2.1 Process Emission Control Devices. Combustion devices, such as flares, thermal or catalytic incinerators, and boilers, are the most prevalent emission control techniques in this industry, especially in the polyolefin segments. Combustion devices are not typically used, however, in either the polystyrene or the polyester processes for VOC emission control and, thus, are not considered as control techniques for the purpose of defining regulatory alternatives for either of these two segments.

The preference of one combustion device over another is dependent on the waste gas characteristics of each vent stream. Flares, the most commonly used control technique for process offgases, are universally applicable in controlling upset emissions from polyolefin plants. They are capable of handling these emergency releases, as well as low volume continuous vent streams from these processes. Thermal incineration in incinerators or boilers also is applicable to polyolefins vent streams and is second only to the flare in its frequency of use in the industry. Thermal incinerators can be used to control continuous streams with a wide range of concentrations or type of VOC. Boilers are used as control devices for continuous flow, high heating value streams. Catalytic incineration may be used for continous flow, low heating value streams. Its lower operating temperatures requires less supplementary fuel than thermal incinerators to achieve the same level of VOC emission reduction.

Condensers, absorbers, and adsorbers are sensitive to changes in VOC flowrate or concentration, and VOC removal efficiencies decrease as the VOC concentration in the offgas decreases. Thus, these devices often are used primarily to recover process materials rather than as an emission reduction technique. These devices also are more chemical specific than other emission control techniques. Absorbers and adsorbers are not widely used in the polymer and resin industry and, thus, are not considered in the regulatory analyses. Condensers, however, can be used as emission control devices in the case of polystyrene and polyester.

6.2.2.2 Fugitive Emission Control Techniques. VOC fugitive emissions control techniques include leak detection and repair programs, and equipment, design, and operational requirements. These programs are discussed in Chapter 4. One combination of these control techniques (see Table 6-11) was chosen as the regulatory alternative for controlling VOC fugitive emissions from the polymers and resins industry. This regulatory alternative was chosen so that this NSPS would be consistent with fugitive regulations for the petroleum refining industry and the synthetic organic chemical manufacturing industry (SOCMI).

6.2.3 Regulatory Alternatives

The following sections present the regulatory alternatives, including baseline control, for each of the model plants. The regulatory alternatives are presented on a process section-by-process section basis. Determination of baseline control (Regulatory Alternative 1) was described previously

Table 6-11. CONTROL SPECIFICATION FOR FUGITIVE EMISSIONS UNDER
REGULATORY ALTERNATIVE 2

Source	Inspection Interval	Equipment Specification
Valves		
Vapor service	Monthly	None
Light liquid	Monthly	None
Heavy liquid	None	None
Pump seals		
Light liquid	Monthly[a] Weekly visual	None
Heavy liquid	None	None
Compressor seals	None	Controlled degassing vents
Safety relief valves		
Vapor service	None	Rupture disks on relief valves
Flanges	None	None
Sampling connections	None	Closed-purge sampling
Open-ended lines (purge, drain, sample lines)	None	Cap

[a]For pumps, instrument monitoring would be supplemented with weekly visual inspections for liquid leakage. If liquid is noted to be leaking from the pump seal, the pump seal would be repaired.

in Section 6.2.1. Regulatory alternatives of increasing stringency were developed, in general, by first implementing fugitive emission controls for each model plant except the PET model plants. The fugitive VOC control program is known to be reasonable and the best technological system for fugitive VOC emissions based on analyses already performed in the SOCMI and Petroleum Refinery Fugitive VOC NSPS's. Therefore, this control is retained in all succeeding regulatory alternatives. This avoids unnecessary combinations of the same process controls with and without fugitive emission control.

Additional regulatory alternatives of increasing stringency were then developed on the basis of controlling the process section remaining with the highest level of potential uncontrolled emissions until all process sections were analyzed. For each model plant, if a whole process section was found to be unreasonable to control as a whole on the basis of emission reduction achieved and the associated cost, that process section was then examined to determine whether an individual stream could be controlled at a reasonable cost.

In developing the regulatory alternatives for the polypropylene and polyethylene model plants, flares and incinerators (thermal and catalytic) were considered the control devices most likely to be used to control continuous streams because of the wide range of applicability, current use in the industry, and favorable costs and were used to develop the regulatory alternatives for these model plants. Boilers, which some plants may choose to use, were not costed specifically for any regulatory alternative because not all polypropylene and polyethylene plants have boilers or the need for steam. However, if a plant has a boiler and a need for the steam, boilers are more cost-effective than either flares or thermal incinerators. For control of intermittent streams in the regulatory alternatives, only flares were used because they are the only control device considered feasible to control intermittent streams.

In developing the regulatory alternatives for the polystyrene and poly(ethylene terephthalate) (PET) plants, recovery techniques are more likely to be used than combustion techniques on streams containing styrene monomer or ethylene glycol. For polystyrene plants, condensers are the most likely recovery technique to be used and are used to develop

the regulatory alternatives for the polystyrene model plant. For the PET plants, ethylene glycol recovery systems are the most likely control techniques to be used in the industry and are used to develop the regulatory alternatives for this model plant. A flare was also analyzed for control of the methanol vent stream found in the PET/DMT plants.

6.2.3.1 Polypropylene - Continuous, Liquid Phase Slurry Process. (Table 6-12.) By applying the general criteria outlined in Section 6.2.1, the process emission streams identified under baseline control (Regulatory Alternative 1) are Stream B (reactor vents), Stream C (decanter vent), Stream D (neutralizer vent), Stream E (slurry filter/vacuum system vent) and Stream F (diluent separation and recovery). These streams correspond to the polymerization reaction section (Stream B) and the material recovery section (Streams C through F), and are assumed to be controlled by combustion (i.e., a flare). Based on a 98 percent VOC destruction efficiency, the baseline control of process emissions is equivalent to an annual process emission reduction of about 5,010 Mg/yr. Fugitive emission baseline control, as discussed in Section 6.2.1, would reduce uncontrolled fugitive emissions from 151 Mg/yr to approximately 106 Mg/yr. Under baseline control, an annual emission reduction of 5,058 Mg/yr is achieved or approximately 89 percent of the total uncontrolled emissions from this model plant.

Regulatory Alternative 2 represents the application of fugitive emission controls in addition to baseline control. Fugitive emission controls result in annual fugitive emissions of 47 Mg/yr. Under this alternative, an annual emission reduction of 5,118 Mg/yr is achieved or approximately 90 percent of the total uncontrolled emissions.

Regulatory Alternative 3 includes the same control as in Regulatory Alternative 2 plus the combustion of the emissions from the product finishing section. This alternative reduces emissions by 5,499 Mg/yr or 97 percent of the total uncontrolled emissions.

Regulatory Alternative 4 includes the same control as in the third regulatory alternative plus the combustion of emissions from the raw material preparation section. This alternative is somewhat more stringent than the third, reducing annual emissions by 5,510 Mg/yr or 97.2 percent from uncontrolled levels.

Table 6-12. REGULATORY ALTERNATIVES FOR THE LIQUID PHASE
POLYPROPYLENE PROCESS

| Regulatory Alternative | Process Emissions | | | Annual Emission Reductions[b] | |
	Process Section(s) Controlled[a]	Control Technique	Fugitive Emissions	Mg/Yr	Percent[c]
1 (Baseline)	PR + MR	Flare	d	5,058	89%
2	PR + MR	Flare	e	5,118	90%
3	PR, MR, plus PF	Combustion[f]	e	5,499	97.0%
4	PR, MR, PF, plus RMP	Combustion[f]	e	5,510	97.2%

[a]Process sections include the following:
RMP = Raw Material Preparation
PR = Polymerization Reaction
MR = Material Recovery
PF = Product Finishing
PS = Product Storage

[b]Represents reduction in annual VOC emissions from the uncontrolled level.

[c]Based on total uncontrolled emissions of 5,667 Mg/yr (5,516 Mg/yr process emissions plus 151 Mg/yr fugitive emissions.)

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-11).

[f]Combustion devices may include flares, thermal incinerators, catalytic incinerators, and boilers.

6.2.3.2 <u>Polypropylene - Gas Phase Process.</u> (Table 6-13.) As
stated in Table 6-2, the mass flowrate of emissions from the material
recovery section is about 460 times smaller than that from the polymerization
reaction section. The intermittent emissions occur less than 1 percent
of the time, and is, for safety purposes, most likely to be sent to a
flare during emergencies or process upsets. The continuous emissions
may or may not be sent to the same flare. Based on information from one
company, the emissions from the material recovery section may be controlled
by the same flare as the emissions from the polymerization reaction
section. However, considering similar processes by other companies, the
continuous stream may not be controlled in the absence of the NSPS.
Thus, the baseline (Regulatory Alternative 1) includes control of only
the intermittent stream, by a flare. This baseline represents a 31 percent
reduction of total uncontrolled VOC emissions for this model plant.

Regulatory Alternative 2 represents the application of fugitive
emission controls in addition to baseline control. Fugitive emission
controls result in annual fugitive emissions of 47 Mg/yr. This alternative
achieves an annual emission reduction of 1,290 Mg/yr, or approximately
32 percent of the total uncontrolled emissions.

Regulatory Alternative 3 includes the same control as in Regulatory
Alternative 2 plus the combustion of the emissions from the material
recovery section. This alternative results in an annual emission reduction
of 3,863 Mg/yr or 97 percent of total uncontrolled emissions.

6.2.3.3 <u>Low Density Polyethylene, High Pressure, Liquid Phase Process.</u>
(Table 6-14.) The baseline control (Regulatory Alternative 1) for this
model plant sends the emissions from the emergency vents other than from
the reactor to a flare. If the high-pressure reactor vent, Stream B, from
this process is sent to a flare, a particulate polymer removal system is
a prerequisite. Based on available information, only one company has a
particulate polymer removal system technology that can handle high-pressure
emergency vent gas. In general, it is expected that the emergency vent
from the reactor will be released through a pressure relief system and
vented to the atmosphere. The exhaust gas contains suspended particulate
polymer that cannot be directly sent to a flare system because of safety
considerations. Unless the particulate removal system technology becomes

Table 6-13. REGULATORY ALTERNATIVES FOR THE GAS PHASE
POLYPROPYLENE PROCESS

| Regulatory Alternative | Process Emissions | | Fugitive Emissions | Annual Emission Reductions[b] | |
	Process Section(s) Controlled[a]	Control Technique		Mg/Yr	Percent[c]
1 (Baseline)	Polymerization Reactor (PR)	Flare	d	1,231	31%
2	PR	Flare	e	1,290	32%
3	PR Plus Material Recovery	Combustion[f]	e	3,863	97%

[a]Process sections include the following:
RMP = Raw Material Preparation
PR = Polymerization Reaction
MR = Material Recovery
PF = Product Finishing
PS = Product Storage

[b]Represents reduction in annual VOC emissions from the uncontrolled level.

[c]Based on total uncontrolled emissions of 3,986 Mg/yr (3,835 Mg/yr process emissions plus 151 Mg/yr fugitive emissions).

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-11).

[f]Combustion devices may include flares, thermal incinerators, catalytic incinerators, and boilers.

Table 6-14. REGULATORY ALTERNATIVES FOR THE HIGH PRESSURE
LIQUID PHASE, LDPE PROCESS

| Regulatory Alternative | Process Emissions | | Fugitive Emissions | Annual Emission Reductions[b] | |
	Process Section(s) Controlled[a]	Control Technique		Mg/Yr	Percent[c]
1 (Baseline)	(Emergency Vents Other than from Reactor)	Flare	d	100	10%
2	Baseline (B)	Flare	e	159	17%
3	B plus PS	Combustion[f]	e	379	39%
4	Baseline, PS plus PF	Combustion[f]	e	434	45%

[a]Process sections include the following:
RMP = Raw Material Preparation
PR = Polymerization Reaction
MR = Material Recovery
PF = Product Finishing
PS = Product Storage

[b]Represents reduction in annual VOC emissions from the uncontrolled level.

[c]Based on total uncontrolled emissions of 963 Mg/yr (812 Mg/yr process emissions plus 151 Mg/yr fugitive emissions).

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-11).

[f]Combustion devices may include flares, thermal incinerators, catalytic incinerators, and boilers.

available to most companies, Stream B is not likely to be controlled. Therefore, Stream B is not included in the baseline or any other regulatory alternative. Under baseline control, process and fugitive emissions would be reduced by 100 Mg/yr or about 10 percent of the total uncontrolled emissions.

Regulatory Alternative 2, which represents fugitive emission controls plus baseline control, results in a total annual emission reduction of 159 Mg/yr or about 17 percent of the total uncontrolled emissions.

Regulatory Alternative 3 represents Regulatory Alternative 2 controls plus the combustion of emissions from the product storage section. This alternative reduces total uncontrolled emissions by about 380 Mg/yr or 39 percent.

Regulatory Alternative 4 is the same as Regulatory Alternative 3 plus the combustion of emissions from the product finishing section. Total uncontrolled emissions under this alternative are reduced by about 434 Mg/yr or 45 percent.

6.2.3.4 <u>Low Density and High Density Polyethylene, Low Pressure, Gas Phase Process</u>. (Table 6-15.) The intermittent streams in the raw materials preparation, the polymerization reaction and the product finishing section and Stream I (product discharge vent) in the product finishing section were determined to be controlled by flaring under the baseline assumptions. Annual emission reductions under baseline control (Regulatory Alternative 1) are about 3,455 Mg/yr or 95 percent of the total uncontrolled emissions from this model plant.

Regulatory Alternative 2 represents the application of fugitive emission controls in addition to baseline control. Fugitive emissions control results in annual fugitive emissions of 47 Mg/yr. Under this alternative, an annual emission reduction of 3,514 Mg/yr is achieved, or approximately 96.5 percent of the total uncontrolled emissions.

Regulatory Alternative 3 represents Regulatory Alternatives 2 controls plus the combustion of continuous emissions from the raw material preparation and polymerization reaction sections. Under Regulatory Alternative 3, emissions are reduced by 3,517 Mg/yr or 96.6 percent.

Regulatory Alternative 4 represents in addition to the control achieved in Regulatory Alternative 3, the combustion of continuous

Table 6-15.   REGULATORY ALTERNATIVES FOR THE LOW PRESSURE,
GAS PHASE LDPE/HDPE PROCESS

Regulatory Alternative	Process Emissions Process Section(s) Controlled[a]	Control Technique	Fugitive Emissions	Annual Emission Reductions[b] Mg/Yr	Percent[c]
1 (Baseline)	RMP (Inter-mittent streams only), PR (Inter-mittent streams only), and PF	Flare	d	3,455	95%
2	Baseline	Flare	e	3,514	96.5%
3	Baseline plus contin-uous streams from RMP and PR	Combustion[f]	e	3,517	96.6%
4	Baseline, continuous streams from RMP and PR, plus PS	Combustion[f]	e	3,524	96.8%

[a]Process sections include the following:
RMP = Raw Material Preparation
PR = Polymerization Reaction
MR = Material Recovery
PF = Product Finishing
PS = Product Storage

[b]Represents reduction in annual VOC emissions from the uncontrolled level.

[c]Based on total uncontrolled emissions of 3,641 Mg/yr (3,490 Mg/yr process emissions plus 151 Mg/yr fugitive emissions).

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-11).

[f]Combustion devices may include flares, thermal incinerators, catalytic incinerators, and boilers.

emissions in the product storage section. This alternative results in an annual emission reduction of 3,524 Mg/yr or 96.8 percent of the total uncontrolled emissions.

6.2.3.5 <u>High Density Polyethylene, Low Pressure, Liquid Phase, Slurry Process</u>. (Table 6-16.) Under baseline control (Regulatory Alternative 1) Stream A (feed preparation) and Stream D (recycle treaters) would be controlled. These streams correspond to the raw material preparation and material recycle sections, respectively. Annual uncontrolled emissions are reduced by 2,700 Mg/yr or 92 percent.

The next level of control, Regulatory Alternative 2, represents baseline control plus the fugitive emission controls listed in Table 6-11. An annual emission reduction of about 2,760 Mg/yr would be achieved, which is a 94 percent reduction from the uncontrolled level of emissions.

Regulatory Alternative 3 requires the same control as Regulatory Alternative 2 plus the combustion of the emissions from the product finishing section. This alternative reduces emissions by about 2,840 Mg/yr or 97 percent of the total uncontrolled emissions from this model plant.

6.2.3.6 <u>High Density Polyethylene, Low Pressure, Liquid Phase, Solution Process</u>. (Table 6-17) For this model plant, baseline control (Regulatory Alternative 1) was assumed to be reflected by the current level of control being practiced by the plant on which the model plant was based. Under this assumption, all the streams in the raw material preparation, polymerization reaction, and material recovery sections were assumed to be flared. Baseline control results in an annual emission reduction of 2,700 Mg/yr or 90 percent of the total uncontrolled emissions.

Regulatory Alternative 2 requires the same control as baseline plus fugitive emission controls. This alternative reduces process and fugitive emissions from the process sections under baseline control by about 2,760 Mg/yr or 92 percent of the total uncontrolled emissions.

Regulatory Alternative 3 includes the controls of Regulatory Alternative 2 as well as the emissions from the product finishing section being controlled by combustion. This regulatory alternative results in an annual emission reduction of 2,891 Mg/yr or 97 percent.

Table 6-16. REGULATORY ALTERNATIVES FOR THE LIQUID PHASE
HIGH DENSITY POLYETHYLENE SLURRY PROCESS

| Regulatory Alternative | Process Emissions | | Fugitive Emissions | Annual Emission Reductions[b] | |
	Process Section(s) Controlled[a]	Control Technique		Mg/Yr	Percent[c]
1 (Baseline)	RMP plus MR	Flare	d	2,700	92%
2	RMP plus MR	Flare	e	2,759	94%
3	RMP, MR plus PF	Combustion[f]	e	2,842	97%

[a]Process sections include the following:
 RMP = Raw Material Preparation
 PR = Polymerization Reaction
 MR = Material Recovery
 PF = Product Finishing
 PS = Product Storage

[b]Represents reduction in annual VOC emissions from the uncontrolled level.

[c]Based on total uncontrolled emissions of 2,945 Mg/yr (2,794 Mg/yr process emissions plus 151 Mg/yr fugitive emissions).

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-11).

[f]Combustion devices may include flares, thermal incinerators, catalytic incinerators, and boilers.

Table 6-17. REGULATORY ALTERNATIVES FOR THE LIQUID PHASE
HDPE SOLUTION PROCESS

| Regulatory Alternative | Process Emissions | | Fugitive Emissions | Annual Emission Reductions[b] | |
	Process Section(s) Controlled[a]	Control Technique		Mg/Yr	Percent[c]
1 (Baseline)	RMP, PR, MR	Flare	d	2,700	90%
2	RMP, PR, MR	Flare	e	2,759	92%
3	RMP, PR, MR plus PF	Combustion[f]	e	2,891	97%

[a]Process sections include the following:
RMP = Raw Material Preparation
PR = Polymerization Reaction
MR = Material Recovery
PF = Product Finishing
PS = Product Storage

[b]Represents reduction in annual VOC emissions from the uncontrolled level.

[c]Based on total uncontrolled emissions of 2,995 Mg/yr (2844 Mg/yr process emissions plus 151 Mg/yr fugitive emissions).

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-11).

[f]Combustion devices may include flares, thermal incinerators, catalytic incinerators, and boilers.

6.2.3.7 Polystyrene, Continuous Process. (Table 6-18.) As discussed earlier, baseline control (Regulatory Alternative 1) is considered to be the use of condensers on the material recovery section steams. This reflects current industry practices so that no process emission reduction is achieved beyond the uncontrolled emission rates reported for this process. Fugitive emission baseline control results in an annual emission reduction of 45 Mg/yr. This results in a 11 percent reduction from uncontrolled emission levels.

Regulatory Alternative 2 represents the application of fugitive emission controls in addition to baseline control. Fugitive emission control results in annual fugitive emissions of 47 Mg/yr. Under this alternative, an annual emission reduction of 104 Mg/yr is achieved, or approximately 26 percent of the total uncontrolled emissions.

Regulatory Alternative 3 represents fugitive emission control under Regulatory Alternative 2 plus further recovery of emissions from the material recovery section by the use of condensers to an emission level of 0.06 kg VOC/Mg product. Under this alternative, emissions are reduced by 331 Mg/yr or 84 percent of the total uncontrolled emissions.

6.2.3.8 Poly(ethylene terephthalate), DMT Process. (Table 6-19). As discussed earlier, baseline control (Regulatory Alternative 1) reflects the recovery of ethylene glycol from the water in the cooling tower servicing the polymerizer vacuum system. This system corresponds to a recovery system currently in use in the industry, thus no emission reduction from uncontrolled emissions occurs under baseline control.

Regulatory Alternative 2 reflects the use of a different recovery system (based on the use of a spent ethylene glycol spray condenser placed between the polymerizers and the vacuum system) that controls emissions from the polymerization reaction section to an emission level of 0.21 kg VOC/Mg product. Under this alternative annual emissions are reduced by about 976 Mg, or about 96 percent from uncontrolled levels.

Regulatory Alternative 3 reflects control under Regulatory Alternative 2 plus the combustion of the methanol emissions from the material recovery section (i.e., methanol recovery section). Under this alternative, annual emissions are reduced by 995 Mg, or about 98 percent of uncontrolled emissions.

Table 6-18 REGULATORY ALTERNATIVES FOR PROCESS EMISSIONS
FOR THE CONTINUOUS POLYSTYRENE PROCESS

| Regulatory Alternative | Process Emissions | | Fugitive Emissions | Annual Emission Reductions[a] | |
	Process Section Controlled	Control Technique		Mg/Yr	Percent[b]
1 (Baseline)	-	-	c	45	11%
2	-	-	d	104	26%
3	Material Recovery	Recovery[e]	d	331	84%

[a]Represents reduction in annual VOC emissions from the uncontrolled level.

[b]Based on total uncontrolled emissions of 394 Mg/yr (243 Mg/yr process emissions plus 151 Mg/yr fugitive emissions).

[c]Fugitive baseline control.

[d]Control equivalent to Regulatory Alternative 2 (see Table 6-11).

[e]Control to an equivalent emission rate of 0.06 kg VOC/Mg product through use of recovery techniques such as condensers.

Table 6-19. REGULATORY ALTERNATIVES FOR PROCESS
EMISSIONS FROM THE DMT POLY(ETHYLENE TEREPHTHALATE) PROCESS

Regulatory Alternative	Process Emissions		Annual Emission Reductions[b]	
	Process Section Controlled[a]	Control Technique	Mg/Yr	Percent[c]
1 (Baseline)	PR	Recovery[d]	0	0
2	PR	Recovery[e]	976	96%
3	PR MR	Recovery[e] Combustion[f]	995	98%

[a]PR = Polymerization Reaction
MR = Material Recovery

[b]Represents reduction in annual VOC emissions from uncontrolled levels.

[c]Based on total uncontrolled emissions of 1,017 Mg/yr.

[d]Recovery of ethylene glycol from the cooling water tower.

[e]Recovery of ethylene glycol to an emission rate of 0.21 kg VOC/Mg product through the use of a more efficient recovery system such as using a spent ethylene glycol spray condenser recovering the ethylene glycol prior to the vacuum system servicing the polymerizers.

[f]Combustion devices include flares, incinerators, and boilers.

6.2.3.9 _Poly(ethylene terephthlate), TPA Process._ (Table 6-20.)
As discussed earlier, baseline control (Regulatory Alternative 1) is
assumed to be equivalent to current industry practice in which ethylene
glycol is recovered from the water in the cooling tower serving the raw
material preparation polymerization reaction sections. The resulting
emissions are the same as the uncontrolled emissions, and, thus, no
emission reduction occurs under baseline control.

Regulatory Alternative 2 reflects the use of a spent ethylene
glycol spray condenser and recovery system controlling emissions from
the polymerization reactors. The spray condensers would be placed
between the reactors and the vacuum system evacuating the reactors.
This regulatory alternative would reduce emissions by 972 Mg/yr or
97 percent from uncontrolled emissions.

6.2.4 _Summary of Regulatory Alternatives_

This section summarizes the regulatory alternatives for each model
plant. The uncontrolled emission rates, annual emission reductions, and
percent control achieved by the regulatory alternatives are summarized
in Table 6-21.

Table 6-20. REGULATORY ALTERNATIVES FOR PROCESS EMISSIONS
FROM THE TPA POLY(ETHYLENE TEREPHTHALATE) PROCESS

| Regulatory Alternatives | Process Emissions | | Annual Emission Reductions[b] | |
	Process Section Controlled[a]	Control Technique	Mg/Yr	Percent[c]
1 (Baseline)	RMP, PR	Recovery[d]	0	0
2	RMP, PR	Recovery[e]	972	97%

[a]RMP = Raw Material Preparation
  PR = Polymerization Reaction

[b]Represents reduction in annual VOC emissions from uncontrolled levels.

[c]Based on total uncontrolled emissions of 988 Mg/yr.

[d]Recovery of elthylene glycol from the cooling water tower servicing
both the raw material preparation and polymerization reaction section.

[e]Control of the raw material preparation section to an emission rate
of 0.04 kg VOC/Mg product through the use of recovery techniques
such as reflux condensers and the polymerization reaction section to
0.21 kg VOC/Mg product through the use of a more efficient recovery
system such as a spent ethylene glycol spray condenser recovering
the ethylene glycol prior to the vacuum system servicing the polymizers.

Table 6-21. SUMMARY OF UNCONTROLLED EMISSIONS AND EMISSION REDUCTIONS FOR REGULATORY ALTERNATIVES BY MODEL PLANT

MODEL PLANT	UNCONTROLLED EMISSIONS (Mg/yr)			REGULATORY ALTERNATIVE	ANNUAL EMISSION REDUCTIONS[a]			
	Process	Fugitive	Total		Process Emissions (Mg/yr)	Fugitive Emissions (Mg/yr)	TOTAL	
							(Mg/yr)	(%)
PP, L	5,516	151	5,667	1	5,013	45	5,058	89%
				2	5,013	104	5,118	90%
				3	5,395	104	5,499	97%
				4	5,406	104	5,510	97.2%
PP, G	3,335	151	3,986	1	1,185	45	1,230	31%
				2	1,185	104	1,290	32%
				3	3,760	104	3,863	97%
LDPE, L	812	151	963	1	55	45	100	10%
				2	55	104	159	17%
				3	275	104	379	39%
				4	330	104	434	45%
LDPE/HDPE, G	3,490	151	3,641	1	3,410	45	3,455	95%
				2	3,410	104	3,514	96.5%
				3	3,417	104	3,517	96.6%
				4	3,420	104	3,524	96.8%
HDPE, SL	2,794	151	2,945	1	2,655	45	2,700	92%
				2	2,655	104	2,759	94%
				3	2,738	104	2,842	97%
HDPE, SO	2,844	151	2,995	1	2,655	45	2,700	90%
				2	2,655	104	2,759	92%
				3	2,787	104	2,891	97%
PS, C	243	151	394	1	0	45	45	11%
				2	0	104	104	26%
				3	227	104	331	34%
PET, DMT	1,017	0	1,017	1	0	0	0	0%
				2	976	0	976	96%
				3	995	0	995	98%
PET, TPA	998	0	998	1	0	0	0	0%
				2	972	0	972	97%

[a]From uncontrolled levels.

KEY: PP = Polypropylene
LDPE = Low Density Polyethylene
HDPE = High Density Polyethylene
PS = Polystyrene
PET = Polyethylene terepthalate

L = Liquid Phase
G = Gas Phase
SL = Slurry Process
SO = Solution Process
C = Continuous Process

7.0 ENVIRONMENTAL IMPACTS

This chapter assesses the environmental impacts of implementing the regulatory alternatives presented in Chapter 6. The assessment discusses these impacts in terms of air quality, water quality, solid waste generation, and energy requirements. Other areas examined include noise impacts, irreversible and irretrievable commitment of resources, and impacts of delaying implementation of the regulatory alternatives.

Process and fugitive VOC emissions from polymers and resins plants operating under Regulatory Alternative II* are projected to be about 84 percent less than uncontrolled emissions and about 25 percent less than estimated baseline emissions (Regulatory Alternative I). VOC emissions under Regulatory Alternative III are almost 96 percent of uncontrolled emission levels and 81 percent of baseline emission levels. VOC emissions under Regulatory Alternative IV are over 96 percent of uncontrolled emissions and 82 percent of baseline emissions.

Secondary air pollutants emitted by VOC emission control devices are anticipated to be minimal in comparison to the quantity of VOC reduced. Water pollution and solid waste disposal impacts of the regulatory alternatives are expected to be minimal in comparison to the amount of liquid and solid wastes generated during polymers and resins manufacturing operations.

The energy required to implement Regulatory Alternative II (fugitive VOC control) is estimated to be 1,340 terajoules (TJ) (218,000 barrels of oil) per year over energy demands in the absence of any VOC control. However, Regulatory Alternative II requires about 78 TJ (13,000 barrels of

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*In this chapter, arabic numerals will be used to describe regulatory alternatives when used in the context of a single model plant, or model plants of a single type; Roman numerals will be used to describe the total values associated with all model plants of a given Regulatory Alternative. For example, all model plants under Regulatory Alternative 1 will be collectively described as Regulatory Alternative I.

oil) per year less than baseline controls (Regulatory Alternative I),
because the energy cost for fugitive VOC control under Regulatory
Alternative 2 for seven model plants and for the more efficient ethylene
glycol recovery system under Regulatory Alternative 2 for the PET model
plants is less than the energy credit resulting from the reduced loss of
VOC under these control measures. Therefore, there is a net reduction
in the energy requirement under Regulatory Alternative II when compared
to Regulatory Alternative I. Regulatory Alternatives III and IV require
about 1,630 TJ and 1,610 TJ, respectively, (265,000 and 262,000 barrels
of oil) per year more than energy demands without VOC controls. Regulatory
Alternatives III and IV require about 213 TJ and 192 TJ, respectively,
(34,000 and 31,000 barrels) per year more than the projected requirement
under current VOC control practices (Regulatory Alternative I) in the
polymers and resins industry.

Projected noise impacts due to implementation of any regulatory
alternative are expected to be minimal. No significant irreversible or
irretrievable commitments of resources are expected to be incurred under
the regulatory alternatives. Delaying implementation of Regulatory
Alternatives II through IV is anticipated to adversely impact air quality.
Detailed discussion of the assessed environmental impacts is presented
in the following sections.

7.1 AIR POLLUTION IMPACTS

The air pollution impact of each regulatory alternative is determined
by comparison of uncontrolled VOC emission rates to residual VOC emission
rates for emission control systems installed on process operations and
residual VOC emission rates for fugitive emission control practices. In
order to analyze the incremental air quality impact of each regulatory
alternative, average annual model plant VOC emission rates are determined
and used to project industrywide air quality impacts of new polymers amd
resins plants.

7.1.1 Average Annual Model Plant VOC Emissions

Annual VOC emission rates for each model plant are determined
through the following equation:

$$E_{ij} = P_{ij} + F_j \qquad (1)$$

where E_{ij} = annual VOC emission rate (Mg/yr) for model plant $\underline{i}$ (for example i = polypropylene/liquid phase; for other model plant types see Table 6-20) under Regulatory Alternative $\underline{j}$ (j = uncontrolled, I, II, III or IV);

P_{ij} = annual VOC process emission rate for model plant $\underline{i}$ under Regulatory Alternative $\underline{j}$;

F_{j} = annual VOC fugitive emission rate for each model plant under Regulatory Alternative $\underline{j}$.

The annual model plant VOC process emission rate (P_{ij}) is a function of the uncontrolled emission rate and the effectiveness of the control technique applied to each VOC stream or process section. Thus the annual average model process VOC emission rate (P_{ij}) can be expressed as:

$$P_{ij} = \sum(U_{ai} \times [1-C_{aij}]) \qquad (2)$$

where U_{ai} = the uncontrolled VOC emission rate (Mg/yr) from process section $\underline{a}$ (a = raw material preparation, polymerization reaction, material recovery, product finishing, or product storage) in model plant $\underline{i}$;

C_{aij} = the VOC emission reduction efficiency of the control system for process section $\underline{a}$ in model plant $\underline{i}$ under Regulatory Alternative $\underline{j}$.

The uncontrolled process emission rates for each process section in each model plant are presented in Tables 6-1 through 6-9. The control techniques employed on appropriate model plant process sections and their emission control efficiencies are presented in Tables 6-11 through 6-19.

The quantity of fugitive VOC emissions are assumed to be the same for each model plant type. The uncontrolled fugitive VOC emission rate is based on equipment component counts and uncontrolled component emission rates presented in Table 6-10. The effectiveness of fugitive VOC emission control through the use of leak detection and repair programs (Regulatory Alternatives II, III, or IV) is based on the leak detection and repair (LDAR) model developed for control of fugitive VOC emissions from SOCMI,[1] and is presented in Table 6-11.

Table 7-1 presents the primary or VOC related air quality impacts of the regulatory alternatives for each model plant. (Table 7-1 and other tables, as appropriate, in this chapter have been separated into

Model Plant	Regulatory Alternative	VOC Emissions per Model Plant [b]			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined	Process	Fugitive	Combined
PP/Liquid	U[a]	5,516	151	5,667	-	-	-
	1	504	106	610	-	-	-
	2	504	47	551	0	56	10
	3	122	47	169	76	56	72
	4	111	47	158	78	56	74
PP/Gas	U[a]	3,835	151	3,986	-	-	-
	1	2,650	106	2,756	-	-	-
	2	2,650	47	2,697	0	56	2
	3	77	47	124	97	56	96
LDPE/Liquid	U[a]	812	151	963	-	-	-
	1	757	106	863	-	-	-
	2	757	47	804	0	56	7
	3	537	47	584	29	56	32
	4	482	47	529	36	56	39
LDPE/HDPE/ Gas	U[a]	3,490	151	3,641	-	-	-
	1	80	106	186	-	-	-
	2	80	47	127	0	56	32
	3	77	47	124	10	56	33
	4	70	47	117	13	56	37
HDPE/Liquid Slurry	U[a]	2,794	151	2,945	-	-	-
	1	139	106	245	-	-	-
	2	139	47	186	0	56	24
	3	56	47	103	60	56	58
HDPE/Liquid Solution	U[a]	2,844	151	2,995	-	-	-
	1	139	106	295	-	-	-
	2	189	47	236	0	56	20
	3	57	47	104	79	56	65
PS	1[a]	243	151	394	-	-	-
	1	243	106	349	-	-	-
	2	243	47	290	0	56	17
	3	16	47	63	93	56	81
PET/DMT	U[a]	1,017	0	1,017	-	-	-
	1	1,017	0	1,017	-	-	-
	2	41	0	41	96	0	96
	3	22	0	22	98	0	98
PET/TPA	U[a]	998	0	998	-	-	-
	1	998	0	998	-	-	-
	2	26	0	26	97	0	97

[a] U=uncontrolled.

[b] From Table 6-21.

Table 7-1b PRIMARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR
POLYMERS AND RESINS PLANTS (tons/yr)

Model Plant	Regulatory Alternative	VOC Emissions per Model Plant[b]			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined	Process	Fugitive	Combined
PP/Liquid	U[a]	6,068	166	6,234	-	-	-
	1	554	117	671	-	-	-
	2	554	52	606	0	56	10
	3	134	52	186	76	56	72
	4	122	52	174	78	56	74
PP/Gas	U[a]	4,227	166	4,393	-	-	-
	1	2,921	117	3,038	-	-	-
	2	2,921	52	2,973	0	56	2
	3	85	52	137	97	56	96
LDPE/Liquid	U[a]	895	166	1,061	-	-	-
	1	834	117	951	-	-	-
	2	834	52	886	0	56	7
	3	591	52	643	29	56	32
	4	530	52	582	36	56	39
LDPE/HDPE/ Gas	U[a]	3,839	166	4,005	-	-	-
	1	88	117	205	-	-	-
	2	88	52	140	0	56	32
	3	85	52	137	10	56	33
	4	77	52	129	13	56	37
HDPE/Liquid Slurry	U[a]	3,080	166	3,246	-	-	-
	1	153	117	270	-	-	-
	2	153	52	205	0	56	24
	3	62	52	114	60	56	58
HDPE/Liquid- Solution	U[a]	3,135	166	3,301	-	-	-
	1	208	117	325	-	-	-
	2	208	52	260	0	56	24
	3	63	52	115	70	56	52
PS	U[a]	268	166	434	-	-	-
	1	268	117	385	-	-	-
	2	268	52	320	0	56	17
	3	18	52	70	93	56	32
PET/DMT	U[a]	1,120	0	1,120	-	-	-
	1	1,120	0	1,120	0	0	0
	2	45	0	45	96	56	96
	3	24	0	24	98	56	98
PET/TPA	U[a]	1,100	0	1,100	-	-	-
	1	1,100	0	1,100	-	-	-
	2	24	0	24	97	0	97

[a] U=uncontrolled.

[b] From Table 6-21.

two tables; the first table reports the information in metric units; the second, in english units.) Process VOC emission reductions under baseline controls (Regulatory Alternative I) range from 0 to 95 percent, depending on whether current polymers and resins industry practices include controls on the type of model plant examined and the extent that these controls are used to treat VOC emissions from a particular model plant process section. For example, under baseline control only a 10 percent reduction in uncontrolled process VOC emissions from a liquid phase, low density polyethylene (LDPE/Liquid) manufacturing plant, because the major emitting process section -- the polymerization reactor -- can not be fitted with a VOC control system because of safety considerations. (Safety problems associated with control of VOC from the LDPE/Liquid polymerization reactor are discussed in Section 6.2.3.3.) Similarly, gas phase polypropylene (PP/Gas) process emissions under baseline control are reduced from uncontrolled levels by 31 percent, because only polymerization reactor exhaust streams are controlled. Fugitive VOC emission control under current industry practices assumes that 75 percent of all gas safety/relief valves and sampling connections, most of the open-ended lines, and 0 percent of all other fugitive emission sources are controlled. This results in a decrease in fugitive emissions from uncontrolled levels of about 30 percent (from 151 Mg/yr to 106 Mg/yr).

Process VOC emission controls and process VOC emission rates under Regulatory Alternative II generally remain unchanged from the baseline levels. However, fugitive emission control practices implemented under Regulatory Alternative II result in a decrease from baseline control of approximately 56 percent in fugitive emissions of VOC from each model plant except for the PET model plants. Regulatory Alternative 2 for the poly(ethylene terepthalate) model plants, reduces baseline emissions through the use of an ethylene glycol spray condenser and recovery system. Under Regulatory Alternative II, overall VOC emission reductions from baseline control (processes and fugitive emission reductions) range from 2 to 97 percent. The differences in emission reduction efficiency over baseline among the model plants is a function primarily of the extent that process controls are currently employed by each industry segment and the proportion of fugitive emissions to overall emissions from the model plant.

Under Regulatory Alternative III, controls are applied to process emissions. Implementation of Regulatory Alternative III would result in a range of overall VOC emission reduction from baseline control of about 32 to 98 percent. (Note that there is no Regulatory Alternative 3 for the PET/TPA Model Plant.)

Under Regulatory Alternative IV, more stringent control requirements for process emissions were analyzed for the following model plants: liquid phase polypropylene (PP/Liquid), liquid phase low density polyethylene (LDPE/Liquid) and gas phase low density/high density polyethylene (LDPE/HDPE/Gas). Combined process and fugitive VOC emission reduction levels of 74, 39, and 37 percent from baseline emissions, respectively, occurred under each model plant's Regulatory Alternative 4.

7.1.2 Industrywide VOC Emission Impacts of New Plants

The effect of the regulatory alternatives on VOC emissions from new plants projected to be built over the next 5 years can be estimated by the equation:

$$N_T = \sum (E_{ij} \times n_i) \tag{3}$$

where N_T = total VOC emissions from all new polymer and resin plants (Mg/yr);

$\quad E_{ij}$ = annual emission rate from model plant $\underline{i}$ under Regulatory Alternative $\underline{j}$ (Mg/yr-from Table 7-1);

$\quad n_i$ = number of new model plant $\underline{i}$ projected to be built by 1988 (from Table 8-38).

The impact of the regulatory alternatives on VOC emissions from the 27 new plants expected to be built are presented in Table 7-2. Baseline emissions from new plants are projected to be 17,910 Mg/yr (19,740 tons/yr), or a 79 percent overall reduction from uncontrolled levels. The additional increment of emission control under Regulatory Alternative II results in overall VOC emissions of about 13,570 Mg/yr (14,950 tons/yr): an 84 percent decrease from the uncontrolled emission rate, or 24 percent incremental reduction in emissions over Regulatory Alternative I. The additional process emission controls required under Regulatory Alternative III are expected to reduce uncontrolled emissions by 96 percent to about 3,430 Mg/yr (3,770 tons/yr): an 75 percent increase in reduction efficiency over Regulatory Alternative II. Additional or more effective process emission controls employed in some plants under Regulatory Alternative IV result

Table 7-2a. INDUSTRYWIDE PRIMARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMERS AND RESINS PLANTS (Mg/yr)

Model Plant	Regulatory Alternative	VOC Emissions per Model Plant			Number of New Plants[c]	Industrywide VOC Emissions For New Plants			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined		Process	Fugitive	Combined	Process	Fugitive	Combined
PP/Liquid	U[a]	5,516	151	5,667	3	16,550	453	17,000	-	-	-
	1	504	106	610		1,510	318	1,830	-	-	10
	2	504	47	551		1,510	141	1,650	0	56	22
	3	122	47	169		365	141	510	76	56	72
	4	111	47	158		330	141	470	78	56	74
PP/Gas	U[a]	3,835	151	3,986	3	11,505	453	11,960	-	-	-
	1	2,650	106	2,756		7,950	318	8,270	-	-	-
	2	2,650	47	2,697		7,950	141	3,090	0	52	52
	3	77	47	124		231	141	370	97	56	96
HDPE/Liquid	U[a]	812	151	963	1	812	151	960	-	-	-
	1	757	106	863		757	106	910	-	-	7
	2	757	47	804		757	47	800	0	56	7
	3	537	47	584		544	47	590	29	56	32
	4	482	47	529		491	47	540	36	56	37
HDPE/HDPE Gas	U[a]	3,400	151	3,641	10	34,900	1,510	36,410	-	-	-
	1	80	106	186		800	1,060	1,860	-	-	32
	2	80	47	127		800	470	1,270	0	56	33
	3	77	47	124		770	470	1,240	10	56	37
	4	70	47	117		700	470	1,170	13	56	37
HDPE/Liquid Slurry	U[a]	2,794	151	2,945	2	5,588	302	5,890	-	-	-
	1	139	106	245		278	212	490	-	-	24
	2	139	47	186		278	94	370	0	56	24
	3	56	47	103		112	94	210	60	56	63
HDPE/Liquid-Solution	U[a]	2,844	151	2,995	3	8,532	453	8,990	-	-	-
	1	189	106	295		567	318	890	-	-	20
	2	189	47	236		567	141	710	0	56	20
	3	57	47	104		171	141	312	70	56	65
PS	U[a]	243	151	394	2	486	302	790	-	-	-
	1	241	106	349		486	212	700	-	-	17
	2	243	47	290		486	94	580	0	56	17
	3	16	47	63		32	94	130	93	56	83

Table 7-2a. INDUSTRYWIDE PRIMARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR
POLYMERS AND RESINS PLANTS (Mg/yr)
(concluded)

Model Plant	Regulatory Alternative	VOC Emissions per Model Plant[b]			Number of New Plants[c]	Industrywide VOC Emissions For New Plants			Percent VOC Emission Reduction from Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined		Process	Fugitive	Combined	Process	Fugitive	Combined
PLT/DMT	0[a]	1,017	0	1,017	1	1,017	0	1,020	-	-	-
	1	1,017	0	1,017		1,017	0	1,020	-	-	-
	2	41	0	41		41	0	40	96	0	96
	3	22	0	22		22	0	20	98	0	98
PLT/IPA	0[a]	998	0	998	2	2,000	0	2,000	-	-	-
	1	998	0	998		2,000	0	2,000	-	-	-
	2	26	0	26		52	0	50	97	0	97
Total	0[a]				27	80,388	3,624	84,010	-	-	-
	I					15,361	2,544	17,910	19	96	24
	II[d]					12,441	1,128	13,570	85	56	41
	III[d]					2,299	1,128	3,430	85	56	41
	IV[d]					2,141	1,128	3,270	86	56	82

[a] Uncontrolled.

[b] From Table 7-1.

[c] From Table 8-38.

[d] Where there is no emission value for a model plant under a particular regulatory alternative, the value of the preceding regulatory alternative for that model plant category is used.

Table 7-2b. INDUSTRYWIDE PRIMARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMERS AND RESINS PLANTS (tons/yr)

Model Plant	Regulatory Alternative	VOC Emissions per Model Plant			Number of New Plants	Industrywide VOC Emissions For New Plants			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined		Process	Fugitive	Combined	Process	Fugitive	Combined
PP/liquid	U[a]	6,068	166	6,234	3	18,200	498	18,500	-	-	-
	1	554	117	671		1,662	351	2,010	-	-	-
	2	554	52	606		1,662	156	1,820	0	56	10
	3	134	52	186		402	156	560	76	56	72
	4	122	52	174		366	156	520	78	56	74
PP/Gas	U[a]	4,227	166	4,393	3	12,681	498	13,180	-	-	-
	1	2,921	117	3,038		8,763	351	9,110	-	-	-
	2	2,921	52	2,973		8,763	156	8,920	0	56	2
	3	85	52	137		255	156	410	97	56	96
LDPE/liquid	U[a]	895	166	1,061	1	895	166	1,060	-	-	-
	1	834	117	951		834	117	950	-	-	-
	2	834	52	886		834	52	890	0	56	7
	3	591	52	643		591	52	640	29	56	32
	4	530	52	582		530	52	580	36	56	39
LDPE/HDPE/Gas	U[a]	3,839	166	4,005	10	38,390	1,660	40,050	-	-	-
	1	88	117	205		880	1,170	2,050	-	-	-
	2	88	52	140		880	520	1,400	0	56	32
	3	85	52	137		850	520	1,370	10	56	33
	4	70	52	129		700	520	1,220	13	56	37
HDPE/liquid Slurry	U[a]	3,080	166	3,246	2	6,160	332	6,490	-	-	-
	1	153	117	270		306	234	540	-	-	-
	2	153	52	205		306	104	410	0	56	24
	3	62	52	114		124	104	230	60	56	53
HDPE/liquid-Solution	U[a]	3,135	166	3,301	3	9,405	498	9,900	-	-	-
	1	208	117	325		624	351	980	-	-	-
	2	208	52	260		624	156	780	0	56	20
	3	63	52	115		189	156	350	70	56	65
PS	U[a]	263	166	434	2	536	332	870	-	-	-
	1	263	117	385		536	234	770	-	-	-
	2	268	52	320		536	104	640	0	56	17
	3	18	52	70		36	104	140	93	56	82

Table 7-2b. INDUSTRYWIDE PRIMARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMERS AND RESINS PLANTS (tons/yr)
(concluded)

Model Plant	Regulatory Alternative	VOC Emissions per Model Plant[b]			Number of New Plants	Industrywide VOC Emissions For New Plants			Percent VOC Emission Reduction from Baseline (Regulatory Alternative I)		
		Process	Fugitive	Combined		Process	Fugitive	Combined	Process	Fugitive	Combined
PE I/DEI	U[a]	1,120	0	1,120	1	1,120	0	1,120	-	-	-
	1	1,120	0	1,120		1,120	0	1,120	-	-	-
	2	45	0	45		45	0	50	96	0	96
	3	24	0	24		24	0	20	93	0	93
PE I/IPA	U[a]	1,100	0	1,100	2	2,200	0	2,200	-	-	-
	1	1,100	0	1,100		2,200	0	2,200	-	-	-
	2	28	0	28		56	0	50	97	0	97
Total	U[a]				27	89,390	3,980	93,370	-	-	-
	I					16,930	2,810	19,740	-	-	-
	II[d]					13,700	1,250	14,950	19	96	24
	III[d]					2,520	1,250	3,770	85	56	81
	IV[d]					2,270	1,250	3,520	86	56	82

a. Uncontrolled.
b. From Table 7-1.
c. From Table 8-38.
d. Where there is no emission value for a model plant under a particular regulatory alternative, the value of the preceding regulatory alternative for that model plant category is used.

in an overall VOC emission rate of 3,270 Mg/yr (3,520 tons/yr): this value represents an emission reduction from uncontrolled levels of slightly greater than 96 percent and a 4 percent increase in emission reduction efficiency over Regulatory Alternative III emission reduction levels.

7.1.3 Secondary Air Quality Impacts of the Regulatory Alternatives

Secondary air pollutants are those emissions which are not usually associated with an uncontrolled process, but which result from the use of pollution control equipment. VOC emission control equipment that may be incorporated into model plant VOC emission reduction systems include flares, thermal incinerators, catalytic incinerators, and condensers. Secondary air pollutants are not expected to be generated by leak detection and repair programs for fugitive emissions of VOC. Secondary air pollutants that may be generated through the improper use or maintenance of VOC emission control systems are not expected to be significant. However, pollutants generated by the combustion of fuel to generate steam for the flares, to incinerate low VOC content streams, or to generate electrical power to operate control devices may adversely impact air quality. Consequently, particulate, sulfur oxide (SO_x), and nitrogen oxide (NO_x) emission rates are estimated, based on energy consumption rates for VOC emission control presented in Chapter 8, by the equation:

$$S_{xj} = \sum [(G_{ij} \times K_{Gx}) + (B_{ij} \times K_{ox})] \, n_i \times 0.454 \cdot \frac{1Mg}{10^3 kg} \qquad (4)$$

where S_{xj} = new plant industrywide emission rate (Mg/yr) of secondary air pollutant $\underline{x}$ (x = particulate, SO_x, NO_x) under Regulatory Alternative $\underline{j}$ (j = 1, 2, 3, 4);

G_{ij} = natural gas requirements to control VOC emissions from model plant $\underline{i}$ under Regulatory Atlernative $\underline{j}$ (10^6 ft^3/yr);

K_{Gx} = pollutant $\underline{x}$ emission factor for natural gas combustion[2] (lb/10^6 ft^3)

B_{ij} = fuel oil required to generate electricity and steam for VOC control equipment in model plant $\underline{i}$ under Regulatory Alternative $\underline{j}$ (gal/yr);

K_{ox} = pollutant $\underline{x}$ emission factor for No. 4 fuel oil combustion[3] (lb/10^3 gal^3);

n_i = number of new plants of model plant type $\underline{i}$ to be built in year 1988.

No. 4 fuel oil is chosen to represent the approximate midrange of fuel oil grades available for use. Fuel oil sulfur content of 2 percent by weight represents the upper limit of sulfur content in No. 4 fuel oil.[4] It is recognized that particulate and SO_x emissions from fuel combustion will be reduced through State Implementation Plan (SIP) requirements which follow the guidelines set forth in EPA regulations on the preparation of implementation plans [40 CFR 51].[5] However, it is useful to estimate uncontrolled secondary pollutant emission rates to construct a worst-case scenario. Potential secondary air pollutant impacts for new polymers and resins plants are presented in Table 7-3.

The 27 new plants that are expected to come on line by 1988 are projected to emit approximately 30 Mg/yr (33 tons/yr) of particulate, 1,323 Mg/yr (1,458 tons/yr) of SO_x, and 258 Mg/yr (284 tons/yr) of NO_x, if these plants operate under current industry practices (Regulatory Alternative I). Secondary air pollutant emissions will not increase appreciably under Regulatory Alternative II, because 24 of the 27 plants will employ fugitive emission control programs which do not require the use of add-on pollution control systems. The new PET plants under Regulatory Alternative 2 are estimated to emit an additional 1.3 Mg/yr (1.4 tons/yr) of particulate, 66 Mg/yr (72.6 tons/yr) of SO_x, and 12.6 Mg/yr (13.9 tons/yr) of NO_x above baseline control.

Particulate emissions generated by new plants during control of VOC containing process streams under Regulatory Alternative III are about 33.2 Mg/yr (36.6 tons/yr), or 11 percent greater than secondary air pollutant emissions under baseline conditions. SO_x emissions under Regulatory Alternative III are projected at 1,400 Mg/yr (1,542 tons/yr): a 6 percent increase in emissions over Regulatory Alternative I. NO_x emissions under Regulatory Alternative III are estimated to be about 299 Mg/yr (330 tons/yr), which represents a 16 percent increase over baseline secondary air pollutant emissions.

Projected particulate, SO_x and NO_x emissions from new plants operating under Regulatory Alternative IV are 32.8 Mg/yr (36.1 tons/yr), 1,380 Mg/yr (1,521 tons/yr) and 295 Mg/yr (325 tons/yr) respectively. Under Regulatory Alternative IV, the 2 percent across-the-board reduction in secondary air pollutant emissions from the levels estimated for Regulatory Alternative III

Table 7-3a. SUMMARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN PLANTS

Model Plant	Regulatory Alternatives	Model Plant Emissions (kg/yr)			Number of New Plants [b]	Industrywide Emissions (Mg/yr)		
		Particulate	SOx	NOx		Particulate	SOx	NOx
PP/Liquid	1	154	6,678	1,347		0.46	20.0	4.04
	2	154	6,678	1,347		0.46	20.0	4.04
	3	154	6,718	1,355	3	0.46	20.1	4.07
	4	7	305	58		0.02	0.91	0.17
PP/Gas	1	18	178	247		0.054	0.53	0.74
	2	18	178	247	3	0.054	0.53	0.74
	3	90	3,440	869		0.27	10.3	2.61
LDPE/Liquid	1	14,190	635,900	121,650		14.2	636	122
	2	14,190	635,900	121,650	1	14.2	636	122
	3	14,210	635,900	122,040		14.2	636	122
	4	14,220	636,900	122,190		14.2	636	122
LDPE/HDPE/Gas	1	130	4,541	1,306		1.3	45.4	13.1
	2	130	4,541	1,306	10	1.3	45.4	13.1
	3	140	4,999	1,394		1.4	50.0	13.9
	4	140	4,999	1,394		1.4	50.0	13.9
HDPE/Liquid Slurry	1	102	4,373	907		0.20	8.75	1.81
	2	102	4,373	907	2	0.20	8.75	1.81
	3	104	4,469	925		0.21	8.94	1.85
HDPE/Liquid Solution	1	106	4,541	939		0.32	13.6	2.82
	2	106	4,541	939	3	0.32	13.6	2.82
	3	823	12,260	10,780		2.47	36.8	32.3
PS	1	c	c	c		c	c	c
	2	c	c	c		c	c	c
	3	0.02	28.2	5.4	2	0.001	0.06	0.01
	4	0.02	28.2	5.4		0.001	0.06	0.01
PET/DMT	1	4,439	199,121	38,049		4.4	200	38.0
	2	4,731	212,226	40,553	1	4.7	212	40.6
	3	4,736	212,226	40,632		4.7	212	40.6
PET/TPA	1	4,439	199,121	38,049	2	8.9	398	76.1
	2	4,731	212,226	40,553		9.4	425	81.1

Table 7-3a. SECONDARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN PLANTS
(concluded)

Model Plant	Regulatory Alternatives	Model Plant Emissions (kg/yr)[a]			Number of New Plants[b]	Industrywide Emissions (Mg/yr)		
		Particulate	SO_x	NO_x		Particulate	SO_x	NO_x
101A	I	--	--	--		29.8 [c]	1,373 [c]	257.9 [c]
	II	--	--	--	27	30.6 [c]	1,400 [c]	265.5 [c]
	III	--	--	--		33.2 [d]	1,400 [d]	299.1 [d]
	IV	--	--	--		32.8 [d]	1,380 [d]	299.3 [d]

[a] Emission factors from References 2 and 3. Fuel consumption under each regulatory alternative is derived from Tables 8-20 to 8-28. Factors used are for combustion of No. 4 fuel oil having 2 percent sulfur content by weight. (Source: Reference 4.)

[b] From Table 8-38.

[c] Estimates of secondary pollutants from polystyrene plants are not available, and are not included in totals for Regulatory Alternatives.

[d] Where there is no emission value for a model plant under a particular regulatory alternative, the emission value of the preceding regulatory alternative for that model plant category is used.

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Table 7-4b. SECONDARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN PLANTS

Model Plant	Regulatory Alternatives	Model Plant Emissions[a] (lbs/yr)			Number of New Plants[b]	Industrywide Emissions (tons/yr)		
		Particulate	SO$_x$	NO$_x$		Particulate	SO$_x$	NO$_x$
PP/Liquid	1	338	14,720	2,970		0.51	22.1	4.45
	2	338	14,720	2,970	3	0.51	22.1	4.45
	3	340	14,810	2,987		0.51	22.2	4.48
	4	15	672	128		0.02	1.01	0.19
PP/Gas	1	39	393	544		0.06	0.59	0.82
	2	39	393	544	3	0.06	0.59	0.82
	3	199	7,580	1,916		0.30	11.3	2.87
LDPE/Liquid	1	31,270	1,402,000	268,200		15.6	701	134
	2	31,270	1,402,000	268,200	1	15.6	701	134
	3	31,330	1,402,000	269,060		15.7	701	135
	4	31,360	1,403,000	269,380		15.7	701	135
LDPE/HDPE/Gas	1	286	10,010	2,800		1.43	50.5	14.0
	2	286	10,010	2,800	10	1.43	50.5	14.0
	3	309	11,020	3,072		1.54	55.1	15.4
	4	309	11,020	3,072		1.54	55.1	15.4
HDPE/Liquid Slurry	1	225	9,640	1,999		0.23	9.64	2.00
	2	225	9,640	1,999	2	0.23	9.64	2.00
	3	230	9,853	2,039		0.23	9.85	2.04
HDPE/Liquid Solution	1	233	10,010	2,069		0.35	15.0	3.10
	2	233	10,010	2,069	3	0.35	15.0	3.10
	3	1,815	27,040	23,760		2.72	40.5	35.6
PS	1	c	c	c		c	c	c
	2	c	c	c	2	c	c	c
	3	1.4	62.2	11.9		0.001	0.06	0.01
	4	1.4	62.2	11.9		0.001	0.06	0.01
PVC/EDW	1	9,787	439,011	83,888		4.8	220	41.9
	2	10,431	467,904	89,409	1	5.2	234	44.7
	3	10,442	467,904	89,583		5.2	234	44.8
PVC/IPA	1	9,787	439,011	83,888		9.8	439	83.9
	2	10,431	467,904	89,409	2	10.4	468	89.4

Table 7-3b. SECONDARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN PLANTS (concluded)

Model Plant	Regulatory Alternatives	Model Plant Emissions[a] (lbs/yr)			Number of New Plants[b]	Industrywide Emissions (tons/yr)		
		Particulate	SO_x	NO_x		Particulate	SO_x	NO_x
TOTAL	I	--	--	--		32.8[c]	1,458[c]	284.2[c]
	II	--	--	--	27	33.8[c]	1,544[c]	292.5[c]
	III[d]	--	--	--		36.6	1,542	329.5
	IV[d]	--	--	--		36.1	1,521	325.3

[a] Emission factors from References 2 and 3. Fuel consumption under each regulatory alternative is derived from Tables 8-20 to 8-28. Factors used are for combustion of No. 4 fuel oil having 2 percent sulfur content by weight. (Reference 4.)

[b] From Table 8-38.

[c] Estimates of secondary pollutants from polystyrene plants are not available, and are not included in totals for Regulatory Alternatives.

[d] Where there is no emission value for a model plant under a particular regulatory alternative, the emission value of the preceeding regulatory alternative for that model plant category is used.

is due to switching from the use of separate flare and thermal incineration control systems to venting all VOC to thermal incinerators in new PP/Liquid plants. Consequently, the energy requirements for VOC emission control systems in PP/Liquid plants under Regulatory Alternative IV are less than the energy requirements for the other regulatory alternatives. The energy requirement aspect of the regulatory alternatives is discussed in detail in Section 7.4.

Comparison of the total mass of secondary pollutant emissions (particulate + SO_x + NO_x) presented in Table 7-3 to total process VOC emissions presented in Table 7-2 indicates that under Regulatory Alternative I, a 66,100 Mg/yr (73,630 ton/yr) reduction in VOC emissions is obtained with a concomitant potential increase in secondary pollutant emissions of 1,611 Mg/yr (1,775 tons/yr). Under Regulatory Alternative III, the alternative having the greatest potential for secondary air pollutant emissions, uncontrolled VOC emissions are reduced by 80,620 Mg/yr (89,650 tons/yr), while increasing the potential amount of secondary pollutants by about 1,730 Mg/yr (1,910/tons/yr). Again, it should be noted that actual secondary air pollutant emissions probably will be less than the values presented due to fuel combustion requirements in many SIP's.[5] To the extent that gas cleaning systems are used (see Section 7.3), actual secondary air pollutant emissions will be reduced further.

7.2 WATER POLLUTION IMPACTS

With the exception of LDPE/Liquid plants, (Section 3.2.2), there are no water effluents from the model plants. In the case of the LDPE/Liquid process, centrifugal scrubbers are used upstream of the VOC control device (catalytic incinerator) to remove particulates that may plug or foul the catalyst bed. The cooling water, used in the scrubber to separate product resin from the waste gases, is routed to a knock-out drum where contained gas is removed. The resin is separated and the water recirculated to the scrubber. Thus, the system essentially is closed and results in no water pollution.

About 4,900 m^3/yr (1.3 x 10^6 gallons/yr) of water are consumed by condensers used to control VOC emissions from a PS production unit (Table 8-26). This represents approximately 4 percent of the consumptive water requirements of a PS plant based on the bulk process.[6] Increased

water use will occur under certain regulatory alternatives because of the ethylene glycol condensers and recovery systems used in PET plants. For Regulatory Alternative 2 at PET/TPA model plants, and Regulatory Alternatives 3 and 4 at PET/DMT model plants, water usage will increase by 58,674 m^3 (15.5 x 10^6 gal/year) each, or about 7 percent of the consumptive water requirements of a continuous polyester resin plant.[6]

Fugitive VOC emission sources can adversely affect water quality, as leaking components that handle liquid hydrocarbon streams may increase the waste level entering wastewater treatment systems. Implementation of leak detection and repair programs (Regulatory Alternative II) should reduce the waste load on wastewater treatment systems by reducing the amount of VOC that may leak from process equipment and enter the wastewater systems.

7.3 SOLID WASTE DISPOSAL IMPACTS

Solid waste impacts are anticipated to arise primarily from the disposal of spent catalyst from catalytic incinerators. Catalytic incinerators are employed in LDPE/Liquid processes under Regulatory Alternative 3 and 4, LDPE/HDPE/Gas processes operating under Regulatory Alternative 4, and HDPE/Liquid solution processes under Regulatory Alternative 3.

The quantity of spent catalyst expected to be generated annually by new plants is estimated by the equation:

$$D_{Tj} = [(R_{aij} \times Q \times 0.02382 \ m^3/ft^3 \times n_i) \div 3] \qquad (5)$$

where D_{Tj} = Annual quantity of spent catalyst generated under
Regulatory Alternative $\underline{j}$ (j = I, II, III or IV);

R_{aij} = Exhaust gas flow rate (scfm) from process section $\underline{a}$
(a = raw materials preparation, polymerization reduction,
material recovery, product finishing, or product storage)
in model plant $\underline{i}$ (i = liquid phase polypropylene and other
process types) under Regulatory Alternative $\underline{j}$ (refer to
Tables 6-1 to 6-8);

Q = 2.25 ft^3 of catalyst/1,000 scfm of catalytic incineration
throughput (Section 8.1.3.1)

N_i = number of new plants of model plant type $\underline{i}$ expected to be
built by 1988 (Table 8-38).

It is assumed that the catalyst bed has a useful life of 3 years (Section 8.1.3.1).

The solid waste disposal impacts of the regulatory alternatives are presented in Table 7-4. The total solid waste disposal impact of the regulatory alternatives for the 14 new plants expected to employ catalytic incinerators is projected to be 2.88 m^3/yr (102 ft^3/yr) under Regulatory Alternative III and 3.01 m^3/yr (135 ft^3/yr) under Regulatory Alternative IV. The volume of solid waste generated by the use of spray towers or liquid jet scrubbers employed upstream of catalytic incineraters to treat corrosive exhaust gases under Regulatory Alternative IV (Section 4.2.2.3) represents about 0.01 percent of the volume of biological sludge generated annually as solid waste by process operations in LDPE/Liquid, LDPE/HDPE/Gas, and HDPE/Liquid processes.[7] Volumes of biological sludge generated by applicable process operations in new plants are presented in Table 7-5. Disposal of spent catalyst is not expected to be affected by the Resource Conservation and Recovery Act of 1976 (RCRA), since catalyst for VOC incinerator is not listed as a hazardous waste under 40 CFR 261.30.[8] Further, the high price of the platinum or palladium catalyst may encourage recycling where feasible.

Solid waste may be generated as a result of implementing the regulatory alternatives in the form of ash collected from control of the secondary air pollutants, discussed in Section 7.1.3. The quantity of fly ash generated is estimated by assuming the installation of a scrubbing system on oil-fired boilers to control both sulfur oxides and particulates that achieve a 60 percent control efficiency for particulate emissions and a 95 percent removal efficiency for SO_x emissions.[9] Applying these efficiences to the secondary pollutants presented in Table 7-3 and assuming a fly ash density factory of 0.72 Mg/m^3 (45 lb/ft^3),[10] the maximum volume of fly ash generated industrywide by new polymers and resins plants is projected to be about 1,157 m^3/yr (40,660 ft^3yr) under Regulatory Alternative III, the alternative having the greatest potential for secondary pollutant emissions. This fly ash generation rate represents 0.8 to 2.2 percent of the total volume of biological sludge estimated to be generated by product manufacturing operations, or a worst-case scenario fly ash generation factor of about 0.01 Mg/Mg process VOC emissions controlled (Table 7-3). Disposal of fly ash should not be affected by RCRA, as fly ash from fuel combustion is specifically exempted from RCRA by 40 CFR 261.4(b)(4).[11]

Table 7-4. INDUSTRYWIDE SOLID WASTE IMPACTS OF THE REGULATORY ALTERNATIVES FOR NEW POLYMER AND RESIN PLANTS[a]

Model Plant	Regulatory Alternative	Solid waste generated $m^3(ft^3)/yr$[b]	Number of New Plants[c]	Solid waste generated industrywide $m^3 (ft^3)/yr$[d]
LDPE/Liquid	1	0		0
	2	0	1	0
	3	0.13 (4.8)		0.13 (4.8)
	4	0.17 (6.0)		0.17 (6.0)
LDPE/HDPE/Gas	1	0		0
	2	0		0
	3	0	10	0
	4	0.09 (3.2)		0.09 (32.0)
HDPE/Liquid-Solution	1	0		0
	2	0	3	0
	3	0.92 (32.4)		2.75 (97.2)
TOTAL	I	–		0
	II	–		0
	III	–	4	2.88 (102)[d]
	IV	–	14	3.01 (135)[d]

[a]Solid waste generated through use of catalytic incinerators.

[b]Based on 2.25 ft^3 of catalyst/1,000 scfm of catalytic incinerator throughput and a three year catalyst life. See Tables 6-1 through 6-13, Section 4.2.2.4.2 and Section 8.1.3.1.

[c]See Table 3-38.

[d]Where there is no value for a model plant under a particular regulatory alternative, the value of the preceeding regulatory alternative for that model plant category is used.

Table 7-5. VOLUME OF BIOLOGICAL SLUDGE GENERATED BY
PROCESS OPERATIONS IN NEW POLYMER AND RESINS PLANTS THAT EMPLOY
FLARES, THERMAL INCINERATION, OR CATALYTIC INCINERATION

Model Plant	Model Plant Annual Production Rate Gg/yr (lb/yr) [a]		Number of New Plants[b]	Industrywide Biological Sludge Production Rate - m³/yr (ft³/yr)[c]
PP/Liquid	150	(331×10^6)	3	4,870-12,200(172,000-431,000)
PP/Gas	105	(231×10^6)	3	4,530-11,370(160,000-402,000)
LDPE/Liquid	280	(617×10^6)	1	3,396-8,528(120,000-302,000)
LDPE/HDPE/ Gas	150	(331×10^6)	10	16,231-40,692(573,000-1,438,000)
HDPE/Liquid Slurry	210	(463×10^6)	2	4,540-11,400(160,000-403,000)
HDPE/Liquid Solution	90	(198×10^6)	3	2,910-7,310(103,000-258,000)
PS	75	(165×10^6)	2	1,618-5,680(57,200-200,000)
PET/DMT	105	(231×10^6)	1	2,840-11,400(100,000-403,000)
PET/TPA	105	(231×10^6)	2	5,680-22,800(200,000-806,000)
Total	-	-	27	47,500-128,000(168,000-453,000)
				22,161-55,592$(1.59 \times 10^6$-$4.51 \times 10^6)$[d]

[a] From Tables 6-1 through 6-9.

[b] From Table 8-38.

[c] Based on 49-123 m^3 sludge/yr/10 x 10^6 lbs. of polypropylene, polyethylene, and polystyrene produced; 123-493 m^3 sludge/yr/10 x 10^6 lbs of polyester produced. (Source: Reference 7)

[d] These totals include only sludge generated by process operations employing catalytic incineration under the regulatory alternatives.

Solid wastes generated by fugitive VOC leak detection and repair programs include replaced mechanical seals, seal packing, rupture disks, and valves. The solid waste impacts of fugitive VOC emission reduction programs are not anticipated to be significant because of the ability to recycle metal solid wastes and the small quantity of wastes generated.

7.4 ENERGY IMPACTS

7.4.1 Model Plant Energy Impacts

The energy impacts of the regulatory alternative for each model plant are calculated as the sum of energy expended to control VOC emissions from process operations and the energy saved by implementing fugitive VOC emission control practices. For the purposes of this analysis, energy impacts of process emission control are assumed to consist of natural gas consumption by flares, thermal incinerators, and catalytic incinerators, electricity consumption by thermal incinerators, catalytic incinerators, and condensers, and steam consumption by flares. The methods used to calculate energy consumption are discussed in Section 8.1. The energy impacts of controlling process emissions of VOC from each model plant under each regulatory alternative are presented in Table 7-6.

Energy impacts associated with process VOC emission control generally increase when moving from a less restrictive regulatory alternative to a more restrictive alternative because additional energy-consuming control techniques are used to reduce VOC emissions. However, energy impacts associated with control of process emissions from PP/Liquid model plants operating under Regulatory Alternative 4 are less than Regulatory Alternative 3 impacts because control techniques that are more energy-efficient are used (thermal incinerators rather than flares).

Energy impacts associated with control of fugitive VOC emissions are presented as negative values (energy credits or positive impacts) in Table 7-6. The energy credit for reduced VOC loss under fugitive VOC control practices is greater in absolute terms than the energy cost of the fugitive VOC control. Energy impacts of fugitive VOC emission control are constant once the leak detection and repair program is implemented, as the leak detection inspection intervals, equipment specifications, and resulting emission rates for each model plant type do not change under Regulatory Alternatives II through IV.

Table 7-6a. ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN MODEL PLANTS

Model Plant	Regulatory Alternative	Process Emissions Control Energy Impact (GJ/yr)				Fugitive Emission Control Energy Impacts		Combined Emission Control Energy Impacts GJ/yr
		Natural Gas[a]	Electricity[b]	Steam[c]	Process Total[d]	Mg VOC reduced	GJ/yr[e]	
PP/Liquid	1	760	-	6,860	7,610	0	0	7,610
	2	760	-	6,860	7,610	59	-2,700	4,910
	3	760	41	6,860	7,660	59	-2,700	4,960
	4	-	310	-	310	59	-2,700	-2,390
PP/Gas	1	2,260	-	180	2,440	0	0	2,440
	2	2,260	-	180	2,440	59	-2,700	- 260
	3	2,260	-	3,530	5,790	59	-2,700	3,090
LDPE/Liquid	1	1,510	-	653,200	654,700	0	0	654,700
	2	1,510	-	653,200	654,700	59	-2,780	651,900
	3	5,580	41	653,200	658,800	59	-2,780	656,000
	4	6,580	310	653,200	660,100	59	-2,780	657,300
LDPE/HDPE/Gas	1	4,660	-	4,660	9,320	0	0	9,320
	2	4,660	-	4,660	9,320	59	-2,780	6,540
	3	4,660	-	5,130	9,790	59	-2,780	7,010
	4	4,660	-	5,130	9,790	59	-2,780	7,010
HDPE/ Liquid Slurry	1	760	-	4,490	5,250	0	0	5,250
	2	760	-	4,490	5,250	59	-2,780	2,470
	3	760	-	4,590	5,350	59	-2,780	2,570
HDPE/ Liquid Solution	1	760	-	4,660	5,420	0	0	5,420
	2	760	-	4,660	5,420	59	-2,780	2,640
	3	89,600	7,810	4,660	102,200	59	-2,780	99,400

Table 7-6a. ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN MODEL PLANTS
(concluded)

Model Plant	Regulatory Alternative	Process Emissions Control Energy Impact (GJ/yr)				Fugitive Emission Control Energy Impacts		Combined Emission Control Energy Impacts GJ/yr
		Natural Gas[a]	Electricity[b]	Steam[c]	Process Total[d]	Mg VOC reduced	GJ/yr[e]	
PS[f]	1	-	-	-	0	0	0	0
	2	-	-	-	0	59	-2,410	-2,410
	3	-	30	-	-9,260[g]	59	-2,410	-11,670
PE1/DMT	1	0	182	2.04×10^5	204,200	0	0	204,200
	2	0	13,650	2.04×10^5	200,000[h]	0	0	200,000
	3	755	13,650	2.04×10^5	200,300[h]	0	0	200,300
PE1/TPA	1	0	182	2.04×10^5	204,200	0	0	204,200
	2	0	13,650	2.04×10^5	200,000[h]	0	0	200,000

[a] 10^6 Btu = 1.0551 GJ. This factor includes the energy conversion efficiency; it is not an energy equivalent. (Reference 12.)

[b] 1 kWh = 9.476 x 10^{-3} GJ. This factor includes the energy conversion efficiency; it is not an energy equivalent. (Reference 12.)

[c] 1,000 lb steam = 1.525 GJ This factor includes the energy conversion efficiency; it is not an energy equivalent. [0.2482 bbl distilled fuel oil/1,000 lb steam].

[d] 1 GJ 0.1628 bbl crude oil. (Reference 13.)

[e] Based on the following: 17,598 Btu/lb of styrene monomer; 19,683 Btu/lb of propylene; 20,276 Btu/lb of ethylene; 7,810 Btu/lb of ethylene glycol. (Reference 14.)

[f] Estimates of energy use under baseline control are not available.

[g] Includes energy credit for recovered styrene monomer.

[h] Includes energy credit for recovered ethylene glycol.

Note: Values on some lines may not total exactly due to rounding. Negative values indicate energy credits.

Table 7-6b. ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN MODEL PLANTS

Model Plant	Regulatory Alternative	Process Emissions Control Energy Impact				Fugitive Emissions Control Energy Impacts		Combined Emission Control Energy Impacts bbl oil/yr
		Natural Gas[a] 10^6 Btu/yr	Electricity[b] kWh/yr	Steam[c] 1,000 lb/yr	Process Total[d] bbl oil/yr	Tons VOC reduced	bbl oil/yr[e]	
PP/Liquid	1	720	-	4,500	1,240	0	0	1,240
	2	720	-	4,500	1,240	65	-440	800
	3	720	4,290	4,500	1,250	65	-440	810
	4	-	33,060	-	50	65	-440	-390
PP/Gas	1	2,140	-	120	400	0	0	400
	2	2,140	-	120	400	65	-440	- 40
	3	2,140	-	2,310	940	65	-440	500
LDPE/Liquid	1	1,430	-	428,300	106,600	0	0	106,600
	2	1,430	-	428,300	106,600	65	-450	106,100
	3	5,280	4,290	428,300	107,200	65	-450	106,800
	4	6,240	33,060	428,300	107,500	65	-450	107,000
LDPE/ HDPE/Gas	1	4,420	-	3,060	1,520	0	0	1,520
	2	4,420	-	3,060	1,520	65	-450	1,070
	3	4,420	-	3,370	1,590	65	-450	1,140
	4	4,420	-	3,370	1,590	65	-450	1,140
HDPE/ Liquid Slurry	1	760	-	2,940	850	0	0	850
	2	720	-	2,940	850	65	-450	400
	3	720	-	3,010	870	65	-450	420
HDPE/ Liquid Solution	1	720	-	3,060	880	0	0	880
	2	720	-	3,060	880	65	-450	430
	3	89,610	834,700	3,060	16,640	65	-450	16,200

Table 7-6b. ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN MODEL PLANTS
(concluded)

Model Plant	Regulatory Alternative	Process Emissions Control Energy Impact				Fugitive Emissions Control Energy Impacts		Combined Emission Control Energy Impacts bbl oil/yr
		Natural Gas[a] 10^6 Btu/yr	Electricity[b] kWh/yr	Steam[c] 1,000 lb/yr	Process Total[d] bbl oil/yr	Tons VOC reduced	bbl oil/yr[e]	
PS[f]	1	–	–	–	0	0	0	0
	2	–	–	–	0	65	-390	-390
	3	–	3,060	–	-150[g]	65	-390	-540[f]
PET/DMT	1	0	19,200	1.34×10^5	33,300	0	0	33,300
	2	0	1.44×10^6	1.34×10^5	32,600[h]	0	0	32,600[g]
	3	708	1.44×10^6	1.34×10^5	32,600[h]	0	0	32,600[g]
PET/TPA	1	0	19,200	1.34×10^5	33,300	0	0	33,300
	2	0	1.44×10^6	1.34×10^5	32,600[h]	0	0	32,600[g]

[a] 10^6 Btu = 1.0551 GJ. This factor includes the energy conversion efficiency; it is not an energy equivalent.

[b] 1 kWh = 9.476×10^{-3} GJ. This factor includes the energy conversion efficiency; it is not an energy equivalent.

[c] 1,000 lb steam = 1.525 GJ This factor includes the energy conversion efficiency; it is not an energy equivalent. [0.2482 bbl distilled fuel oil/1,000 lb steam].

[d] 1GJ = 0.1628 bbl crude oil. (Reference 13)

[e] Based on the following: 17,598 Btu/lb of styrene monomer; 19,683 Btu/lb of propylene; 20,276 Btu/lb of ethylene; 7,810 But/lb of ethylene glycol.

[f] Estimates of energy use under baseline control are not available.

[g] Includes energy credit for recovered styrene monomer.

[h] Includes energy credit for recovered ethylene glycol.

Note: Values on some lines may not total exactly due to rounding. Negative values indicate energy credits.

Table 7-6 shows that the energy impact of process VOC emission reduction in a model plant generally increases in moving from a less stringent to a more stringent regulatory alternative. The magnitude of the total energy impact value is less than the impact of controlling only VOC emissions from process operations because the energy credit resulting from fugitive VOC emission control work practices at least partially offsets the energy impact of process emission controls.

7.4.2 Industrywide Energy Impacts

Model plant energy impacts presented in Table 7-6 are used to project industrywide energy impacts of the regulatory alternatives on new polymers and resins plants expected to be built by the year 1988. The industrywide energy impacts of process and fugitive VOC emissions control in new plants operating under each regulatory alternative are presented in Table 7-7.

Under Regulatory Alternative I, industrywide energy demand is estimated to be about 1,418 TJ/yr (231,000 bbl crude oil/yr). Under Regulatory Alternative II, industrywide energy demand is estimated to be about 1,340 TJ/yr (218,000 bbl oil/yr). The total energy demand decreases when shifting from baseline control (Regulatory Alternative I) to Regulatory Alternative II, because implementation of fugitive VOC control practices and increased recovery from process emissions in the polystyrene and poly(ethylene terephthalate) plants saves about 118 TJ/yr (19,000 bbl oil/yr). Energy impacts under Regulatory Alternative III are approximately 1,630 TJ/yr (265,000 bbl oil/yr); the increase is due to a greater energy demand in controlling process emissions. Total energy impacts on the industry under Regulatory Alternative IV are about 20 TJ/yr (3,000 bbl oil/yr) less than Regulatory Alternative III levels because of the lower energy demands to control process emissions from PP/Liquid plants.

For seven of the nine model plants, the incremental energy requirements of the most stringent regulatory alternatives are either net decreases or an increase of less than 0.1 percent of the energy required under baseline (Regulatory Alternative 1) control. For the polypropylene, gas phase and high density polyethylene solution process model plants, the incremental percent increase associated with the most stringent regulatory alternative over baseline control is large. However, the absolute size of the energy requirement compared to the total energy used in making

Table 7-73. INDUSTRYWIDE ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES NEW POLYMER AND RESIN PLANTS

Model Plant	Regulatory Alternatives	Model Plant Energy Impacts (1J/yr)[a]			Number of New Plants[b]	Industrywide Energy Impacts (1J/yr)		
		Process	Fugitive[c]	Combined[c]		Process	Fugitive[c]	Combined[c]
PP/Liquid	1	7.61	0	7.61	3	22.8	0	22.8
	2	7.61	-2.70	4.91		22.8	-8.1	14.7
	3	7.66	-2.70	4.96		23.0	-8.1	14.9
	4	0.31	-2.70	-2.39		0.93	-8.1	-7.2
PP/Gas	1	2.44	0	2.44	3	7.32	0	7.32
	2	2.44	-2.70	-0.26		7.32	-8.1	0.8
	3	5.79	-2.70	3.09		17.37	-8.1	9.3
LDPE/Liquid	1	654.7	0	654.7	1	654.7	0	654.7
	2	654.7	-2.78	651.92		654.7	-2.78	651.9
	3	658.8	-2.78	656.02		658.8	-2.78	656.0
	4	660.1	-2.78	657.32		660.1	-2.78	657.3
LDPE/HDPE/Gas	1	9.32	0	9.32	10	93.2	0	93.2
	2	9.32	-2.78	6.54		93.2	-27.8	65.4
	3	9.79	-2.78	7.01		97.9	-27.8	70.1
	4	9.79	-2.78	7.01		97.9	-27.8	70.1
HDPE/ Liquid-Slurry	1	5.25	0	5.25	2	10.5	0	10.5
	2	5.25	-2.78	2.47		10.5	-5.56	4.9
	3	5.35	-2.78	2.57		10.7	-5.56	5.1
HDPE/ Liquid Solution	1	5.42	0	5.42	3	16.3	0	16.3
	2	5.42	-2.78	2.64		16.3	-8.34	7.9
	3	102.2	-2.78	99.42		306.6	-8.34	298.3
PS/Continuous	1	-	0	0	2	0	0	0
	2	-	-2.41	-2.41		0	-4.82	-4.82
	3	-9.26	-2.41	-11.67		-18.52	-4.82	-23.34
PET/DMT	1	204.2	0	204.2	1	204.2	0	204.2
	2	200.0	0	200.0		200.0	0	200.0
	3	200.3	0	200.3		200.3	0	200.3
PET/TPA	1	204.2	0	204.2	2	408.4	0	408.4
	2	200.0	0	200.0		400.0	0	400.0

Table 7-7a. INDUSTRYWIDE ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES NEW POLYMER AND RESIN PLANTS
(concluded)

Model Plant	Regulatory Alternatives	Model Plant Energy Impacts[a] (TJ/yr)			Number of New Plants[b]	Industrywide Energy Impacts (TJ/yr)		
		Process	Fugitive[c]	Combined[c]		Process	Fugitive[c]	Combined[c]
Total[d]	I	–	–	–	27	1417.5	0	1417.5
	II	–	–	–		1404.8	-65.5	1339.3
	III[e]	–	–	–		1696.2	-65.5	1630.7
	IV[e]	–	–	–		1675.4	-65.5	1609.9

[a] From Table 7-6.
[b] From Table 8-38.
[c] Negative values indicate energy credits.
[d] Totals do not include baseline control energy use for polystyrene plants.
[e] Where there is no value for a plant under a particular Regulatory Alternative, the value of the preceeding Regulatory Alternative for the model plant category is used.
NOTES: Dashes indicate that the entry is not applicable.
Values on some lines may not total exactly due to rounding.

Table 7-7b. INDUSTRYWIDE ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES NEW POLYMER AND RESIN PLANTS

Model Plant	Regulatory Alternatives	Model Plant Energy Impacts[a] (1,000 bbl oil/yr)			Number of New Plants[b]	Industrywide Energy Impacts[c] (1,000 bbl oil/yr)		
		Process	Fugitive[c]	Combined[c]		Process	Fugitive[c]	Combined[c]
PP/Liquid	1	1.24	0	1.24	3	3.72	0	3.72
	2	1.24	-0.44	0.80		3.72	-1.32	2.40
	3	1.25	-0.44	0.81		3.75	-1.32	2.43
	4	0.05	-0.44	-0.39		0.15	-1.32	-1.17
PP/Gas	1	0.40	0	0.40	3	1.2	0	1.20
	2	0.40	-0.44	-0.04		1.2	-1.32	-0.12
	3	0.94	-0.44	0.50		2.82	-1.32	1.50
LDPE/Liquid	1	106.6	0	106.6	1	106.6	0	106.6
	2	106.6	-0.45	106.2		106.6	-0.45	106.2
	3	107.2	-0.45	106.8		107.2	-0.45	106.8
	4	107.5	-0.45	107.0		107.5	-0.45	107.0
LDPE/HDPE/Gas	1	1.52	0	1.52	10	15.2	0	15.2
	2	1.52	-0.45	1.07		15.2	-4.5	10.7
	3	1.59	-0.45	1.14		15.9	-4.5	11.4
	4	1.59	-0.45	1.14		15.9	-4.5	11.4
HDPE/ Liquid-Slurry	1	0.85	0	0.85	2	1.70	0	1.7
	2	0.85	-0.45	0.40		1.70	-0.90	0.80
	3	0.87	-0.45	0.42		1.74	-0.90	0.84
HDPE/ Liquid Solution	1	0.88	0	0.88	3	2.64	0	2.64
	2	0.88	-0.45	0.43		2.64	-1.35	1.29
	3	16.64	-0.45	16.19		49.92	-1.35	48.57
PS/Continuous	1	-	0	0	2	0	0	0
	2	-	-0.39	-0.39		0	-0.78	-0.78
	3	0.005	-0.39	-0.39		0.01	-0.78	-0.77
PET/DMT	1	33.3	0	33.3	1	33.3	0	33.3
	2	35.5	-0.18	35.3		35.5	-0.18	35.3
	3	35.6	-0.18	35.4		35.6	-0.18	35.4
PET/TPA	1	33.3	0	33.3	2	66.6	0	66.6
	2	35.5	0	35.5		71.0	0	71.0

Table 7-7b. INDUSTRYWIDE ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES NEW POLYMER AND RESIN PLANTS
(concluded)

Model Plant	Regulatory Alternatives	Model Plant Energy Impacts[a] (1,000 bbl oil/yr)			Number of New Plants[b]	Industrywide Energy Impacts (1,000 bbl oil/yr)		
		Process	Fugitive[c]	Combined[c]		Process	Fugitive[c]	Combined[c]
Total[d]	I	-	-	-		231	0	231
	II	-	-	-	27	229	-10.7	218
	III[e]	-	-	-		276	-10.7	265
	IV[e]	-	-	-		273	-10.7	262

[a] From Table 7-6.
[b] From Table 8-38
[c] Negative values indicate energy credits
[d] Totals do not include baseline control energy use for polystyrene plants.
[e] Where there is no value for a plant under a particular Regulatory Alternative, the value of the preceeding Regulatory Alternative for the model plant category is used.
NOTES: Dashes indicate that the entry is not applicable.
Values on some lines may not total exactly due to rounding.

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the polymer is small. For example, the HDPE, solution process model plant would use approximately 23.8×10^6 Gg/yr (3.88×10^9 bbl oil/yr).[15] Total energy consumed by process emission controls under the most stringent alternative for this model plant is 102 TJ/yr (16.6×10^6 bbl oil/yr), less than 0.5 percent of the energy required for the production of the polymer.

7.5 OTHER ENVIRONMENTAL IMPACTS

7.5.1 Noise Impacts

Flares can be a source of noise pollution. Noise generated during flaring results from unsteadiness in the combustion process and steam injection.[16,17] Noise levels in excess of 100 decibels (dB) have been recorded within 100 meters (330 feet) of the flare tip,[18] but attenuate with increasing distance from the flare.[19]

. Noise from combustion by a flare can be reduced by partially enclosing the flare tip with acoustical materials, though this practice may not be feasible for elevated flares: the enclosure's useful working life may be shortened because of direct flame impingement, high exit gas velocities, and openness to weather conditions. Noise from steam injection can be minimized by use of properly positioned multiport steam jets, partial enclosure of the elevated flare tip, or use of shrouding and acoustical barriers for the steam jets and injectors.[17] Noise impacts also can be mitigated by placing the flare as far as practicable from the plant boundaries, though this practice does not attenuate lower frequency noise (less than 500 HZ) as effectively as higher frequency noise.[19] Thus, by employing the proper flare design and site selction, potential noise impacts on community areas surrounding each affected polymers and resins plant should be minimal.

7.5.2 Irreversible and Irretrievable Commitment of Resources

In general, Regulatory Alternative II will require no process VOC emission controls beyond those required under current practices (Regulatory Alternative I). As indicated in Section 7.4, the fugitive VOC leak detection and repair program implemented under Regulatory Alternative II will result in energy savings. Thus, Regulatory Alternative II appears to require no irreversible and irretrievable commitment of resources beyond what is currently committed under industry practices.

Regulatory Alternatives III and IV require installation of additional pollution control equipment to reduce VOC emissions from process operations. Consequently, demand for metals, refractory, electrical equipment, and other raw materials needed to manufacture VOC emission control equipment will increase, though it is unlikely that the amount of resources used by the 27 new plants to meet Regulatory Alternatives III and IV will be significant in comparison to industrywide consumption of the same resources. It is possible that some materials used to construct the additional VOC controls ultimately will be salvaged and recycled at the end of the control device's useful life, thereby reducing the amount of resources permanently committed to control of VOC emissions. Guywire-supported flares and settling ponds for scrubber wastewater may require considerable land space,[20-22] but are not expected to commit significant additional areas of land than what is currently used for manufacturing operations at polymers and resins plant sites. Therefore, no irreversible and irretrievable commitment of resources is expected in meeting the require- ·ments of Regulatory Alternatives III and IV.

7.5.3 Environmental Impacts of Delayed Regulatory Action

The 5-year impact of delaying implementation of any regulatory action regarding control of VOC from new polymers and resins plants depends on the regulatory alternative considered. Assuming that current industry VOC control practices (Regulatory Alternative I) are maintained through year 1988, nationwide VOC emissions may increase by as much as 4,340 mg/yr (4,790 tons/yr) in the absence of Regulatory Alternative II, and as much as 14,640 mg/yr (16,220 tons/yr) in the absence of Regulatory Alternative IV. Consequently, a delay in implementation of any regulatory action beyond that prescribed under Regulatory Alternative I may effectively increase the nationwide level of VOC emissions from polymers and resins manufacturing by 20 to 65 percent.[23]

Delay of any regulatory action for VOC emissions from the polymers and resins industry probably will have a negative impact on water quality, as leak detection and repair of fugitive VOC emission sources can reduce the amount of VOC contained in runoff.[24] No negative solid waste impacts are anticipated by delay of regulatory action: as discussed in Section 7.3, the quantity of solid waste generated increases as a result of implementing the regulatory alternatives.

Table 7-7 indicates that a delay in implementing Regulatory Alternative II may result in an adverse energy impact of 78 TJ/yr (13,000 bbl crude oil/yr). However, delay in implementing Regulatory Alternatives III and IV may result in energy savings of 213 to 192 TJ/yr (34,000 to 31,000 bbl oil/yr) when compared to baseline control.

7.6 REFERENCES

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8.0 COSTS

This chapter presents assumptions, procedures, and results of the analysis to estimate the costs of controlling volatile organic compounds (VOC) emissions from the polymers and resins industry. The results are estimates of capital costs, annualized costs, and incremental costs of emission reductions, both from the baseline and from each succesively less stringent regulatory alternative described in Chapter 6 and summarized in Table 8-1. The cost impacts of environmental regulations other than the NSPS are also discussed.

8.1 COST ANALYSIS OF REGULATORY ALTERNATIVES

The cost analysis consists of two steps for each control system: designing a system that will reliably maintain the desired efficiency and estimating capital and operating costs for such a system. Designing a control system for process VOC emissions requires an analysis of the waste gas characteristics of the combined stream to each control device specified in a regulatory alternative. The stream characteristics along with mass and energy balances are the basis for determining the equipment sizes, operating parameters, and operating requirements (e.g., fuel). For fugitive VOC emissions, requirements and costs of a leak detection and repair (LDAR) program were developed for an assumed equipment count[1] based on the analysis for control of fugitive emissions from the synthetic organic chemical manufacturing industry (SOCMI).[2]

Once these control system parameters have been determined, then the capital and annual costs can be calculated. The capital cost estimates for the implementation of the regulatory alternatives include purchase and installation of the control or monitoring devices and piping systems necessary for proper control of process and fugitive VOC emissions from each model plant.

All process VOC control capital costs are converted to June 1980 dollars using the plant cost indices published in the Chemical

Table 8-1. SUMMARY OF REGULATORY ALTERNATIVES FOR THE MODEL PLANTS

Model Plant/ Regulatory Alternative	Process Sections Controlled[1,2]						Fugitive Emission Regulatory Control	Annual Emission Reduction[3]	
	RMP	PR	HR	PF	PS	Other		Mg/yr	Percent
Polypropylene, liquid phase									
1 (Baseline)		F	F				No	5,060	89
2		F	F				Yes	5,120	90
3		C	C	C			Yes	5,500	97
4	C	C	C	C			Yes	5,510	97.2
Polypropylene, gas phase									
1 (Baseline)		F					No	1,230	31
2		F					Yes	1,290	32
3		C	C				Yes	3,365	97
Low density, polyethylene, liquid phase									
1 (Baseline)						F	No	100	10
2						F	Yes	160	17
3					C	C	Yes	380	39
4				C	C	C	Yes	435	45
Low density and high density polyethylene, gas phase									
1 (Baseline)	F[4]	F[4]		F			No	3,455	95
2	F[4]	F[4]		F			Yes	3,515	96.5
3	C	C		C			Yes	3,520	96.7
4	C	C		C	C		Yes	3,525	96.8
High density polyethylene, slurry process									
1 (Baseline)	F		F				No	2,700	92
2	F		F				Yes	2,750	94
3	C		C	C			Yes	2,840	96
High density polyethylene, solution process									
1 (Baseline)	F	F	F				No	2,700	90
2	F	F	F				Yes	2,750	92
3	C	C	C	C			Yes	2,890	96

Table 8-1. SUMMARY OF REGULATORY ALTERNATIVES FOR THE MODEL PLANTS (Concluded)

Model Plant/	Process Sections Controlled[1,2]						Fugitive Emission Regulatory Control	Annual Emission Reduction[3]	
Regulatory Alternative	RMP	PR	MR	PF	PS	Other		Mg/yr	Percent
Polystyrene, continuous process									
1 (Baseline)		R					No	45	11
2		R					Yes	105	26
3		R	RR				Yes	330	84
Poly(ethylene terephthalate), dimethyl terephthalate									
1 (Baseline)	R	R					N.A.[5]	0	0
2	R	RR					N.A.	975	96
3	R	RR	C				N.A.	995	98
Poly(ethylene terephthalate), terephthalic acid									
1 (Baseline)	R	R					N.A.	0	0
2	RR	RR					N.A.	970	97

[1]Process Sections:

RMP - raw materials preparation (including esterification in poly(ethylene terephthalate) plants

PR - polymerization reaction
PF - product finishing
PS - product storage

Other - all emergency vents other than from the reactor

[2]Control Devices:

F = flare.
C = combustion devices including flares, incinerators, boilers.
R = recovery devices used under baseline as part of process.
RR = additional recovery devices or more efficient recovery system.

[3]From uncontrolled levels.

[4]Flaring of intermittent streams only.

[5]N.A. = not applicable.

Engineering Economic Indicators. The installed capital costs for process controls represent the total investment, including indirect costs such as engineering and contractors' fees and overhead, required for purchase and installation of all equipment and materials for the control systems. These are battery-limit costs and do not include any provisions for bringing utilities, services, or roads to the site, or for any backup facilities, land, research and development required, or for any process piping and instrumentation interconnections that may be required within the process generating the waste gas. The installation factors assumed for the various process control devices are presented in Table 8-2. Actual direct and indirect cost factors depend upon the plant specific conditions and may vary with the size of the system. The annualized costs consist of the direct operating and maintenance costs, including labor, utilities, fuel, and materials for the control system, and indirect costs for overhead, taxes, insurance, administration, and the capital recovery charges. The utilities considered include natural gas and electricity. The annualized cost factors that are used to analyze all of the process (and fugitive) VOC control systems are summarized in Table 8-3.

Although fugitive control costs are in May 1980 rather than June 1980 dollars, the costs are considered, for all practical purposes, to be on the same time basis as the ones for process VOC controls.[2] The fugitive control capital cost includes costs for monitoring instruments, caps for open-ended lines, piping for the connections to an existing enclosed combustion device or vapor recovery header, rupture disk assemblies, closed-loop sampling connections, and initial leak repair.[2] The derivation of annualized labor, administrative, maintenance, and capital costs for fugitive control are presented in Section 8.1.6.

The following sections outline the design and costing procedures developed for flares, thermal incinerators, catalytic incinerators, condensers, ethylene glycol recovery systems, and fugitive control systems. Details of these procedures are given in Appendix E. This section presents an overview of the procedures and their important features. The results of the cost analysis for the various regulatory alternatives are also presented.

Table 8-2. INSTALLATION COST FACTORS

Installation Cost Component	Flare	Thermal Incinerator	Catalytic Incinerator[a]	Condenser	EGRS (Baseline)[b]	EGRS (Alternative)[c]
Major Equipment Purchase Price (P)	1.0	1.0	1.0	1.0	1.0	1.0
Unspecified Equipment / Total Equipment (A)	0.0/1.0	0.2/1.2	0.0/1.0	0.0/1.0	0.0/1.0	0.0/1.0
Installation Factors (Multiples of A)						
Foundations	0.06	0.10	0.03	–	0.03	0.06
Structures	0.01	0.03	–	–	0.20	0.30
Equipment Erection	0.15	0.26	0.05	0.02	0.15	0.25
Piping	0.20	0.34	0.10	0.10	0.40	0.60
Insulation	–	0.10	0.04	–	0.05	0.10
Paint	0.02	0.09	0.01	0.01	0.03	0.05
Fire Protection	–	0.02	0.01	–	0.01	0.01
Instruments	0.05	0.26	0.05	–	0.15	0.25
Electrical	0.05	0.09	0.05	–	0.10	0.15
Sales Tax	0.06	0.08	0.06	0.05	0.08	0.09
Freight	0.10	0.16	0.08	0.06	0.16	0.16
Contractor's Fee	0.16	0.39	0.12	0.05	0.34	0.53
Engineering	0.09	0.25	0.08	0.03	0.27	0.35
Contingencies	0.15	0.17	0.08	0.03	0.27	0.35
TOTAL	2.10 A / 2.10 P	3.33 A / 4.00 P	1.82 A / 1.82 P	1.39 A / 1.39 P	3.24A / 3.24P	4.26A / 4.26P

[a] Skid mounted - requiring little foundation and erection allowance.

[b] Baseline ethylene glycol recovery system.

[c] Alternative ethylene glycol recovery system.

TABLE 8-3. ANNUALIZED COST FACTORS FOR POLYMERS AND RESINS NSPS
(June 1980 Dollars)

Direct Cost Factors

Operating labor price: \$18/hr (including overhead)[a]

Operating labor requirements (including supervisory labor):

- = 620 labor hours/yr for flare[b]
- = 1200 labor hours/yr for thermal incinerator without heat recovery[c]
- = 1500 labor hours/yr for thermal incinerator with heat recovery[c]
- = 620 labor hours/yr for catalytic incinerator without heat recovery[b]
- = 930 labor hours/yr for catalytic incinerator with heat recovery[d]
- = 60 labor hours/yr for condenser[e]
- = 4300 labor hours/yr for ethylene glycol recovery baseline system[f]
- = 8600 labor hours/yr for ethylene glycol recovery regulatory alternative system[g]

Electricity price: \$0.049/kwh[h]

Natural gas price: \$5.67/GJ (\$5.98/MMBtu)[i]

Steam price: \$13.62/Mg (\$6.18/1000 lb)[j]

Water price: \$0.079/m^3 (\$0.30/1000 gal)[k]

Styrene recovery credit: \$0.788/kg (\$0.357/lb)[l]

Ethylene glycol recovery credit: \$0.33/kg (\$0.15/lb), for 80 percent purity[m]

Ethylene glycol recovery credit: \$0.60/kg (\$0.27/lb), for high purity[n]

Indirect Cost Factors

Interest rates:

    8.5 percent (after taxes)
    10 percent (before taxes)

Equipment life, N:[o]

    15 years for flare, ethylene glycol recovery system
    10 years for thermal incinerator, catalytic incinerator, condenser, piping

$$\text{Capital recovery charge factor} = \frac{i(1 + i)^N}{(1 + i)^N - 1}$$

TABLE 8-3. ANNUALIZED COST FACTORS FOR POLYMERS AND RESINS NSPS
(June 1980 Dollars)
(Concluded)

___

Indirect Cost Factors (Con't)

0.131 for flare
0.163 for thermal incinerator, catalytic incinerator, condenser, ethylene glycol recovery system, piping
0.244 overall for fugitive (varies with individual equipment components)

Taxes, insurance, and administration: 0.04 x Total installed capital cost[p]

Maintenance cost: 0.05 x Total installed capital cost[q]
0.085 x Total installed capital cost, overall for fugitive VOC control (varies from zero to estimates of actual materials costs, although the 0.05 factor is used in general)

Operating hours: 8600 hours/yr

___

[a]Includes wages plus 40 percent for labor-related administrative and overhead costs.

[b]0.5 man-hours/shift x 8600 hrs/yr ÷ 8 hrs/shift + 15 percent of the operating labor for supervisory costs.

[c]Blackburn, J.W. Control Device Evaluation: Thermal Oxidation, Report No. 1 in Organic Chemical Manufacturing, Volume 4. U.S. Environmental Protection Agency. Research Triangle Park, N.C. Publication No. EPA-450/3-80-026. December 1980.

[d]0.75 man-hours/shift x 8600 hrs/yr ÷ 8 hrs/shift + 15 percent of the operating labor for supervisory costs.

[e]1 man-hour/week x 8600 hrs/yr ÷ 8 hrs/shift ÷ 21 shifts/week + 15 percent of operating labor for supervisory costs.

[f]4 man-hours/shift x 8600 hrs/yr ÷ 8 hrs/shift.

[g]Number of man-hours/year provided by an industry source.

[h]Memo from Chasko and Porter, EPA, September 17, 1980. Guidance for developing CTGD Cost Chapters.

[i]Memo from Al Wehe, to Information Analysis Working Group for the Industrial Boiler Working Group. April 23, 1981. IFCAM Modification:

Projected 1985 price in 1978 dollars is $4.91 + $.60 delivery charge per MMBtu.

Projected 1990 price in 1978 dollars is $5.55 + $0.61 delivery charge per MMBtu.

By linear interpolation between $4.91 and $5.55/MMBtu; 1988 price in 1978 dollars = $5.29/MMBtu.

Using GNP implicit price deflator index: 4th quarter 1978 of 154.99 and 2nd quarter 1980 of 175.28; 1988 price in 1980 dollars = 175.28/154.99 x 5.29 = $5.98/MMBtu.

Assumed higher heating value of 1040 Btu/scf at 16°C(60°F).

[j]Neverill, R.B. Capital and Operating Costs of Selected Air Pollution Control Systems. U.S. Environmental Protection Agency. Research Triangle Park, N.C. Publication No. EPA-450/5-80-002. December 1978. p. 3-12:

$5.04/1000 lb steam, 4th quarter 1977.

Using GNP implicit price deflator index: 4th quarter 1977 of 142.91 and 2nd quarter 1980 of 175.28; updated steam price = 175.28/142.91 x $5.04 = $6.18/1000 lb steam.

[k]Peters, M.S. and K.D. Timmerhaus. Plant Design and Economics for Chemical Engineers. McGraw-Hill Book Co. New York, N.Y. Third Edition. 1980. p. 881.

[l]90 percent of styrene price given in Chemical Marketing Reporter

[m]Assumed based on polymer raw material (high purity) ethylene glycol price of 27¢/lb.

[n]Polymer raw material grade of $0.27/lb of ethylene glycol given in Chemical Marketing Reporter.

[o]Average equipment lives given by Neverill in reference cited in i., p. 3-16.

[p]Fugitive Emission Sources of Organic Compounds -- Additional Information on Emissions, Emission Reductions, and Costs. U.S. Environmental Protection Agency. Research Triangle Park, N.C. Publication No. EPA-450/3-82-010. April 1982. p. 5-16.

[q]Per reference cited in footnote n:

9 percent of total installed capital costs for maintenance and miscellaneous charges - 4 percent of total installed capital costs for taxes, insurance and administration equivalent to miscellaneous).

8.1.1 Flare Design and Cost Basis

Elevated flares were costed based upon state-of-the-art industrial design. Associated piping and ducting from the process sources to a header and from a header to the flare were conservatively designed for costing purposes. Operating costs for utilities were based on industry practice.

8.1.1.1 Flare Design. Design of flare systems for the combinations of waste streams was based on standard flare design equations for diameter and height presented by IT Enviroscience.[3] These equations were simplified to functions of the following waste gas characteristics: volumetric flow rate, lower heating value, temperature, and molecular weight. The diameter equation is based on the equation of flow rate with velocity times cross-sectional area. A minimum commercially available diameter of 2 inches was assumed. The height correlation premise is design of a flare that will not generate a nonlethal radiative heat level (1500 Btu/ft^2 hr, including solar radiation[4]) at the base of the flare (considering the effect of wind). Heights in 5-foot multiples with a minimum of 30 ft. were used.[5]

Supplemental fuel, natural gas, is added to increase the heating value to 115 Btu/scf to ensure combustion.[6] For flares with diameters of 24-inches or less, this natural gas was assumed to be premixed with the waste gas and to exit out the stack. For larger flares, a gas ring at the flare tip was assumed because such separate piping is more economical than increasing the flare stack size for large diameter.

Purge gas also may be required to prevent air intrusion and flashback. A purge velocity requirement of 1 fps was assumed during periods of continuous flow for standard systems without seals.[7] For flares handling only intermittent flows, purge gas requirements were assumed to be negligible according to the industry practice of not purging or perhaps purging before a planned intermittent release.[8] For combined streams with very large turndown ratios (intermittent flow : continuous flow), supplying purge gas to maintain an adequate continuous flow in a large flare (designed for the intermittent flow) can become more expensive than designing a second separate flare for the continuous flare. In such cases, a fluidic seal, which requires a greatly reduced purge rate, was used.

Natural gas consumption at a rate of 80 scfh per pilot flame to ensure ignition and combustion was assumed. The number of pilots was based on diameter according to available commercial equipment.[9]

Steam was added to produce smokeless combustion through a combined mixing and quenching effect. A steam ring at the flare tip was used to add steam at a rate of 0.4 lb steam/lb of hydrocarbons (VOC plus methane and ethane) in the continuous stream (or the intermittent stream if no continuous flow was present).[10] Availability and deliverability of this quantity of steam was assumed.

Piping (for flows less than 700 scfm) or ducting (for flows equal to or greater than 700 scfm) was designed from the process sources to a header combining the streams (via "source legs") and from the header to the base of the flare (via "pipelines"). Since it is usual industry practice, adequate pressure (approximately 3 to 4 psig) was assumed available to transport all waste gas streams without use of a compressor or fan. The source legs were assumed to be 70 feet in length,[11] while the length of pipelines to the flare was based on the horizontal distance required to provide the safe radiation level for continuous working (440 Btu/hr-ft^2, including solar radiation[12]). For flows less than 700 scfm, an economic pipe diameter was calculated based on an equation in the Chemical Engineer's Handbook[13] and simplified as suggested by Chontos.[14,15,16] The next larger size (inner diameter) of schedule 40 pipe was selected unless the calculated size was within 10 percent of the size interval between the next smaller and next larger standard sizes. For flows of 700 scfm and greater, duct sizes were calculated assuming a velocity of 2,000 fpm for flows of 60,000 acfm or less and 5,000 fpm for flows greater than 60,000 acfm. Duct sizes that were multiples of 3-inches were used. (See Section E.7 for detailed design and cost procedures for piping and ducting.)

8.1.2.1 Flare Costing. Flare purchase costs were based on costs for diameters from 2 to 24 inches and heights from 20 to 200 feet provided by National Air Oil Burner, Inc., (NAO) during November 1982.[9] A cost was also provided for one additional case of 60 inch diameter and 40 feet height.[10] These costs are October 1982 prices of self-supporting flares without ladders and platforms for heights of 40 feet and less and of guyed flares with ladders and platforms for heights of 50 feet and

greater. Flare purchase costs were estimated for the various regulatory alternatives by either choosing the value provided for the required height and diameter or using two correlations developed from the NAO data for purchase cost as a function of height and diameter. (One correlation for heights of 40 feet and less and one for heights of 50 feet and greater.) An installation factor of 2.1 (see Table 8-2) was used to estimate installed flare costs.

Piping costs were based on those given in the Richardson Engineering Services Rapid Construction Estimating Cost System[17] as combined for 70 ft. source legs and 500 ft. and 2,000 ft. pipelines for the cost analysis of the Distillation NSPS.[18] Ducting costs were calculated based on the installed cost equations given in the GARD Manual.[19] Installed costs were put on a June 1980 basis using the following Chemical Engineering Plant Cost Indices: the overall index for flares; the pipes, valves, and fittings index for piping; and the fabricated equipment index for ducting. Annualized costs were calculated using the factors presented in Table 8-3.

8.1.2 Thermal Incinerator Design and Cost Basis

For costing purposes thermal incinerator designs were based on heat and mass balances for combustion of the waste gas and any required auxiliary fuel, considering requirements of total combustion air. Associated piping, ducting, fans, and stacks were also costed.

8.1.2.1 <u>Thermal Incineration Design</u>. Designs of thermal incineration systems for the various combinations of waste gas streams were developed using a procedure based on heat and mass balances and the characteristics of the waste gas in conjunction with some engineering design assumptions. In order to ensure a 98 percent VOC destruction efficiency, thermal incinerators were designed to maintain a 0.75 second residence time at 870°C (1600°F).[20] The design procedure is outlined in this section.

In order to prevent an explosion hazard and satisfy insurance requirements, dilution air was added to any individual or combined waste stream with both a lower heating value between 13 and 50 Btu/scf at 0°C (32°F) (about 25 and 100 percent of the lower explosive limit) and an oxygen concentration of 12 percent or greater by volume. Dilution air was added to reduce the lower heating value of the stream to below

13 Btu/scf. (Adding dilution air is a more conservative assumption than the alternative of adding natural gas and is probably more realistic as other streams often have enough heat content to sustain the combustion of the combined stream for the regulatory alternative.)

The combustion products were then calculated assuming 18 percent excess air for required combustion air, but 0 percent excess air for oxygen in the waste gas, i.e., oxygen thoroughly mixed with VOC in waste gas. The procedure includes a calculation of auxiliary fuel requirements for streams (usually with heating values less than 60 Btu/scf) unable to achieve stable combustion at 870°C (1600°F) or greater. Natural gas was assumed as the auxiliary fuel as it was noted by vendors as the primary fuel now being used by industry. Natural gas requirements were calculated using a heat and mass balance assuming a 10 percent heat loss in the incinerator. Minimum auxiliary fuel requirements for low heating value streams were set at 5 Btu/scf to ensure flame stability.[21]

For streams able to maintain combustion at 870°C (1600°F), fuel was added for flame stability in amounts that provided as much as 13 percent of the lower heating value of the waste gas for streams with heating values of 650 Btu/scf or less. For streams containing more than 650 Btu/scf, flame stability fuel requirements were assumed to be zero since coke oven gas, which sustains a stable flame, contains only about 590 Btu/scf. In order to prevent damage to incinerator construction materials, quench air was added to reduce the combustion temperature to below the incinerator design temperature of 980°C (1800°F) for the cost curve given by IT Enviroscience.[22]

The total flue gas was then calculated by summing the products of combustion of the waste gas and natural gas along with the dilution air. The required combustion chamber volume was then calculated for a residence time of 0.75 sec, conservatively oversizing by 5 percent according to standard industry practice.[23] The design procedure assumed a minimum commercially available size of 1.01 m^3 (35.7 ft^3) based on vendor information[24] and a maximum shop-assembled unit size of 205 m^3 (7,238 ft^3).[25]

The design procedure would allow for pretreating of combustion air, natural gas, and when permitted by insurance guidelines, waste gas using a recuperative heat exchanger in order to reduce the natural gas required

to maintain a 870°C (1600°F) combustion temperature. However, all streams to thermal incinerators for polymers and resins regulatory alternatives had sufficient waste gas heating values to combust at 870°C (1600°F) without preheating the input streams. If a plant had a use for it, heat could be recovered. (In fact, a waste heat boiler can be used to generate steam, generally with a net cost savings.)

8.1.2.2 <u>Thermal Incinerator Costing</u>. Thermal incinerator purchase costs were taken directly from the IT Enviroscience graph for the calculated combustion chamber volume.[23] (Essentially equivalent purchase costs would be obtained by using data from the GARD manual.[20]) An installation cost factor of 4.0 was used based on the Enviroscience document (see Table 8-2).[26] The installed cost of one 150-ft. duct to the incinerator and its associated fan and stack were also taken directly from the IT Enviroscience study.[27] A minimum cost of $70,000 (December 1979 dollars) was assumed for waste gas streams with flows below 500 scfm. The costs of piping or ducting from the process sources to the 150-ft. duct costed above were estimated as for flares. Installed costs were put on a June 1980 basis using the following Chemical Engineering Plant Cost Indices: the overall index for thermal incinerators; the pipes, valves, and fittings index for piping; and the fabricated equipment index for ducts, fans, and stacks. Annualized costs were calculated using the factors in Table 8-3. The electricity required was calculated from an equation developed for the Distillation NSPS cost analysis.[28]

8.1.3 <u>Catalytic Incinerator Design and Cost Basis</u>

Catalytic incinerators are generally cost effective VOC control devices for low concentration streams. The catalyst increases the chemical rate of oxidation allowing the reaction to proceed at a lower energy level (temperature) and thus requiring a smaller oxidation chamber, less expensive materials, and much less auxiliary fuel (especially for low concentration streams) than required by a thermal incinerator. The primary determinant of catalytic incinerator capital cost is volumetric flow rate. Annual operating costs are dependent on emission rates, molecular weights, VOC concentration, and temperature. Catalytic incineration in conjunction with a recuperative heat exchanger can reduce overall fuel requirements.

8.1.3.1 <u>Catalytic Incinerator Design</u>. The basic equipment components of a catalytic incinerator include a blower, burner, mixing chamber, catalyst bed, an optional heat exchanger, stack, controls, instrumentation, and control panels. The burner is used to preheat the gas to catalyst temperature. There is essentially no fume retention requirement. The preheat temperature is determined by the VOC content of the combined waste gas and combustion air, the VOC destruction efficiency, and the type and amount of catalyst required. A sufficient amount of air must be available in the gas or be supplied to the preheater for VOC combustion. (All the gas streams for which catalytic incinerator control system costs were developed are dilute enough in air and therefore require no additional combustion air.) The VOC components contained in the gas streams include ethylene, n-hexane, and other easily oxidizable components. These VOC components have catalytic ignition temperatures below 315°C (600°F). The catalyst bed outlet temperature is determined by VOC gas content. Catalysts can be operated up to a temperature of 700°C (1,300°F). However, continuous use of the catalyst at this high temperature may cause accelerated thermal aging due to recrystallization.

The catalyst bed size required depends upon the type of catalyst used and the VOC destruction efficiency desired. Heat exchanger requirements are determined by gas inlet temperature and preheater temperature. A minimum practical heat exchanger efficiency is about 30 percent; 65 percent was assumed for this analysis. Gas temperature, preheater temperature, gas dew point temperature, and gas VOC content determine the maximum possible heat exchanger efficiency. A stack is used to vent the flue gas to the atmosphere.

Fuel gas requirements were calculated based on the heat required for a preheat temperature of 315°C (600°F), plus 10 percent for auxiliary fuel. The fuel was assumed to be natural gas, although oil (No. 1 or 2) can be used. Electricity demand was based on pressure drops of 4 inches water for systems without heat recovery and 10 inches water for systems with heat recovery, a conversion rate of 0.0001575 hp/in. water, 65 percent motor efficiency, and 10 percent additional electricity required for instrumentation, controls, and miscellaneous. A catalyst requirement of 2.25 ft^3/1,000 scfm was assumed for 98 percent efficiency.[29] Catalyst replacement every three years was assumed.

8.1.3.2 <u>Catalytic Incinerator Costing</u>. Calculations for capital cost estimates were based on equipment purchase costs obtained from vendors for all basic components and the application of direct and indirect cost factors.[29,30,31] Purchase cost equations were developed based on vendor third quarter 1982 purchase costs of catalyst incinerator systems with and without heat exchangers for sizes from 1,000 scfm to 50,000 scfm. The cost data are based on carbon steel material for incinerator systems and stainless steel for heat exchangers. Catalytic incinerator systems of gas volumes higher than 50,000 scfm can be estimated by considering two equal volume units in the system. Heat exchangers for small size systems are costly and may not be practical. The direct and indirect cost component factors used for estimating capital costs of catalytic incinerator systems with no heat exchangers and for heat exchangers were presented in Table 8-2. Installed costs of piping, ducts, fans, and stacks were estimated by the same procedure as for thermal incinerators. Installed costs were put on a June 1980 basis using the following Chemical Engineering Plant Cost indicies: the overall index for catalytic incinerators; the pipes, valves, and fittings index for piping; and the fabricated equipment index for ducts, fans, and stacks. Annualized costs were calculated using the factors in Table 8-3.

8.1.4 <u>Condenser Design and Cost Basis</u>

This section outlines the procedures used for sizing and estimating the costs of surface condenser systems applied to the gaseous streams from the continuous process polystyrene model plant.

8.1.4.1 <u>Surface Condenser Design</u>. The condenser system evaluated consists of a shell and tube heat exchanger with the hot fluid in the shell side and the cold fluid in the tube side. The condenser system, which condenses the vapors by isothermal condensation, is sized based on the total heat load and the overall heat transfer coefficient which is established from individual heat transfer coefficients of the gas stream and the coolant.

Total heat load was calculated using the following procedure: the system condensation temperature was determined from the total pressure of the gas and vapor pressure data for styrene and steam. As the vapor

pressure data are not readily available, the condensation temperature was estimated by trial-and-error using the Clausius Clapeyron equation which relates the stream pressures to the temperatures. The total pressure of the stream is equal to the vapor pressures of individual components at the condensation temperature. Once the condensation temperature was known, the total heat load of the condenser was determined from the latent heat contents of styrene and steam. The coolant is selected based on the condensation temperature.

For our purpose, no detailed calculations were made to determine the individual and overall heat transfer coefficients. Since the streams under consideration contain low amounts of styrene, the overall heat transfer coefficient was estimated based on published data for steam. Then the total heat transfer area was calculated from the known values of total heat loads and overall heat transfer coefficient using Fourier's general equation.

8.1.4.2 <u>Surface Condenser Costing</u>. Since the gas volumes of the two streams are very low, the calculated heat transfer areas are also very low (about 10 ft^2). The heat exchanger costs for each stream were obtained for minimum available size of 20 ft^2 from vendors.[32,33,34,35] An installation factor of 1.39 (See Table 8-2) was used to estimate installed condenser costs. No additional piping was costed since the condenser unit is so small (1-2 ft. diameter) that it should be able to be installed adjacent to the source.

8.1.5 <u>Ethylene Glycol Recovery System Design and Cost Basis</u>

This section briefly describes the procedures used to estimate the cost for recovering ethylene glycol (EG) using a baseline system and using an alternative ethylene glycol recovery system that is composed of an ethylene glycol spray condenser and associated equipment.

8.1.5.1 <u>Ethylene Glycol Recovery System Design</u>. The equipment selected to comprise the ethylene glycol spray condenser and both recovery systems was obtained from information provided by industry sources. The baseline system recovers ethylene golycol from the downstream of the cooling water tower. Ethylene glycol emissions from the polymerization reactor and from the distillation column recovering EG from the esterifier emissions accumulate in the cooling water and are emitted from the

cooling tower. The regulatory alternative system recovers ethylene glycol from the polymerization reactor through use of an EG spray condenser and recovers ethylene glycol from the esterifier through use of a reflux condenser. Thus, very little EG is emitted to be collected in the cooling water and later emitted to the atmosphere from the cooling water tower. Industry sources also provided basic operating and maintenance parameters. The design and operating parameters for the baseline system were used to design a system for the model plant. The information for the EG spray condenser system was provided for a PET/TPA plant of larger capacity than the model plant. The basic equipment was considered to be the same regardless of process (DMT or TPA). (The flow from the DMT process can be as much as 35 percent less than that from a TPA process; however, differences depend on the completeness of the polymerization reaction and the part of the process where VOC is emitted. Both of these factors are plant-specific.)

8.1.5.2 EG Recovery System Costing. The costs of the baseline system were estimated based on the preliminary design using standard engineering procedures as well as some industry information. The costs of the ethylene glycol spray condenser and recovery system for new plants using the DMT and the TPA processes were derived from a base set of costs provided by an industry source for a similar system on a larger capacity plant. Details of the cost estimates are given in Appendix E (Section E.6). Installation cost factors and annualized cost factors are given in Tables 8-2 and 8-3, respectively.

8.1.6 Fugitive Emission Control Program Design and Cost Basis

As noted in Section 6.2.4, a leak detection and repair program and equipment specifications was specified as the regulatory alternative beyond baseline control. This regulatory alternative was selected so that fugitive emission control in the polymers and resins industry would be consistent with other fugitive VOC emission regulations for the petroleum refining industry and the synthetic organic chemical manufacturing industry.

The following sections outline the specific requirements of the regulatory alternative and the procedures used to estimate its capital and annual costs. Detailed descriptions are found in Appendix F.

8.1.6.1 Design of Fugitive VOC Regulatory Alternative. The following equipment in the polymers and resins industry were considered for regulation: in-line process valves, pumps, compressors, safety relief valves, flanges, sampling connections, and open-ended lines. Those VOC emissions resulting from the transfer, storage, treatment, and disposal of process wastes are not covered by this regulation.

The selected leak detection and repair program and equipment specifications for the regulatory alternative are listed in Table 8-4. The control specifications are a combination of Regulatory Alternatives III and V for VOC fugitive emissions in the petroleum refining industry.[2]

The technical parameters for the polymers and resins model plant for fugitive emissions are given in Table 8-5.

8.1.6.2 Fugitive VOC Emission Control Costs. This section presents the cost estimates and the input parameters affecting the cost estimates for each of the fugitive emission sources within the polymers and resins industry. All the cost information presented here is obtained from the SOCMI BID for Model Unit B. All costs are updated to represent second quarter 1980 dollars. Table 8-6 summarizes the total cost associated with the fugitive VOC regulatory alternative and Table 8-7 summarizes the costs to control the individual equipment.

The following parts describe the cost estimation procedures and assumptions used to derive these costs as well as more detailed cost tables for each fugitive VOC emission source.

Valves

The fugitive emission control specifications for valves include monthly monitoring and leak repairs. Therefore, the annual costs associated with controls include initial leak repair costs and recurring monthly monitoring and leak repair costs.

The costs for leak detection and repair programs for valves are based on the following factors: (1) monitoring time, (2) repair time for on-line and off-line repair, and (3) fractions of leaks repaired on-line and off-line. The following estimates were used for the above:

(1) Monitoring time: The monitoring time estimate of 2 man-minutes per valve was used.

Table 8-4. FUGITIVE VOC REGULATORY ALTERNATIVE CONTROL SPECIFICATIONS[a]

| Emission source | Control Specification | |
	Inspection/ monitoring interval[b]	Equipment
Valves		
Gas	Monthly	None
Light liquid	Monthly	None
Heavy liquid	None	None
Pumps seals		
Light liquid	Monthly[c] Weekly Visual	None
Heavy liquid	None	None
Safety/relief valves		
Gas	None	Rupture disks on relief valves
Open-ended lines (purge, drain, sample lines)	None	Cap
Compressors	None	Controlled degassing vents
Sampling connections	None	Closed-purge sampling
Flanges	None	None

[a]The regulatory alternative is a combination of Regulatory Alternatives V and III for VOC Fugitive Emissions in Petroleum Refining Industry - Background Information.

[b]Fraction of sources found to be leaking by monitoring would be repaired.

[c]For pumps, monthly instrument monitoring would be supplemented with weekly visual inspections for liquid leakage. If liquid is noted to be leaking from the pump seal, the pump seal would be repaired.

Table 8-5. FUGITIVE VOC EMISSION DATA FOR THE SOURCES IN POLYMERS AND RESINS (P&R) MODEL PLANTS[a,b]

Emission source	Number of sources[f]	Uncontrolled emission factor, kg/hr/source	Annual uncontrolled emissions, Mg/yr[d]	Control method[e]	Control efficiency	Annual controlled emissions, Mg/yr[d]	Annual emission reductions due to control, Mg/yr[g]
Valves							
Gas	402	0.0056	19.72	Leak detection and repair	0.73	5.32	14.40
Light liquid	524	0.0070	32.59		0.59	13.36	19.23
Heavy liquid	524	0.00023	1.06		--	1.06	0
Pump seals							
Light liquid	29	0.0494	12.55	Leak detection and repair	0.61	4.92	7.63
Heavy liquid	30	0.0214	5.62		--	5.62	0
Safety/relief valves							
Gas	11	0.1040	10.02	Rupture disk	1.0	0	10.02
Open-ended lines	0	0.0017	0	Plugs and caps	1.0	0	0
Compressors	2	0.228	3.99	Vented seal areas	1.0	0	3.99
Sampling connections	26	0.0150	3.42	Closed purge systems	1.0	0	3.42
Flanges	2,400	0.00083	17.45	No control	--	17.45	0
Total			106.42			47.73	58.69

[a]P&R model unit is assumed to be the same as Model B of the SOCMI BID.

[b]The data are obtained from the report, "Fugitive Emission Sources of Organic Compounds--Additional Information on Emissions, Emission Reductions and Costs." U.S. Environmental Protection Agency. Report Number EPA-450/3-82-010. April 1982. p. B-2 and B-3.

[c]Number of sources left uncontrolled under baseline control level.

[d]Based on 8,760 hours in one operating year.

[e]Specified by the regulatory alternative. See Table 8-4.

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Table 3-6. FUGITIVE VOC REGULATORY ALTERNATIVE COSTS
FOR POLYMERS AND RESINS MODEL UNITS

Cost item	May 1980 dollars
1. Installed capital cost	66,800
2. Annual cost of operation	
a) Operating labor including administration and support	20,700
b) Maintenance	5,700
c) Miscellaneous[a]	2,600
d) Annualized capital costs	16,300
Total	45,300
3. VOC recovery credits[b]	(29,600)[c]
4. Net annualized cost	15,700

[a]Taxes, insurance, and administration.

[b]Based on 56.12 Mg of VOC recovered annually at a credit of $528 per Mg. (This amount, 56.12 Mg, of VOC recovered represents 96 percent of the total VOC reduction under the fugitive regulatory alternative, i.e., a 96 percent recovery efficiency is assumed.)

[c]Values in parentheses denote credits.

Table 8-7. SUMMARY OF FUGITIVE VOC EMISSION CONTROL COSTS
FOR THE SOURCES IN POLYMERS AND RESINS MODEL UNIT
(May 1980 dollars)

Cost item	VOC emission sources						
	Gas and light liquid valves	Light liquid pump seals	Gas safety/ relief valves	Compressors	Sampling connections	Leak monitoring equipment	Total
1. Installed capital cost	2,300	1,400	33,700	6,400	13,800	9,200	66,800
2. Annualized capital cost							
a) Control equipment	0	1,600	6,600	1,000	2,200	4,200	15,700
b) Initial leak repair	400	200	0				600
3. Annual operating cost							
a) Operating labor							
1. Monitoring and leak repair	10,100	4,700					14,800
2. Administration & Support	4,000	1,900					5,900
b) Maintenance			1,700	300	700	3,000	5,700
c) Miscellaneous			1,300	300	600	400	2,600
4. Total annual cost	14,500	8,400	9,600	1,600	3,500	7,600	45,300
5. VOC recovery credit	(17,700)[a]	(4,000)	(5,300)	(800)	(1,800)	0	(29,600)
6. Net annualized cost	(3,200)	4,400	4,300	800	1,700	7,600	15,700
7. Cost effectiveness per Mg of VOC	95	575	430	485	520	-	265

[a] Values in parentheses denote credit.

(2) Repair time: The repair time estimate of 10 man-minutes per valve for on-line repair of valves and 4 man-hours per valve for off-line repair of valves was used.

(3) Fraction of leaks repaired on-line and off-line: Seventy-five percent of all valves were estimated to be repaired on-line, while 25 percent were estimated to be repaired off-line.

Based on the results of the leak detection and repair (LDAR) model,[2] the annual costs of monitoring and repairing valves have been estimated for the monthly monitoring program. The cost calculations are presented in Tables 8-8 through 8-11. The input parameters, e.g., occurrence rate, initial leak frequency, etc., are discussed in the _Emission Reduction_ section (Section 4) of the SOCMI BID.[37]

Pumps

The fugitive VOC emission control specifications for light liquid pumps include monthly monitoring supplemented by weekly visual inspections and leak repairs. Therefore, the annual costs associated with controls include initial leak repair, including pump seal replacement costs, and recurring monthly monitoring and leak repair, including seal replacement costs. The factors affecting the costs of a leak detection and repair program are (1) monitoring time, (2) repair time, and (3) cost of replacement seal. The cost estimates were based on 10 man-minutes for instrument monitoring, 0.5 man-minutes for visual monitoring, and 16 man-hours per pump for repair. Every month 4.2 percent of all pump seals will be replaced as routine maintenance. On the average, half of routinely maintained seals, i.e., 2.1 percent of all seals, are assumed to be leaking seals.

Based on the LDAR model, the annual costs of leak detection and repair have been calculated for monthly programs for pumps. The calculations are presented in Tables 8-12 through 8-15. The input parameters, e.g., occurrence rate, initial frequency, and others, are discussed in the _Emission Reduction_ section (Section 4) of the SOCMI BID.

Safety/Relief Valves

The fugitive VOC emission control specifications for safety/relief valves include installation of rupture disk systems on the relief valves.

Table 8-8. INITIAL LEAK REPAIR LABOR-HOURS REQUIREMENT
FOR VALVES FOR THE MODEL UNIT

No. of valves per model	Initial leak frequency	Estimated no. of initial leaks	Repair time, man-hours[a]	Labor-hours required, man-hours
402	0.114	45.8	1.13	51.8
524	0.065	34.1	1.13	38.5
				90.3

[a]Based on 75 percent valves repaired on-line in 10 man-minutes and
25 percent repaired off-line in 4 man-hours.

Table 8-9. TOTAL ANNUAL COSTS FOR INITIAL LEAK REPAIR
FOR VALVES FOR THE MODEL UNIT
(May 1980 Dollars)

Initial leak repair labor charges $13/hour	1,630
Admin. & Support costs, 40 percent of labor charges	650
Total costs	2,280
Annualized charges for initial leak repair 16.3 percent of total cost[a]	370

[a]Capital recovery factor is 0.163 based on initial leak repair costs
amortized over 10 years at 10 percent interest.

Table 8-10. ANNUAL MONITORING AND LEAK REPAIR LABOR REQUIREMENTS FOR VALVES FOR THE MODEL UNIT
(Monthly Leak Detection and Repair Program)

No. of valves per model	Type of monitoring	Monitoring time,[a] man-min	Fraction screened	Monitoring labor required, man-hours	No. of leaks per year[b]	Repair time, man-hours	Leak repair labor-hours required, man-hours
402	Instrument	2	11.79	158	76.4	1.13	86.3
524	Instrument	2	11.79	205.9	99.6	1.13	112.5
Total				363.9			198.8

[a] Instrument monitoring time is 1 minute for a 2-man team.

[b] Average number of leaks found over turnaround 2 from the LDRP model, based on fraction of sources operated on equal to 19 percent.

Table 8-11. ANNUAL MONITORING AND LEAK REPAIR COSTS FOR
MONTHLY MONITORING OF VALVES FOR THE MODEL UNIT
(May 1980 Dollars)

Labor charges of $18 per hour for total 562.7 labor-hours,[a] $	10,130
Admin. & Support costs at 40 percent of labor charges, $	4,050
Annualized charge for initial leak repair[b], $	370
Total annual costs, $	14,550
Annual product recovery credit,[c] $	(17,750)[d]
Net annualized costs, $	($3,200)
Cost Effectiveness, $/Mg	(95)

[a]Total of 363.9 monitoring labor hours and 198.8 leak repair labor hours from Table 8-10.

[b]From Table 8-9.

[c]Product recovery credit is calculated at $528/Mg. The emission reductions are 14.4 Mg/yr from vapor service valves and 19.23 Mg/yr from light liquid service valves (see Table 8-5).

[d]Figures in parenthesis indicate credits.

Table 8-12. INITIAL LEAK REPAIR LABOR-HOURS REQUIREMENT
FOR PUMP SEALS FOR THE MODEL UNIT

No. of pump seals per model unit	Initial leak frequency	Estimation no. of initial leaks	Repair time, man-hours	Labor-hours required, man-hours
29	0.088	2.6	16	41.6

Table 8-13. TOTAL ANNUAL COSTS FOR INITIAL LEAK REPAIR
FOR PUMP SEALS FOR THE MODEL UNIT
(May 1980 Dollars)

Initial leak repair labor charges, at $18/hour	750
Admin. & Support costs, 40 percent of labor charges	300
Seal costs, $139.4/single seal[a,b]	360
Total Costs	1,410
Annualized charges for initial leak repair, 16.3 percent of total costs[c]	230

[a]Seal cost is $139.4. The value has been obtained by updating a cost value of $113 for last quarter 1978. The cost includes 50 percent credit for old seal.

[b]Calculation = No. of initial leaks (from Table 8-12) x seal cost
= 2.6 x 139.4 = $362

[c]Capital recovery factor is 0.163 based on initial leak repair costs amortized over 10 years at 10 percent interest.

Table 8-14. ANNUAL MONITORING AND LEAK REPAIR LABOR REQUIREMENTS FOR PUMP SEALS OF THE MODEL UNIT
(Monthly Leak Detection and Repair Program)

No. of pump seals per model unit	Type of monitoring	Monitoring time, man-min[a]	Times monitored per year	Monitoring labor required, man-hours	No. of leaks per year[b]	Repair time, man-hours	Leak repair labor required, man-hours
29	Instrument	10	12	58.0	11.8	16	188.8
	Visual	0.5	52	12.5			
				70.5			

[a]Instrument monitoring time is 5 minutes for a 2-man team.

[b]Based on 5.5 percent monthly occurrence rate less 2.1 percent credit for routine maintenance. These values were computed using the LDAR model. Fractions of leaking pumps were included in both cost and emission estimates.

Table 8-15. ANNUAL MONITORING AND LEAK REPAIR COSTS FOR
MONTHLY MONITORING OF PUMP SEALS FOR THE MODEL UNIT
(May 1980 Dollars)

Labor charges of $18 per hour for total 259.3 labor-hours,[a] $	4,670
Admin. & Support costs at 40 percent of labor charges, $	1,870
Annualized charge for initial leak repair,[b] $	230
Annual replacement cost of leaking seals, $	<u>1,650</u>
Total annual costs	$8,420
Annual product recovery credit,[d] $	(4,030)[e]
Net annualized cost, $	4,390
Cost effectiveness, $/Mg	575

[a]Total of 70.5 monitoring hours and 188.8 leak repair labor hours from Table 8-14.

[b]From Table 8-13.

[c]Based on $139.4/seal and 11.8 leaking seals per year obtained from Table 8-14.

[d]Product recovery credit is calculated at $528/Mg. The emission reductions obtained are 7.63 Mg/yr from pump seals (see Table 8-5).

[e]Figures in parentheses indicate a credit.

There are no requirements for leak detection and repair. It is assumed that leaks would be corrected by routine maintenance with no additional labor requirement. Therefore, the only annual costs associated with controls for safety/relief valves are equipment costs.

Equipment Costs: Costs were computed for the installation of a rupture disk upstream of a safety/relief valve in gas service. These costs were based on estimates from the Hydroscience (now IT Enviroscience) report.[38] The cost estimates were based on the following assumptions: No piping modification was required, and the disk and its holder simply could be inserted between the flanges of the relief valve and the system it protects. To allow in-service disk replacement, a block valve or a 3-way valve was assumed to be installed upstream of the rupture disk. In addition, to prevent damage to the relief valve by disk fragments, it was assumed that an off-set mounting would be required. The rupture disk life was assumed to be 2 years.

Equipment cost estimates for control of fugitive emissions from safety/relief valves were calculated for two different systems: (1) rupture disk with block valve and (2) rupture disk with 3-way valve. In computing total costs, half of the safety/relief valve sources (i.e., 5.5 from Table 8-5) were assumed to be installed with rupture disks with block valves and the remaining half with the rupture disks with 3-way valves. These costs are shown in Table 8-16. The total estimated installed cost of a new rupture disk system with block valve was $1,995 in second quarter 1980. The total estimated installed cost of a new rupture disk with 3-way block valve was $4,137 in second quarter of 1980. Based on equipment installation costs, total capital and annualized costs for control of emissions from the model unit's safety/relief valves were calculated and are presented in Table 8-17.

Open-ended lines

As there are no fugitive VOC emission sources, cost data were not generated for open-ended lines.

Compressors

The fugitive VOC control specifications for compressors require control of degassing vents. The costs of control are presented below.

Table 8-16. RELIEF VALVE CONTROL COSTS FOR RUPTURE DISK SYSTEMS
WITH BLOCK VALVES AND THREE-WAY VALVES
(May 1980 dollars)

	Capital costs
(1) Rupture disk systems with block valve	
Rupture disk and assembly	
One 7.6 cm stainless steel rupture disk	230
One 7.6 cm carbon steel rupture disk holder	384
One 0.6 cm dial face pressure gauge	18
One 0.6 cm carbon steel bleed valve	30
One 7.6 cm gate valve	700
One 10.2 cm tee and one 10.2 cm elbow	21
Installation at 34 total hours and $18/hr	612
Total cost for second quarter 1980	$1,995

	Annualized costs
Rupture disk at 58 percent of total costs[a]	$133
Assembly at 16.3 percent of total costs[b]	288
Maintenance & Miscellaneous.[c]	180
Total $/year	$601

	Capital costs
(2) Rupture disk systems with 3-way valve.	
Rupture disk and assembly	
One 7.6 cm stainless steel rupture disk	230
One 7.6 cm carbon steel disk holder	384
One 0.6 cm dial face pressure guage	18
One 0.6 cm carbon steel bleed valve	30
One 7.6 cm safety/relief valve	1,456
Two 7.6 cm elbows	30
One 10.2 cm tee and one 10.2 cm elbow	21
Subtotal	2,169
Three-way valve	
One 7.6 cm, 3-way, 2 port valve	1,320
Total Installation	648
Subtotal	1,968
Total cost for second quarter 1980	$4,137

	Annualized costs
Rupture disk at 58 percent of total costs[a]	$133
Assembly at 16.3 percent of total costs[b]	637
Maintenance & Miscellaneous.[c]	372
Total ($/year)	$1,142

[a] Capital recovery factor is 0.58 based on 2-year equipment life and 10 percent interest.

[b] Capital recovery factor is 0.163 based on 10-year equipment life and 10 percent interest.

[c] Based on 9 percent of total capital costs.

Table 8-17. CAPITAL AND NET ANNUALIZED COSTS FOR CONTROL OF
EMISSIONS FROM SAFETY/RELIEF VALVES FOR THE MODEL UNIT
(May 1980 Dollars)

Costs	Rupture disk system[a]
Installed capital cost,[b] $	33,730
Annualized costs,[c] $	6,550
Annual operating cost, $	
Maintenance at 5 percent of capital cost, $	1,690
Miscellaneous at 4 percent of capital cost, $	1,350
Total annual cost, $	9,590
Annual product recovery credit,[d] $	(5,290)
Net annualized costs, $	4,300
Cost effectiveness, $/Mg	430

[a]One half of the total systems with block valves and the remaining half with 3-way valves.

[b]Based on $3,066 per system and 11 sources. The cost of one system is computed from Table 8-16 as follows:
0.5 ($1,995 for rupture disk system with block valve and assembly + $4,137 for rupture disk system with 3-way valve and assembly).

[c]Based on annualized cost data in Table 8-16 per system and 11 sources. The cost of one system is computed as follows: 0.5 [($133 + 288) annualized cost for rupture disk system with block valve and assembly + ($133 + 637) annualized cost for rupture disk system with 3-way valve and assembly).

[d]Product recovery credit is calculated at $528/Mg and an emission reduction rate of 10.02 Mg/yr (see Table 8-5).

Equipment Costs: The cost of control equipment for compressors was based on installation of closed vents for degassing reservoirs of compressors. The estimate was based on information contained in the Hydroscience (now ITE) report[38] and was for the following items per compressor:

122 m length of 5.1-cm diameter schedule 40 carbon steel pipe	$6,400
Three 5.1-cm cast steel plug valves and one metal gauge flame arrestor	$1,600
Total for second quarter 1980	$8,000

The above costs include connection of the degassing reservoir to an existing enclosed combustion device or vapor recovery header. The cost of a control device, added specifically to control the degassing vents, is, therefore, not included.

Total capital and net annualized costs for control of emissions from compressor seals for the model units were developed. These cost data are presented in Table 8-18.

Sampling Systems

Equipment costs were computed for closed loop sampling connections. The cost estimates were based on information from the Hydroscience (now ITE) report[38] and was for the following items per sampling system. Table 8-19 presents capital and annual costs for control of emissions from sampling systems in the model plant.

One 6-m length of 25-cm diameter schedule 40 carbon steel pipe and three 2.5-cm carbon steel ball valves.	$210
Installation (at 18 hours and $18/hr)	$320
Total cost (second quarter 1980 dollars)	$530

8.1.7 Cost Analysis Results

The results of the cost analyses for all regulatory alternatives and model plants are summarized in this section. The installed capital costs by control device and the total operating and other annualized costs are presented for each regulatory alternative in Tables 8-20 to 8-28, for the nine model plants. Tables 8-29 to 8-37 show the costs and

Table 8-18. CAPITAL AND NET ANNUALIZED COSTS FOR CONTROL
OF EMISSIONS FROM COMPRESSOR SEALS FOR THE MODEL UNIT
(May 1980 Dollars)

Costs	Degassing reservoir vent
Installed capital cost,[a,b] $	6,400
Annualized cost at 16.3 percent of capital cost,[c] $	1,040
Annual operating cost for Maintenance at 5% of capital cost, $	320
Miscellaneous at 4% of capital cost, $	260
Total annual cost, $	1,620
Annual product recovery credit[b,d]	(850)[e]
Net annualized costs, $	770
Cost effectiveness, $/Mg	485

[a] Based on two compressor sources in the model plant and the assumption that controls apply to only 40 percent of the sources. (60 percent of compressors within the industry are known to be controlled and need not be considered for the purpose of cost analysis.)

[b] Capital cost per compressor is $8,000.

[c] Capital recovery factor, 0.163, is based on the equipment amortized over 10 years at 10 percent interest.

[d] Product credit is calculated based on $528/Mg and emission credit of 1.6 Mg/yr (i.e., 40 percent of the 3.99 Mg/yr from the model from Table 8-5).

[e] Values within parentheses denote credit.

Table 8-19. CAPITAL AND NET ANNUALIZED COSTS FOR CONTROL
OF EMISSIONS FROM SAMPLING SYSTEMS IN MODEL UNIT
(May 1980 Dollars)

Costs	Closed loop sampling connections
Installed capital cost,[a] $	13,800
Annualized cost at 16.3 percent of capital cost,[b] $	2,250
Annual operating costs for Maintenance at 5% of capital cost, $ Miscellaneous at 4% of capital cost, $	690 550
Total annual cost, $	3,490
Annual product recovery credit,[c] $	(1,800)[d]
Net annualized costs, $	1,690
Cost effectiveness, $/Mg	520

[a]Based on $530 per sampling system and 26 sampling systems in the model plant.

[b]Capital recovery factor of 16.3 percent, is based on the equipment amortized over 10 years at 10 percent interest.

[c]Product recovery credit is calculated based on $528/Mg and 3.24 Mg/yr of VOC recovered annually (see Table 8-5).

[d]Values within parentheses denote credit.

Table 3-20. POLYPROPYLENE, LIQUID PHASE MODEL PLANT REGULATORY
ALTERNATIVES COSTS (JUNE 1980 DOLLARS)

	Regulatory Alternative			
	1	2	3	4
INSTALLED CAPITAL COST ($)				
Flare(s)	12,500	12,500	12,500	
Flare Ducting	36,400	36,400	36,400	
Thermal Incinerator			218,000	368,000
Thermal Incinerator Ducts, Fans & Stack			95,200	125,000
Catalytic Incinerator				
Catalytic Incinerator Ducts, Fans & Stack				
Condenser				
Fugitive Leak Detection and Repair (LDAR)		66,800	66,800	66,800
Total	49,000	116,000	429,000	561,000
ANNUALIZED COST ($/yr)				
Direct				
Operating Labor	11,200	31,900	53,500	42,300
Operating Materials (e.g., catalyst)				
Maintenance Materials and Labor	2,450	8,150	23,800	30,400
Utilities Natural gas	4,280	4,280	4,290	
Electricity			210	1,620
Steam	27,800	27,800	27,800	
Subtotal	45,700	72,100	110,000	74,300
Indirect				
Taxes, Insurance, & Administration[a]	1,960	4,560	17,100	19,800
Capital Recovery	7,580	23,900	74,800	96,600
Subtotal	9,540	28,400	91,900	116,000
Recovery Credit		29,600	29,600	29,600
Total (Direct + Indirect − Credit)[b]	55,200	70,900	171,800	161,100

[a] Includes miscellaneous operating costs for fugitive leak detection and repair program.

[b] Some totals do not add up exactly due to rounding.

Table 8-21. POLYPROPYLENE, GAS PHASE MODEL PLANT REGULATORY
ALTERNATIVES COSTS (JUNE 1980 DOLLARS)

	Regulatory Alternative		
	1	2	3
INSTALLED CAPITAL COST ($)			
Flare(s)	199,000	199,000	206,000
Flare Ducting	166,000	166,000	181,000
Thermal Incinerator			
Thermal Incinerator Ducts, Fans & Stack			
Catalytic Incinerator			
Catalytic Incinerator Ducts, Fans & Stack			
Condenser			
Fugitive Leak Detection and Repair (LDAR)		66,800	66,800
Total	366,000	432,000	454,000
ANNUALIZED COST ($/yr)			
Direct			
Operating Labor	11,200	31,900	31,900
Operating Materials (e.g., catalyst)			
Maintenance Materials and Labor	18,300	24,000	25,100
Utilities			
Natural gas	12,800	12,800	12,800
Electricity			
Steam	740	740	14,300
Subtotal	43,000	69,400	84,100
Indirect			
Taxes, Insurance & Administration[a]	14,600	17,200	18,100
Capital Recovery	53,300	69,600	72,900
Subtotal	67,900	86,800	90,900
Recovery Credit		29,600	29,600
Total (Direct + Indirect - Credit)	110,900	126,600	145,400

[a]Includes miscellaneous operating costs for fugitive leak detection
and repair program.

Table 8-22. LOW DENSITY POLYETHYLENE, LIQUID PHASE MODEL PLANT REGULATORY ALTERNATIVES COSTS (JUNE 1980 DOLLARS)

	Regulatory Alternative			
	1	2	3	4
INSTALLED CAPITAL COST ($)				
Flare(s)	88,400	88,400	88,400	88,400
Flare Ducting	99,400	99,400	99,400	99,400
Thermal Incinerator				
Thermal Incinerator Ducts, Fans & Stack				
Catalytic Incinerator			219,000	257,000
Catalytic Incinerator Ducts, Fans & Stack			139,000	152,000
Condenser				
Fugitive Leak Detection and Repair (LDAR)		66,800	66,800	66,800
Total	188,000	255,000	613,000	663,000
ANNUALIZED COST ($/yr)				
Direct				
Operating Labor	11,200	31,900	48,600	48,600
Operating Materials (e.g., catalyst)			17,200	21,400
Maintenance Materials and Labor	9,390	15,100	33,000	35,500
Utilities				
Natural gas	3,550	8,550	31,600	37,300
Electricity			210	1,620
Steam	2,676,200[b]	2,547,100[b]	2,647,100[b]	2,647,100[b]
Subtotal	2,696,900	2,702,600	2,782,700	2,796,500
Indirect				
Taxes, Insurance & Administration[a]	7,510	10,100	24,400	26,400
Capital Recovery	27,800	44,100	102,000	111,000
Subtotal	35,300	54,200	127,000	137,000
Recovery Credit		29,600	29,600	29,600
Total (Direct + Indirect - Credit)	2,711,500	2,727,200	2,879,800	2,903,900

[a]Includes miscellaneous operating costs for fugitive leak detection and repair program.

[b]Excess steam may be available at zero cost to the plant.

[c]Some totals may not add up exactly due to rounding.

Table 8-23. LOW AND HIGH DENSITY POLYETHYLENE, GAS PHASE MODEL
PLANT REGULATORY ALTERNATIVES COSTS
(June 1980 Dollars)

	Regulatory Alternative			
	1	2	3	4
INSTALLED CAPITAL COST ($)				
Flare(s)	158,000	158,000	158,000	137,000[a]
Flare Ducting	921,000	921,000	927,000	938,000
Thermal Incinerator				
Thermal Incinerator Ducts, Fans & Stack				
Catalytic Incinerator				
Catalytic Incinerator Ducts, Fans & Stack				
Condenser				
Fugitive Leak Detection and Repair (LDAR)		66,800	66,800	66,800
Total	1,080,000	1,146,000	1,151,000	1,142,000
ANNUALIZED COST ($/yr)				
Direct				
Operating Labor	11,200	31,900	31,900	31,900
Operating Materials (e.g., catalyst)				
Maintenance Materials and Labor	54,000	59,700	59,900	59,500
Utilities				
Natural gas	17,100	17,000	17,000	88,600
Electricity				
Steam	18,200	18,200	18,200	19,500
Subtotal	100,500	126,900	127,100	199,400
Indirect				
Taxes, Insurance & Administration[b]	43,200	43,200	43,400	43,000
Capital Recovery	170,700	187,000	187,800	187,000
Subtotal	213,900	232,800	233,800	232,600
Recovery Credit		29,600	29,600	29,600
Total (Direct + Indirect - Credit)[c]	314,300	333,000	331,400	402,400

[a]Flare capital cost lower as no seal necessary to prevent inflow with higher continuous flow.

[b]Includes miscellaneous operating costs for fugitive leak detection and repair program.

[c]Some totals do not add up exactly due to rounding.

Table 8-24. HIGH DENSITY POLYETHYLENE, LIQUID PHASE-SLURRY MODEL PLANT
REGULATORY ALTERNATIVES COSTS
(June 1980 Dollars)

	Regulatory Alternative		
	1	2	3
INSTALLED CAPITAL COST ($)			
Flare(s)	16,700	16,700	19,900
Flare Ducting	35,100	35,100	49,800
Thermal Incinerator			
Thermal Incinerator Ducts, Fans & Stack			
Catalytic Incinerator			
Catalytic Incinerator Ducts, Fans & Stack			
Condenser			
Fugitive Leak Detection and Repair (LDAR)		66,800	66,300
Total	51,800	118,600	136,500
ANNUALIZED COST ($/yr)			
Direct			
Operating Labor	11,200	31,900	31,900
Operating Materials (e.g., catalyst)			
Maintenance Materials and Labor	2,590	8,290	9,180
Utilities			
Natural gas	4,280	4,280	4,280
Electricity			
Steam	18,200	18,200	18,600
Subtotal	36,200	62,600	63,900
Indirect			
Taxes, Insurance, & Administration[a]	2,070	4,670	5,390
Capital Recovery	7,910	24,200	27,000
Subtotal			
Recovery Credit		29,600	29,600
Total (Direct + Indirect - Credit)[b]	46,200	61,900	66,700

[a]Includes miscellaneous operating costs for fugitive leak detection and repair program.

[b]Some totals may not add up exactly due to rounding.

Table 3-25. HIGH DENSITY POLYETHYLENE, LIQUID PHASE-SOLUTION
MODEL PLANT REGULATORY ALTERNATIVES COSTS
(JUNE 1980 DOLLARS)

	Regulatory Alternative		
	1	2	3
INSTALLED CAPITAL COST ($)			
Flare(s)	12,500	12,500	12,500
Flare Ducting	30,100	30,100	30,100
Thermal Incinerator			
Thermal Incinerator Ducts, Fans & Stack			
Catalytic Incinerator			1,070,000
Catalytic Incinerator Ducts, Fans & Stack			348,000
Condenser			
Fugitive Leak Detection and Repair (LDAR)		66,800	66,800
Total	42,700	109,000	1,530,000
ANNUALIZED COST ($/yr)			
Direct			
Operating Labor	11,200	31,900	43,600
Operating Materials (e.g., catalyst)			131,300
Maintenance Materials and Labor	2,130	7,830	73,700
Utilities			
Natural gas	4,280	4,280	507,900
Electricity			40,900
Steam	18,900	18,900	18,900
Subtotal	36,400	62,300	825,800
Indirect			
Taxes, Insurance, & Administration[a]	1,710	4,310	61,000
Capital Recovery	6,550	22,800	254,000
Subtotal	8,260	27,200	315,000
Recovery Credit		29,600	29,600
Total (Direct + Indirect - Credit)[b]	44,700	60,400	1,111,800

[a]Includes miscellaneous operating costs for fugitive leak detection and repair program.

[b]Some totals do not add up exactly due to rounding.

Table 8-26. POLYSTYRENE-CONTINUOUS MODEL PLANT REGULATORY
ALTERNATIVES COSTS (JUNE 1980 DOLLARS)

| | Regulatory Alternative | | |
	1	2	3
INSTALLED CAPITAL COST ($)			
Flare(s)			
Flare Ducting			
Thermal Incinerator			
Thermal Incinerator Ducts, Fans & Stack			
Catalytic Incinerator			
Catalytic Incinerator Ducts, Fans & Stack			
Condenser			2,780
Fugitive Leak Detection and Repair (LDAR)		66,800	66,800
Total	0	66,800	69,600
ANNUALIZED COST ($/yr)			
Direct			
Operating Labor		20,700	21,800
Operating Materials (e.g., catalyst)			
Maintenance Materials and Labor		5,700	5,840
Utilities			
Natural gas			
Electricity			150
Steam			
Water			390
Subtotal	0	26,400	28,200
Indirect			
Taxes, Insurance, & Administration[a]		2,600	2,710
Capital Recovery		16,300	16,800
Subtotal	0	18,900	19,500
Recovery Credit		29,600	190,600
Total (Direct + Indirect - Credit)[b]	0	15,700	(142,900)[c]

[a] Includes miscellaneous operating costs for fugitive leak detection and repair program.

[b] Some totals do not add up exactly due to rounding.

[c] Parentheses denote a negative total net control cost.

Table 8-27. POLYESTER (PET) - DMT PROCESS MODEL PLANT REGULATORY
ALTERNATIVES COSTS (JUNE 1980 DOLLARS)

	Regulatory Alternative		
	1	2	3
INSTALLED CAPITAL COST ($)			
Flare(s)			12,500
Flare Ducting			8,600
EGRS[a] (Baseline)[b]	234,000		
EGRS[a] (Alternative)[c]		2,514,900	2,514,900
TOTAL	234,000	2,514,900	2,536,100
ANNUALIZED COST ($/yr)			
Direct			
Operating Labor	77,400	154,800	166,000
Operating Materials (e.g., catalyst)			
Maintenance Materials and Labor	10,400	27,700	22,800
Utilities			
Natural gas			4,280
Electricity	940	70,600	70,600
Steam	828,100	828,100	828,200
Water		4,650	4,650
Subtotal	916,800	1,079,800	1,096,400
Indirect			
Taxes, Insurance, & Administration	9,360	100,600	101,400
Capital Recovery	38,100	408,900	423,900
Subtotal	47,500	510,500	525,300
Recovery Credit	407,100	1,292,000	1,292,000
Total (Direct - Indirect - Credit)[d]	557,200	298,400	329,700

[a]EGRS - ethylene glycol recovery system.

[b]The baseline system recovers ethylene glycol (EG) from downstream of the cooling water tower. Ethylene glycol emissions from the polymerization reactor and from the distillation column recovering EG emitted from the esterifiers accumulate in the cooling water and are emitted from the cooling tower.

[c]The regulatory alternative system recovers ethylene glycol from the polymerization reactor through use of an EG spray condenser and recovers ethylene glycol from the esterifier through use of a reflux condenser. Thus, very little EG is emitted to be collected in the cooling water and later emitted to the atmosphere from the cooling water tower.

[d]Some totals do not add up exactly due to rounding.

Table 8-28. POLYESTER (PET) - TPA PROCESS MODEL PLANT REGULATORY
ALTERNATIVES COSTS (JUNE 1980 DOLLARS)

| | Regulatory Alternative | | | |
	1	2	3	4
INSTALLED CAPITAL COST ($)				
EGRS[a] (Baseline)[b]	234,000			
EGRS[a] (Alternative)[c]		2,514,900		
Total	234,000	2,514,900		
ANNUALIZED COST ($/yr)				
Direct				
Operating Labor	77,400	154,800		
Operating Materials (e.g., catalyst)				
Maintenance Materials and Labor	10,400	21,700		
Utilities				
Natural gas				
Electricity	940	70,600		
Steam	828,100	828,100		
Water		4,650		
Subtotal	916,800	1,079,800		
Indirect				
Taxes, Insurance, & Administration	9,360	100,600		
Capital Recovery	38,100	409,900		
Subtotal	47,500	510,500		
Recovery Credit	407,100	1,292,000		
Total (Direct & Indirect - Credit)	557,200	298,400		

[a]EGRS - ethylene glycol recovery system.

[b]The baseline system recovers ethylene glycol (EG) from downstream of the cooling water tower. Ethylene glycol emissions from the polymerization reactor and from the distillation column recovering EG emitted from the esterifiers accumulate in the cooling water and are emitted from the cooling tower.

[c]The regulatory alternative system recovers ethylene glycol from the polymerization reactor through use of an EG spray condenser and recovers ethylene glycol from the esterifier through use of a reflux condenser. Thus, very little EG is emitted to be collected in the cooling water and later emitted to the atmosphere from the cooling water tower.

Table 8-29. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR POLYPROPYLENE, LIQUID PHASE PROCESS

Regulatory Alternative	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	$ 15,700	60	$260	$ 15,700	60	$ 260
3	$117,000	440	$265	$ 101,000	380	$ 255
4	$106,000	450	$235	$ (10,700)[a]	10	$ (1,070)[a]

[a]Parentheses indicate a negative incremental cost.

Table 8-30. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR POLYPROPYLENE, GAS PHASE PROCESS

Regulatory Alternative	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	$ 15,700	60	$260	$ 15,700	60	$260
3	$ 34,500	2,635	$ 13	$ 18,300	2575	$ 7

Table 8-31. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR LDPE, HIGH PRESSURE, LIQUID PHASE PROCESS

Regulatory Alternative	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	$ 15,700	60	$260	$ 15,700	60	$260
3	$168,300	275	$610	$152,600	215	$710
4	$192,400	335	$570	$ 24,100	60	$400

Table 8-32. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR LDPE/HDPE LOW PRESSURE, GAS PHASE PROCESS

Regulatory Alternative	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	$ 15,700	60	$260	$15,700	60	$260
3	$ 17,100	63	$270	$1,400	3	$470
4	$ 88,100	71	$1,240	$71,000	8	$8,800

Table 8-33. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR HDPE, LIQUID PHASE SLURRY PROCESS

Regulatory Alternative	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	$15,700	60	$260	$15,700	60	$260
3	$20,500	140	$150	$ 4,800	30	$ 60

Table 8-34. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR HDPE, LIQUID PHASE SOLUTION PROCESS

	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
Regulatory Alternative	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	$ 15,700	60	$ 260	$ 15,700	60	$ 260
3	$1,067,100	190	$5,620	$1,051,400	130	$8,090

Table 8-35. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR POLYSTYRENE, CONTINUOUS PROCESS

Regulatory Alternative	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	$ 15,700	60	$260	$ 15,700	60	$260
3	($142,900)[a]	285	($500)[a]	($158,600)[a]	225	($700)[a]

[a]Parentheses denote a negative net or incremental cost.

Table 8-36. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR PET/DMT PROCESS

Regulatory Alternative	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	($258,800)[a]	975	($270)[a]	($267,800)[a]	975	($270)[a]
3	($277,500)[a]	995	($230)[a]	$31,300	20	$1,570

[a]Parentheses indicate a negative net or incremental cost.

Table 8-37. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY
ALTERNATIVES FOR PET/TPA PROCESS

Regulatory Alternative	Implementation of Alternative from Baseline Control			Implementation of Alternative from Next Less Stringent Alternative		
	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)	Annualized Cost ($)	VOC Reduction (Mg/yr)	Cost of Reduction ($/Mg)
1 (Baseline)	-	-	-	-	-	-
2	($258,800)[a]	970	($270)	($258,800)[a]	$970	(270)[a]

[a]Parentheses indicate a negative net or incremental cost.

associated emission reductions of regulatory alternatives for the nine
model plants for the implementation of an alternative both from the
baseline control level and from the next less stringent alternative.

National control costs are estimated, as shown in Table 8-38, based
on 27 projected new facilities, described extensively in Chapter 9
"Economic Impact". The total nationwide fifth-year net annualized cost
ranges from $8.3 million for baseline control to $3.7 million for the
maximum achievable control level. The maximum achievable control level
is equivalent to a 14,950 Mg VOC reduction from baseline per year.

8.2 OTHER COST CONSIDERATIONS

8.2.1 Water Pollution Control Regulations

8.2.1.1 Federal Water Pollution Control Act (FWPCA). Polymers and
resins industry (a subcategory of SIC 2821) facilities are required by
the FWPCA to comply with effluent limitation guidelines. Under the
guidelines, existing sources must apply the best practical control
technologies available (BPCTA) and new sources must apply best available
demonstrated control technology (BADCT).

The Clean Water Act of 1977 amended the FWPCA and required that the
best available technology economically achievable (BATEA) be implemented
by 1984 for nonconventional and toxic pollutants. For conventional
pollutants, best conventional technology (BCT) is required. The development
of BATEA and BCT guidelines take into account different cost considerations.

EPA has developed water quality criteria documents for 64 toxic
water pollutants. These documents contain recommended maximum permissible
pollutant concentrations for the protection of aquatic organisms, human
health, and some recreational activities. These documents do not consider
treatment technology, costs, or other feasibility factors.

The National Pollution Discharge Elimination System (NPDES) authorizes
States to issue discharge permits. Approximately 85 percent of the
chemical products industry (SIC 28) is in compliance with the Federal
water pollution reporting regulations required under NPDES.

The capital cost to the plastics and synthetics industry of water
pollution control totalled $308 million from 1972 through 1977 (updated
to second quarter 1980 dollars from second quarter 1977 dollars using

Table 8-38. TOTAL FIFTH-YEAR NET ANNUALIZED COST OF PROCESS
AND FUGITIVE EMISSION CONTROLS FOR POLYMERS AND RESINS
FACILITIES AFFECTED BY NSPS

Polymer/ Process	Regulatory Alternative Number	Annualized Per Facility Costs, $/yr	Number of Facilities	Total Nationwide Fifth-Year Net Annualized Process and Fugitive Costs, $/yr
				($)
PP				
Liquid	4	105,900	3	317,700
Gas	3	34,500	3	103,500
LDPE				
Liquid	4	192,400	1	192,400
Gas	4	88,100	5	440,500
HDPE				
Liquid Slurry	3	20,500	2	41,000
Liquid, Solution	3	1,067,100	3	3,201,300
Gas	4	88,100	5	440,500
PS				
Continuous[a]	3	(142,900)[b]	2	(285,800)[b]
PET				
DMTa	3	(227,500)[b]	1	(227,500)[b]
TPA[a]	2	(258,800)[b]	2	(517,600)[b]
TOTAL			27	3,706,000

[a]From uncontrolled baseline, others from controlled baseline

[b]Parentheses denote a negative net or incremental cost.

Source: Tables 8-20 through 8-28 of this report.

the fixed nonresidential investment part of implicit price deflator of the gross national product). The cumulative capital costs from 1977 through 1986 were projected to be $470 million. The total annualized costs for 1972 - 1977 was $117 million and the projection for 1977 - 1986 was $850 million.[39]

8.2.1.2 <u>Safe Drinking Water Act (SDWA)</u>. The Safe Drinking Water Act requires EPA to establish primary and secondary drinking water standards. Primary regulations are aimed at protecting public health. They establish maximum allowable containinant levels in drinking water and provide for water supply system operation. Secondary regulations are designed to protect public welfare and to control the taste, odor, and appearance of drinking water. The Act also controls underground injection through permitting. In establishing maximum control levels (MCL), the technological and economic feasibility is considered as well as the health effects. Currently, the MCL for VOC in groundwater is being developed; therefore, control costs are unknown. Since there are very few MCLs at this time, States have the option of controlling toxic pollutants when a MCL does not exist.

8.2.2 <u>Occupational Safety and Health Regulations</u>

The Occupational Safety and Health Administration (OSHA) is responsible for protecting workers against hazardous materials found in the work place. There are two types of OSHA regulations affecting the organic chemical industry. The first type requires general work practice and engineering controls for hazardous substances. If engineering controls and work practice standards are not capable of achieving full compliance, protective equipment is to be used.

A second type of OSHA regulation, for more significant hazardous air pollutants, involves comprehensive requirements for administrative practices and engineering controls specific to a particular pollutant.

The average cost of OSHA regulations on the entire chemical industry is estimated to be $208.40 per worker per year. The type of worker protection is dependent on the chemical produced at each distillation facility. In those facilities where only general controls are required, the costs would vary with the control method(s) employed by each facility.

OSHA also has specific regulations, under Section 29 CFR 1910.106, for chemical facilities which handle, store, or use flammable and combustible liquids with a flash point less than 90°C (200°F). OSHA develops these standards for toxicity levels and not based upon cost criteria.

8.2.3 Toxic Substance Control Regulations

Toxic Substance Control Act (TSCA) requirements are based on the need to provide necessary information concerning the toxicity of new and existing chemicals. In order to develop a chemical inventory, TSCA requires reporting of the manufacturing, importing, or processing of any chemical substance used for a commercial purpose. Any substance not on the inventory will be considered new and premanufacture notice and testing will be required. Reporting and premanufacture notification (PMN) requirements include: (1) the cost of using screening and testing to gain appropriate information for new chemicals, (2) the cost of testing existing chemicals, and (3) the cost of the delay caused by the testing/reporting process. PMN could have a significant impact on the entire chemical industry, with cost estimates ranging from $78.5 million to $2 billion.

Small companies will probably suffer more than the larger firms since small firms have minimal access to the information necessary to develop a PMN. The impact of PMN requirements also will be greater to the small firms because the cost per unit product will be higher for low volume, low revenue chemicals. The cost of preparing notices for new chemicals is estimated to be between $820 and $7400 per chemical.

EPA has been concentrating its efforts on new chemicals being developed rather than on existing chemicals; therefore, the actual cost of meeting TSCA for the polymers and resins industry is unknown.

8.2.4 Solid and Hazardous Waste Regulations

8.2.4.1 Resource Conservation and Recovery Act (RCRA). RCRA establishes a national program to improve solid waste management including the control of hazardous waste, the promotion of resource conservation and recovery, and the establishment of a solid waste disposal program.

The hazardous waste program regulates wastes from generation to disposal ("cradle to grave") requiring EPA to produce standards for generators, transporters, and those who transport, store, and dispose

(TSD facilities). The wastes are identified and listed by industry. At the time of generation, a manifest is issued to record the movement of the wastes from cradle to grave.

The management of nonhazardous wastes is essentially a State and local function implemented under State and regional solid waste plans.

As the cost of handling wastes increases, some firms will reduce their costs by changing their process to eliminate wastes or by recycling or reclaiming the waste. New plant and equipment expenditures for solid waste control were $42-$45 million for the entire chemical industry in both 1978 and 1979. The annual cost imposed by RCRA on 45 organic chemical plants generating hazardous wastes is estimated to be $10.9 million or an average annual cost of $240,000 per plant. These estimates are based on model plants.

8.2.4.2 <u>Superfund</u>. The Comprehensive Environmental Response, Compensation, and Liability Act, or Superfund, regulates the cleanup of hazardous waste dumpsites and chemical spills. Superfund provides adequate funding, liability, standards, and authority to the government to recover costs from the responsible parties. Any person in charge of a facility is required to report any "release" of a specified quantity of hazardous waste into the environment immediately. The emphasis of the regulation is to report the release of the wastes and to clean them up first and then to recover costs later. The Act also authorizes a tax on all hazardous wastes received at a disposal facility that is to be deposited in a trust fund for use after the facility closes. The fee to the chemical industry for this trust fund is less than 2 percent of their profits.

8.2.5 <u>Clean Air Act</u>

There are two regulatory actions begun since 1977 under Sections 111 and 112 of the Clean Air Act that will, if promulgated, impose costs on firms with new, modified, or reconstructed polymerization facilities. These actions are the potential NSPS for control of VOC emissions from both process and fugitive sources (the subject of this study), and the potential NSPS for control of VOC emissions from raw materials storage at the production site. The control costs for the former have been tabulated above (Section 8.1); the control costs for the latter are

estimated, as discussed below, primarily from information contained in the draft BID for volatile organic liquid (VOL) storage tanks.[40]

For the number and type of affected facilities discussed in Section 9.1, the sum of the potential fifth-year net annualized costs of the VOL storage is estimated (Table 9-15). The control cost estimation for VOC emissions from monomer and raw materials storage requires estimations of (1) the number of polymers and resins plants at which storage tanks are required, (2) the number of storage tanks required at each plant, and (3) the VOC control costs per storage tank.

First, the number of plants at which storage tanks are required is estimated. The polymerization plant is either a part of a larger chemical complex at which the monomer is both produced and consumed or it is an isolated plant. Where the polymer is produced at a major chemical complex storage tanks serve as both monomer storage following production and the onsite monomer storage for the polymerization process. The costs for these storage emission controls are not assigned here to the polymerization facilities because the costs would be incurred even in the absence of the polymerization facilities. The proportion of the new plants requiring onsite storage is that proportion of plants shown in Table 9-1 that are not owned by large oil or chemical companies. The results are shown in Table 8-39.

Second, the required number of storage tanks is estimated. It is assumed that the required storage capacity is equal to a 30-day supply or about 8.4 percent of annual capacity and that 2.4 tanks are required for each gigagram of plant production capacity. The number of tanks required per gigagram is based on data for the model producer/consumer contained in the draft VOL storage BID, indicated above, and was calculated by dividing the number of tanks for the model by the plant production capacity.[41] Thus, the number of tanks, shown by type of polymer in Table 8-40, totals 116.

Third, the cost for emission control per storage tank is calculated from cost data contained in the draft VOL storage BID.[40] These costs, given in first quarter 1980 dollars, are converted to June 1980 dollars by use of the Engineering News Record Construction Index (101.8). At the time the draft EIS was prepared, EPA recommended Regulatory

Table 8-39. NUMBER AND CAPACITY OF PLANTS TO BE AFFECTED BY BOTH
VOL STORAGE AND POLYMERS AND RESINS STANDARDS THROUGH 1988[a]

Polymer	Projected number of new plants	Percent of these expected to be nonmajor chemical or oil company	Number of new nonmajor company plants	Capacity of the plants (Gg/y)
PP		20		
Liquid	3		0.6	90
Gas	3		0.6	63
LDPE		20		
Liquid	1		0.1	28
Gas	5		0.5	75
HDPE		10		
Liquid,				
Slurry	2		0.2	42
Liquid,				
Solution	3		0.3	27
Gas	5		0.5	75
PS		25		
Continuous	2		0.5	38
PET		10		
DMT	1		0.1	10
TPA	2		0.2	20

[a]This table is based on data contained in Section 9.1 of this report.

Table 8-40. STORAGE TANK COMPLIANCE COSTS FOR PLANTS TO BE AFFECTED
BY BOTH VOL STORAGE AND POLYMERS AND RESINS
STANDARDS THROUGH 1988

Polymer	Storage,[a] Gg	Number[b] tanks	Credit @ $255/tank
PP			
Liquid	8	19	4,800
Gas	5	12	3,100
LDPE			
Liquid	2	5	1,300
Gas	6	14	3,600
HDPE			
Liquid, Slurry	4	10	2,600
Liquid, Solution	2	5	1,300
Gas	6	14	3,600
PS			
Continuous	3	7	1,800
PET			
DMT	1	2	0[c]
TPA	2	5	0[c]
TOTAL	39	93	22,100

[a]Production capacity requiring storage (from Table 8-39) times 8.4
percent (see text).

[b]Storage times 2.4 (see text).

[c]PET/DMT and PET/TPA is excluded since only heavy liquids and solids are
used in these processes; heavy liquids and solids are excluded from
control requirements.

Alternative III. This option would require each storage vessel storing a volatile organic liquid (VOL) with a true vapor pressure of less than 76.6 kPa to be equipped with a contact internal floating roof with a liquid-mounted primary seal and a continuous secondary seal. A vapor control system would be required for all storage vessels storing a VOL with a true vapor pressure of greater than or equal to 76.6 kPa. Thus, all polymers and resins model plants except PET/DMT and PET/TPA, which use heavy liquids and solids, will be covered.

The VOC control costs for monomer and raw materials storage are estimated from costs developed for SOCMI in the following manner. Alternative III for all of the SOCMI was estimated to result in a total annualized credit of $6,220,000 due to product recovery credits[42] In order to obtain the credit per tank, the annualized credit was divided by the total SOCMI storage tank population of 24,287.[43] Finally, the emission control costs for these storage tanks are assumed to be comparable to these required for the polymers and resins industry. The control costs for the polymers and resins industry are shown in Table 3-40. The resulting fifth-year cost to the polymers and resins industry under the above assumptions is a negative $22,100.

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9. ECONOMIC IMPACT

9.1 INDUSTRY CHARACTERIZATION

The five polymers and resins segments selected for potential NSPS development account for about 75 percent of the current total estimated VOC process emissions from 16 major polymers and resins manufacturing operations. The five are:

1. Polypropylene (PP),
2. High Density Polyethylene (HDPE),
3. Low Density Polyethylene (LDPE),
4. Polystyrene (PS), and
5. Poly(ethylene Terephthalate) (PET) or Polyester Resin

9.1.1 Industry Structure

Polymers are chemically prepared through polymerization, a process that converts monomers or intermediate materials obtained from the synthetic organic chemical manufacturing industry (SOCMI) into polymer products. Such products include plastic materials, synthetic resins, synthetic rubbers, and synthetic organic fibers. It is for this polymerization process with its large volume emission of volatile organic compounds (VOC) that the potential NSPS is being developed.

LDPE, HDPE, and PS--polymers used in plastics--have characteristics similar to those of PET, a polymer used in fibers. PP, additionally, is a polymer used both as a fiber and as a plastics substance.

9.1.1.1 Industries. The fibers portion of the industry, producing both cellulosic (rayon, acetate) and noncellulosic (polyester) fibers, is composed of relatively few plants, and those that produce

Note: Last minute changes in some of the process parameters and regulatory alternatives, and in the plant lists in Chapter 3, have not yet been incorporated in this chapter. These changes, however, do not significantly alter the economic analysis nor change the conclusions of this chapter.

noncellulosic fibers often carry out both the polymerization process and the process of converting the polymers into noncellulosic fibers in the same plant. The plastic materials portion of the industry is much larger and more diverse than the fibers portion and includes two SIC industries.

In the polymers and resins industry, products are typically manufactured in four stages (Figure 9-1): (1) polymerization, (2) compounding, (3) processing, and (4) and fabricating and finishing. As indicated above, polymerization is the initial stage in which monomers are converted to polymers and resins, and again as indicated above, it is this stage that is the primary focus of this analysis. This stage involves the five process steps described in Chapter 3.

After polymerization, the polymers and resins are usually combined, in the second stage, with various compounding materials, frequently coloring agents, plasticizers for flexibility, fillers (inert mineral powders) for firmness and rigidity, and to decrease production costs, reinforcing agents for product strength and abrasive resistance. Other additives may also be used to promote product flame retardance and curing. For some polymers and resins, the compounding stage is omitted.

The third stage in polymers and resins manufacturing is referred to as processing. This stage involves the molding, casting, calendaring (pressing), or extruding of the polymers and resins and their compounding additives into films, sheets, fiber, or rigid plastics. Such product reinforcement materials as fiberglass or various synthetic fibers may also be added during this stage.

The fourth and final stage in polymers and resins products manufacturing is that of fabricating and finishing. At this stage, the sheets, rods, tubes, and special shapes are finished or fabricated into final products such as packing materials, containers, housings, pipes, and toys. After this stage, the products are shipped to the major end-use markets--construction, packaging, transportation, electronics, appliances, textiles, furniture, and housewares.

Although the potential NSPS development is germane only to the polymerization stage of production, the other stages are discussed in this report because of their interrelationships. Polymerization takes place in plants of varying characteristics. In some, only the

polymerization stage is completed. In others, however, polymerization and one or all of the three additional stages may constitute plant operations. Of all polymers and resins produced, approximately 20 percent is processed (stage three) in the same plants in which the polymers and resins are produced. Thirty percent is processed by about 5,600 independent custom and proprietary processors and fabricators and finishers. The remaining 50 percent is processed and fabricated at 9,100 plants that additionally manufacture such nonplastics products as automobiles, appliances, textiles, and housewares.[1]

The plants that produce the five polymers and resins selected for the potential NSPS development are predominantly classified in three SIC Industries:

SIC 2821, Plastics materials and resins

SIC 2824, Organic fibers, noncellulosic   ·

SIC 3079, Miscellaneous plastics products

Industry 2821 comprises those plants that produce polymers and resins for sale or shipment to other plants. These plants limit production to stage 1. Plants combining stage 1 with any other stage are classified in either SIC 3079 or SIC 2824. Industry 3079, comprised of the various types of processors, includes the largest number of plants in the plastics industry. Industry 2824 includes those PET plants producing polyester fibers. PET plants that produce polyester for film and other nonfiber materials are classified in either SIC 2821 or SIC 3079.

9.1.1.1.1 Ownership. Table 9-1 lists the producers of the five considered polymers and resins and indicates the number of U.S. plants owned by each. About a fourth of the plants producing the five polymers and resins are subsidiaries of large petroleum manufacturers. Exxon with total sales in excess of $100 billion in 1981 had plastics sales of about $500 million in its subsidiary, the Exxon Chemical Company.[2] This company operates three plants that produce polymers. Gulf Oil's subsidiary, Gulf Oil Chemicals, operates six polymer plants, most of which produce polyethylene. The Mobil Chemical Company, a subsidiary of Mobil Oil, owns four polymer plants, three of which produce polystyrene. Other major oil companies owning separate subsidiaries that operate polymer

Table 9-1. NUMBER OF POLYMER AND RESIN PLANTS BY MANUFACTURER AND TYPE; January 1, 1982 [a]

Manufacturer	PP	LDPE	HDPE	PS	PET	Total
			---No.---			
A&E Plastics				1		1
Akzona					2	2
Allied Corporation			1		1	2
American Hoechst			1	3	2	6
American Petrofina				4		4
Atlantic Richfield	1	1	1	1		4
Avtex Fibers					1	1
BASF Wyandotte				2		2
Chemplex		1	1			2
Cities Service		1				1
Crest Container				2		2
Dart Industries				1		1
Dow Chemical		2	2	5		9
Du Pont		2	2		10	14
Eastman Chemicals	1	1			4	6
El Paso Natural Gas	2	2				4
Exxon		2				3
Fiber Industries					5	5
Firestone Tire & Rubber					1	1
Goodyear Tire & Rubber	1				2	3
Carl Gordon				2		2
Gulf Oil	1	2	1	2		6
Hercules	2		1			3
Huntsman-Goodson				1		1
ICI Americas					1	1
InterNorth, Inc.	1	1				2
Kama				1		1
Minnesota Mining					2	2
Mobil Chemical			1		3	4
Monsanto				3	1	4
National Distillers		2				2
National Petrochemical			1			1
Phillips Petroleum	1		1			2
Polysar Group				3		3
Rohm and Haas					1	1
Shell Chemical	2			1		3
Soltex Polymer	1		1			2
Standard Oil (Indiana)	1		1	3		5
Texstyrene Plastics				1		1
Union Carbide		3	1	1		5
United Foam		1				1
USS Chemicals	2		1			3
Vititek				1		1
TOTAL	16	22	15	42	33	128

[a] Source: Compiled from data contained in Tables 3-1 to 3-5.

plants are Standard Oil (Indiana), Atlantic Richfield, and Shell Oil; and two major oil companies--Phillips Petroleum and Cities Service--operate plants that are integral parts of their parent corporate structures.

About 10 percent of the polymers and resins plants are owned by companies in industries as diverse as food processing, textiles, packaging, and photographic equipment manufacturing. For the most part, these plants operate as subsidiaries or as parts of subsidiaries owned by the major companies.

The remaining two-thirds of the plants are owned by corporations classified in the basic chemical industry and are generally organized as divisions of the parent company. Only a few plants operate as subsidiaries. Du Pont, the largest chemical manufacturer, with total sales exceeding $20 billion in 1981 and plastics sales (including fibers) of over $3 billion, operates 14 plants, most of which produce PET.[2] Dow Chemical, the second largest chemical manufacturer, owns nine plants producing polymers and resins. All of Dow's plants produce either polyethylene or PS; none produces PET. Other large chemical companies are Union Carbide, American Hoechst, Dart Industries, and Monsanto. Enka, a subsidiary of Akzona, operates two polymer plants producing PET fibers. All but a few polymer plants are owned by large, multi-plant corporations.

9.1.1.1.2 <u>Capacity share</u>. Table 9-2 lists the 1981 production capacities of those firms that produced the five polymers and resins.

The total capacity of the plants producing the five polymers and resins exceeded 15 thousand gigagrams in 1981. The basic chemical manufacturers provided approximately one third; major oil companies another third; and the remaining third was accounted for by plants owned by such diverse companies as Eastman Chemicals, USS, Bordens, and National Distillers.

The largest, single manufacturer of PP was Hercules, which accounted for nearly 25 percent of the total capacity. Oil companies produced about 40 percent.

Phillips Petroleum had the largest production capacity (about 25 percent) of HDPE. Oil companies, in general, accounted for approximately 60 percent of the HDPE market, and chemical companies provided 30 percent.

Table 9-2. CAPACITY OF POLYMER AND RESIN PLANTS BY MANUFACTURER AND TYPE; January 1, 1982 [a]

Manufacturer	PP	LDPE	HDPE	PS	PET	Total
	----------Gg----------					
A&E Plastics				15		15
Akzona					75	75
Allied Corporation			300		40	340
American Hoechst			135	285	295	715
American Petrofina				225		225
Atlantic Richfield	180	180	160	215		735
Avtex Fibers					20	20
BASF Wyandotte				215		215
Chemplex		190	120			310
Cities Service		330				330
Crest Container				20		20
Dart Industries				70		70
Dow Chemical		700	205	425		1,330
Du Pont		320	205		900	1,425
Eastman Chemicals	65	165			480	710
El Paso Natural Gas	135	250				385
Exxon	180	455				635
Fiber Industries					680	680
Firestone Tire & Rubber					25	25
Goodyear Tire & Rubber					135	135
Carl Gordon				70		70
Gulf Oil	180	390	260	160		990

Manufacturer	PP	LDPE	HDPE	PS	PET	Total
	----------Gg----------					
Hercules	600		5			605
Huntsman-Goodson				10		10
ICI Americas					35	35
InterNorth	90	275				365
Kama				10		10
Minnesota Mining					35	35
Mobil Chemical		135		90		225
Monsanto				315	90	405
National Distillers		325				325
National Petrochemical			285			285
Phillips Petroleum	90		680			770
Polysar Group				125		125
Rohm and Haas					90	90
Shell Chemical	270			135		405
Soltex Polymer	90		340			430
Standard Oil (Indiana)	235		170	195		600
Texstyrene Plastics				25		25
Union Carbide		960	90	5		1,055
United Foam		75				75
USS Chemicals	235			10		245
Vititek				5		5
TOTAL	2,350	4,750	2,955	2,625	2,900	15,580

[a] Source: Compiled from data contained in Tables 3-1 to 3-5.

9-6

Oil companies accounted for about 30 percent of the total LDPE capacity. The chemical companies accounted for close to 30 percent with the two largest chemical producers of this polymer producing about 20 percent of this amount.

The basic chemical companies accounted for close to 40 percent of the PS market with the two largest producers providing about 30 percent of the capacity--about the same as that of all oil companies.

PET capacity was essentially--close to 90 percent--devoted to manufacturing fiber. Du Pont had the largest capacity for production--30 percent of the total. About 10 percent of the PET capacity was in film and bottle resins, and 40 percent of that was produced by du Pont.

9.1.1.1.3 <u>Vertical integration</u>. Vertical integration is the operation of a single firm at more than one stage of production. For the purpose of this analysis, it is an indicator of the ability of firms to invest in pollution control, and to use or produce substitutes that could involve less air pollution. Also, vertically integrated firms may find it easier to pass compliance costs forward or backward. Vertical integration in polymers and resins follows two distinct patterns: "backward-to-suppliers" is characteristic of oil companies; "forward-to-final-product markets" is typical of chemical companies and of the subsidiaries of other types of manufacturers.

<u>Backward-to-suppliers integration</u>. Over a third of the polymers and resins production capacity is owned by oil companies that provide organic chemical inputs to production. The greatest portion of this backward integration involves large petroleum producers such as Exxon and Gulf. A relatively small portion is produced by industrial organic chemical producers that produce both the monomers and polymers. Essentially all of the backward integration is employed by the manufacturers of PP, HDPE, LDPE and PS; very few PET manufacturers are so organized.

<u>Forward-to-final product market integration</u>. Forward integration, which involves two or more of the stages of manufacturing, occurs among all of the different types of firms that produce one or all of the five polymers and resins. Firms that employ such integration account for

one-third of all polymers and resins production capacity. Chemical firms frequently employ forward integration: Dow Chemical, for instance produces polymers and resins and such end products as food wrap and insulated cups; Union Carbide carries its polymers and resins production through to food wrap products. Forward integration is also characteristic of particular divisions or subsidiaries of firms in such diverse industries as food processing, textiles, photographic equipment, and aerospace. In addition to its PET production, for instance, Eastman Chemicals manufactures plastics materials for molding camera housings. (Additionally, Eastman Chemicals maintains approximately 30 percent of all polyester film capacity.) Dart Industries supplies polymer to its household products manufacturing subsidiaries. Forward vertical integration is found also among other PET fiber producers: Firestone and Goodyear produce both PET fibers and the tires in which they are used; Monsanto produces the PET fibers used in its production of artificial turf. Texfi previously produced PET fibers for its textile mills; however, it discontinued this production in 1980.

9.1.1.1.4 <u>Horizontal integration--product diversification.</u>
Horizontal integration can be found among (1) firms that are essentially petroleum companies concerned with polymers and resins production, (2) firms that are primarily chemical manufacturers, and (3) firms within other, more diverse industries, e.g., food processing and textiles. Some 60 percent of all polymers and resins plants are owned by diversified and/or horizontally integrated firms which produce two or more polymers and resins.

All of the major petroleum corporations have horizontally integrated facilities for the production of at least two polymers. Atlantic Richfield and Gulf Oil operate separate plants to produce each of the polymers and resins except PET. Among chemical firms, all major corporations produce two or more polymers and resins in separate plants. National Distillers with such diverse operations as distilling, metal fabrication, and textiles is typical of other firms whose operations include the horizontally integrated production of two or more polymers.

The polymers and resins industry's single-plant operations are generally those producing PET. These constitute some 70 percent of the single-plant operations in PET.

9.1.1.1.5 Total production and capacity. Total production of the five polymers and resins increased from 11.3 thousand gigagrams in 1980 to 11.7 thousand in 1981.[3] (These data vary from data contained in Table 9-9 because of difference in the scopes of the sources.) Total polyester fiber production accounted for about 45 percent of all synthetic fibers produced. Of the five polymers, LDPE accounted for the greatest capacity--close to 30 percent while PET accounted for the lowest at 16 percent. In 1981, capacity utilization varied from a high of 90 percent in the production of PET (fibers) to a low of 70 percent for PS (Table 9-9).

9.1.1.1.6 Value of shipments. Values of shipments for the five polymers and resins are included in the total shipments data for three SIC industries: Plastics materials and resins (SIC 2821), Organic fibers, noncellulosic (SIC 2824), and Miscellaneous plastics products (SIC 3079). Values of shipments for SICs 2821 and 2824 are listed below for the period 1977-1981. Shipments for SIC 3079 are not shown, because the five polymers and resins are a minor portion of that sector's shipments and cannot be separated from the total.

The total value of shipments for industry 2821 increased from $10.8 billion (nominal) in 1977 to $17.5 billion (nominal) in 1981.[4] SIC 2821 does not show shipments by individual type polymers, but it does show them by categories--thermosetting and thermoplastic resins. Shipments for thermoplastic resins (SIC 28213), those which include PP, HDPE, LDPE, PS and polyvinyl chloride, amounted to $9.3 billion in 1977.[5] Although no values of shipments for thermoplastics have been published since 1977, estimates of the 1981 value can be made by assuming that thermoplastics grew at the same rate as did overall shipments for SIC 2821. Based on this assumption, the 1981 value of shipments for thermoplastics is estimated to be about $15.0 billion. Although the value of shipments of thermoplastics includes polyvinyl chlorides, the value of the four polymers produced in the plastics products industry (SIC 3079) are assumed to be about the same as that of the polyvinyl chloride;

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consequently, $15.0 billion is a reasonable estimate of the total SIC 2281 and 3079 shipments in 1981 for PP, HDPE, LDPE, and PS.

Shipments for SIC 2824 Organic fibers, noncellulosic increased from $6.4 billion in 1977 to $11.0 billion in 1981.[6] Assuming that the growth of PET approximated that for the total industry, the shipments for PET fibers (SIC 28244) are estimated to have increased to $3.8 billion in 1981. The sum of the shipments of thermoplastics and PET fibers, as shown below, is $18.8 billion, an approximation of the value of the five polymers and resins.

The industries' values of shipments for 1977 and estimated values (of the five polymers) for 1978, 1979, 1980, and 1981 are tabulated below:

Industry	1977	1978	1979	1980	1981
	----------(million nominal dollars)---------				
SIC 2821 Plastic materials and resins	10,818	11,998	14,282	15,570	17,500
SIC 28213 Thermoplastic materials and resins	9,266	10,277[a]	12,233[a]	13,340[a]	14,990[a]
SIC 2824 Organic fibers noncellulosic	6,380	6,921	8,227	8,811	11,035
SIC 28244 PET fibers	2,187	2,373[a]	1,821[a]	3,021[a]	3,783[a]

[a] Estimated by EPA

9.1.1.2 <u>Plants.</u> The considered polymers and resins are produced in over 100 plants with varying characteristics. Some of the more important characteristics--size, age, location, and employment--are discussed in this section.

9.1.1.2.1 <u>Size.</u> The annual production capacities of plants producing the polymers and resins vary from less than 10 gigagrams for the

smaller plants, which produce PET fiber and film, to more than 400 gigagrams for the largest polyethylene plants. Except for the large du Pont and Fiber Industries plants with capacities in excess of 100 gigagrams, most PET plants have capacities of less than 50 gigagrams.

Table 9-3 shows the size distribution of the polymer and resin plants. The distribution of the LDPE and HDPE plants is similar--about 70 percent of the plants of each type have capacities in excess of 100 gigagrams.

9.1.1.2.2 Age.[7] Data are not published on the ages of individual plants listed in Table 9-1; consequently, plant age distributions cannot be developed. Information is available that indicates when the different types of polymers and resins were first introduced. PS was introduced in 1938 and used in manufacturing housewares. LDPE was introduced in 1942 for use in packaging. HDPE and PP were introduced in 1957; PET was introduced in 1970.

9.1.1.2.3 Location. The 128 known plants producing the five polymers and resins are located in 21 states and Puerto Rico (Table 9-4). The high capacity plants producing the polyolefins (polyethylene and PP) are predominately located in the petroleum producing regions of Texas and Louisiana. The PS plants are located primarily in industrial states: Illinois, Massachusetts, Ohio, and New Jersey. The PET plants are located in large textile mill areas in order to supply fibers to that industry, and over 60 percent are located in the Southeast.

9.1.1.2.4 Employment.[4,6] The 1980 employment data for plants producing the five considered polymers and resins are included into the SIC data.

As shown in Table 9-5, in 1981 an estimated 66,100 workers were employed in plants producing the five polymers and resins: 34,100 in plants producing polymers only; 10,700 in processing plants captively producing polymers; and 21,300 in PET fiber plants. Total employment in thermoplastics increased slightly from 1980 to 1981; however, employment in PET fibers decreased significantly a condition reflecting an overall reduction in textile industry production. The ratio of production workers to all workers is significantly greater in the fiber plants (SIC 2824) than in the plastics plants (SIC 2821), a difference that reflects the labor intensiveness of these plants.

Table 9-3. SIZE DISTRIBUTION OF POLYMER AND RESIN PLANTS
BY TYPE AND CAPACITY; January 1, 1982[a]

(percent)

| Product | Capacity in gigagrams | | | | |
	100 and below	101 to 200	201 to 300	301 and above	Total
PP	43.8	37.5	12.5	6.3	100.0
LDPE	10.0	40.0	35.0	15.0	100.0
HDPE	25.0	50.0	12.5	12.5	100.0
PS	75.7	21.6	2.7	-	100.0
PET	59.3	22.2	14.8	3.7	100.0

[a] Source: Compiled from data contained in Tables 3-1 to 3-5.

Table 9-4. NUMBER OF POLYMER AND RESIN PLANTS
BY TYPE AND LOCATION; January 1, 1982[a]

Location	PP	LDPE	HDPE	PS	PET	Total
Texas	11	13	11	7		42
Illinois	1	2		7		10
South Carolina					10	10
Louisiana	2	4	3			9
North Carolina				1	8	9
Ohio				6	1	7
California				6		6
Massachusetts				5		5
New Jersey	1			3	1	5
Alabama				1	3	4
Tennessee					4	4
Pennsylvania				2	1	3
Virginia				1	2	3
Iowa		1	1			2
Kentucky		1		1		2
West Virginia	1				1	2
Colorado					1	1
Connecticut				1		1
Michigan				1		1
New York					1	1
Puerto Rico		1				1
TOTAL	16	22	15	42	33	128

[a] Source: Compiled from data contained in Tables 3-1 to 3-5.

Table 9-5. EMPLOYMENT IN POLYMER AND RESIN PLANTS IN 1977 AND 1981 [4, 5]

Industry	SIC Code	1977	1981	Assumptions
1. All plastics materials and resins	2821	57,200	53,500	
2. Thermoplastic	28213	46,700	43,700	The proportion of the number of employees in thermoplastics to the total number in SIC 2821 is the same in 1981 as in 1977.
3. Polymers (PP, HDPE, LDPE, PS)	28213	N.A.	34,100	Excludes employment in PVC which is 22 percent of the total in SIC 28214; this is based on the proportion of PVC production to the total production.
4. Processors captively producing polymers	3079	N.A.	10,700	The applicable employment in SIC 3079 is 20 percent of the total employment in SIC 2821.
5. Organic fiber	2824	74,000	69,000	
6. Polyester fibers (PET)	28244	22,800	21,300	The proportion of the number of employers in PET to the total number in SIC 2824 is the same in 1981 as in 1977.
Total (PP, LDPE, HDPE, PS, PET) (3+4+6)			66,100	

N.A. = not available.

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9.1.1.3 <u>Markets.</u>[8] The versatility of plastics has made them
suitable for use in such diverse markets as furniture, automobiles,
housing, packaging, toys, and electronics. Their special properties and
high-volume processing capabilities combine to give them a competitive
edge over wood, metals, paper, and glass, and most especially in the
packaging and consumer products industries. The largest product by volume
of the plastics and resins is LDPE, a product used as film in the
packaging industry. PP has realized the most rapid production growth in
recent years--generally at the expense of PS--with significant applica-
tions in both the auto industry and the fibers market.

9.1.1.3.1 <u>Substitutes.</u> Plastics materials and fibers derived from
the five polymers and resins are used as substitutes for wood, metal,
natural fibers, paper, glass, and other plastics and synthetic fabrics in a
variety of industries. PET derived fibers, for instance, are widely used
substitutes for cotton fabrics; PP is used as a nylon substitute for
automotive fabrics and home furnishing (especially carpeting) materials;
LDPE is an extensively used packaging material substitute for paper; and
HDPE substitutes for steel and other plastics in the fabrication of pipes.
One of the most rapidly developing markets is the automotive industry where
the emphasis on fuel efficiency has lead to the use of lighter-weight
substitutes for metal; PP, for instance, is now used extensively in the
fabrication of such automotive body components as fenders and doors.

9.1.1.3.2 <u>Imports-Exports.</u> The U.S. balance of payments has been
strengthened by the consistent export surplus of both plastics materials
and resins and synthetic organic fibers. The 1981 export surplus in
plastics was about $3.0 billion; that in synthetic organic fibers was $576
million.[9] However, as U.S. raw materials prices gravitate toward the
world prices as a result of natural gas and oil deregulation, domestic
producers will lose much of their past competitive advantage over foreign
producers. Additionally, developing countries with extensive gas and oil
exports are planning to bring petrochemical complexes on line in the 1980s
to process their raw materials. Saudi Arabia, Mexico, and some Far
Eastern nations, for instance, plan major world-scale plants. Among the
more industrialized nations, Great Britain, Norway, and Canada plan to
develop processing plants to exploit their newly discovered oil and gas
supplies.[8]

Table 9-6 shows the exports of the five polymers and resins for the period 1976-1981.[9, 10] The greatest increase in exports for the period occurred for PET fibers--over 500 percent. The next greatest increase is shown for LDPE--an increase of about 76 percent. No increase is shown for PS. The greatest volume of exports--over 400 gigagrams exported in 1981--was of LDPE, and the volumes of PP and PET exports were 311 and 400, respectively. The lowest exports were for PS--fewer than 100 gigagrams exported.

Polymer imports have been relatively minor--less than 100 gigagrams in 1981--and most of it was of HDPE. In PET fibers, imports reached about 15 gigagrams, an increase over the approximately 8 gigagrams reported in 1980.[10]

9.1.1.3.3 <u>Product differentiation</u>. The 1981 consumption of polymers and resins by type of processing is shown in Table 9-7. About 3,000 gigagrams of LDPE were processed with over 60 percent of this amount extruded into film. Of the 1,400 gigagrams of PP processed, about a third was extruded as fibers, another third processed by injection molding, and the remainder was processed into film and other products. The largest portion of the 1,900 gigagrams of HDPE processed was molded, 40 percent by blow molding and about 25 percent by injection molding. About one-half of the PS processed was injection molded. The primary use of PET was for fibers and filaments.

9.1.1.3.4 <u>End-use markets</u>. Shipments to end-use markets in 1981 of PP, HDPE, LDPE, PS, and nonfiber PET are shown in Table 9-8. Over 4,000 gigagrams of these polymers and resins were shipped to the packaging industry for fabrication into such products as bottles, jars, food containers, refuse bags, and baskets and LDPE represented almost 50 percent of the plastics so shipped. Another 1,000 gigagrams entered the consumer (toys, kitchenware) and institutional markets: PS accounted for one-third of this, and the remaining two-thirds was divided about equally among the polyolefins. Approximately 190 gigagrams were supplied to the transportation industry with PP accounting for about 60 percent of this amount. About 400 gigagrams entered the furniture and home furnishings market with nearly 90 percent of this as PP, essentially for use in textile products such as carpets and drapes. HDPE contributed about half of the 400 gigagrams utilized in the building and construction industry.

Table 9-6. U.S. EXPORTS OF POLYMERS AND RESINS, BY TYPE AND YEAR,
1976-1981 [9, 10]
(gigagrams)

Type [a]	Year					
	1976	1977	1978	1979	1980	1981
PP	161	128	168	324	308	311
LDPE	254	260	343	407	520	446
HDPE	166	193	223	276	239	209
PS	70	53	71	73	76	68
PET	63	80	150	238	330	400

[a] Exports represent sales of plastic resins for all types except PET; exports of polyester indicated are fibers. Less than 20 gigagrams of polyester resins were exported in 1981.

Table 9-7. DOMESTIC CONSUMPTION OF POLYMERS AND RESINS BY END-USE MARKET AND PROCESS TYPE, 1981 [9, 10]
(gigagrams)

Process; End-use market	PP	LDPE	HDPE	PS	PET
Extrusion:					
Film	137	1,849	127	NR	110
Sheet	16	6	24	NR	NR
Wire and cable	a	157	50	NR	NR
Pipe and conduit	a	11	185	NR	NR
Fiber and filaments	482	NR	NR	NR	1,490
Other extruded products	33	284	24	NR	NR
Total	668	2,307	410	518	1,600
Injection molding	564	238	452	743	0
Blow molding	29	27	769	0	170
All Others	191	391	284	305	0
Total	1,452	2,963	1,915	1,566	1,770

a Included in "other extruded products."

NR = Not Reported.

Table 9-8. SHIPMENTS OF POLYMERS AND RESINS BY MAJOR MARKET, 1981 [9, 10]
(gigagrams)

Market	PP	LDPE	HDPE	PS	PET [a]
Transportation	121	24	29	7	10
Packaging	368	2,087	1,074	678	202
Building; Construction	b	67	220	124	d
Electrical; Electronic	112	196	66	163	15
Furniture; Furnishing	369	7	b	28	d
Consumer; Institutional [c]	221	271	219	343	5
Industrial; Machinery	25	b	46	b	d
Adhesives; Inks & Coatings	b	b	b	b	d
All Other	236	311	261	224	d
Exports	311	446	209	68	19
Total	1,763	3,409	2,124	1,635	264

a   Non-fiber PET only.

b   Included in "all other."

c   Consumer and institutional products include kitchenware, coolers, health care products, sporting goods, toys, etc.

d   Withheld to avoid disclosure.

Not shown in Table 9-8 is the shipment of polyester fibers. Approximately 90 percent of these fibers were shipped to the textile industry for processing into yarn for the apparel and related fabric industries in 1981. PET continued to be the predominant noncellulosic fiber. It accounted for over 50 percent of the noncellulosic output and over 30 percent of the total of all fibers, both man-made and natural, consumed in the textile industry.[11]

9.1.2 Industry Profile

The initial commercial development of the major thermoplastics occurred in the period 1930-1940 with the introduction of such plastics as PS and LDPE. In the 1950's HDPE and PP were introduced. Large scale productions of these reduced their cost and they began to compete with materials such as wood, cotton, paper, metal, and glass. The current general production marketing and financial characteristics of the industry are examined below.

9.1.2.1 Capacity and Production.[9, 10] Yearly production and capacity data for the period 1976-1981 are shown in Table 9-9. During the period, PP had the highest average annual rates of increase both in capacity, about 13 percent, and production, about 10 percent. PP capacity notably increased about 40 percent in 1978-1979; however, it decreased about 5 percent 1980-1981. HDPE averaged annual increases of about 9 and 10 percent in production and capacity, respectively. Production and capacity in LDPE increased at average rates of about 6 and 8 percent, respectively, and the rate of average annual increase for PS was 2 and 1 percent for production and capacity, respectively. PET fibers had the lowest increase during the period with an average annual 3 percent increase in production and less than one percent in capacity. PET fiber capacity decreased 4 percent during 1980-1981--essentially because of the closing of the Texfi operation.[12]

Utilization rates were generally highest for LDPE, varying from a low of 79 percent in 1981 to a high of 93 percent in 1979. The lowest utilization rates occurred in the manufacturing of PS--varying from a low of 64 percent in 1976 to 77 percent in 1979. Such low rates reflect the competition in this relatively mature sector.

Table 9-9. PRODUCTION, CAPACITY, AND CAPACITY UTILIZATION OF POLYMERS AND RESINS [9, 10]

		PP	LDPE	HDPE	PS	PET [a]
1976	Production (Gg)	1,174	2,640	1,417	1,450	1,630
	Capacity (Gg)	1,230	3,000	1,680	2,260	2,070
	Utilization (%)	97	88	84	64	79
1977	Production (Gg)	1,246	2,935	1,657	1,563	1,640
	Capacity (Gg)	1,450	3,220	1,815	2,385	2,040
	Utilization (%)	86	91	89	66	80
1978	Production (Gg)	1,394	3,226	1,906	1,730	1,720
	Capacity (Gg)	1,540	3,610	2,150	2,360	2,090
	Utilization (%)	91	89	99	73	82
1979	Production (Gg)	1,740	3,530	2,270	1,820	1,890
	Capacity (Gg)	2,180	3,810	2,470	2,360	2,180
	Utilzlation (%)	80	93	92	77	87
1980	Production	1,650	3,310	2,000	1,600	1,810
	Capacity (Gg)	2,400	3,950	2,540	2,450	2,180
	Utilization (%)	69	84	79	65	83
1981	Production	1,790	3,490	2,130	1,640	1,890
	Capacity	2,290	4,430	2,720	2,350	2,100
	Utilization (%)	78	79	78	70	90

[a] Fiber only.

In 1980, the utilization rate for PP dropped to a low of 69 percent as a result of that year's recession in the automobile and housing industries. In 1981, the rate increased to 78 percent, an increase brought about by increased production and a decrease in capacity from plant closings. The utilization rates for PET varied between 79 and 90 percent during the six-year period. As in the case of PP, the high rate of 1981 reflects both a decrease in capacity and an increase in production.

9.1.2.2 Prices. Polymer and resin prices have fluctuated considerably during the last ten years. As measured by the Producer Price Index for SIC 2821, Plastics materials and resins, nominal prices increased 56 percent during 1973-74 as inventory buildups occurred as a result of the OPEC oil embargo and anticipated ensuing shortages. Prices remained relatively stable from 1974 to 1978. In 1979, however, they jumped by 18 percent as oil prices increased.

The 1978 and 1981 prices are compared below for each of the polymers and resins. The prices, based on data contained in Chemical and Engineering News, indicate price fluctuations during each of the years.[13,14,15,16] The greatest increases occurred in the prices for PP and PS with increases of about 45 percent for each. These increases primarily reflect the post-1977 rising demand for these polymers in the automobile and home furnishings industries. The price increase for the other polymers amounted to about 30 percent.

Price ranges [13,14,15,16]

(nominal)

Product	1978	1981
	-------------($/kg)-------------	
PP	.60-.73	1.10-1.23
LDPE	.64-.73	1.01-1.06
HDPE	.68-.73	.99-1.08
PS	.56-.66	1.10-1.14
PET	1.43-1.54	1.85-2.58

9.1.2.3 <u>Finances</u>.[8] Plastics have experienced relatively good profits over the past two decades because of their exceptionally high demand as a replacement for such materials as metal, wood, glass, and rubber; indeed, this demand has fueled relatively high growth rates in production of all types of plastics. The Federal Reserve Board's production index indicates that the annual growth for all plastics materials averaged 9.5 percent per year between 1973 and 1979 compared with the 3.1 percent growth rate for all industrial products. However, in 1980, plastics production fell 9 percent while the overall industrial output declined by only 3.5 percent. This comparatively large drop reflected the severely depressed state of two key plastics markets--the automobile and construction industries. Because many of the plastics markets are expected to mature within the next ten years, long-term growth is not expected to exceed 6 percent, an amount well below past growth rates. Recent trends in sales and profits for the five polymers and resins are outlined below.

9.1.2.3.1 <u>Sales and value of production</u>. The trends in sales (nominal dollars) and value of production (nominal dollars) over the past several years is shown in Table 9-10. The sales data reflect the actual sale of polymers and resins on the open market and exclude interplant transfers and captive consumption; however, data for the estimated value of production of polymers and resins produced (for sale and inventory) do include captive consumption and interplant transfers.

The highest domestic dollar sales during the 1978-81 period for the four polymers and resins (sales for PET fibers are not available) are shown for LDPE. Its sales increased from $1.5 billion in 1978 to over $1.9 billion in 1980. Sales for this polymer then increased only slightly from 1980 to 1981, from $1.9 billion to about $2.1 billion. The greatest sales increase during the period occurred for HDPE with the level increasing from less than $1 billion dollars in 1978 to over $1.4 billion in 1981.[17]

The value of production for all five polymers increased significantly between 1978 and 1979, stabilized between 1979 and 1980 and increased between 1980 and 1981. The greatest value of production is shown for PET fibers. The value for this polymer increased from $2.4 billion in 1978 to

Table 9-10. SALES AND VALUE OF PRODUCTION OF POLYMERS AND
RESINS [12,13,14,15,16,18,19,20] 1978-1981 (million nominal dollars)

Item	1978	1979	1980	1981
Domestic sales				
PP	612	695	793	1,015
LDPE	1,492	1,901	1,974	2,077
HDPE	828	1,148	1,237	1,414
PS	978	1,387	1,358	1,446
PET [a]	NA	NA	NA	NA
Value of production				
PP	800	1,250	1,250	2,000
LDPE	2,000	2,800	2,800	3,300
HDPE	1,200	1,700	1,750	2,250
PS	1,000	1,500	1,500	1,800
PET	2,400	3,400	3,500	3,850

NA = Not available

[a] Domestic sales data are not available for polyester fibers.

close to $3.9 billion in 1981. The increases in values for the polymers generally varied between 10 and 20 percent from 1980 to 1981 except for PP. Its value increased from $1.25 billion in 1980 to $2 billion in 1981--a 60 percent increase.[13,14,15,16,18,19,20]

9.1.2.3.2 Profits.[21] The profitability of firms classified in SIC Industry 2821, Plastics materials and resins, is shown below for the period 1974-1981. These before-tax profit margins are based on Robert Morris Associates Annual Statement Studies, a publication that incorporates the financial data of about 130 firms in the plastics industry; however, because this publication excludes data from firms having assets in excess of $50 million, the profitability values shown do not reflect the profiles of the larger firms that produce polymers, e.g., du Pont, Dow Chemical, and the large petroleum companies. It may, however, cover some of the small subsidiaries of these large corporations. For the most part the returns reflect only the profits of the smaller, single-plant operations.

BEFORE-TAX PROFIT MARGINS [21]

(Profit as a percent of sale)

1974	1975	1976	1977	1978	1979	1980	1981
8.6	6.0	7.8	3.2	4.8	3.7	3.6	3.6

Composite returns in 1974 approached 9 percent of sales and reflected the effects of the general price increases and inventory reductions in 1973. In 1975, the profits dropped to 6 percent of sales. Since 1976, profits have varied between 3 and 5 percent.

9.1.3 Five-Year Projections

Capacity increases are projected below for the period January, 1984 to January, 1989 and are expressed in terms of required new plants. Twenty-seven new plants equivalent in size to the model plants (defined in Chapter 6) are projected. The numbers of plants by type of polymer and process are:

Polymer/Process	Number of plants
PP	
Liquid phase	3
Gas phase	3
LDPE	
Liquid phase	1
Gas phase	5
HDPE	
Liquid phase-slurry	2
Liquid phase-solution	3
Gas phase	5
PS	
Continuous	2
PET	
DMT	1
TPA	2

The methodology used to project capacity increases expressed in terms of the number of new model plants is discussed below. No explicit locational component was included in these projections, for the model plant construction and the baseline control estimation (See Chapter 8) took into account the current geographic distribution (SIP and non SIP states) of the existing plants and assumed new plants would mirror the same distribution, capacity, and emission controls. The numbers of plants by process result from a growth analysis of each polymer, as described below. The projection is subject to some uncertainty because a few markets, as indicated in Section 9.1.2 above, are currently unstable, and such instability begets caution. Chemical industry opinion indicates that there may be few new grass roots polymers and resins facilities built over the next several years.[18] Such assumptions imply that unstable markets mean that some growth in production will come from the upgrading or demothballing of existing production units in ways that may not technically be considered modifications under the Clean Air Act, and, therefore, not subject to regulation by an NSPS. (See Chapter 5 for a definition of modification.) If this assumption proves correct and the projection of 27 new plants is excessive, the projected national annualized cost of the proposed standards may be overstated. However, in Section 9.2 the economic effects of the individual regulatory alternatives were found to be so small that a more

detailed study of future growth is not warranted in order to guide the selection of proposed regulatory alternatives.

New capacity and plants are projected using the two equations:

$$NC = FP \left(\frac{1}{CU} \right) - CC + RC - IC \tag{1}$$

where

NC = new capacity

FP = 1988 production

CU = 1988 capacity utilization rate

CC = current capacity (1981)

RC = retired capacity, 1984-88

IC = interim capacity, 1982-83

and

$$NP = NC \div MP \tag{2}$$

where

NP = new plants

NC = new capacity

MP = model plant size

The factors and assumptions used and the results of the calculations, tabulated in Tables 9-11 and 9-12, are discussed below.

9.1.3.1 Projected Production. The 1988 production of each of the five polymers and resins was projected by applying the 1981 domestic production and net exports growth rates as determined from published projections.[23,24] These published growth rates were developed from historical trends and end uses data, from conversations with industry personnel, and from considerations of worldwide developments.

In general, new production capacity announced for Canada and other petroleum producing countries has led some analysts to conclude that U.S.

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plastics exports will decrease as worldwide supply and consequent U.S. imports increase. In addition, price decontrols have decreased the feedstock advantage of U.S. producers. Assessing the import-export situation is difficult because it can change quickly in response to governmental policy decisions, especially trade restrictions.[25]

Domestically, the continuing decline in demand since 1979 has resulted in the industry's carefully examining its capacity and adjusting it by temporary and permanent closures.

In order to determine the appropriate annual growth rates used in projecting the 1988 production of each of the polymers and resins, growth rates published within the last three years were assessed. To more completely specify the projected demand, domestic concensus growth rates were determined for both domestic markets and net exports. (Consensus rates were used in order that the NSPS cost estimates would be conservative · and not underestimated.) The annual growth rates assumed for each polymer and resin are discussed individually below.

PP. Published annual growth rates for domestic consumption of PP for the period to 1990 vary from 5.4 to 8.0 percent depending in part on the time at which the projections were made.[23] An intermediate value of 7.0 percent was used as the basis for the projection of domestic demand.

Published net exports growth rates to 1990 were -4.7 percent and -7.2 percent. For this analysis the higher value of -4.7 percent was used in order that the projections would be conservative and the NSPS cost estimates would not be underestimated.

LDPE. A complicating factor in the projections for LDPE is the production of the relatively new polymer linear low density polyethylene (LLDPE). For these projections a combined LDPE and LLDPE production was considered. In assessing these projections, it should be noted that with but a small capital investment, LDPE producers can switch to LLDPE production.

Published growth rates for the combined LDPE and LLDPE consumption to 1990 varied from 5.0 to 6.0 percent.[23] For this analysis a value of 5.6 was selected as most representative of values given.

Annual growth rates for net exports ranged from -19.9 to -1.1 percent.[19] Due to the variation, a growth rate for net exports of -10.0 was selected for use in the analysis.

HDPE. Published annual growth rates for HDPE range from 6 to 10 percent with most analysts projecting values slightly less than 8.[23,26] A growth rate of 7.9 percent was used. In assessing the growth of HDPE, the secondary capacity in LLDPE plants is acknowledged.[27]

Since the published growth rates for net exports ranged from -15.0 to +8.4 percent, net exports were assumed to remain constant in this analysis.

PS Since less than five percent of the polystyrene production is exported and imports are negligible, no differential annual growth rates were determined. PS is a mature industry and is not expected to show a high growth rate. A concensus annual growth rate of 4.0 percent was assumed on the basis of published projections ranging from 3.6 to 4.5 percent.[23]

PET. Growth in the polyester fiber portion of the PET industry is predicted to turn positive again in 1984 and increase at an annual rate of about 4 percent through 1990.[28] Polyester resins for films and bottles account for about 10 percent of the total PET production. This portion of the industry is expected to increase somewhat more rapidly.[24] For this analysis no growth was assumed to 1984. From 1984 on, a growth rate of 4 percent was used.

The situation for net exports is not clear cut. The traditional export outlet for PET fibers has been Western Europe, but since that area now experiences considerable overcapacity, some limitations have been placed on imports. In 1981 exports to the People's Republic of China were large, but declined in 1982. Some industry analysts indicate that the 1982 decrease in U.S. exports probably signals a long-term trend. However, due to the uncertainties in the import/export situation, net exports were assumed to remain constant in order to arrive at a conservative projection.

A sensitivity analysis was not conducted in light of the low level of price effects as discussed below.

9.1.3.2 Projected Capacity. Projected capacity required in 1988 was calculated by applying a utilization rate to the projected 1988 production.

In order to project the required capacity, a capacity utilization rate that will trigger additional capacity must be assumed. The capacity utilization rates used were estimated on the basis of published information[29] and discussions with industry representative.[30] These sources indicate that because the industry is generally profitable at an overall industry utilization rate of 85 percent, a utilization rate in the high 80's will trigger capacity expansion. Thus, the arbitrary assumption was made that an 88 percent utilization rate would trigger expansion in the PP, LDPE, HDPE and PET industries. Since PS is a mature industry, but one with new specialty uses being developed, a slightly lower utilization rate of 85 percent was assumed to trigger expansion. It is well to note, of course, that individual companies do not base their expansion plans solely on consideration of overall industry utilization rates, for other company-specific factors also affect expansion decisions. However, for these industry projections the overall industry utilization rates were used.

The projected 1988 capacities for the five polymers and resins, calculated using the assumed utilization rates, are shown in Table 9-11.

9.1.3.3 <u>Retired Capacity</u>. In calculating the retired capacity, the useful life of all plants was assumed to be 20 years. Therefore, any capacity built between January, 1964 and January, 1969, was assumed to be retired during the five-year period examined in the analysis. The portion of this retired capacity that will be replaced by plants coming under the potential NSPS varies by polymer process. For this analysis, all 20-year old PP, LDPE, HDPE and PS plants were assumed to be replaced. For the PET industry where there are no known plans to rebuild a portion of the retired capacity, it was assumed that only half the retired capacity would be replaced by 1989. The assumed retired capacities, then, are those units constructed between January, 1964 and January, 1969. These values, tabulated from the "Construction Alerts" published in the April and October 1964-1968 issues of <u>Chemical Engineering</u>, are shown in Table 9-11.[31]

9.1.3.4 <u>Interim Capacity</u>. Interim capacities are the capacities of plants to be completed in 1982 and 1983. Values for PS, LDPE, HDPE and PS were taken from the <u>Facts and Figures of the U.S. Plastics</u>

Table 9-11. ESTIMATED REQUIRED NEW CAPACITY FOR POLYMERS AND RESINS, 1988

Product	Growth rate, domestic	Growth rate, net imports	Projected 1988 production	Utilization rate	Projected 1988 capacity	Interim capacity 1982-83	Retired capacity 1984-1988	Total required new capacity
	%	%	Gg	%	Gg/yr	Gg/yr	Gg/yr	Gg/yr
PP	7.0	-4.7	2,590	88	2,940	0	115	765
LDPE	5.6	-10.0	4,665	88	5,300	390	520	1,000
HDPE	7.9	-0-	3,510	88	3,990	195	330	1,410
PS	4.0	4.0	2,160	85	2,540	160	120	150
PET	4.0[a]	-0-	2,140	88	2,440	90	180[b]	340

[a]Zero through 1984; 4% 1985-1988.

[b]Assumed only half the retired capacity replaced. (See text p. 9-33.)

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Industry.[9] Values for PET were taken from Textile Organon.[10] The interim capacities are shown in Table 9-11.

9.1.3.5 New Capacity. New capacities are calculated by using equation (1). The 1988 production for all polymers, except PS, was estimated by applying the applicable growth rate to the 1981 consumption and net exports, and summing the two. For PS, the 1981 production level was increased by its growth rate. These values are tabulated in Table 9-11.

In projecting the new capacity required in 1988 it should be noted that new capacity does not necessarily translate into new plants; indeed, it may not translate completely to other affected facilities, i.e., modifications and reconstructions. Additional capacity may be obtained in a variety of ways including:

- debottlenecking existing plants
- adding a process train
- converting facilities now or formerly used to produce other polymers to the production of high demand polymers
- bringing back on stream plants now on standby or mothballed
- modifying production processes in order to enhance conversion rates, e.g., changing the catalyst
- modifying existing plants
- reconstructing existing plants, and
- constructing new facilities

Only new plants and older plants modified or reconstructed (as defined by the Clean Air Act) will be affected by this potential NSPS and, thus, only they are of interest in this study. For purposes of this study it was assumed that all affected facilities will be new plants, i.e., no modifications and reconstructions are included. (See Chapter 5.)

9.1.3.6 New Plants. The projected new capacity requires the equivalent of 27 new model plants. However, industry sources indicate that not all the new capacity will be in the form of new grass roots plants—some capacity will stem from adding new process trains to existing plants. Insufficient information is available, however, to allow the projection of

the amount of new capacity that will be added by new plants and that which will be added by additional process trains at existing plants.

Another factor to take into consideration in projecting plant growth by process is that technology is now available to shift both conventional HDPE and high-pressure LDPE plants over to low-pressure operations to make either LLDPE or HDPE. Thus, data on plant source for the two traditional types of polyethylene, LDPE and HDPE, are becoming increasingly unreliable. In the next few years, individual sources of high or low density polyethylene will probably be difficult to ascertain. Capacity reporting could change to reflecting combined polyethylene data, and eventually the density differentiation could end as have property distinctions among other polymers.[32]

Additionally, although some new capacity might be expected to come from modifications and reconstructions, the analysis makes no projections of modifications or reconstruction, i.e., changes in a plant that would bring an existing plant under the potential NSPS (See Chapter 5).

For purposes of this analysis, therefore, the projected new capacities are expressed in terms of new model plants.

9.1.3.7 <u>Projections by Process</u>. In order to categorize these projected plants by process, it is necessary to determine the probable proportion of these plants that will be devoted to each process. The assumptions underlying those determinations, based on discussions with industry and trade associations representatives, are discussed below. The projected number of new plants by process are tabulated in Table 9-12.

9.1.3.7.1 <u>PP plants</u>. Both liquid-phase and gas-phase process plants will be built in the future. Companies will construct gas-phase plants if they already have the appropriate proprietary technology; however, if they do not, they are more likely to build plants using liquid-phase process technology. It was assumed that approximately equal use will be made of the two processes; therefore, three new gas phase and three new liquid phase process plants are projected. The capacity of these six plants, 765 Gg/yr, is equal to the projected required new capacity.

9.1.3.7.2 <u>LDPE plants</u>. Most new construction will employ gas-phase, low pressure, Unipol or similar proprietary process technology;

Table 9-12. PROJECTED NUMBER OF NEW POLYMERS AND RESINS PLANTS, BY PROCESS TO 1988

Product	New capacity required[a] Gg/yr	Process	Model plant capacity Gg/yr	Number of new plants built 1984-1989
PP	765	Liquid phase	150	3
		Gas phase	105	3
LDPE	1,000	Liquid phase	280	1
		Gas phase	150	5
HDPE	1,410	Liquid phase slurry	210	2
		Liquid phase solution	90	3
		Gas phase	150	5
PS	150	Continuous	75	2
PET	340	DMT	105	1
		TPA	105	2

a See Table 9-11.

however, the use of the high pressure liquid process cannot be ruled out. The consequent projections assume that one liquid-phase plant will be built and that the remainder of the capacity will be added with gas-phase plants. These assumptions result in one high-pressure, liquid-phase plant and five low-pressure, gas-phase plants having a combined capacity of 1030 Gg/yr, somewhat more than the 1000 Gg/yr required capacity projected.

9.1.3.7.3 HDPE plants. Both liquid-phase and gas-phase process plants will be constructed. The gas-phase process plants are cheaper to build and operate since the process catalyst remains in the polymer. However, if companies do not have the appropriate proprietary, gas phase technology, they will build liquid phase plants. The choice between these two depends upon technology licensing costs. The projections assume equal capacity for each process and result in five gas phase plants.

Further, the liquid phase plants may be either slurry or solution. Essentially equal capacity was assumed for each process to yield two liquid phase slurry plants and three liquid phase solution plants.

The projected new plants have a combined capacity of 1440 Gg/yr; the projected required capacity is 1410 Gg/yr.

9.1.3.7.4 PS plants. Two types of PS plants exist today. Production of bulk plastics, i.e., large quantities of one formulation, is most efficiently accomplished in a continuous plant. On the other hand, if a firm's market requires many grades and specialized materials, then a batch plant is more suitable. Since indications are that most new capacity will probably be added in the form of continuous process plants, only a continuous process model plant is described. The projected new capacity would require two plants.

9.1.3.7.5 PET plants. New PET plants may use either the terephthalic acid (TPA) process or dimethyl terephthalate (DMT) process. Industry representatives indicate the TPA process is more likely; therefore, since the TPA and DMT model plants are the same size, two TPA process plants and one DMT plant are projected. The combined capacity of the PET plants is 315 Gg/yr, slightly less than the 330 Gg/yr projected. However, considerations of the current industry conditions in which plants are being closed down[33] make it unlikely that a fourth new plant would be built by January, 1989.

Overall PET plant capacity remained constant in 1979 and 1980 and was reduced in 1981 from 2,180 Gg to 2,100 Gg. Although the PET industry has experienced growth in the bottle segment and more rapid growth is projected for that segment, it constitutes less than 10 percent of the 1981 capacity and was operated at less than 50 percent of capacity utilization. Industry representatives indicate that by modifying present facilities, capacity can be increased to meet demand without new plants until 1986.[34] The projection assumes, therefore, that some new capacity required for PET production would be obtained by increases in capacity that would not be subject to the potential NSPS. By 1988, three new plants should meet the additional demand.

9.2 ECONOMIC IMPACT ANALYSIS

Section 9.2 discusses the economic impact analysis methodology and the potential impacts of regulatory alternatives controlling VOC emissions from new source polymers and resins manufacturing processes. Generally speaking and as the impact analysis methodology discussion will show, the potential economic impacts of VOC emission controls, though real and of measurable impact, will have little significant impact on the 27 considered plants.

Total additional annualized costs of controls in 1988 (Table 9-16), the fifth year of controls, for the projected 27 polymers and resins manufacturing plants are estimated to be but $4.9 million dollars under the combination of the most stringent regulatory alternatives for each plant. Thus, detailed "Regulatory Impact Analysis" as prescribed by Executive Order 12291 is not required.

The potential economic impacts of the regulatory alternatives are expected to be very minor in view of the small price increases anticipated as a result of the control costs. Assuming that the incremental costs can be passed forward by each of the new source facilities, price increases required by the facilities for any of the alternatives would be less than four-tenths of a percent, except for the one HDPE liquid phase, solution facility. The costs of Regulatory Alternative 3 for this plant is substantially higher than those for the other projected plants. However, even though these costs are significantly higher, the maximum price

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increase required to fund the most stringent regulatory alternative would not exceed 2 percent; price increases required for the least stringent alternative for this plant is less than 0.1 percent. Because of the minor increases required by this and the other plants, no significant economic impact of the potential NSPS are expected.

Section 9.2.1 will discuss this study's revenue and price impact assessment methodology. It will be followed by 9.2.2--the economic impacts projected by that methodology.

9.2.1 Economic Impact Assessment Methodology: Revenue and Price

As explained in Chapter 8, each regulatory alternative is associated with incremental levels of investment, operating costs, recovery credits, and VOC emission control levels. The incremental costs alter the total cost structure of each affected plant and potentially affect its pricing, profitability, and economic viability. In this analysis, the expected economic effects of these regulatory alternatives are analyzed.

The methodology used in this analysis of expected economic impacts on the polymers and resins industry involves first a quantitative financial analysis utilizing a model plant approach to compare prices before and after implementation of the standards, and second, a qualitative analysis of expected industry and macroeconomic effects. The price impact methodology is based on a simplified price impact analysis. This methodology calculates the revenue and price increases required by model plants to maintain the same net present values (NPV) before and after the installation of emission-control equipment. This revenue calculation relies on a single derived equation that requires several types of input data for each model plant. The macroeconomic analysis consists of an evaluation of aggregate industry and macroeconomic impacts based on an understanding of the market structure and dynamics in the polymers and resins industry, the background of which is discussed in Section 9.1.

The purpose of this analysis is to determine the revenue increase that exactly offsets emission control costs so that the NPV of the model plant remains constant or the NPV of the incremental cash flow is zero at the stated weighted average cost of capital. In this analysis, capital costs, operating and maintenance costs, recovery credits, investment life, income taxes, and inflation need to be taken into account. A nominal discount

rate is used for the analysis because most equity capital cost data are available in nominal terms. The use of a nominal discount rate requires that revenues and operating costs be properly inflated. The revenue increase that the analysis calculates is expressed in base-year dollars, in this case June, 1980. The required revenue increase is converted to a required unit price increase by dividing the revenue by the annual sales volume. This step requires the assumption that annual sales volume is constant, indicating perfectly inelastic demand.

The derivation of the basic formula for the price impact analysis requires the following assumptions:

- Emission control investments have zero salvage value.
- No differential inflation occurs among the cost or revenue items.
- The weighted average cost of capital and the marginal income tax rate remain constant during the life of the investment.
- Depreciation is based on the 1981 Economic Recovery Act, Accelerated Cost Recovery System (ACRS) five-year rates.
- A 10 percent investment tax credit is applicable on emissions control investment and is realized in the year following the investment.
- NPV of the model plant remains constant at the stated weighted cost of capital.
- Where control equipment lifetime is less than that of the process unit, replacement investment in control equipment can occur automatically and the basic formula derived below will hold. Theoretically, a problem could arise if the last control equipment replacement outlives the process unit itself. However, such a situation is not predictable and is unlikely to enter into the initial decisions to invest in control equipment and to adjust product price to recoup that investment.

The derivation of the basic formula of the methodology is presented in the context of the standard definition of NPV:

$$ NPV = \sum_{y=1}^{n} \frac{CF_y}{(1+d)^y} - I_0 \qquad (1) $$

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where,

NPV = net present value (in base year dollars)

y   = time period

n   = investment life

CF_y = projected operating cash flows in period y

d   = nominal interest rate or cost of capital used for discounting

I_o = investment (period $y=0$)

Cash flow, CF, is defined as revenues, R, less operating and maintenance costs, OM, less income taxes, T.

$$CF_y = R_y - OM_y - T_y \qquad (2)$$

Revenue and operating costs will inflate at the same rate, thus, nominal revenue (and O&M) equals the product of the inflation factor, $(1+inf)^y$, and constant dollar value in $y=0$ dollars. That is,

$$R_y = (R_o) \ (1+inf)^y \qquad (3)$$

and,

$$OM_y = (OM_o) \ (1+inf)^y \qquad (4)$$

where, inf = annual inflation rate.

Income taxes, T_y, are computed on the basis of nominal current income. Income taxes are defined to equal the tax rate, t, times taxable net income. The investment tax credit, ITC_y, is subtracted directly from income taxes. Taxable income equals revenue less operating and maintenance expense, and less depreciation. Thus,

$$T_y = t(R_y - OM_y - D_y) - ITC_y \qquad (5)$$

Substituting the income tax Equation (5) into the cash flow Equation (2) yields the following expanded expression for cash flow.

$$CF_y = R_y - OM_y - T_y \tag{6a}$$

$$= R_y - OM_y - t(R_y - OM_y - D_y) + ITC_y \tag{6b}$$

$$= R_y - OM_y - tR_y + tOM_y + tD_y + ITC_y \tag{6c}$$

$$= (1-t)R_y - (1-t)OM_y + tD_y + ITC_y \tag{6d}$$

Further, by substitution of Equations (3) and (4) into (6d), operating cash flow, CF_y, may be expressed by:

$$CF_y = (1-t)R_0(1+inf)^y - (1-t)OM_0(1+inf)^y + tD_y + ITC_y \tag{7}$$

Net present value can now be defined in terms of revenue, operating and maintenance costs, and income tax effects, including depreciation and the investment tax credit.

By substitution of Equation (7) into Equation (1),

$$NPV = -I_0 + \sum_{y=1}^{n} \frac{(1-t)R_0(1+inf)^y - (1-t)OM_0(1+inf)^y + tD_y + ITC_y}{(1+d)^y} \tag{8}$$

Equivalently, each term under summation can be summed individually so that

$$NPV = -I_0 + \sum_{y=1}^{n} \frac{(1-t)R_0(1+inf)^y}{(1+d)^y} - \sum_{y=1}^{n} \frac{(1-t)OM_0(1+inf)^y}{(1+d)^y} + \sum_{y=1}^{n} \frac{ITC_y + tD_y}{(1+d)^y} \tag{9}$$

By imposing the constraint that incremental NPV=0, it is possible to solve for the annual revenue requirement, R_0, (in base year dollars) that is equivalent to incremental emissions control investment, I_0, and annual

operating and maintenance costs, OM_0. Rearranging terms on the right hand side of Equation (9) so that revenue, operating costs and investment related items are grouped, and setting NPV=0 yields

(10)

$$0 = \sum_{y=1}^{n} \frac{(1-t)R_0(1+inf)^y}{(1+d)^y} - \sum_{y=1}^{n} \frac{(1-t)OM_0(1+inf)^y}{(1+d)^y} - I_0 + \sum_{y=1}^{n} \frac{ITC_y + tD_y}{(1+d)^y}$$

Next, the terms related to emissions control investment are isolated. Assuming a 10 percent investment tax credit in the year after investment and five-year ACRS rates, the investment, investment tax credit, and depreciation terms on the right hand side of Equation (10) can be expressed as a product of a constant, TAXF, and I_0.

(11)

$$I_0 - \sum_{y=1}^{n} \frac{ITC_y + tD_y}{(1+d)^y} = I_0 \left[1 - \frac{(.1+.15t)}{(1+d)} - \frac{.22t}{(1+d)^2} - \frac{.21t}{(1+d)^3} - \frac{.21t}{(1+d)^4} \right.$$

$$\left. - \frac{.21t}{(1+d)^5} \right]$$

$$= I_0 \{TAXF\}$$

(12)

where TAXF is the sum of terms in brackets. TAXF is a constant that can be repeatedly applied to different investments so long as the tax rate, interest rate, and ACRS life remain the same.

Next, I_0 {TAXF} is substituted into Equation (10), the constant terms are moved outside their summations, and R_0 is isolated on the left hand side of the equation. R_0 is then expressed as

$$R_O = \dfrac{(1-t)OM_O \displaystyle\sum_{y=1}^{n} \dfrac{(1+inf)^y}{(1+d)^y} + I_O \ \{TAXF\}}{(1-t)\displaystyle\sum_{y=1}^{n}\dfrac{(1+inf)^y}{(1+d)^y}} \tag{13}$$

Simplifying Equation (13) results in

$$R_O = \dot{O}M_O + I_O \dfrac{TAXF}{(1-t)\displaystyle\sum_{y=1}^{n}\dfrac{(1+inf)^y}{(1+d)^y}} \tag{14}$$

Equation (14) is the equation used to calculate the annual revenue increase that exactly offsets NSPS capital and operating costs so that the NPV of the firm remains constant.

The above equation is applicable in those cases in which the total investment has a single economic life of n. When the investments cover capital equipment with different lives, the equation needs to be expanded to reflect capital costs associated with each life. For example, when a portion of the investment has a life of 15 years and another portion 10 years, R_O is then expressed as:

$$R_O = OM_O + I_{15}\dfrac{TAXF}{(1-t)\displaystyle\sum_{y=1}^{15}\dfrac{(1+inf)^y}{(1+d)^y}} + I_{10}\dfrac{TAXF}{(1-t)\displaystyle\sum_{y=1}^{10}\dfrac{(1+inf)^y}{(1+d)^y}} \tag{15}$$

where, I_{15} = investment with an economic life of 15 years

I_{10} = investment with an economic life of 10 years.

The investment lives of the various equipment items are indicated in Chapter 8.

The price increase that the model plant needs to realize this annual revenue increase is

$$PI = \frac{R_o}{Q_o}$$

(16)

where,   PI = the unit price increase in base year dollars

R_o = the required annual revenue increase in base year dollars

Q_o = annual sales volume in units

Finally, the percentage increase in unit price is calculated using the formula

$$PPI = \frac{PI}{P_o} \times 100$$

(17)

where,   PPI = percent unit price increase

PI = unit price increase in base year dollars

P_o = pre-emission control unit price in base year dollars

Equations (15), (16), and (17) were used to calculate the revenue and price effects of emission controls on the affected polymers and resins plants.

9.2.2 Economic Impact of VOC Potential NSPS Regulatory Alternatives - Polymers and Resins

Industry and macroeconomic impacts of the potential NSPS regulatory alternatives include effects on prices, profitability, plant viability, employment, and other economic measures. The price analysis here is quantitative. The other characteristics are analyzed qualitatively on the basis of the industry price analysis. A more involved analysis of

aggregate effects is not warranted, since the quantitative price impacts described in the next section are minor.

9.2.2.1 <u>Required Revenue and Maximum Price Increases</u>. On the plant level, the net present value (NPV) method is used to estimate required revenue increases and maximum percentage product price increases for each model plant and regulatory alternative. These measures are based on the assumption that incremental emission control costs are fully passed forward.

The model plants represent the new facilities that are expected to be constructed during the analysis' five-year period. The model plant product type, plant capacity, and processes were discussed in Chapters 6 and 8. The numbers of new plants by type, size, and process were projected in Section 9.1. The number, capacity, production, and annual sales for the projected new model polymers and resins plants are summarized in Table 9-13.

The model plants represent ten product-process combinations. Twenty-seven affected plants are projected for the first five years, i.e., 1984-1988. The sales level for each model is the product of the average 1980 polymer price and the model plant's production. The production levels of the model plants are based on a capacity utilization rate of 85 percent, the assumed optimal rate for new facilities.

The average prices for the model plant products were derived from one of two sources. Prices for PP, LDPE, HDPE, and PS were determined from data contained in the "Plastics Resins Domestic Merchant Sales" table in <u>Facts and Figures</u>.[35] That table reflects both quantities of sales and <u>net</u> 1980 dollar values which represents actual selling prices after deductions for discounts; consequently, the division of the dollar value by the quantity of the products sold provides a valid measure of typical plant average annual prices. PET fiber prices were not available from the above source.

It is assumed that the PET new facility will be a fibers plant. Determining an average annual price for the PET fibers model plant was more involved than determining that for the other plants, because the PET plant produces a mix of product fibers of varying prices. <u>Chemical and Engineering News</u> (C&EN)[36] reports 1980 prices for three polyester fibers:

Table 9-13. OPERATIONAL CHARACTERISTICS OF POLYMERS AND RESINS MODEL PLANTS
BUILT 1984 THROUGH 1988 (JUNE, 1980 DOLLARS)

Product	Process	Projected number of affected plants	Annual capacity per plant (Gg)	Utilization rate [a] (%)	Annual production per plant [b] (Gg)	Price [c] ($/Kg)	Annual sales per plant [d] ($1000)
PP	Liquid phase	3	150	85	127.5	.70	89,300
	Gas phase	3	105	85	89.3	.70	62,500
LDPE	Liquid phase	1	280	85	238.0	.82	195,200
	Gas phase	5	150	85	127.5	.82	104,600
HDPE	Liquid phase slurry	2	210	85	178.5	.79	141,000
	Liquid phase solution	3	90	85	76.5	.79	60,408
	Gas phase	5	150	85	127.5	.79	100,700
PS	Continuous	2	75	85	63.8	1.02	65,000
PET	DMT	1	105	85	89.3	1.80	160,700
	TPA	2	105	85	89.3	1.80	160,700
		27					

a   Capacity utilization rates are assumed to be 85 percent.
b   Annual production equals annual capacity times utilization rate.
c   See text for derivation of prices.
d   Annual sales equals annual production times price.

staple, textured filament, and feeder filament. To establish a price for the plant's product mix, the analysis assumed that its sales were similar to those of total shipments shown in the 1977 Census of Manufactures[5] for these three types of PET fibers (on a percentage basis). Consequently, the model plant product price used to determine annual sales is the average price of the three types of fibers reported in C&EN weighted by the product shipments of these fibers as reported in the Census of Manufactures.

Required revenue and maximum price increases for polymer and resin model plants to comply with potential NSPS regulatory alternatives are shown in Table 9-14. These measures are shown with the associated estimated investment, operating and maintenance costs, and recovery credits for each regulatory alternative. For each model plant, Regulatory Alternative 1 is baseline control. Other regulatory alternatives reflect different levels of control and associated cost as described in Chapter 8.

For these computations, the corporate income tax rate on marginal income was assumed to be 50 percent. Inflation was assumed to be 8 percent per annum. The nominal weighted cost of capital was estimated to be 14 percent. The after-tax cost of capital reflects a 12.5 percent nominal interest rate, a capital structure of 35 percent debt and 65 percent equity, and a 17.9 percent nominal return on equity. Because of the capital structure and tax effect, the size of the interest rates accounts for only a small portion of the total cost of capital, and changes in the rate result in relatively minor changes in the cost of capital. For example, changing the interest rate by 5 percentage points (10 to 15 percent) changes the cost of capital by only one percentage point. The interest rate is based on a 2 percent real risk-free component which was estimated from an analysis of the historical real yields on U.S. Treasury securities (10 year maturities) and an 8 percent inflation component. The product of these two components (1.02 X 1.08) represents the nominal risk-free interest rate--10 percent. This rate is increased by 25 percent (to approximate the interest on a Baa (Moody's) corporate bond) in order to incorporate an appropriate risk component. The capital structure and cost of equity are based on an analysis of data reported in Value Line for nine chemical firms involved in the manufacture of plastics.[37] The cost of equity was derived by the dividend yield method.[38]

Table 9-14.   POLYMERS AND RESINS MODEL PLANT CONTROL COSTS[a] AND MAXIMUM PRICE INCREASES BY PLANT PRODUCT, PROCESS, AND REGULATORY ALTERNATIVE (JUNE, 1980 DOLLARS)

Model plant categories by product and process	Regulatory alternative	Investment[b]	Recovery[c] credit	O&M[d]	Required[e] revenue increase	Maximum[f] price increase
		——————————————————($1000)——————————————————				———(%)———
PP						
● Liquid phase	2	66.3	29.6	29.0	16.0	.02
	3	380.0	29.6	79.0	113.6	.13
	4	510.9	29.6	46.5	101.4	.11
● Gas phase	2	56.8	29.6	29.0	16.0	.03
	3	88.3	29.6	44.5	34.6	.06
LDPE						
● Liquid phase	2	66.8	29.6	29.0	16.0	.01
	3	424.8	29.6	123.7	165.1	.08
	4	475.3	29.6	138.8	187.9	.10
● Gas phase	2	66.8	29.6	29.0	16.0	.02
	3	29.8	29.6	27.5	8.8	.01
	4	330.8	29.6	129.8	156.9	.15
HDPE						
● Liquid phase slurry	2	66.8	29.6	29.0	16.0	.01
	3	84.7	29.6	31.0	20.6	.01
● Liquid phase solution	2	66.8	29.6	29.0	16.0	.03
	3	1,484.8	29.6	848.7	1,051.8	1.74
● Gas phase	2	66.8	29.6	29.0	16.0	.02
	3	29.8	29.6	27.5	8.8	.01
	4	330.8	29.6	129.8	156.9	.16
PS						
● Continuous	2	66.8	29.6	29.0	16.0	.02
	3	69.6	185.5	30.9	(137.6)	-.21
	4	69.6	189.1	30.9	(141.2)	-.22
PET						
● DMT	2	67.0	29.6	29.0	16.0	.01
	3	2,347.7	914.6	283.1	(268.2)	-.17
	4	2,368.8	914.6	311.1	(237.5)	-.15
● TPA	2	2,280.9	884.9	254.2	(284.0)	-.18

[a]   Control costs and recovery credits are based on data contained in Tables 8-20 to 8-28.

[b]   Investment for each alternative represents the sum of the incremental capital costs (over baseline, Alternative 1) for the various types of equipment as aggregated by economic life (costs were rounded to the nearest $100).   For example, the investment for Alternative 3 of PP gas phase is $88,800 (shown above as 88.8).   This is obtained from data in Table 8-21 and represents the sum of a $7,000 incremental cost for flares (with an economic life of 15 years as indicated in Table 8-3), a $15,000 incremental cost for flare ducting with an economic life of 10 years, and $66,800 for LDAR with an economic life of 5½ years.   Because of rounding, the sum of the incremental costs (88,800) does not equal the total incremental cost of $88,000 obtained by subtracting $366,000 from $454,000 (See Table 8-21).

[c]   Recovery credits are taken directly from the tables in Chapter 8 for each alternative (since there are no credits under baseline).

[d]   O&M represents the sum of the incremental direct costs and indirect costs (excluding capital recovery). From Table 8-21, the incremental direct cost for PP gas phase Alternative 3 is $41,100 (84,100 less 43,000) and the incremental indirect cost is $3,500 (18,000 less 14,600). the total incremental O&M cost is $44,600 (shown as 44.6 above).

e   The calculation of the required revenue increase is based on Equation (15) (from the text). With the substitution of the parameters indicated in the text, Equation (15) simplifies to

$$R_0 = OM_0 - .115I_1 - .152I_2 - .248I_3,$$

where

OM_0 = O&M (from above) minus recovery credit
I_1 = Incremental capital costs of equipment with economic life of 15 years.
I_2 = Incremental capital costs of equipment with economic life of 10 years.
I_3 = Incremental capital costs of LADR equipment (with average economic life of $5\frac{1}{4}$ years).

For PP gas phase Alternative 3, the calculations are:

$$R_0 = (44.6 - 29.6) + .115(7.0) - .152(150) - .248(66.8)$$
$$= 34.6$$

Parentheses indicate negative required revenue increases.

f   The calculation of the maximum price increase is based on Equations (16) and (17) from the text. By substituting Equation (16) into Equation (17), the latter simplifies to:

$$PP1 = \frac{R_0}{P_0 Q_0} \times 100 \quad \text{or} \quad \frac{\text{Required revenue increase}}{\text{dollar sales volume}} \times 100$$

The dollar sales volume for each model plant is shown in Table 9-13

Required revenue decreases and price decreases indicate situations in which recovery credits exceed annual capital and operating and maintenance costs. These include Regulatory Alternatives 3 and 4 for both the PS and PET/DMT model plants and Alternative 2 for PET/TPA model.

Required revenue increases and maximum price increases for all other alternatives of the model plants are relatively insignificant except in the single case indicated below. The maximum price increase for Alternative 2 of all model plants is less than .05 percent. Increases for the remaining alternatives are less than .2 percent for all model plants except the HDPE liquid phase solution model. The price increase under Alternative 3 for this model is 1.7 percent; its after-tax annualized required revenue increase is $1.1 million.

9.2.2.2 <u>Expected Price and Profitability Impacts</u>. The elasticity of demand determines the extent to which cost increases can be passed forward to the consumer in terms of higher prices. An elastic demand implies that sales revenue will be reduced if costs are passed forward. In such a case, the producers can be expected to absorb some of the costs in order to minimize the impact on profits. On the other hand, an inelastic demand implies that costs can be passed forward. In general, the demand for plastic products tends to be inelastic in the major end-use markets discussed in Section 9.1. Although, as discussed in Section 9.1, substitutes exists for plastics in many of the markets (e.g., building, packaging, and transportation), there are no competitively priced substitutes available in the short run. In a number of the markets (primarily consumer and institutional) plastic products represent a small portion of the end user's budget. Both of these determinants suggest an inelastic demand. In addition to these determinants, other factors exist in the industry that would affect the ability of the industry to pass costs forward. For example, the extensive vertical integration that exists within the large petroleum and chemical firms facilitates substantial cost pass-through. All of these demand determinants and market factors indicate that costs imposed industry-wide can very likely be passed forward. However, in the case of the potential NSPS costs, the ability of the 27 projected new plants to pass the costs forward may be limited in some instances because production in these plants constitutes a minor

portion of the total industry's output. The potentially impacted PP, LDPE, and HDPE plants constitute about 30 percent of their respective projected 1988 polymer and resin capacities, while the PS and PET plants constitute less than eight percent of theirs. Consequently the PS and PET plants could be expected to absorb a portion of the cost of the regulatory alternatives (this will depend also, on the relative costs of production of the new and existing plants). Even though these plants do absorb the costs, the costs are for the most part insignificant as shown above and the impact on profitability would be insignificant.

    9.2.2.3 <u>Other Economic Effects</u>. Since no significant impacts on prices or profits are anticipated, the potential NSPS is not expected to have any significant effects on the industry or economy. Although capital availability may be of considerable concern to the industry in its efforts to modernize, the small incremental costs of NSPS controls are not expected to have any effect on the generation of the required capital for the regulatory alternatives. Little or no postponement of plant construction is expected to occur. The potential NSPS is not expected to have significant aggregate effects on output, employment, competition, industry structure, productivity, or foreign trade.

9.3 POTENTIAL SOCIOECONOMIC AND INFLATIONARY IMPACTS

    The socioeconomic and inflationary impacts of the potential NSPS are examined in terms of the fifth year costs and benefits to society of each regulatory alternative, the impacts on small facilities, the level of inflation, and the balance of trade.

9.3.1 <u>Fifth Year Costs and Benefits</u>

    The total annualized costs to society for each regulatory alternative in the fifth year of implementation, 1988, are presented in Table 9-15. Projected regulatory costs are customarily summed for the fifth year to facilitate comparison of cost impacts among various environmental standards. The need for and effects of regulations are reconsidered every four years. Costs were determined as the product of the annualized costs (or net annualized costs where there are product recovery credits) and the projected number of plants expected to be affected in the fifth year. These costs to society are based on a 10 percent real social interest rate

Table 9-15. FIFTH YEAR NET ANNUALIZED COST [a] TO SOCIETY OF REGULATORY ALTERNATIVES BY MODEL PLANT PRODUCT AND PROCESS (JUNE, 1980 DOLLARS)

Model plant categories by product and process	Regulatory alternative	Net annualized cost [b] ($1000)	Number of plants	Total annualized cost [b] ($1000)
PP				
● Liquid phase	2	15.7	3	47.1
	3	116.6		349.8
	4	105.9		317.7
● Gas phase	2	15.7	3	47.1
	3	34.5		103.5
LDPE				
● Liquid phase	2	15.7	1	15.7
	3	168.3		168.3
	4	192.4		192.4
● Gas phase	2	15.7	5	78.5
	3	8.2		41.0
	4	159.5		797.5
HDPE				
● Liquid phase slurry	2	15.7	2	31.4
	3	20.5		41.0
● Liquid phase solution	2	15.7	3	47.1
	3	1,067.1		3,201.3
● Gas phase	2	15.7	5	78.5
	3	8.2		41.0
	4	159.5		797.5
PS				
● Continuous	2	15.7	2	31.4
	3	(137.9)		(275.8)
	4	(141.5)		(283.0)
PET				
● DMT	2	7.0	1	7.0
	3	(298.0)		(298.0)
	4	(279.0)		(279.0)
● TPA	2	(306.0)	2	(612.0)

[a] These costs represent the costs of the regulatory alternatives over and above the baseline as described in Chapter 8.

[b] Parenthesis indicate negative required revenue increases.

as opposed to the 14 percent nominal weighted cost of capital used in determining the price changes (Section 9.2). The 10 percent rate represents costs to society and is generally used in benefit cost studies. The highest alternative cost for each of the model plants is shown in Table 9-16. The total of these costs, which provides a view of the upper boundary, is $4.9 million.

Executive Order 12291 specifies that a regulatory action, to the extent permitted by law, must not be undertaken unless the potential benefits to society from the regulation outweigh the potential costs to society. An exhaustive benefit-cost analysis is not appropriate here because the potential NSPS will not constitute a major rule within the meaning of the Executive Order since the cost of the standards and their overall impacts on the economy are minor. A qualitative enumeration of the benefits follows.

The potential standards will reduce the rate of VOC emissions to the atmosphere. These compounds are precursors of photochemical oxidants, particularly ozone. The EPA publication, AIR QUALITY CRITERIA FOR OZONE AND OTHER PHOTOCHEMICAL OXIDANTS (EPA-600/8-78-004, April 1978), explains the effects of exposure to elevated ambient concentrations of oxidants. (The problem of ozone depletion of the upper atmosphere and its relation to this standard are not addressed here.) These effects include:

- Human health effects. Ozone exposure has been shown to cause increased rates of respiratory symptoms such as coughing, wheezing, sneezing, and shortness of breath; increased rates of headache and of eye and throat irritation; and physiological damage to red blood cells. One experiment links ozone exposure to those human cell damages known as chromosomal aberrations.
- Vegetation effects. Some rates can result in reduced crop yields from damages to leaves and/or plants have been shown for several crops including citrus, grapes, and cotton. The reduction in crop yields was shown to be linked to both the level and duration of ozone exposure.

Table 9-16. UPPER BOUNDARY OF TOTAL ANNUALIZED FIFTH YEAR COST
TO SOCIETY (JUNE, 1980 DOLLARS) [a]

Model plant categories by product and process	Number of plants	Regulatory alternative	Total annualized cost ($)
PP			
● Liquid Phase	3	3	349,800
● Gas Phase	3	3	103,500
LDPE			
● Liquid Phase	1	4	192,400
● Gas Phase	5	4	797,500
HDPE			
● Liquid Phase Slurry	2	3	41,000
● Liquid Phase Solution	3	3	3,201,300
● Gas Phase	5	4	797,500
PS			
● Continuous	2	2	31,400
PET			
● DMT	1	2	7,000
● TPA	2	2	(612,000)
TOTAL			4,909,400

[a] Annualized fifth year costs are based on the highest alternative cost for each model plant as shown in Table 9-15.

- Materials effects. Ozone exposure has been shown to accelerate the deterioration of such organic materials as plastics and rubber (elastomers), textile dyes, fibers, and certain paints and coatings.

- Ecosystem effects. Continued ozone exposure has been shown to be linked to structural changes in forests--the disappearance of certain tree species (Ponderosa and Jeffrey pines) and the death of predominant vegetation. Continued ozone exposure, hence, causes a stress on the ecosystem.

In addition to the evidence of the physical and biological effects enumerated above, a reduction of VOC emissions is likely to improve the aesthetic and economic value of the environment through: (1) beautification of natural forests and undeveloped land through increased vegetation; (2) increased visibility; (3) reduced incidence of noxious odors; (4) increased length of life for works of art, including paintings, sculpture, architecturally important buildings, and historic monuments; (5) improved appearance of structures, sculptures, and paintings, and (6) the improved productivity of workers, especially farm laborers.

9.3.2 Impacts on Small Facilities

The Regulatory Flexibility Act (Public Law 96-354, September 19, 1980) directs Federal agencies to pay close attention to minimizing any potentially adverse impacts of a standard on small businesses, small governments, and small organizations. This standard will have no known effects on small governments and small organizations. It may affect some small businesses but the impacts will be few and minor. Essentially, all firms that will be required to comply with the standard either are not small businesses, or are subsidiaries of large firms. The businesses that are expected to own or operate polymers and resins producing plants during the first five years following proposal are those currently in the field. (See Table 9-2). The Small Business Administration (SBA) classifies small businesses in SIC 2821 (PP, LDPE, HDPE, and PS) as those with 750 or fewer employees and in SIC 2824 (PET) and SIC 3079 as those with 1,000 and 250 or fewer, respectively. These levels were set as criteria for extending SBA loans and related assistance (13 CFR Part 121, Schedule A). Only two of

the firms in Table 9-2 are believed to be small businesses, and both of these produce PET. Because of the competitive nature of the industries and the relatively high levels of capital required to construct polymers and resins manufacturing plants, it is unlikely that any small businesses would undertake construction of new plants. Furthermore, as the analysis in Section 9.2 explains, any potential adverse economic impacts on new plants, regardless of whether they are large or small businesses, would be minor.

9.3.3 Other Impacts

No other impacts are expected as a result of the costs of the regulatory alternatives. There should be no significant pressure on the level of inflation. As the construction of the new source plants will not be adversely affected, neither will the employment level in the industry. No effects are expected on the balance of trade.

9.4 REFERENCES FOR CHAPTER 9

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APPENDIX A

EVOLUTION OF THE PROPOSED STANDARDS

APPENDIX A

EVOLUTION OF THE PROPOSED STANDARDS

The purpose of this study was to develop new source performance standards for the polymers and resins industry. Work on the study was begun in January 1980 by Energy and Environmental Analysis, Inc., (EEA) under the direction of the Office of Air Quality Planning and Standards (OAQPS), Emission Standards and Engineering Division (ESED). In June 1982, this study was transferred from EEA to Pacific Environmental Services, Inc. (PES). The decision to develop this standard was made on the recommendation of EEA based on a source category survey study. In performing the standard development, previous EPA study reports, responses to requests for information under Section 114 of the Clean Air Act, plant visit information and industry comments were used.

The following chronology lists the important events which have occurred in the development background information for the new source performance standards for the polymers and resins industry.

Date	Activity
May 19, 1980	Meeting of CPB, EMB, SDB, EAB, and EEA representatives to discuss recommendations for the development of the NSPS for polymers and resins industry.
July 15, 1980	Meeting of EPA and EEA representatives to discuss scope of the polymers and resins NSPS.
August 15, 1980	Plant visit to USS Novamont's polyproylene facility at LaPorte, Texas.
August 19, 1980	Plant visit to Soltex Polymer Corporation's high-density polyethylene facility at Deer Park, Texas.

Date	Activity
August 28, 1980	Plant visit to Phillips Chemical Company's high-density polyethylene facility at Pasadena, Texas.
September 2, 1980	Meeting between EEA, EPA, and Union Carbide in South Charleston, West Virginia.
September 2, 1980	Union Carbide Corporation's response to Section 114 request for information on Union Carbide's low-density polyethylene facility at Port Lavaca, Texas.
September 9, 1980	Mobil Chemical Company's response to Section 114 request for information on Mobil's styrenics plant at Santa Ana, California.
September 11, 1980	Plant visit to Union Carbide Corporation's low-density polyethylene facility at Port Lavaca, Texas.
September 15, 1980	Union Carbide's response to information requested at September 2, 1980 meeting on low pressure and high pressure polyethylene.
September 16, 1980	Plant visit to Mobil Chemical Company's styrenics facility at Santa Ana, California.
September 22, 1980	USS Novamont's response to Section 114 request for information on USS Novamont's polypropylene facility at LaPorte, Texas.
September 29, 1980	Gulf Oil Chemicals Company's reply to EPA's September 2, 1980, letter requesting information on Gulf's gas phase, high-density polyethylene process at Orange, Texas.
October 2, 1980	Plant visit to Tennessee Eastman Company's polyester resin facility at Kingsport, Tennessee.
October 9, 1980	American Hoechst Corporation's response to Section 114 request for information on American Hoechst's polyester resin facility at Greer, South Carolina.
November 7, 1980	Tennessee Eastman Company's response to Section 114 request for information on Tennessee Eastman's poly(ethylene terephthalate) facility at Kingsport, Tennessee.
November 21, 1980	Northern Petrochemical Company's response to Section 114 letter request for information on Northern Petrochemical's low-density polyethylene plant at Morris, Illinois.
December 16, 1980	Meeting of CMA, EPA, and EEA representatives on status of the polymers and resins NSPS development.

Date	Activity
December 18, 1980	Phillips Chemical Company's response to EPA request for additional information on Phillips Chemical's high-density polyethylene facility at Pasadena, Texas.
March 17, 1981	Standard Oil Company (Indiana) response to Section 114 request for information on Amoco Chemicals Corporation's gas phase polypropylene process.
July 6, 1981	Meeting of CMA, EPA, and EEA representatives on status of the polymers and resins NSPS development.
August 24, 1981	Meeting of EPA and EEA representatives on status of and future plans for cost analysis for polymers and resins NSPS.
September 1981	Model Plant parameters package was sent to CMA for comments.
November 17, 1981	Meeting of CPB, EAB, SDB, and EEA representatives to review the proposed model plant parameters, process baseline control and regulatory alternatives and fugitive regulatory alternatives for the polymers and resins NSPS.
February 26, 1982	BID Chapters 3 to 6 were sent to the industry for comments.
March 4, 1982	Meeting of Allied Chemical Company, EPA, and EEA representatives to discuss vacuum system design alternatives to reduce air emissions in polyester - TPA process plant.
March 26, 1982	Meeting of CPB, EAB, SDB, and EEA representatives to discuss the regulatory approach and recommendations for the standard.
April 2, 1982	Phillips Chemical Company's comments on draft BID Chapters 3-6.
April 13, 1982	Texas Chemical Council's comments on draft BID Chapters 3-6.
April 13, 1982	Union Carbide Corporation's comments on draft BID Chapters 3-6.
April 14, 1982	Gulf Oil Chemicals Company's comments on draft BID Chapters 3-6.
April 14, 1982	Northern Petrochemical Company's comments on draft BID Chapters 3-6.

Date	Activity
April 14, 1982	Tennessee Eastman's comments on draft BID Chapters 3-6.
April 15, 1982	Allied Fibers and Plastics' comments on the draft BID Chapters.
April 15, 1982	Chemical Manufacturing Association's comments on draft BID Chapters 3-6.
April 15, 1982	USS Corporation's comments on draft BID Chapters 3-6.
April 19, 1982	DuPont's comments on draft BID Chapters 3-6.
April 19, 1982	Monsanto's comments on draft BID Chapters 3-6.
June 1982	Project transferred from EEA to PES.
June 30, 1982	Plant visit to Soltex Polymer Corporation's high-density polyethylene facility at Deer Park, Texas.
July 1, 1982	Plant visit to DuPont Corporation's high-density polyethylene facility at Orange, Texas.
September 14, 1982	Plant visit to Gulf Oil Chemicals Company's polystyrene facility at Marietta, Ohio.
September 15, 1982	Plant visit to Monsanto Plastics and Resins Company's polystyrene facility at Port Plastics (Addyston), Ohio.
September 29, 1982	Plant visit to Fiber Industries' polyester facilities at Salisbury, North Carolina.

APPENDIX B

INDEX TO ENVIRONMENTAL CONSIDERATIONS

This appendix consists of a reference system which is cross-indexed
with the October 21, 1974, Federal Register (39 FR 37419) containing
the Agency guidelines for the preparation of Environmental Impact
Statements. This index can be used to identify sections of the document
which contain data and information germane to any portion of the
Federal Register guidelines.

APPENDIX B

INDEX TO ENVIRONMENTAL IMPACT CONSIDERATIONS

Agency Guidelines for Preparing Regulatory Action Environmental Impact Statements (39 FR 37419)	Location Within the Background Information Document
Background and summary of regulatory alternatives	The regulatory alternatives are summarized in Chapter 1, Section 1.1, pages 1-1 through 1-2.
Statutory basis for proposing standards	The statutory basis for the proposed standards is summarized in Chapter 2, Section 2.1, pages 2-1 through 2-4.
Relationship to other regulatory agency actions	The various relationships among the regulatory agency actions are discussed in Chapters 2, 3, 5, 7, and 9.
Affected industry	A discussion of the industry affected by the regulatory alternatives is presented in Chapter 3, Section 3.1. Details of the "business/economic" nature of the industry are presented in Chapter 9, pages 9-1 through 9-36.
Affected processes and facilities	The specific sources and processes to be affected by the regulatory alternative are discussed in Chapter 3, Sections 3.2 and 3.3, pages 3-11 through 3-57.
Availability of control technology	Information on the types and availability of control technology is given in Chapter 4, Sections 4.1 and 4.2, pages 4-1 through 4-36.

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Agency Guidelines for Preparing Regulatory Action Environmental Impact Statements (39 FR 37419)	Location Within the Background Information Document
Environmental, energy, and economic impacts of regulatory alternatives	Various regulatory alternatives are discussed in Chapter 6, Section 6.2, pages 6-13 through 6-33.
Environmental impacts	The environmental impacts of the various regulatory alternatives are presented in Chapter 7, Sections 7.1, through 7.3, pages 7-2 through 7-23.
Energy impacts	The energy impacts of the various regulatory alternatives are discussed in Chapter 7, Section 7-4, pages 7-23 through 7-33.
Cost analysis	Cost analysis of the various regulatory alternatives on model plants are discussed in Chapter 8, Section 8.1, pages 8-1 through 8-54.
Economic impacts	The economic impacts of the various regulatory alternatives are presented in Chapter 9, Sections 9.2 and 9.3, pages 9-36 through 9-55.

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APPENDIX C
EMISSION SOURCE TEST DATA
AND
FUGITIVE EMISSION SOURCE COUNTS

APPENDIX C: EMISSION SOURCE TEST DATA
AND FUGITIVE EMISSION SOURCE COUNTS

The purpose of this appendix is to describe the test results of
flare and thermal incinerator volatile organic compounds (VOC) emissions
reduction capabilities and the equipment inventory used in the development
of fugitive emission estimates for the background information document
(BID) for this industry. Background data and detailed information
which support the emission levels, reduction capabilities and the
fugitive emission model are included.

Section C.1 of this appendix presents the VOC emissions test data
including individual test descriptions for control of process sources
by flaring. Sections C.2 and C.3 present the VOC emissions test data
for control of process sources by thermal incineration and vapor
recovery system, respectively. Section C.4 consists of comparisons of
various VOC test results and a discussion exploring and evaluating the
similarities and differences of these results. Section C.5 contains
the available fugitive emission inventories for the polymers and
resins industry and discusses the selection of a single model plant to
best represent the characteristics of the fugitive emissions from the
industry.

C.1   FLARE VOC EMISSION TEST DATA

The design and operating conditions and results of the five
experimental studies of flare combustion efficiency that have been
conducted were summarized in Section 4.1.1.1.1. This section presents
more detailed results of the first flare efficiency emissions test to
encompass a variety of "non ideal" conditions that can be encountered
in an industrial setting. These results represent only the first
phase of an extended study of which a final report should be available
in 1983.

The aforementioned experimental study was performed during a three week period in June 1982 to determine the combustion efficiency for both air- and steam-assisted flares under different operating conditions. The study was sponsored by the U.S. Environmental Protection Agency and the Chemical Manufacturers Association (CMA). The test facility and flares were provided by the John Zink Company. A total of 23 tests were conducted on the steam-assisted flares and 11 tests on air-assisted flares. The values of the following parameters were varied: flow rate of flare gas, heating valve of flare gas, flow rate of steam, and flow rate of air. This section describes the control device and the sampling and analytical technique used and test results for the steam-assisted flare.

C.1.1 Control Device.

A John Zink standard STF-S-8 flare tip was used for the steam-assisted flare test series. This flare tip has an inside diameter of 0.22 m (8 5/8 in.) and is 3.7 m (12 ft. 3.5 in.) long with the upper 2.2 m (7 ft 3 in) constructed of stainless steel and the long 1.5 m (5 ft 0.5 in) constructed of carbon steel. Crude propylene was used as the flare gas. The maximum capacity of the flare tip was approximately 24,200 kg/hr (53,300 lb/hr) for crude propylene at 0.8 Mach exit velocity. Variations in heating valves of flare gas were obtained by diluting the propylene with inert nitrogen.

C.1.2 Sampling and Analytical Techniques

An extractive sampling system was used to collect the flare emission samples and transport these samples to two mobile analytical laboratories. Figure C-1 is a diagram of the sampling and analysis system. A specially designed 8.2 m (27 ft) long sampling probe was suspended over the flare flame by support cables from a hydraulic crane.

Gaseous flare emission samples entered the sampling system via the probe tip, passed through the particulate filter, and then were carried to ground level. The sampling system temperature was maintained above 100°C (212°F) to prevent condensation of water vapor. The flare emission sample was divided into three possible paths. A fraction of the sample was passed through an EPA Reference Method 4 sampling train to determine moisture content of the sample. A second fraction was
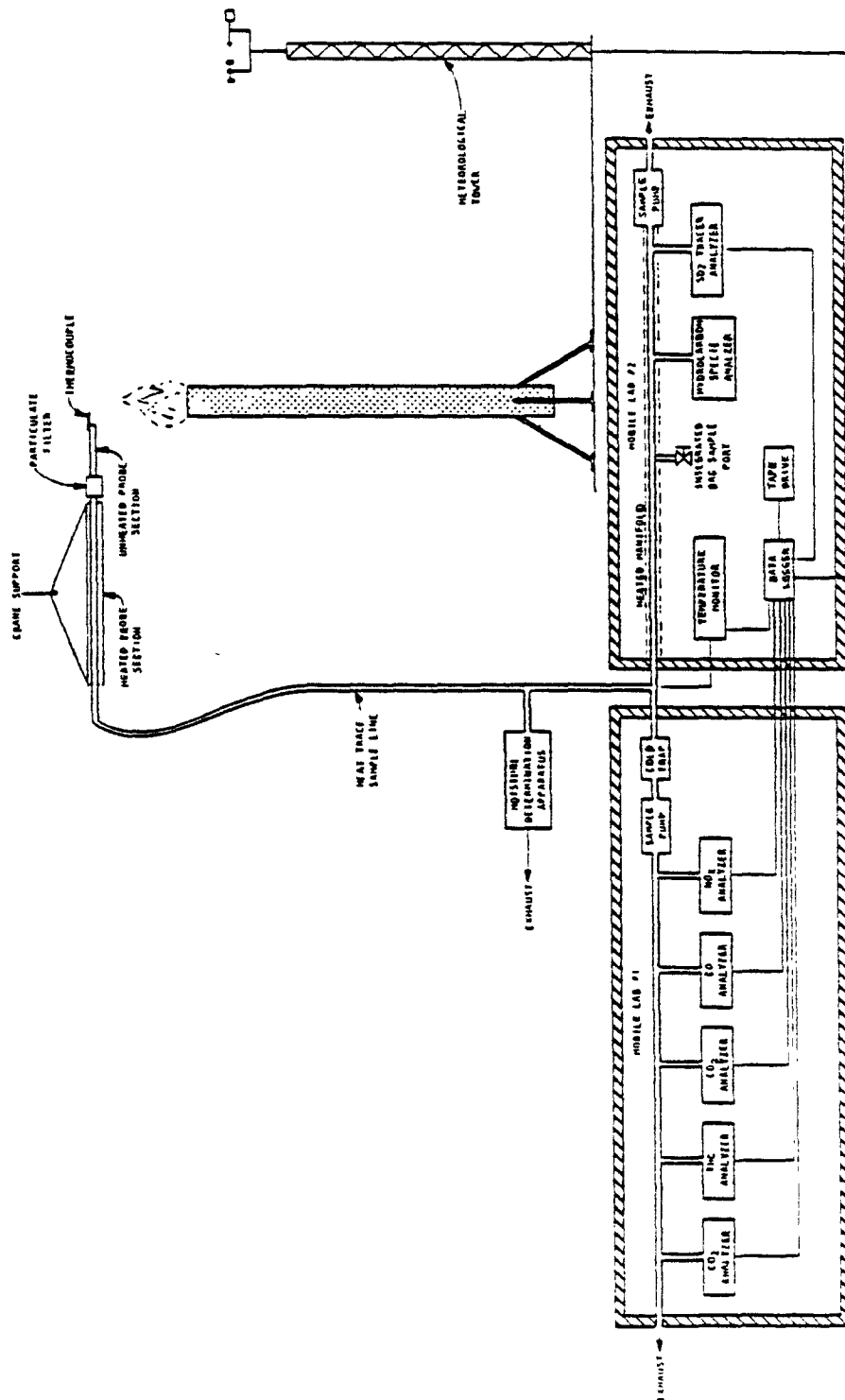
C-3

Figure C-1. Flare sampling and analysis system

directed through a moisture removal cold trap and thence, into a sampling manifold in one of the mobile laboratories. Sample gas in this manifold was analyzed by continuous monitors for O_2, CO, CO_2, NO_x and THC on a dry sample basis. A third sample was directed into a sampling manifold in the other mobile laboratory. Sample gas in this manifold was analyzed for SO_2 and hydrocarbon species on a wet basis.

Data collection continued for each test for a target period of 20 minutes. Ambient air concentrations of the compounds of interest were measured in the test area before and after each test or series of tests.

Flare emission measurements of carbon monoxide (CO), carbon dioxide (CO_2), oxygen (CO_2), oxides of nitrogen (NO_x), total hydrocarbons (THC) and sulfur dioxide (SO_2) were measured by continuous analyzers that responded to real time changes in concentrations. Table C-1 presents a summary of the instrumentation used during the tests.

C.1.3 <u>Test Results</u>

Twenty three tests were completed on the steam-assisted flare. Table C-2 summarizes the results of these tests. The results indicate that the combustion efficiencies of the flare plume are greater than 98 percent under varying condition of flare gas flow rate, including velocities as high as 18.2 m/s (60 fps) flare gas, heat content over 11.2 MJ/m³ (300 Btu/scf), and steam flow rate below 3.5 units of per unit of flare gas. The concentrations of NO_x emissions which were also measured during the testing ranged from 0.5 to 8.16 ppm.

C.2 THERMAL INCINERATOR VOC EMISSION TEST DATA

The results of six emission tests and one laboratory study were reviewed to evaluate the performance of thermal incinerators under various operating conditions in reducing VOC emissions from the different process waste streams generated during the manufacture of polymers and several synthetic organic chemicals. The variable parameters under which the incinerator tests were performed include combustion temperature and residence time, type of VOC, type and quantity of supplemental fuel, and feedstocks (solid, liquid, and gaseous waste streams). The test results, which are summarized in Table C-3, in combination with a theoretical analysis indicate that high VOC reduction efficiencies (by weight) can be achieved by all new incinerators.

Table C-1. EMISSION ANALYZERS AND INSTRUMENTATION UTILIZED FOR JOINT EPA/CMA FLARE TESTING

Make and Model	Parameter	Primary Operating Range	Operating Principle
Thermo Electron Model 10	NO_x	0-25 ppm	Chemiluminescence
Horiba PIR 2000	CO	0-1,000 ppm	Infrared absorption
Horiba PIR 2000	CO_2	0-5%	Infrared absorption
Teledyne 320 AX	O_2	0-25%	Electro catalysis
Scott 116	Total hydrocarbon	0-100 ppm	Flame ionization
Carle 211 Gas Chromatograph	Hydrocarbon Species	N/A	Flame ionization
Meloy SA 285 (ES Modified)	Tracer (SO_2)	0-5 ppm	Flame photometry
Climatronics Electronic Weather Station	Wind Speed Wind Direction Ambient Temperature	0-50 mph 0-540° 40-120°F	Photo chopper Precision potentiometer Thermistor
Omega Thermocouple Assembly	Probe Temperature	-300° to 2300°F	Chromel-Alumel exposed bead thermocouple

Table C-2. STEAM-ASSISTED FLARE TESTING SUMMARY

Test Number	FLARE GAS[a] Flow (scfm)	Velocity (ft/min)	Lower Heating Value (Btu/scf)	Propylene Flow (lbs/hr)	Propylene Flow (scfm)	Nitrogen Flow (lbs/hr)	Nitrogen Flow (scfm)	Steam Flow (lbs/hr)	Steam-to-Relief Gas Ratio (lb/lb)	Combustion Efficiency Percent	Comments
1	473	2,523	2,183	3,138	473	-	-	2,159	0.688	99.96	
2	464	2,475	2,183	3,078	464	-	-	1,564	0.508	99.82	
3	456	2,432	2,183	3,077	465	-	-	1,355	0.448	99.82	Incipient smoking flare
4	283	1,509	2,183	1,875	283	-	-	-	-	98.80	Smoking flare
3	157	837	2,183	1,044	157	-	-	-	-	98.81	Smoking flare
7	154	821	2,183	1,019	154	-	-	722	0.757	99.84	Incipient smoking flare
5	149	795	2,183	991	149	-	-	1,543	1.56	99.94	
67	148	789	2,183	980	148	-	-	711	0.725	-	Sampling probe in flame
17	24.5	131	2,183	162	24.5	-	-	150	0.926	99.84	
50	24.4	130	2,183	162	24.4	-	-	498	3.07	99.45	
56	24.5	131	2,183	163	24.5	-	-	562	3.45	99.70	
61	25.0	133	2,183	166	25.0	-	-	941	5.67	82.18	Steam-quenched flame
55	24.7	132	2,183	164	24.7	-	-	1,125	6.86	68.95	Steam-quenched flame
57	703	3,749	294	629	94.8	2,663	608	497	0.150	99.90	
11(a)	660	3,520	305	612	92.2	2,489	568	-	-	99.93	
11(b)	599	3,195	342	623	93.9	2,210	505	-	-	99.86	
11(c)	556	2,965	364	616	92.8	2,028	463	-	-	99.82	
59(a)	591	3,152	192	345	52	2,361	539	-	-	98.11	
59(b)	496	2,645	237	350	52.7	1,942	443	-	-	99.32	
60	334	1,781	298	212	32	1,325	302	-	-	98.92	
51	325	1,733	309	305	46	1,222	279	256	0.168	98.66	
16(a)	320	1,707	339	329	49.6	1,182	270	-	-	99.74	No smoke
16(b)	252	1,344	408	313	46.2	897	205	-	-	99.75	No smoke
16(c)	194	1,035	519	307	46.2	650	148	-	-	99.74	Incipient smoking flare
16(d)	159	848	634	307	46.3	496	113	-	-	99.78	Smoking flare
54	0.356	1.90	209	0.226	0.0341	1.41	0.322	-	-	99.90	
23	0.494	2.63	267	0.451	0.0680	2.13	0.487	-	-	100.01	
52	0.556	2.96	268	0.452	0.0682	2.14	0.488	201	77.5	98.82	
53	0.356	1.90	209	0.226	0.0341	1.41	0.322	201	123	99.40	

[a] All values at standard conditions of 68°F and 29.92 in Hg.

[b] The flare gas flow rates ranged from 1/3 scfm (approximately 60 percent flare capacity) to 0.35 scfm (purge flow rate).

Table C-3. SUMMARY OF THERMAL INCINERATOR EMISSION TEST RESULTS

Company and Location	Type of Product	Waste Incinerated	Temp. (°F)	Residence Time (seconds)	Efficiency, by weight (%)	Entity	Sampling Runs
Arco Polymers, Deer Park, TX	Polypropylene	Waste Gas[a]	1600	1.5	99.83[b]	EPA	2
		Waste Gas[a]	1800	1.5	99.77[c]		2
		Atactic Waste[a]	1600	1.5	99.996[b]		3
		Atactic Waste[a]	1800	1.5	99.985[c]		3
		Atactic Waste[a]	2000	1.5	99.986[c]		1
		Waste Gas + Atactic Waste[a]	1600	1.5	99.995[c]		2
		Waste Gas + Atactic Waste[a]	1800	1.5	99.997[c]		2
		Waste Gas + Atactic Waste[a]	2000	1.5	99.996[c]		2
Denka, Houston, TX	Maleic Anhydride	Waste Gas[d]	1400	0.6	98.5	EPA	5
Rohm & Haas, Deer Park, TX	Acrylic Acid & Esters	Waste Gas[d]	1425	1.0	82.6	EPA	3
		Waste Gas[d]	1510	1.0	98.3		4
		Waste Gas	1545	1.0	99.7		1
Union Carbide Corp., Taft, IA	Acrylic Acid & Acrylate Esters	Waste Gas[a]	1160	2-3	96.1	EPA	6
		Waste Gas[a]	1475	2-3	99.9		3
Petro-tex Chemical Corp., Houston, TX	Butadiene	Waste Gas[a]	1400	0.6	99.6[e]	Petro-tex	2 sets
Monsanto Chemical Intermediates, Co., Alvin, TX	Acrylonitrile	Waste Gas[a]	Confidential	Confidential	>99	Monsanto	2 units

[a]Supplemented with natural gas to improve combustion.

[b]Average of mean test results measured by each of four methods: proposed EPA Method 18 (on-site) for total hydrocarbons, proposed EPA Method 18 (off-site) for individual hydrocarbons, Byron Instruments Models 90 and 401 in combination for total hydrocarbons, and Byron Instruments Models 90 and 401 for total nonmethane hydrocarbons.

[c]Average of mean test results measured by each of three methods: proposed EPA Method 18 (on-site) for total hydrocarbons, Byron Instruments Models 90 and 401 in combination for total hydrocarbons, and Byron Instruments Models 90 and 401 for total nonmethane hydrocarbons.

[d]Supplemented with gas to improve combustion.

[e]Average efficiency for tests after completion of major modifications to improve mixing.

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Three sets of test data are available. *These are emission tests* conducted on (1) incinerators at polymers and resins plants by EPA, (2) incinerators for waste streams from air oxidation processes conducted by EPA or the chemical companies, and (3) laboratory unit data from tests conducted by Union Carbide Company on incinerated streams containing various pure organic compounds. (No adequately documented data were found for tests of incinerators at polymers and resins plants that were conducted by the companies.)

The EPA test studies represent the most in-depth work available. These data show the combustion efficiencies for full-scale incinerators on process vents at four chemical plants. The tests measured inlet and outlet VOC, by compound, at different incineration temperatures. The reports include complete test results, process rates, and descriptions of the test method. The four plants tested by the EPA are:

1. ARCO Polymers, Deer Park, Texas, polypropylene unit,
2. Denka Chemicals, Houston, Texas, maleic anhydride unit,
3. Rohm and Haas, Deer Park, Texas, acrylic acid unit, and
4. Union Carbide, Taft, Louisiana, acrylic acid unit.

The data from ARCO Polymers include test results based on three different incinerator temperatures and three different waste stream combinations. The data from Rohm and Haas also include results for three temperatures. The data from Union Carbide include test results based on two different incinerator temperatures. In all tests, bags were used for collecting integrated samples and a gas chromatagraph with flame ionization detector (GC/FID) was used for obtaining an organic analysis.

C.2.1 Environmental Protection Agency (EPA) Polymers Test Data[2]

EPA conducted emission tests at the incinerator at the ARCO Polymers, Inc., LaPorte polypropylene plant in Deer Park, Texas (listed as ARCO Chemical, Co., in LaPorte, Texas, in the 1982 Directory of Chemical Producers[3]) to assess emission levels and VOC destruction efficiency.

The ARCO polypropylene facility has a nameplate capacity of 131,000 Mg/yr (400 million lbs/yr).[3] The facility produces polypropylene resin by a liquid phase polymerization process. The facility includes two "plants" (Monument I and Monument II) comprised of a total of six

process trains producing a variety of polypropylene resins. Both plants discharge their gaseous, liquid, and solid process wastes to the same incinerator system where they undergo thermal destruction. The wastes in the plants occur from:

a) processing chemicals and dilution solvents for the catalyst,

b) spent catalyst,

c) waste polymeric material (by-product atactic polymer), and

d) nitrogen-swept propylene from the final stages (product resin purge columns) of the process.

The feed rates of these wastes to the incinerator vary according to which trains are running and what startups are occurring in the two plants. Feed rate variations were observed during the two weeks of the incinerator test.

The waste heat boiler associated with the incinerator provides a major portion of the process steam needed by the two polymer plants. Natural gas is used as an auxiliary fuel to fire the incinerator. If necessary, fuel oil can also be used. Under full production conditions, the atactic waste provides approximately 50 percent of the energy needed to produce the steam, and natural gas use is reduced.

C.2.1.1 Control Device. The incinerator and associated equipment were designed by John Zink, Company. The system was put into operation on August 16, 1978. The incinerator's two main purposes are to destroy organic waste from the polymer processes (primary) and to provide heat to generate steam (secondary). Figure C-2 depicts a flow diagram of the incinerator and associated equipment. Each inlet stream has its own nozzle inside the incinerator. Combustion air is fed into the incinerator at the burner nozzles located approximately 4 feet beyond the incinerator entrance. The combustion air flow rate is regulated manually. The quench air enters the incinerator within 3 feet of the burner nozzles. It is used to maintain a constant temperature and provide excess combustion air. The quench air flow rate is automatically regulated by an incinerator temperature controller.

During normal operation with all waste streams entering the incinerator, the natural gas is cut back and the atactic waste becomes the major fuel source. The purge gas, which has a low fuel value because
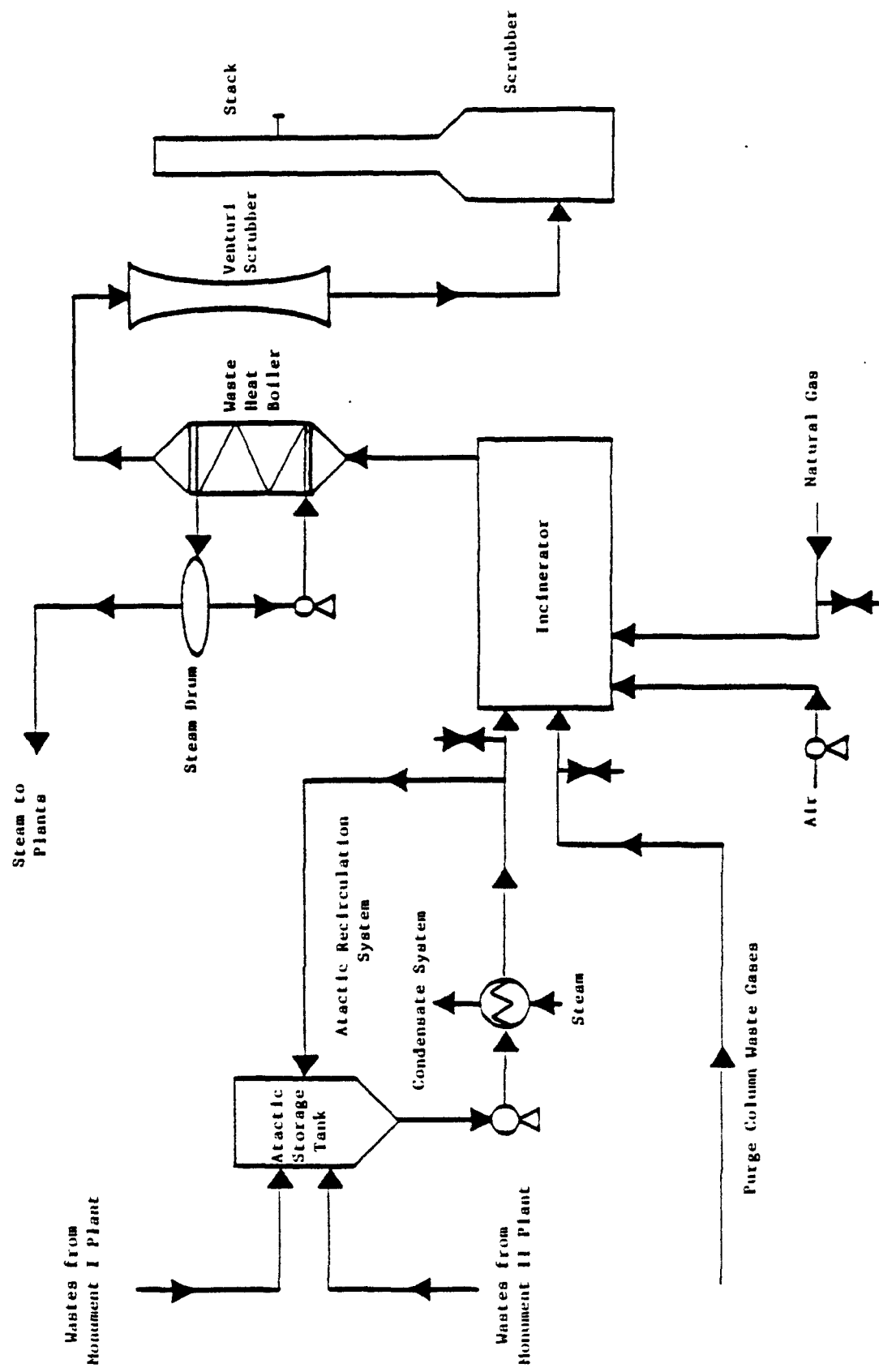
Figure C-2.   Schematic of incineration system at ARCO polypropylene facility

it is 95 percent nitrogen, is fed continuously to the incinerator for destruction of the VOC since there is no gas storage capacity in the system. During an upset of the incinerator this stream is sent to a flare. ARCO provided data to illustrate normal operating parameters of the incinerator. These are listed in Table C-4 and represent the averages for the month of August 1981. The following are considered design parameters:

a) heat input $\approx$2.18 MJ/s (7.45 x 10^6 Btu/hr),

b) air supply $\approx$15.1 standard m³/s at 0°C (33,900 scfm, at 60°F)

c) firebox temperature $\approx$980°C average and 1,200°C maximum (1,800°F average and 2,200°F maximum),

d) firebox residence time $\approx$1.5 seconds, and

e) pressure $\approx$19 kPa (78 in. H_2O).

C.2.1.2 <u>Sampling and Analytical Techniques</u>. A secondary purpose of the ARCO incinerator test was to compare results of different analytical methods for to the measurement of VOC emissions. During the testing phase of this program, three different methods were used for the collection and analysis of hydrocarbons. These were:

a) EPA Method 25,

b) Proposed EPA Method 18 (both on-site and off-site analyses performed), and

c) Byron instruments Model 90 sample collection system and Model 401 hydrocarbon analyzer sampling system and instrument combination.

To characterize the VOC destruction efficiency across the thermal incinerator, liquid, solid, and gas phase sampling was performed. The sampling locations were:

a) Incinerator inlet - waste gas stream
                      - natural gas stream
                      - atactic waste stream

b) Waste heat boiler outlet, and

c) Scrubber stack outlet (volumetric flow rate).

The sampling system used for Method 25 consisted of a mini-impinger moisture knockout, a condensate trap, flow control system, and a sample tank. Both pre- and post-sampling leak tests were performed to ensure sample integrity. In the case of Method 18, samples were collected using a modification of EPA Method 110 for benzene. This

Table C-4. TYPICAL INCINERATOR PARAMETERS FOR ARCO POLYMERS EMISSION TESTING BASED ON DATA FROM
AUGUST 1981
(PROVIDED BY ARCO CHEMICAL COMPANY FROM PROCESS DATA)[3]

Parameter		Waste Liquid to Incinerator from Monument I Plant	Waste Liquid to Incinerator from Monument II Plant	Waste Gas to Incinerator from Monument I & II Plants	Atactic Waste Recirculation from Monument I & II Plants
Temperature:	Range	60-68°C(140-155°F)	52-60°C(125-140°F)	107-116°C(225-240°F)	66-71°C(150-160°F)
Pressure:	Average	-	-	-	-
	Range	960-1,030 kPa(140-150 psig)	998-1,070 kPa(145-155 psig)	21-41 kPa(3-6 psig)	585 kPa(85 psig)
	Maximum	-	-	-	860 kPa(125psig)
Steam Rate:	Average	-	-	0	-
	Maximum	-	-	-	-
Air Flow:	Average	-	-	-	-
	Maximum	-	-	-	-
Nitrogen Flow:	Range	-	-	0.047-0.062 m^3/sec (6,000-8,000 scfh)	-
Organic Solids:	Average	41.2 g/sec(327 lb/hr)	51.8 g/sec(411 lb/hr)	-	-
Organic Liquids:	Average	32.7 g/sec(259 lb/hr)	32.7 g/sec(259 lb/hr)	-	126 g/sec(1,000 lb/hr)
Organic Gases:	Average	3.9 g/sec(31 lb/hr)	4.4 g/sec(35lb/hr)	11-13 g/sec(90-100 lb/hr) (propylene)	-
	Range	-	-	-	-

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Table C-4. TYPICAL INCINERATOR PARAMETERS FOR ARCO POLYMERS EMISSION TESTING BASED ON DATA FROM AUGUST 1981 (Concluded)
(PROVIDED BY ARCO CHEMICAL COMPANY FROM PROCESS DATA)[3]

Parameter		Atactic Waste to Incinerator	Combustion and Quench Air to Incinerator	Natural Gas to Incinerator[a]	Steam Production from Waste Heat Boiler	Waste Heat Boiler Outlet -VOC Sampling Point-
Temperature:	Average	–	41°C(105°F)	29°C(85°F)	193°C(380°F)	211°C(412°F)
	Range	66-68°C(150-155°F)	–	–	–	–
Pressure:	Average	413 kPa(60 psig)	–	–	1,240 kPa(180 psig)	–
	Range	–	–	34-69 kPa(5-10 psig)	–	8.5-18 kPa(34-71 in.H$_2$O)[b]
	Maximum	–	–	–	–	–
Steam Rate:	Average	–	–	–	3,250 g/sec(25,800 lb/hr)	–
	Maximum	–	–	–	6,680 g/sec(53,000 lb/hr)	–
Air Flow:	Average	–	11.34 m^3/sec(24,000 scfm)	–	–	–
	Maximum	–	16.02 m^3/sec(34,000 scfm)	–	–	–
Nitrogen Flow:	Range	–	–	–	–	–
Organic Solids:	Average	–	–	–	–	–
Organic Liquids:	Average	–	–	–	–	–
Organic Gases:	Average	–	–	–	–	–
	Range	–	–	0.079-0.118 m^3/sec (10,000-15,000 scfh)[c]	–	–

[a] Number 6 fuel oil can be substituted for natural gas.

[b] During test period, this pressure was 10" H$_2$O.

[c] Cut back during Atactic Waste incineration.

modification was necessary due to the high moisture content of the incinerator gases and the positive pressure of the emissions. To ensure that a representative, integrated sample was collected using the modified Method 18, three validation tests for sample flow rate and sample volume into the Tedlar bag were performed.

The principle underlying the Byron method is the same as EPA Method 25. However, rather than using a modified standard GC, the Byron method uses a process analyzer. This instrument speciates C_2 from higher hydrocarbons, but gives a single value for all nonmethane hydrocarbons. After separation, all carbonaceous material is combusted to CO_2 which is then converted to CH_4 before being measured by an FID. Thus, the variable response of the FID to different types of organics is eliminated in the Byron 401 as it is in EPA Method 25.

The oxides of nitrogen (NO_x) content of the flue gas was determined using the methodology specified in EPA Method 7. A detailed description of all these sampling and analytical techniques can be found in the ARCO test report.

The total flue gas flow rate was determined two or three times daily using procedures described in EPA Method 2. Based on this method, the volumetric gas flow rate was determined by measuring the cross-sectional area of the stack and the average velocity of the flue gas. The area of the stack was determined by direct measurements.

The work performed during this program incorporated a comprehensive quality assurance/quality control (QA/QC) program as an integral part of the overall sampling and analytical effort. The major objective of the QA/QC program was to provide data of known quality with respect to completeness, accuracy, precision, representativeness, and comparability.

C.2.1.3 Test Results. The VOC measurements were made by at least four of five independent methods for each of eight different combinations of incinerator temperature and waste streams. Table C-5 summarizes the results of measured destruction efficiencies (DE's) for each of these conditions.

The results indicate that the values for the DE's by Method 25 are consistently lower and of poorer quality. The poorer quality is indicated by the imprecision reflected by the much larger standard

Table C-5.   ARCO POLYMERS INCINERATOR DESTRUCTION EFFICIENCIES FOR EACH SET OF CONDITIONS

Conditions[b]	Percent Destruction Efficiency[a] Calculated for Each Method				
	Method 18 (on-site) HC[c]	Byron THC[d]	Byron NMHC[e]	Method 25[f]	Method 18 (off-site) Speciated HC[g]
AW/NG WG 1,000°F	99.99777 ± .00008	99.994 ± .002	99.997 ± .002	99.844 ± .006	
AW NG/WG 1,300°F	>99.9979 ± .0004	99.996 ± .001	99.998 ± .001	99.3 ± .4	
AW/NG/WG 1,600°F	>99.99721 ± .00009	99.9961 ± .0003	99.9957 ± .0002	99.6 ± .2	
NG WG 1,300°F	99.3 ± .1	99.9 ± .1	99.6 ± .4	76 ± 20	
NG WG 1,600°F	>99.76 ± .07	99.3 ± .10	99.88 ± .04	66 ± 10	99.38 ± .04
AW NG 1,000°F	99.99674 ± .00007	99.9941 ± .0001	99.99796 ± .00005	96.32 ± .08	
AW NG 1,300°F	>99.990 ± .004	99.983 ± .007	99.983 ± .007	98 ± 3	
AW/NG 1,600°F	>99.9975 ± .0001	99.994 ± .002	99.995 ± .003[h]	99 ± 1	99.9979 ± .0001

[a] Percent destruction efficiency $= 100 - \dfrac{gC \text{ in Stack gas}}{(gC \text{ in Atactic Waste} + gC \text{ in Waste Gas})}$

where: gC = grams of organic carbon
The number following the ± sign is the standard deviation (statistically expected true value would fall between the reported value minus the standard deviation and the reported value plus the standard deviation).

[b] Conditions of test given are materials burned and the temperature of the incinerator. Material codes are AW = Atactic Waste, NG = Natural Gas, and WG = Waste Gas. Incinerator design parameters are about 2.18 MJ/s (7.45 MMBtu/hr), 15.1 sm³/s (33,900 scfm) air supply, 980°C; 1200°C(1800°F; 2200°F maximum) firebox temperature, 1.5 seconds residence time, and 19 kPa (78 in. H₂O pressure).

[c] Measured using proposed EPA Method 18 (on-site) for hydrocarbons (HC) utilizing gas chromatography (GC) with a flame ionization detector (FID). The values with "greater than" signs (>) indicate that the VOC was below the detectable limit and the detection level was used to calculate the DE's.

[d] Measured using the Byron Instruments Model 90 sample collection system and the Bryon Model 401 Hydrocarbon Analyzer sampling system and instrument combination (utilizing reduction to methane and FID) in the total hydrocarbon (THC) mode.

[e] Measured using the Byron Models 90 and 401 combination (utilizing reduction to methane and FID) in the nonmethane hydrocarbon mode.

[f] Measured using EPA Method 25 for total gaseous nonmethane organics (TGNMO) utilizing GC-FID. Data not believed to represent true values.

[g] Measured using proposed EPA Method 18 (off-site) for individual hydrocarbon species utilizing GC-FID.

[h] Difficulties with analysis - Based on most probable value.

deviations for this measurement method. The accuracy and representativeness of these values obtained from Method 25 is, thus, questionable. If Method 25 results are disregarded, the DE's for all testing combinations are found to be consistently above 99 percent.

C.2.2 Environmental Protection Agency (EPA) Air Oxidation Unit Test Data

The EPA test study represents the most in-depth work available for full-scale incinerators on air oxidation vents at three chemical plants. Data includes inlet/outlet tests on three large incinerators. The tests measured inlet and outlet VOC concentrations by compound for different incinerator temperatures. The referenced test reports include complete test results, process rates, and test method descriptions. The three plants tested are Denka's maleic anhydride unit in Houston, Texas, Rohm and Haas's acrylic acid unit in Deer Park, Texas, and Union Carbide's acrylic acid unit in Taft, Louisiana. The data from Union Carbide include test results for two different incinerator temperatures. The data from Rohm and Haas include results for three temperatures. In all tests, bags were used for collecting integrated samples and a GC/FID was used for organic analysis.

C.2.2.1 Denka Test Data.[4] The Denka maleic anhydride facility has a nameplate capacity of 23 Gg/yr (50 million lbs/yr). Maleic anhydride is produced by vapor-phase catalytic oxidation of benzene. The liquid effluent from the absorber, after undergoing recovery operations, is about 40 weight percent aqueous solution of maleic acid. The absorber vent is directed to the incinerator. The thermal incinerator has a primary heat recovery system to generate process steam and uses natural gas as supplemental fuel. The plant was operating at about 70 percent of capacity when the sampling was conducted. The plant personnel did not think that the lowered production rate would seriously affect the validity or representativeness of the results.

1. Control Device. The size of the incinerator combustion chamber is 204 m^2 (2,195 ft^2). There are three thermocouples used to sense the flame temperature, and these are averaged to give the temperature recorded in the control room. A rough sketch of the combustion chamber is provided in Figure C-3.
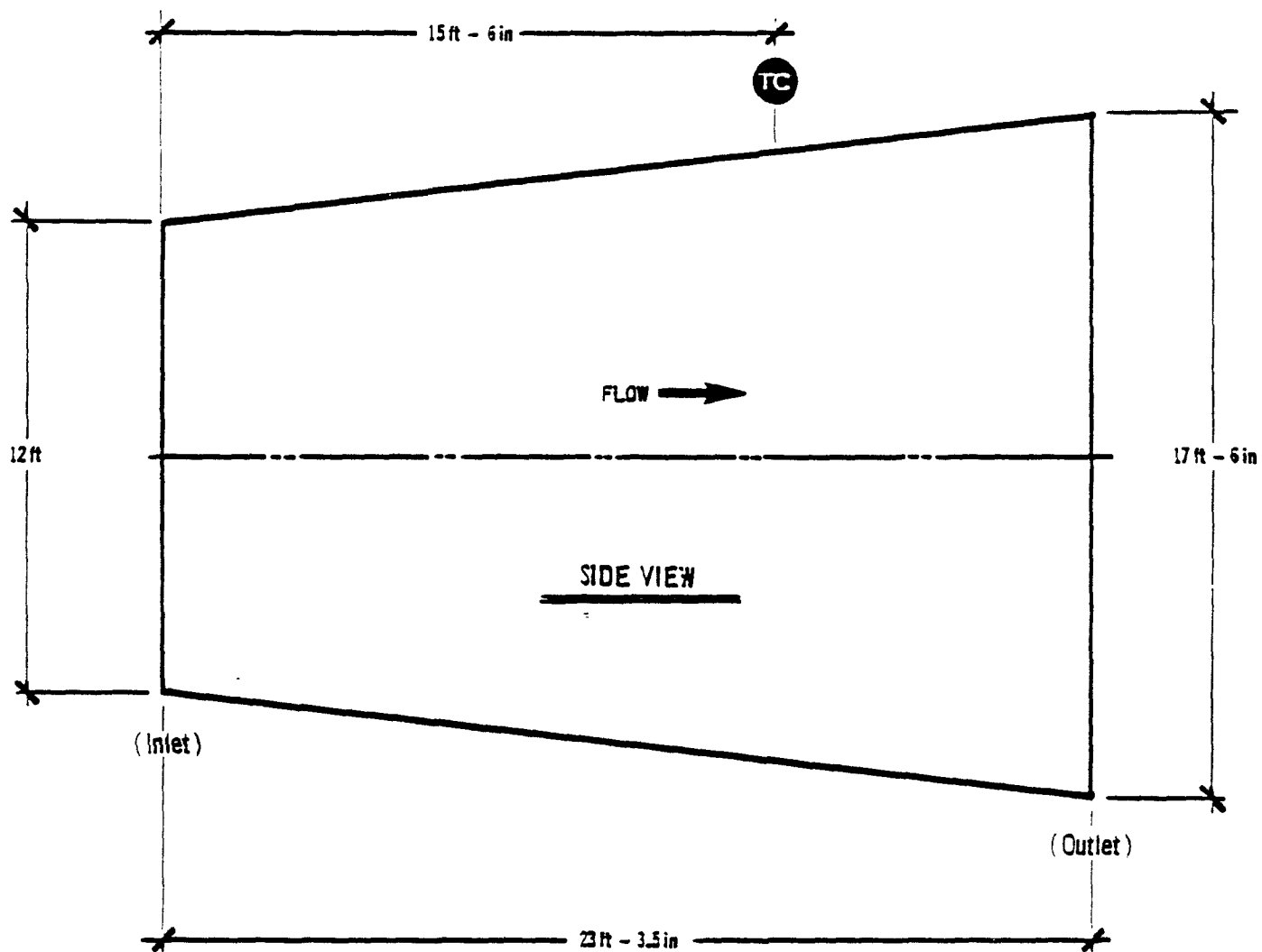
There are Three Thermocouples Spaced Evenly Across the Top of the Firebox.

The Width of the Firebox is 6 ft – 6 in.

Figure C-3. Incinerator Combustion Chamber

2. <u>Sampling and Analytical Techniques.</u> Gas samples of total hydrocarbons (THC), benzene, methane, and ethane were obtained according to the September 27, 1977, EPA draft benzene method. Seventy-liter aluminized Mylar[R] bags were used to collect samples over periods of two to three hours for each sample. The insulated sample box and bag were heated to approximately 66°C (150°F) using an electric drum heater. During Run 1-Inlet, the rheostat used to control the temperature malfunctioned so the box was not heated for this run. A stainless steel probe was inserted into the single port at the inlet and connected to the gas bag through a "tee." The other leg of the "tee" went to the total organic acid (TOA) train. A Teflon[R] line connected the bag and the "tee." A stainless steel probe was connected directly to the bag at the outlet. The lines were kept as short as possible and not heated. The boxes were transported to the field lab immediately upon completion of sampling. They were heated until the GC analyses were completed.

A Varian model 2440 gas chromatograph with a Carle gas sampling valve, equipped with matched 2 cm³ loops, was used for the integrated bag analysis. The SP-1200/Bentone 34 GC column was operated at 80°C (176°F). The instrument has a switching circuit which allows a bypass around the column through a capillary tube for THC response. The response curve was measured daily for benzene (5, 10, and 50 ppm standards) with the column and in the bypass (THC) mode. The THC mode was also calibrated daily with propane (20, 100, and 2000 ppm standards). The calibration plots showed moderate nonlinearity. For sample readings that fell within the range of the calibration standards, an interpolated response factor was used from a smooth curve drawn through the calibration points. For samples above or below the standards, the response factor of the nearest standard was assumed. THC readings used peak height and column readings used area integration measured with an electronic "disc" integrator.

Analysis for carbon monoxide was done on samples drawn from the same integrated gas sample bag used for the THC, benzene, methane, and ethane analyses. Carbon monoxide analysis was done following the GC analyses using EPA Reference Method 10 (<u>Federal Register</u>, Vol. 39,

No. 47, March 8, 1974). A Beckman Model 215 NDIR analyzer was used to analyze both the inlet and outlet samples.

Duct temperature and pressure values were obtained from the existing inlet port. A thermocouple was inserted into the gas sample probe for the temperature while a water manometer was used for the pressure readings. These values were obtained at the conclusion of the sampling period.

Temperature, pressure, and velocity values were obtained for the outlet stack. Temperature values were obtained by a thermocouple during the gas sampling. Pressure and velocity measurements were taken according to EPA Reference Method 2 (Federal Register, Vol. 42, No. 160, August 18, 1977). These values also were obtained at the conclusion of the sampling period.

3. Test Results - The Denka incinerator achieved greater than 98 percent reduction at 760°C (1400°F) and 0.6 second residence time. These results suggest that 98 percent control is achievable by properly maintained and operated incinerators under operating conditions less stringent than 870°C (1600°F) and 0.75 second. Table C-6 provides a summary of these test results.

C.2.2.2 Rohm and Haas Test Data[5]. The Rohm and Haas plant in Deer Park, Texas, produces acrylic acid and ester. The capacity of this facility has been listed at 181 Gg/yr (400 million lbs/yr) of acrylic monomers. Acrylic esters are produced using propylene, air, and alcohols, with acrylic acid produced as an intermediate. Acrylic acid is produced directly from propylene by a vapor-phase catalytic air oxidation process. The reaction product is purified in subsequent refining operations. Excess alcohol is recovered and heavy end by-products are incinerated. This waste incinerator is designed to burn offgas from the two absorbers. In addition, all process vents (from extractors, vent condensers, and tanks) that might be a potential source of gaseous emissions are collected in a suction vent system and normally sent to the incinerator. An organic liquid stream generated in the process is also burned, thereby providing part of the fuel requirement. The remainder is provided by natural gas.

Table C-6. AIR OXIDATION UNIT THERMAL INCINERATOR FIELD TEST DATA

Company & location	Type of Process	Waste Gas Flow Rate (scfm)	Set Number (Number of Tests)	Test Date	Supplemental Fuel and Amount Used (scfm)	Residence Time (seconds)	Temp. °F	VOC Concentration (ppm) Inlet	VOC Concentration (ppm) Outlet	VOC Destruction Efficiency, by weight (%)
					Natural Gas					
Denka, Houston, TX	Maleic Anhydride	33,000[b]	(3)	3/21/78	1060 (gas)	0.6	1400	950	13	98.5
				3/22/78	1060	0.6	1400	950	13	98.5
				3/23/78	1060	0.6	1400	950	13	98.5
Rohm & Haas, Deer Park, TX	Acrylic Acid & Esters	52,500[c]	Set 1 (3)	3/78	900 (gas)	1.0	1425	TVF[c] 2,580 / OXV[c] 11,600	1330	82.6
			Set 2 (4)	3/78	900	1.0	1510	TVF[c] 2,600 / OXV[c] 12,800	150	98.3
			Set 3 (1)	3/78	900	1.0	1545	TVF[c] 2,410 / OXV[c] 12,200	25	99.7
Union Carbide Corp. Taft, LA	Acrylic Acid & Acrylate Esters	20,600	Set 1 (6)	12/78	Natural Gas	2-3	1160	11,900	243	96.1
			Set 2 (3)	12/78		2-3	1475	11,900	10	99.9
Petro-Tex Chemical Corp., Houston, TX	Butadiene	7,250[d]	Set 1	5/25/77	1400	0.6	1400	10,300	1000	70.3
		15,617[d]	Set 2	9/09/77	1467	0.6	1400	10,650	215	94.1
		20,750[d]	Set 3	12/01/77	900	0.6	1400	10,650	215	94.1
		15,867[d]	Set 4	4/19/78	1175	0.6	1400	10,300	10	99.6
		12,500[d]	Set 5	9/27/78	1176	0.6	1400	10,300	10	99.6
Monsanto Chemical Intermediates Co., Alvin, TX	Acrylonitrile	75,000[f]	Unit 1	12/16/77	Natural Gas	NA[e]	Confidential	Confidential	25	99
			Unit 2	12/16/77	Natural Gas	NA[e]	dential	dential	47	99

[a] At production capacity.

[b] At 79 percent rather than 100 percent of total capacity.

[c] Total waste gas flow rate of 52,500 scfm composed of 12,500 scfm from tank farm vent (TVF) and 40,000 scfm from oxidizer vent (OXV).

[d] Average combustion air was 49,333 scfm.

[e] Not available.

[f] Average.

1. Control Device - Combustion air is added to the incinerator in an amount to produce six percent oxygen in the effluent. Waste gases are flared during maintenance shutdowns and severe process upsets. The incinerator unit operates at relatively shorter residence times (0.75-1.0 seconds) and higher combustion temperatures (650° - 850°C) [1200°-1560°F] than most existing incinerators.

The total installed capital cost of the incinerator was $4.7 million. The estimated operating cost due to supplemental natural gas use is $0.9 million per year.

2. Sampling and Analytical Techniques - Samples were taken simultaneously at a time when propylene oxidations, separations, and esterifications were operating smoothly and the combustion temperature was at a steady state. Adequate time was allowed between the tests conducted at different temperatures for the incinerator to achieve steady state. Bags were used to collect integrated samples and a GC/FID was used for organic analysis.

3. Test Results - VOC destruction efficiency was determined at three different temperatures and a residence time of 1.0 second at each temperature. The test results are summarized in Table C-6. Efficiency is found to increase with temperature and, except for 774°C (1425°F), is above 98 percent. Theoretical calculations show that greater efficiency would be achieved at 870°C (1600°F) and 0.75 second than at the longer residence times but lower temperatures represented in these tests.

C.2.2.3 Union Carbide Corporation (UCC) Test Data[6]. The total capacity for the UCC acrylates facilities is about 90 Gg/yr (200 million lbs/yr) of acrolein, acrylic acid, and esters. Acrylic acid comprises 60 Gg/yr (130 million lbs/yr) of this total. Ethyl acrylate capacity is 40 Gg/yr (90 million lbs/yr). Total heavy ester capacities (such as 2-ethyl-hexyl acrylate) are 50 Gg/yr (110 million lbs/yr). UCC considers butyl acrylate a heavy ester.

The facility was originally built in 1969 and utilized British Petroleum technology for acrylic acid production. In 1976 the plant was converted to a technology obtained under license from Sohio.

1. Control Device - The thermal incinerator is one of the two major control devices used in acrylic acid and acrylate ester manufacture.

The UCC incinerator was installed in 1975 to destroy acrylic acid and acrolein vapors. This unit was constructed by John Zink Company for an installed cost of $3 million and incorporates a heat recovery unit to produce process steam at 4.1 MPa (600 psig). The unit operates at a relatively constant feed input and supplements the varying flow and fuel value of the streams fed to it with inversely varying amounts of fuel gas. Energy consumption averages 15.5 MJ/s (52.8 million Btu/hr) instead of the designed level of 10.5 to 14.9 MJ/s (36 to 51 million Btu/hr). The operating cost in 1976, excluding capital depreciation, was $287,000. The unit is run with nine percent excess oxygen instead of the designed three to five percent excess oxygen. The combustor is designed to handle a maximum of four percent propane in the oxidation feed.

The materials of construction of a nonreturn block valve in the 4.1 MPa (600 psig) steam line from the boiler section require that the incinerator be operated at 650°C (1200°F) instead of the designed 980°C (1800°F). The residence time is three to four seconds.

2. Sampling and Analytical Techniques - The integrated gas samples were obtained according to the September 27, 1977, EPA draft benzene method.

Each integrated gas sample was analyzed on a Varian Model 2400 gas chromatograph with FID, and a heated Carle gas sampling valve with matched 2-cm^3 sample loops. A valved capillary bypass is used for total hydrocarbon (THC) analyses and a 2 m long, 3.2 mm (1/8-in.) outer diameter nickel column with PORAPAK^R P-S, 80-100 mesh packing is used for component analyses.

Peak area measurements were used for the individual component analyses. A Tandy TRS-80, 48K floppy disc computer interfaced via the integrator pulse output of a Linear Instruments Model 252A recorder acquired, stored, and analyzed the chromatograms.

The integrated gas samples were analyzed for oxygen and carbon dioxide by duplicate Fyrite readings. Carbon monoxide concentrations were obtained using a Beckman Model 215A nondispersive infrared (IR) analyzer using the integrated samples. A three-point calibration (1000, 3000, and 10,000 ppm CO standards) was used with a linear-log curve fit.

Stack traverses for outlet flowrate were made using EPA Methods 1 through 4 (midget impingers) and NO$_x$ was sampled at the outlet using EPA Method 7.

3. Test Results - VOC destruction efficiency was determined at two different temperatures. Table C-6 provides a summary of these test results. Efficiency was found to increase with temperature. At (800°C) 1475°F, the efficiency was well above 99 percent. These tests were, again, for residence times greater than 0.75 second. However, theoretical calculations show that even greater efficiency would be achieved at 870°C (1600°F) and 0.75 second than at the longer residence times but lower temperatures represented in these tests.

All actual measurements were made as parts per million (ppm) of propane with the other units reported derived from the equivalent values. The values were measured by digital integration.

The incinerator combustion temperature for the first six runs was about 630°C (1160°F). Runs 7 through 9 were made at an incinerator temperature of about 800°C (1475°F). Only during Run 3 was the acrolein process operating. The higher temperature caused most of the compounds heavier than propane to drop below the detection limit due to the wide range of attenuations used, nearby obscuring peaks, and baseline noise variations. The detection limit ranges from about 10 parts per billion (ppb) to 10 ppm, generally increasing during the chromatogram, and especially near large peaks. Several of the minor peaks were difficult to measure. However, the compounds of interest, methane, ethane, ethylene, propane, propylene, acetaldehyde, acetone, acrolein, and acrylic acid, dominate the chromatograms. Only acetic acid was never detected in any sample.

The probable reason for negative destruction efficiencies for several light components is generation by pyrolysis from other components. For instance, the primary pyrolysis products of acrolein are carbon monoxide and ethylene. Except for methane and, to a much lesser extent, ethane and propane, the fuel gas cannot contribute hydrocarbons to the outlet samples.

A sample taken from the inlet line knockout trap showed 6 mg/g of acetaldehyde, 25 mg/g of butenes, and 100 mg/g of acetone when analyzed by gas chromatography/flame ionization detection (GC/FID).

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C.2.3 Chemical Company Air Oxidation Unit Test Data

These data are from tests performed by chemical companies on incinerators at two air oxidation units: the Petro-Tex oxidative butadiene unit at Houston, Texas, and the Monsanto acrylonitrile unit at Alvin, Texas. Tests at a third air oxidation unit, the Koppers maleic anhydride unit at Bridgeville, Pennsylvania,[7] were disregarded as not accurate because of poor sampling technique.[8]

C.2.3.1 Petro-Tex Test Data[9]. The Petro-Tex Chemical Corporation conducted emission testing at its butadiene production facility in Houston, Texas, during 1977 and 1978. This facility was the "Oxo" air oxidation butadiene process. The emission tests were conducted during a period when Petro-Tex was modifying the incinerator to improve mixing and, thus, VOC destruction efficiency.

1. Control Device - The Petro-Tex incinerator for the 'Oxo' butadiene process is designed to treat 48,000 scfm waste gas containing about 4000 ppm hydrocarbon and 7000 ppm carbon dioxide. The use of the term hydrocarbon in this discussion indicates that besides VOC, it may include nonVOC such as methane. The waste gas treated in this system results from air used to oxidize butene to butadiene. After butadiene has been recovered from air oxidation waste gas in an oil absorption system, the remaining gas is combined with other process waste gas and fed to the incinerator. The combined waste gas stream enters the incinerator between seven vertical Coen duct burner assemblies. The incinerator design incorporates flue gas recirculation and a waste heat boiler. The benefit achieved by recirculating flue gas is to incorporate the ability to generate a constant 100,000 lbs/hr of 750 psi steam with variable waste gas flow.[10] The waste gas flow can range from 10 percent to 100 percent of the design production rate.

The incinerator measures 72 feet by 20 feet by 8 feet, with an average firebox cross-sectional area of 111 square feet. The installed capital cost was $2.5 million.

The waste gas stream contains essentially no oxygen; therefore, significant combustion air must be supplied. This incinerator is fired with natural gas which supplies 84 percent of the firing energy. The additional required energy is supplied by the hydrocarbon content of the waste gas stream. Figure C-4 gives a rough sketch of this unit.
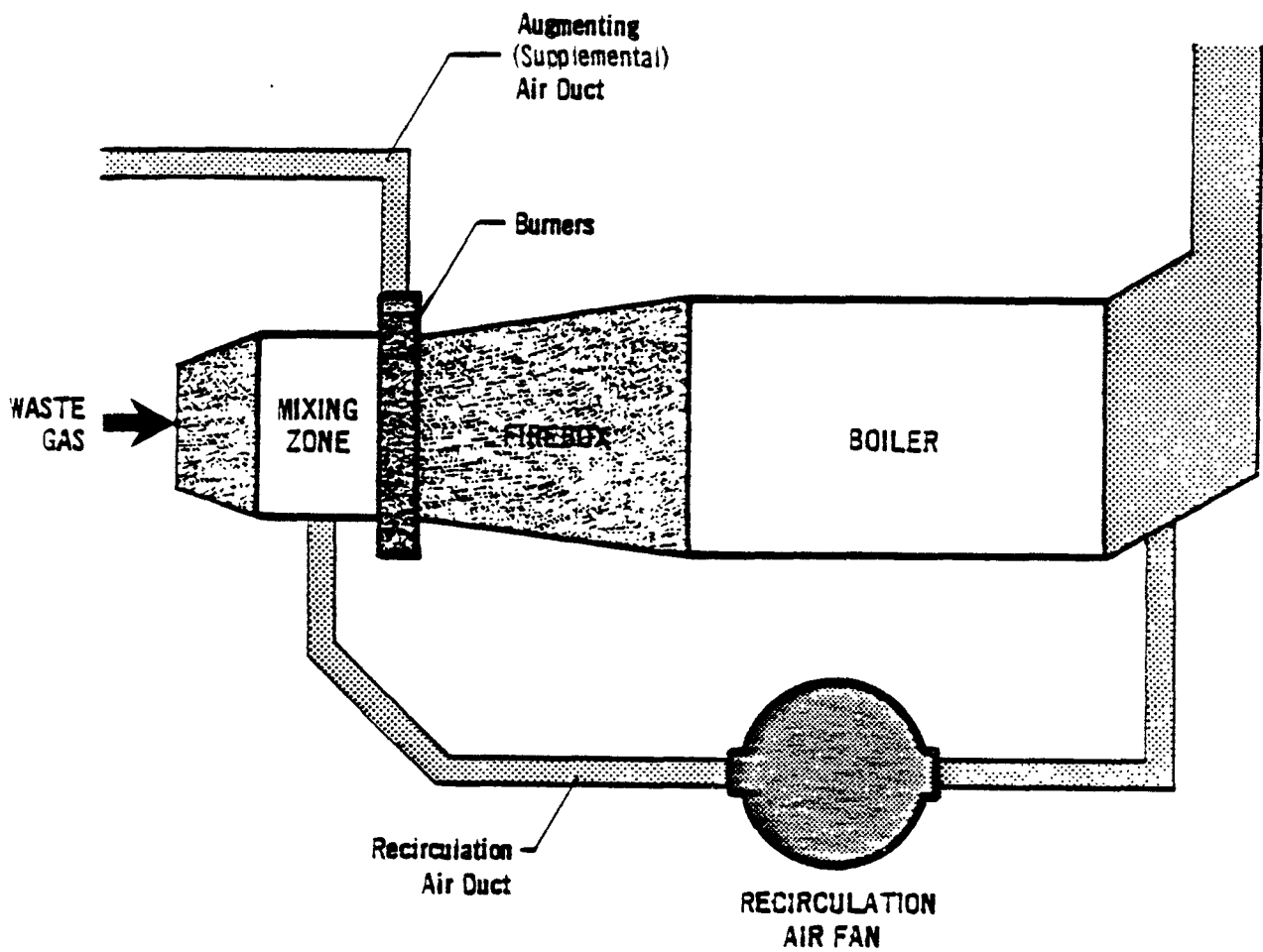
Figure C-4.   Petro-Tex oxo unit incinerator.

2. <u>Sampling and Analytical Techniques.</u> Integrated waste gas samples were collected in bags. The analysis was done on a Carle analytical gas chromatograph having the following columns:

1. 6-ft OPN/PORASIL^R (80/100).

2. 40-ft 20 percent SEBACONITRILE^R on gas chrom. RA 42/60.

3. 4-ft PORAPAK^R N 80/100.

4. 6-ft molecular sieve bx 80/100.

Stack gas samples were collected in 30 to 50 cc syringes via a tee on a long stainless steel probe, which can be inserted into the stack, at nine different locations. They were then transferred to a smaller 1 cm^3 syringe via a small glass coupling device sealed at both ends with a rubber grommet. The 1-cm^3 samples were injected into a Varian 1700 chromatograph for hydrocarbon analysis. The chromatograph has a 1/8-in. x 6-ft column packed with 5A molecular sieves and a 1/4-in. x 4-ft column packed with glass beads connected in series with a bypass before and after the molecular sieve column, controlled by a needle valve to split the sample. The data are reported as ppm total HC, ppm methane, and ppm non-methane hydrocarbons (NMHC). The CO content in the stack was determined by using a Kitagawa sampling probe. The O_2 content in the stack was determined via a Teledyne O_2/combustible analyzer.

3. <u>Test Results.</u> Petro-Tex has been involved in a modification plan for its 'Oxo' incinerator unit after startup. The facility was tested by the company after each major modification to determine the impact of these changes on the VOC destruction efficiency. The incinerator showed improved performance after each modification and the destruction efficiency increased from about 70 percent to above 99 percent. Table C-4 provides a summary of these test results. The modifications made in the incinerator are described below.

<u>November 1977</u>

Test data prior to these changes showed the incinerator was not destroying hydrocarbons as well as it should (VOC destruction efficiency as low as 70 percent), so the following changes were made:

1. Moved the duct burner baffles from back of the burner to the front;

2. Installed spacers to create a continuous slot for supplemental air to reduce the air flow through the burner pods;

3. Installed plates upstream of the burners so that ductwork matches burner dimensions;

4. Cut slots in recycle duct to reduce exit velocities and improve mixing with Oxo waste gas;

5. Installed balancing dampers in augmenting (supplemental) air plenums, top and bottom;

6. Installed balancing dampers in three of the five sections of the recycle duct transition; and

7. Cut opening in the recirculation duct to reduce the outlet velocities.

March 1978

After the November changes were made, a field test was made in December 1977, which revealed that the incinerator VOC destruction efficiency increased from 70.3 percent to 94.1 percent. However, it still needed improvement. After much discussion and study the following changes were made in March 1978:

1. Took the recirculation fan out of service and diverted the excess forced draft air into the recirculation duct;

2. Sealed off the 14-cm (5-1/2-in.) wide slots adjacent to the burner pods and removed the 1.3 cm (1/2-in.) spacers which were installed in November 1977;

3. Installed vertical baffles between the bottom row of burner pads to improve mixing;

4. Installed perforated plates between the five recirculation ducts for better waste gas distribution in the incinerator; and

5. Cut seven 3-in. wide slots in the recycle duct for better secondary air distribution.

July 1978

After the March 1978 changes, a survey in April 1978, showed the Oxo incinerator to be performing very well (VOC destruction efficiency of 99.6 percent) but with a high superheat temperature of 450°C (850°F). So, in July 1978, some stainless steel shields were installed over the superheater elements to help lower the superheat temperature. A

subsequent survey in September 1978, showed the incinerator to be still destroying 99.6 percent of the VOC and with a lower superheat temperature of 400°C (750°F).

This study pointed out that mixing is a critical factor in efficiency and that incinerator adjustment after startup is the most feasible and efficient means of improving mixing and, thus, the destruction efficiency.

C.2.3.2 Monsanto Test Data.[11] Acrylonitrile is produced by feeding propylene, ammonia, and excess air through a fluidized, catalytic bed reactor. In the air oxidation process, acrylonitrile, acetonitrile, hydrogen cyanide, carbon dioxide, carbon monoxide, water, and other miscellaneous organic compounds are produced in the reactor. The columns in the recovery section separate water and crude acetonitrile as liquids. Propane, unreacted propylene, unreacted air components, some unabsorbed organic products, and water are emitted as a vapor from the absorber column overhead. The crude acrylonitrile product is further refined in the purification section to remove hydrogen cyanide and the remaining hydrocarbon impurities.

The organic waste streams from this process are incinerated in the absorber vent thermal oxidizer at a temperature and residence time sufficient to reduce stack emissions below the required levels. The incinerated streams include (1) the absorber vent vapor (propane, propylene, CO, unreacted air components, unabsorbed hydrocarbons), (2) liquid waste acetonitrile (acetonitrile, hydrogen cyanide, acrylonitrile), (3) liquid waste hydrogen cyanide, and (4) product column bottoms purge (acrylonitrile, some organic heavies). The two separate acrylonitrile plants at Chocolate Bayou, Texas, employ identical thermal oxidizers.

1. Control Device - The Monsanto incinerator burns both liquid and gaseous wastes from the acrylonitrile unit and is termed the absorber vent thermal oxidizer. Two identical oxidizers are employed. The primary purpose of the absorber vent thermal oxidizers is hydrocarbon emission abatement.

Each thermal oxidizer is a horizontal, cylindrical, saddle-supported, end-fired unit consisting of a primary burner vestibule attached to the main incinerator shell. Each oxidizer measures 18 feet in diameter by 36 feet in length.

The thermal oxidizer is provided with special burners and burner guns. Each burner is a combination fuel-waste liquid unit. The absorber vent stream is introduced separately into the top of the burner vestibule. The flows of all waste streams are metered and sufficient air is added for complete combustion. Supplemental natural gas is used to maintain the operating temperature required to combust the organics and to maintain a stable flame on the burners during minimum gas usage. Figure C-5 gives a plan view of the incinerator.

2. Sampling and Analytical Techniques. The vapor feed streams (absorber vent) to the thermal oxidizer and the effluent gas stream were sampled and analyzed using a modified analytical reactor recovery run method. The primary recovery run methods are Sohio Analytical Laboratory procedures.

The modified method involved passing a measured amount of sample gas through three scrubber flasks containing water and catching the scrubbed gas in a gas sampling bomb. The samples were then analyzed with a gas chromatograph and the weight percent of the components was determined.

Figure C-6 shows the apparatus and configuration used to sample the stack gas. It consisted of a sampling line from the sample valve to the small water-cooled heat exchanger. The exchanger was then connected to a 250 ml sample bomb used to collect the unscrubbed sample. The bomb was then connected to a pair of 250 ml bubblers, each with 165 ml of water in it. The scrubbers, in turn, were connected to another 250 ml sample bomb used to collect the scrubbed gas sample which is connected to a portable compressor. The compressor discharge then was connected to a wet test meter that vents to the atmosphere.

After assembling the apparatus, the compressor was turned on drawing the gas from the stack and through the system at a rate of 90 m/s (0.2 ft^3/min). Sample gas was drawn until at least 0.28 m^3 (10 ft^3) passed through the scrubbers. After the 0.28 m^3 (10 ft^3) was scrubbed, the compressor was shutdown and the unscrubbed bomb was analyzed for CH_4, C_2's, C_3H_6, and C_3H_8, the scrubbed bomb was analyzed for N_2, air, O_2, CO_2, and CO, and the bubbler liquid was analyzed for acrylonitrile, acetonitrile, hydrogen cyanide, and total organic carbon. The gaseous samples were analyzed by gas chromatography.
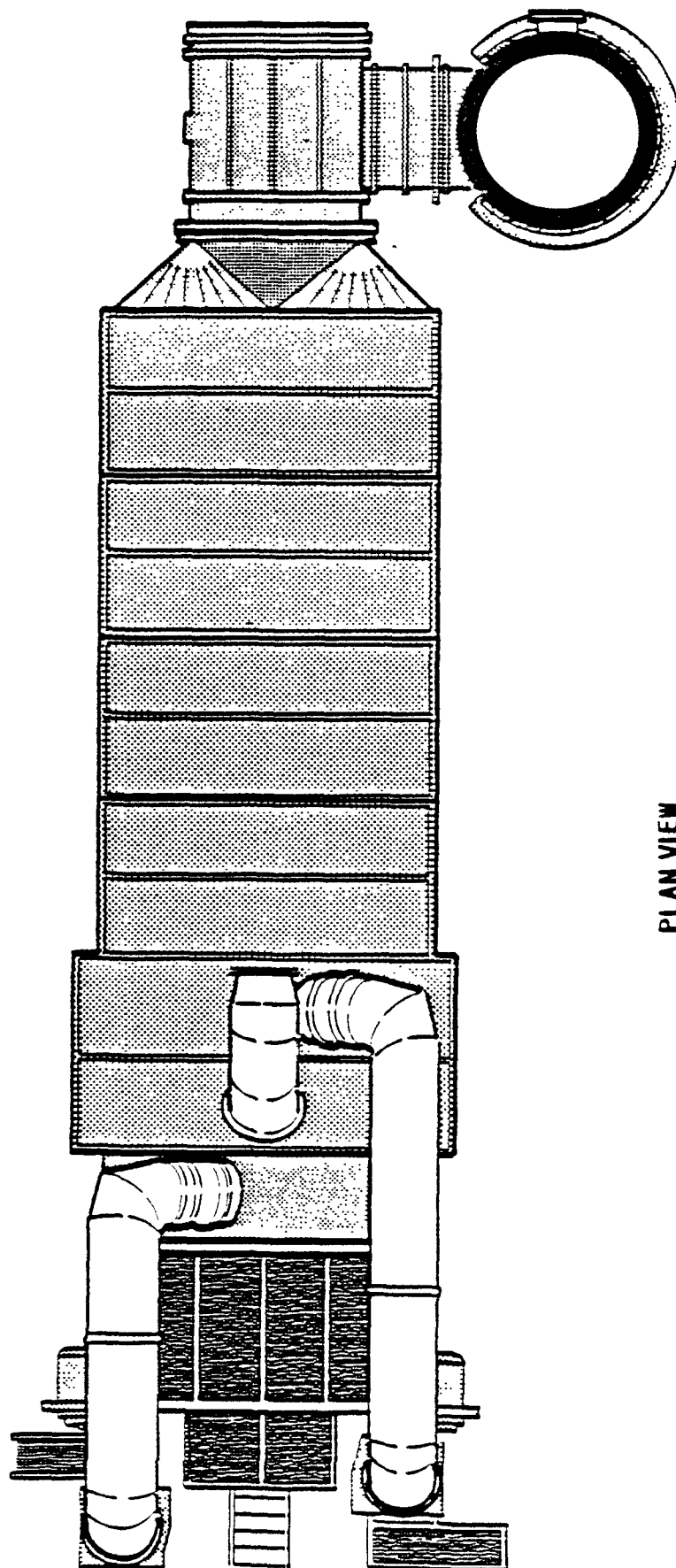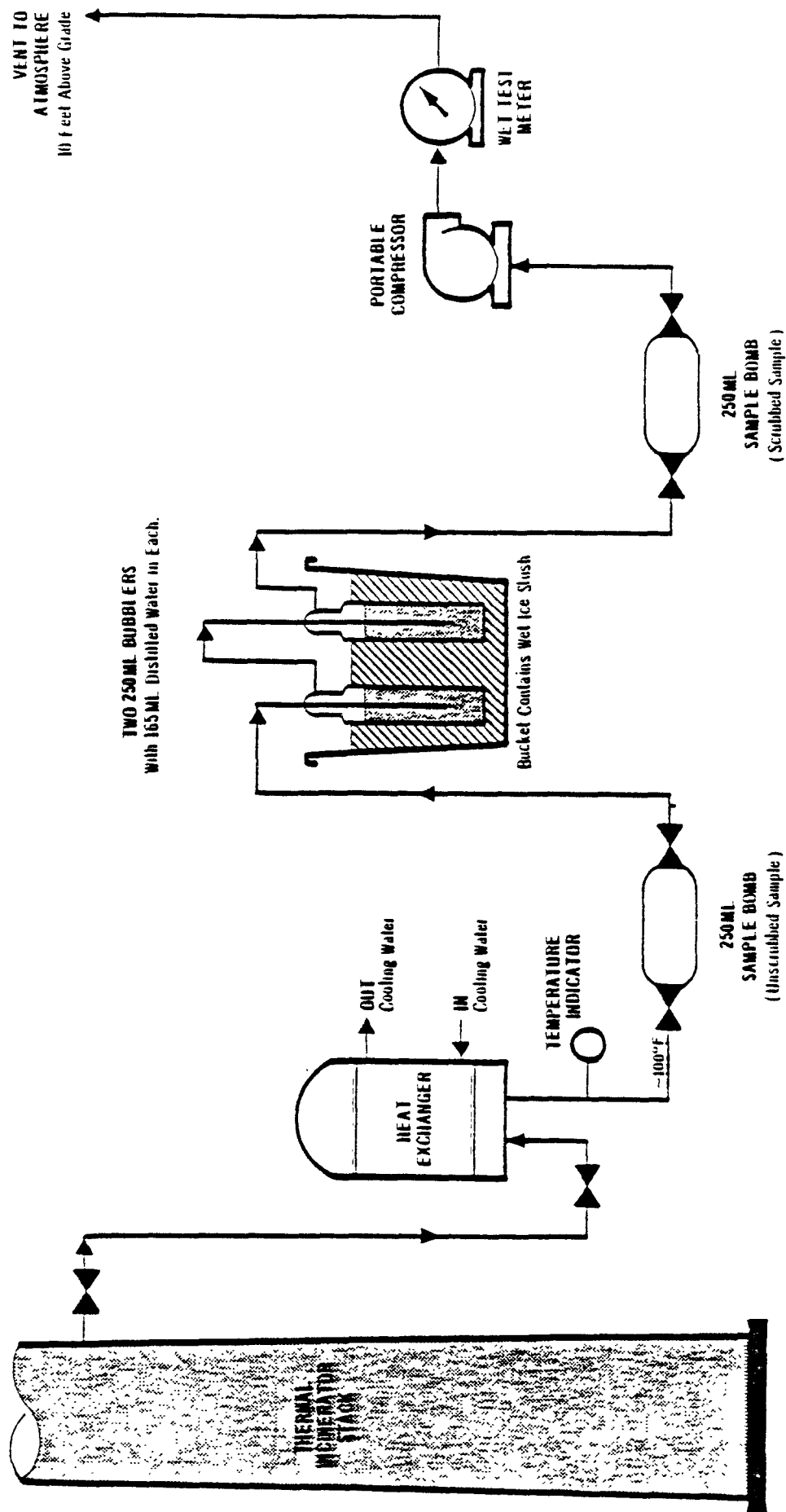
PLAN VIEW

Figure C-5. Off-gas incinerator, Monsanto Co., Chocolate Bayou Plant.

Figure C-6.   Thermal incinerator stack sampling system.

NOTE. From Exchanger Process Outlet, All Lines are Vacuum Tubing.

3. <u>Test Results.</u> The Monsanto Chemical Intermediate Company conducted emissions testing at its Alvin (Chocolate Bayou), Texas, acrylonitrile production facility during December 1977. The VOC destination efficiency reported was 99 percent. (Residence time information was not available and the temperature of the incinerator is considered confidential information by Monsanto.)

C.2.4 <u>Union Carbide Lab-Scale Test Data</u>[12]

Union Carbide test data show the combustion efficiencies achieved on 15 organic compounds in a lab-scale incinerator operating between 430° and 830°C (800° and 1500°F) and 0.1 to 2 seconds residence time. The incinerator consisted of a 130 cm, thin bore tube, in a bench-size tube furnace. Outlet analyzers were done by direct routing of the incinerator outlet to a FID and GC. All inlet gases were set at 1000 ppmv.

In order to study the impact of incinerator variables on efficiency, mixing must first be separated from the other parameters. Mixing cannot be measured and, thus, its impact on efficiency cannot be readily separated when studying the impact of other variables. The Union Carbide lab work was chosen since its small size and careful design best assured consistent and proper mixing.

The results of this study are shown in Table C-7. These results show moderate increases in efficiency with temperature, residence time, and type of compound. The results also show the impact of flow regime on efficiency.

Flow regime is important in interpreting the Union Carbide lab unit results. These results are significant since the lab unit was designed for optimum mixing and, thus, the results represent the upper limit of incinerator efficiency. As seen in Table C-7, the Union Carbide results vary by flow regime. Though some large-scale incinerators may achieve good mixing and plug flow, the worst cases will likely require flow patterns similar to complete backmixing. Thus, the results of complete backmixing would be relatively more comparable to those obtained from large-scale units.

C.3 VAPOR RECOVERY SYSTEM VOC EMISSION TEST DATA[13]

On July 14, 1980, Mobil Company collected samples of hydrocarbon emissions from the exhaust vent of the Vapor Recovery/Knockdown System

Table C-7. DESTRUCTION EFFICIENCY UNDER STATED CONDITIONS
BASED ON RESULTS OF UNION CARBIDE LABORATORY TESTS[a]

| Flow Regime[b] | Temperature (°F) | Destruction Efficiency of Compound in Percent at Residence Time | | | | |
| | | 0.75 second | | | | 0.5/1.5 sec |
		Ethyl Acrylate	Ethanol	Ethylene	Vinyl Chloride	Ethylene
Two-stage Backmixing	1300	99.9	94.6	92.6	78.6	87.2/97.6
	1400	99.9	99.6	99.3	99.0	98.6/99.8
	1500	99.9	99.9	99.9	99.9	99.9/99.9
	1600	99.9	99.9	99.9	99.9	99.9/99.9
Complete Backmixing	1300	98.9	86.8	84.4	69.9	78.2/91.5
	1400	99.7	96.8	95.6	93.1	93.7/97.8
	1500	99.9	99.0	98.7	98.4	98.0/99.0
	1600	99.9	99.7	99.6	99.6	99.4/99.8
Plug Flow	1300	99.9	99.9	99.5	90.2	97.3/99.9
	1400	99.9	99.9	99.9	99.9	99.9/99.9
	1500	99.9	99.9	99.9	99.9	99.9/99.9
	1600	99.9	99.9	99.9	99.9	99.9/99.9

[a]The results of the Union Carbide work are presented as a series of equations. These equations relate destruction efficiency to temperature, residence time, and flow regime for each of 15 compounds. The efficiencies in this table were calculated from these equations.

[b]Three flow regimes are presented: two-stage backmixing, complete backmixing, and plug flow. Two-stage backmixing is considered a reasonable approximation of actual field units, with complete backmixing and plug flow representing the extremes.

at its Santa Ana, California polystyrene plant. The samples were
taken using a MDA-808 Accuhaler[R] pump while velocity was determined
using a Kurz[R] Model 441 air velocity meter. Samples were taken while
the plant was in normal operation. One set of samples was taken while
a vacuum was drawn on dissolver tanks. Another set of samples was
taken while a vacuum was drawn on the flash tank. Both sets of samples
were analyzed for styrene and ethylbenzene by an independnet laboratory.
Computations for emission rates were made based on velocity, sample
volume and sample time. The test results, submitted by the company,
indicate that 0.942 kg/day of ethylbenzene and 10.018 kg/day of styrene
are emitted from the exhaust vent of the vapor recovery/knockdown
system. No more information was provided regarding the sampling and
analysis procedure used by Mobil or the laboratory. It is assumed
that standard industrial practices were used, thus generating valid
estimates of emissions. However, the data should not be used as a
significant basis for emission limitation.

C.4 DISCUSSION OF TEST RESULTS AND THE TECHNICAL BASIS OF THE POLYMERS AND RESINS VOC EMISSIONS REDUCTION REQUIREMENT

This section discusses test results as well as available theoretical
data and findings on flare and incinerator efficiencies, and presents
the logic and the technical basis behind the choice of the selected
control level.

C.4.1 Discussion of Flare Emission Test Results

The results of the five flare efficiency studies summarized in
Section 4.1.1.1.1 showed a 98 percent VOC destruction efficiency
except in a few tests with excessive stream, smoking, or sampling
problems. The results of the Joint CMA-EPA study, summarized in
Table C-2, confirmed that 98 percent VOC destruction efficiency was
achievable for all tests (including when smoking occurred) except when
steam quenching occurred within the range of flare gas velocities and
heating values tested. Therefore, flare gas velocity as high as
18.2 m/s (60 fps) and lower heating values as low as 11.2 MJ/m³ (300 Btu/scf)
were selected as the range of operating conditions that would ensure
achievement of the 98 percent VOC reduction efficiency.

C.4.2 Discussion of Thermal Incineration Test Results

Both the theoretical and experimental data concerning combustion
efficiency of thermal incinerators are discussed in this section. A

theoretical consideration of VOC combustion kinetics leads to the conclusion that at 870°C (1600°F) and 0.75 second residence time, mixing is the crucial design parameter.[14] Published literature indicates that any VOC can be oxidized to carbon dioxide and water if held at sufficiently high temperatures in the presence of oxygen for a sufficient time. However, the temperature at which a given level of VOC reduction is achieved is unique for each VOC compound. Kinetic studies indicate that there are two rate-determining (i.e., critically slow) steps in the oxidation of a compound. The first slow step of the overall oxidation reaction is the initial reaction in which the original compound disappears. The initial reaction of methane (CH_4) has been determined to be slower than that of any other nonhalogenated organic compound. Kinetic calculations show that, at 870°C (1600°F), 98 percent of the original methane will react in 0.3 seconds. Therefore, any nonhalogenated VOC will undergo an initial reaction step within this time. After the initial step, extremely rapid free radical reactions occur until each carbon atom exists as carbon monoxide (CO) immediately before oxidation is complete. The oxidation of CO is the second slow step. Calculations show that, at 870°C (1600°F), 98 percent of an original concentration of CO will react in 0.05 second. Therefore, 98 percent of any VOC would be expected to undergo the initial and final slow reaction steps at 870°C (1600°F) in about 0.35 second. It is very unlikely that the intermediate free radical reactions would take nearly as long as 0.4 seconds to convert 98 percent of the organic molecules to CO. Therefore, from a theoretical viewpoint, any VOC should undergo complete combustion at 870°C (1600°F) in 0.75 second. The calculations on which this conclusion is based have taken into account the low mole fractions of VOC and oxygen which would be found in the actual system. They have also provided for the great decrease in concentration per unit volume due to the elevated temperature. However, the calculations assume perfect mixing of the offgas and combustion air. Mixing has been identified as the crucial design parameter from a theoretical viewpoint.

The test results both indicate an achievable control level of 98 percent at or below 870°C (1600°F) and illustrate the importance of

mixing. Union Carbide results on lab-scale incinerators indicated a minimum of 98.6 percent efficiency at 760°C (1400°F). Since lab-scale incinerators primarily differ from field units in their excellent mixing, these results verify the theoretical calculations and suggest that a full-size field unit can maintain similar efficiencies if designed to provide good mixing. The tests cited in Table C-6 are documented as being conducted on full-scale incinerators controlling offgas from air oxidation process vents of a variety of types of plants. To focus on mixing, industrial units were selected where all variables except mixing were held constant or accounted for in other ways. It was then assumed any changes in efficiency would be due to changes in mixing.

The case most directly showing the effect of mixing is that of Petro-Tex incinerator. The Petro-Tex data show the efficiency changes due to modifications on the incinerator at two times after startup. These modifications (see Section C.2.3.1, 3. Test Results) increased efficiency from 70 percent to over 99 percent, with no significant change in temperature.

A comparison of the Rohm and Haas test versus the Union Carbide lab test, as presented in Table C-8, indirectly shows the effect of mixing. The UCC lab unit clearly outperforms the R&H unit. The data from both units are based on the same temperature, residence time, and inlet stream conditions. The more complete mixing of the lab unit is judged the cause of the differing efficiencies.

The six tests of in-place incinerators do not, of course, cover every feedstock. However, the theoretical discussion given above indicates that any VOC compound should be sufficiently destroyed at 870°C (1600°F). More critical than the type of VOC is the VOC concentration in the offgas. This is true because the kinetics of combustion are not first-order at low VOC concentrations. The Petro-Tex results are for a butadiene plant, and butadiene offgas tends to be lean in VOC. Therefore, the test results support the achievability of 98 percent VOC destruction efficiency by a field incinerator designed to provide good mixing, even for streams with low VOC concentrations.

Table C-8. COMPARISONS OF EMISSION TEST RESULTS FOR UNION CARBIDE
LAB INCINERATOR AND ROHM & HAAS FIELD INCINERATOR[a]

Compound	Rohm and Haas Incinerator Inlet (lbs/hr)	Rohm and Haas Incinerator Outlet (lbs/hr)	Union Carbide Lab Incinerator Inlet (lbs/hr)	Union Carbide Lab Incinerator Outlet (lbs/hr)
Propane	900	150	71.4	0.64
Propylene	1800[b]	150[b]	142.9	5.6
Ethane	10	375	0.8	3.9
Ethylene	30	190	2.4	3.4
TOTAL	2740	865	217.5	13.54
Overall VOC Destruction Efficiency:	68.4%		93.8%	

[a]Table shows the destruction efficiency of the four listed compounds for the Rohm & Haas (R&H) field and Union Carbide (UC) lab incinerators. The R&H results are measured; the UC results are calculated. Both sets of results are based on 1425°F combustion temperature and one second residence time. In addition, the UC results are based on complete backmixing and a four-step combustion sequence consisting of propane to propylene to ethane to ethylene to CO_2 and H_2O. These last two items are worst case assumptions.

[b]Are not actual values. Actual values are confidential. Calculations with actual values give similar results for overall VOC destruction efficiency.

The EPA tests at Union Carbide and Rohm and Haas were for residence times greater than 0.75 second. However, theoretical calculations show that greater efficiency would be achieved at 870°C (1600°F) and 0.75 second than at the longer residence times but lower temperatures represented in these two tests. The data on which the achieveability of the 98 percent VOC destruction efficiency is based is test data for similar control systems: thermal incineration at various residence times and temperatures. If 98 percent VOC reduction can be achieved at a lower temperature, then according to kinetic theory it can certainly be achieved at 870°C (1600°F), other conditions being equal.

A control efficiency of 98 percent VOC reduction, or 20 ppm by compound, whichever is less stringent, has been considered to be acheivable control level for all new incinerators, considering available technology, cost and energy use.[14] This is based on incinerator operation at 870°C (1600°F) and on adjustment of the incinerator after start-up. The 20 ppm (by compound) level was chosen after three different incinerator outlet VOC concentrations, 10 ppm, 20 ppm, and 30 ppm, were analyzed. In addition to the incinerator tests cited earlier in this Appendix, data from over 200 tests by Los Angeles County (L.A.) on various waste gas incinerators were considered in choosing the 20 ppm level. However, the usefulness of the L.A. data was limited by three factors: (1) the incinerators tested are small units designed over a decade ago; (2) the units were designed, primarily, for use on coating operations; and (3) the units were designed to meet a regulation requiring only 90 percent VOC reduction.

The 10 ppmv level was judged to be too stringent. Two of the six non L.A. tests and 65 percent of the L.A. tests fail this criteria. Consideration was given to the fact that many of the units tested were below 870°C (1600°F) and did not have good mixing. However, due to the large percent that failed, it is judged that even with higher temperatures and moderate adjustment, a large number of units would still not meet the 10 ppmv level.

The 20 ppm level was judged to be attainable. All of the non L.A. and the majority of the L.A. units met this criteria. There was concern over the large number of L.A. tests that failed, i.e. 43 percent. However, two factors outweighed this concern.

First, all of the non L.A. units met the criteria. This is significant since, though the L.A. units represent many tests, they represent the same basic design. They all are small units designed over a decade ago to meet a rule for 90 percent reduction. They are for similar applications for the same geographic region designed in many cases by the same vendor. Thus, though many failed, they likely did so due to common factors and do not represent a widespread inability to meet 20 ppm.

Second, the difference between 65 percent failing 10 ppmv and 43 percent failing 20 ppm is larger than a direct comparison of the percentages would reveal. At 20 ppm, not only did fewer units fail, but those that did miss the criteria did so by a smaller margin and would require less adjustment. Dropping the criteria from 10 ppm to 20 ppm drops the failure rate by 20 percent, but is judged to drop the overall time and cost for adjustment by over 50 percent.

The difference between the two levels is even greater when the adjustment effort for the worst case is considered. The crucial point is how close a 10 ppm level pushes actual field unit efficiencies to those of the lab unit. Lab unit results for complete backmixing indicate that a 10 ppm level would force field units to almost match lab unit mixing. A less stringent 20 ppm level increases the margin allowed for nonideal incinerator operation, especially for the worst cases. Given that an exponential increase may occur in costs to improve mixing enough for field units to approach lab unit efficiencies, a drop from 10 ppm to 20 ppm may decrease costs to improve mixing in the worst case by an order of magnitude.

The 30 ppm level was judged too lenient. The only data indicating such a low efficiency was from L.A. All other data showed 20 ppm. The non-L.A. data and lab data meet 20 ppm and the Petro-tex experience showed that moderate adjustment can increase efficiency. In addition, the L.A. units were judged to have poor mixing. The mixing deficiencies were large enough to mask the effect of increasing temperature. Thus, it is judged that 20 ppm could be reached with moderate adjustment and that a 30 ppm level would represent a criteria not based on the best available control technology cost, energy, and environmental impact.

C.5 FUGITIVE EMISSION EQUIPMENT INVENTORY

The fugitive VOC emission and control cost analysis for the polymers and resins industry is based upon the analysis for the synthetic organic chemical manufacturing industry (SOCMI). The SOCMI analysis is reported in EPA-450/3-80-033a, Background Information for Proposed Standards for VOC Fugitive Emissions in Synthetic Organic Chemicals Manufacturing Industry, and EPA-450/3-82-010, Fugitive Emission Sources of Organic Compounds - Additional Information on Emissions, Emission Reductions and Costs. Table C-9 summarizes the model units and emissions estimates developed in these studies. The available fugitive emission equipment inventory from the polymers and resins industry, corresponding emission estimates (based on SOCMI emission factors), and plant capacities are presented in Table C-10. The majority of the plants for which data are available are most similar to SOCMI Model Unit B. One SOCMI model unit (Model Unit B) was chosen to represent all polymers and resins facilities with regard to fugitive VOC emissions.[15]

Table C-9. EQUIPMENT COUNTS AND EMISSIONS FOR FUGITIVE VOC EMISSION SOURCES IN SOCMI MODEL UNITS

| | Revised SOCMI Fugitive Analysis[b] | | | |
| | Equipment count for Model Unit[c] | | | Average Revised Emission Factor (kg/hr/source) |
	Model Unit A	Model Unit B	Model Unit C	
Pump Seals	15	60	185	
Light Liquid Service	8	30	92	0.0494
Heavy Liquid Service	7	30	93	0.0214
Valves	362	1,450	4,468	
Vapor Service	99	402	1,232	0.0056
Light Liquid Service	131	524	1,618	0.0071
Heavy Liquid Service	132	524	1,618	0.00023
Safety/relief valves	13[d]	50[d]	157[d]	
Vapor Service	11	42	130	0.104
Light Liquid Service	1	4	13	
Heavy Liquid Service	1	4	14	
Open-ended lines	104[e]	415[e]	1,277[e]	0.0017
Vapor Service				
Light Liquid Service				
Heavy Liquid Service				
Compressor seals	1	2	8	0.228
Sampling connections	26[f]	104[f]	320[f]	0.0150
Flanges	600	2,400	7,400	0.0083
Total Emissions (kg/hr)				
- baseline:	3.2	12.2	37.9	
- uncontrolled:	4.5	17.3	53.3	

[a]Equipment components in VOC service only.

[b]From *Fugitive Emission Sources of Organic Compounds - Additional Information on Emissions, Emission Reductions, and Costs*, EPA-450/3-82-010, April 1982.

[c]52 percent of existing SOCMI units are similar to Model Unit A;
33 percent of existing SOCMI units are similar to Model Unit B;
15 percent of existing SOCMI units are similar to Model Unit C.

[d]Seventy-five percent of gas safety/relief valves are assumed to be controlled at baseline; therefore, the baseline emissions estimates are based on the following counts: A,33; B,11; C,33.

[e]All open-ended lines are considered together with a single emission factor; 100 percent controlled at baseline.

[f]Seventy-five percent of sampling connections are assumed to be controlled at baseline; therefore, the baseline emissions estimates are based on the following counts: A,7; B,26; C,80.

Table C-10. EQUIPMENT INVENTORIES AND EMISSION ESTIMATES FOR FUGITIVE VOC EMISSION SOURCES IN POLYMERS AND RESINS PLANTS

Equipment Component[a]	Polymers and Resins Plant Data Type of Plant											Assumed For Polymers And Resins Analysis
	A:PP-LP	B:LDPE-HP	C:LDPE-LP	D:HDPE-GP	E:HDPE-GP	F:HDPE-LP	G:HDPE-LP	H:HDPE-LP	I:PS-Cont	J:PET-DMT	K:PET-TPA	
Pump Seals												
Light Liquid Service	40[b]	28[b]	3[b]	7	1[d]	2[d]	6[d]	67[d]	17[d]	110[d]	11[d]	29
Heavy Liquid Service												30
Valves												
Vapor Service	2,300[c]	1,724	1,500[c]	490[f]	-	293	2,313[c]	-	100[c]	1,500[c]	43[c]	1,450
Light Liquid Service		943				78						402
Heavy Liquid Service		781				215						524
Safety/relief valves												
Vapor Service	11[d,c]	205[e]	104[d,c]	19[d,e]	-	7[d,e]	9[d,e]	270[d,e]	-	-	42[e]	524
Light Liquid Service		143[f]				6[g]					42	
Heavy Liquid Service		62				1						
Open-ended lines												
Vapor Service	10[f]	676[f]	100[f]	-	-				unknown	250[f]	-	415[f]
Light Liquid Service		457[f]										
Heavy Liquid Service		219[f]										
Compressor seals	5	8	6	3	2	3	3	3	0	0	0	2
Sampling connections	30[g]	7[g]	30[g]	-	-	3[j]			14[g]	100[g]	-	104[g]
Flanges	7,000	5,580	7,500	485	-	695		2,000	100	4,000	-	2,400
Total Estimated Emissions (kg/hr)[h]												
- baseline:	24.0	22.4	20.1	5.1	0.5	3.5	15.9	5.7	1.8	25.2	0.82	12.1
- uncontrolled:	24.1	34.8	28.7	6.9	0.5	4.0	16.4	5.7	2.7	47.8	0.82	17.2

[a] Equipment components in VOC service only.

[b] Assumed all pumps in light liquid service.

[c] Assumed one-half of valves in vapor service and one-half in light liquid service.

[d] Assumed all safety/relief valves in vapor service.

[e] Seventy-five percent of gas safety/relief valves are assumed to be controlled at baseline; therefore, the baseline emissions estimates are based on the following counts: A,33; B,11; C,33.

[f] All open-ended lines are considered together with a single emission factor; 100 percent controlled at baseline.

[g] Seventy-five percent of sampling connections are assumed to be controlled at baseline; therefore, the baseline emissions estimates are based on the following counts: A,7; B,26; C,80.

[h] Calculated using SOCMI average revised emission factors for each equipment component given in Table C-10.

C.6 REFERENCES FOR APPENDIX C

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APPENDIX D: EMISSION MEASUREMENT AND PERFORMANCE TEST METHODS
I. Process VOC Sources

APPENDIX D: EMISSION MEASUREMENT AND PERFORMANCE TEST METHODS
I.   Process VOC Sources

I-D.1   EMISSION MEASUREMENT
I-D.1.1   Introduction

A new source performance standard for process sources in the
polymers and resins manufacturing industry could be in three different
formats.   A regulation could be based on emission concentration,
emission rate, or percentage emission reduction.   The purpose of this
appendix is to discuss and recommend measurement methods acceptable
for determination of VOC concentration and emission flow rates, and
procedures for calculation of emission reduction efficiency.

I-D.1.2   VOC Concentration Measurement

Numerous methods exist for the measurement of organic emissions.
Among these methods are gas chromatograph (GC, draft Method 18),
direct flame ionization detection (FID, Method 25A), and EPA Reference
Method 25 (EPA 25)-Determination of Total Gaseous Nonmethane Organic
Emissions as Carbon.   Each method has advantages and disadvantages.
Of the three procedures, GC has the distinct advantage of identifying
and quantifying the individual compounds present.   Disadvantages are
that GC systems are expensive and determination of the column required
and analysis of samples can be time consuming.

The FID technique is the simplest procedure.   However, the FID
responds differently to various organic compounds and can yield highly
biased results depending upon the compounds involved.   Another dis-
advantage of the FID is that a separate methane measurement is required
to determine nonmethane organics.   The direct FID procedure does not
identify or quantify individual compounds.

Method 25 sampling and analysis provides a single nonmethane
organic measurement on a carbon basis; this is convenient for establishing
control device efficiencies on a consistent basis.   However, EPA 25

does not provide any qualitative or quantitative information on individual compounds present.

I-D.1.3 Emission Test Experience

The EPA conducted an emission test(1) using these three methods at a polypropylene production facility. This facility was equipped with an incinerator-heat recovery boiler system that was fired with a process waste liquid for primary fuel. A gaseous vent stream was directed to this unit for emission control. The results of testing show that Method 25 resulted in the highest results for outlet concentration and lowest emission reduction efficiency. It is believed that the CO_2 separation column specified in Method 25 was not removing completely the high levels of CO_2 in the exhaust samples prior to oxidation of the VOC fractions. This incomplete separation would yield high results. The results of Method 18 and Method 25A testing are similar. No difficulties were experienced in the performance of Methods 25A or 18.

I-D.2 RECOMMENDED TEST METHODS

The EPA Method 18 is recommended as the test procedure for determining the VOC concentration in emissions from polymers and resins facilities. If a mass flow rate is required, this result can be multiplied by appropriate molecular weights to obtain mass concentrations. EPA Method 2 can be used to measure exhaust flow rates so that VOC mass rates can be calculated. EPA Method 2A, process flow instrumentation or, if appropriate, material balances can be used to calculate inlet vapor flow rates which, when combined with inlet VOC concentration, will allow calculation of the emission reduction efficiency of the control device.

The cost of performing an emission test will vary depending on the format of a regulation. If it is assumed that the emission reduction efficiency must be measured, then a test is estimated to cost from $10,000 to $15,000 per source.

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II. Fugitive VOC Sources

II-D.1 Emission Measurement Methods

II-D.1.1 General Background

A test method was not available when EPA began the development of control technique guidelines, new source performance standards, and hazardous pollutant standards for fugitive volatile organic compounds from industrial categories such as petroleum refineries, synthetic organic chemical manufacturing, and other types of processes that handle organic materials.

During the development and selection of a test method, EPA reviewed the available methods for measurement of fugitive leaks with emphasis on procedures that would provide data on emission rates from each source. To measure emission rates, each individual piece of equipment must be enclosed in a temporary cover for emission containment. After containment, the leak rate can be determined using concentration change and flow measurements. This procedure has been used in several studies[1,2] and has been demonstrated to be a feasible method for research purposes. It was not selected for this study because direct measurement of emission rates from leaks is a time-consuming and expensive procedure, and is not feasible or practical for routine testing.

Procedures that yield qualitative or semi-quantitative indications of leak rates were then reviewed. There are essentially two alternatives: leak detection by spraying each component leak source with a soap solution and observing whether or not bubbles were formed; and, the use of a portable analyzer to survey for the presence of increased organic compound concentration in the vicinity of a leak source. Visual, audible, or olefactory inspections are too subjective to be used as indicators of leakage in these applications. The use of a portable analyzer was selected as a basis for the method because it would have been difficult to establish a leak definition based on bubble formation rates. Also, the temperature of the component, physical configuration, and relative movement of parts often interfere with bubble formation.

Once the basic detection principle was selected, it was then necessary to define the procedures for use of the portable analyzer.

Prior to performance of the first field test, a procedure was reported that conducted surveys at a distance of 5 cm from the components.(3) This information was used to formulate the test plan for initial testing.(4) In addition, measurements were made at distances of 25 cm and 40 cm on three perpendicular lines around individual sources. Of the three distances, the most repeatable indicator of the presence of a leak was a measurement at 5 cm, with a leak definition concentration of 100 or 1000 ppmv. The localized meteorological conditions affected dispersion significantly at greater distances. Also, it was more difficult to define a leak at greater distances because of the small changes from ambient concentrations observed. Surveys were conducted at 5 cm from the source during the next three facility tests.

The procedure was distributed for comment in a draft control techniques guideline document.(5) Many commentors felt that a measurement distance of 5 cm could not be accurately repeated during screening tests. Since the concentration profile is rapidly changing between 0 and about 10 cm from the source, a small variance from 5 cm could significantly affect the concentration measurement. In response to these comments, the procedures were changed so that measurements were made at the surface of the interface, or essentially 0 cm. This change required that the leak definition level be increased. Additional testing at two refineries and three chemical plants was performed by measuring volatile organic concentrations at the interface surface.

A complication that this change introduces is that a small mass emission rate leak ("pin-hole leak") can be totally captured by the instrument and a high concentration result will be obtained. This has occurred occasionally in EPA tests, and a solution to this problem has not been found.

The calibration basis for the analyzer was evaluated. It was recognized that there are a number of potential vapor stream components and compositions that can be expected. Since all analyzer types do not respond equally to different compounds, it was necessary to establish a reference calibration material. Based on the expected compounds and the limited information available on instrument response factors, hexane was chosen as the reference calibration gas for EPA test programs. At the 5 cm measurement distance, calibrations were conducted at

approximately 100 or 1000 ppmv levels. After the measurement distance was changed, calibrations at 10,000 ppmv levels were required. Commentors pointed out that hexane standards at this concentration were not readily available commercially. Consequently, modifications were incorporated to allow alternate standard preparation procedures or alternate calibration gases in the test method recommended in the Control Techniques Guideline Document for Petroleum Refinery Fugitive Emissions.

Since that time, studies have been completed that measured the response factors for several instrument types[6,7,8]. The results of these studies show that the response factors for methane and hexane are similar enough for the purposes of this method to be used interchangeably. Therefore, in later NSPS, the calibration materials were hexane or methane.

The alternative of specifying a different calibration material for each type stream and normalization factors for each instrument type was not intensively investigated. There are at least four instrument types available that might be used in this procedure, and there are a large number of potential stream compositions possible. The amount of prior knowledge necessary to develop and subsequently use such factors would make the interpretation of results prohibitively complicated. Additionally, based on EPA test results, the measured frequency of leak occurrence in a process unit was not significantly different when the leak definition was based on meter reading using a reference material and when response factors were used to correct meter readings to actual concentrations for comparison to the leak definition. The variation in response factor is not a significant problem because ambient concentrations around leaks are usually much higher than the leak definition and much lower when no leak exists.

An alternative approach to leak detection was evaluated by EPA during field testing.[9,10] The approach used was an area survey, or walkthrough, using a portable analyzer. The unit area was surveyed by walking through the unit positioning the instrument probe within 1 meter of all valves and pumps. The concentration readings were recorded on a portable strip chart recorder. After completion of the walkthrough, the local wind conditions were used with the chart data

to locate the approximate source of any increased ambient concentrations. This procedure was found to yield mixed results. In some cases, the majority of leaks located by individual component testing could be located by walkthrough surveys. In other tests, prevailing dispersion conditions and local elevated ambient concentrations complicated or prevented the interpretation of the results. Additionally, it was not possible to develop a general criteria specifying how much of an ambient increase at a distance of 1 meter is indicative of a 10,000 ppm concentration at the leak source. Because of the potential variability in results from site to site, routine walkthrough surveys were not selected as a reference or alternate test procedure.

II-D.1.2 Emission Testing Experience

During the development of the new source performance standard for fugitive VOC emissions in the synthetic organic manufacturing industry, a screening program using Method 21 was conducted at 24 process units.[11] Several of the process units included in this survey were monomer production and polymerization sections of plants included in the polymers and resins industry category. The instruments used were flame ionization, catalytic oxidation, and, in one case, photoionization. The flame ionization and catalytic oxidation instruments were calibrated with methane standards. The photoionization instrument was calibrated with isobutylene. The response factors for these compounds are similar for use in this application.

II-D.2 CONTINUOUS MONITORING SYSTEMS AND DEVICES

Since the leak determination procedure is not a direct emission measurement technique, there are no continuous monitoring approaches that are directly applicable. Continual surveillance is achieved by repeated monitoring or screening of all affected potential leak sources. A continuous monitoring system or device could serve as an indicator that a leak has developed between inspection intervals. The EPA performed a limited evaluation of fixed-point monitoring systems for their effectiveness in leak detection.[9,12,13] The systems consisted of both remote sensing devices with a central readout and a central analyzer system (gas chromatograph) with remotely collected samples. The results of these tests indicated that fixed point systems were not capable of sensing all leaks that were found by individual component

testing. This is to be expected since these systems are significantly affected by local dispersion conditions and would require either many individual point locations, or very low detection sensitivities in order to achieve similar results to those obtained using an individual component survey.

It is recommended that fixed-point monitoring systems not be required since general specifications cannot be formulated to assure equivalent results, and each installation would have to be evaluated individually.

II-D.3 PERFORMANCE TEST METHOD

The recommended fugitive emission detection procedure is Reference Method 21. This method incorporates the use of a portable analyzer to detect the presence of volatile organic vapors at the surface of the interface where direct leakage to atmosphere could occur. The approach of this technique assumes that if an organic leak exists, there will be an increased vapor concentration in the vicinity of the leak, and that the measured concentration is generally proportional to the mass emission rate of the organic compound.

An additional procedure provided in Reference Method 21 is for the determination of "no detectable emissions." The portable VOC analyzer is used to determine the local ambient VOC concentration in the vicinity of the source to be evaluated, and then a measurement is made at the surface of the potential leak interface. If a concentration change of less than 5 percent of the leak definition is observed, then a "no detectable emissions" condition exists. The definition of 5 percent of the leak definition was selected based on the readability of a meter scale graduated in 2 percent increments from 0 to 100 percent of scale, and not necessarily on the performance of emission sources.

Reference Method 21 does not include a specification of the instrument calibration basis or a definition of a leak in terms of concentration. Based on the results of EPA field tests and laboratory studies, methane or hexane are recommended as the reference calibration bases for fugitive emission sources in the polymers and resins manufacturing industries.

There are at least four types of detection principles currently available in commercial portable instruments. There are flame ionization, catalytic oxidation, infrared absorption (NDIR), and photoionization. Two types (flame ionization and catalytic oxidation) are known to be available in factory mutual certified versions for use in hazardous atmospheres.

The recommended test procedure includes a set of design and operating specifications and evaluation procedures by which an analyzer's performance can be evaluated. These parameters were selected based on the allowable tolerances for data collection, and not on EPA evaluations of the performance of individual instruments. Based on manufacturers' literature specifications and reported test results, commercially available analyzers can meet these requirements.

The estimated purchase cost for an analyzer ranges from about $1,000 to $5,000 depending on the type and optional equipment. The cost of an annual monitoring program per unit, including semiannual instrument tests and reporting is estimated to be from $3,000 to $4,500. This estimate is based on EPA contractor costs experienced during previous test programs. Performance of monitoring by plant personnel may result in lower costs. The above estimates do not include any costs associated with leak repair after detection.

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13. "Emission Test Report: Union Carbide Corp., Torrance, CA." EMB Report No. 78-OCM-12A, November 1980.

APPENDIX E: DETAILED DESIGN AND COST ESTIMATION PROCEDURES

E.1 GENERAL

This appendix consists of a more detailed presentation of the
bases, assumptions, and procedures used to estimate equipment designs
and corresponding capital and operating costs for flares, thermal
incinerators, catalytic incinerators, shell-and-tube condensers, and
piping and ducting. The basis of design and cost estimates are presented
in the following sections: E.2, flares; E.3, thermal incinerators;
E.4, catalytic incinerators; E.5, shell-and-tube condensers; and E.6,
piping and ducting. Sufficient detail was presented in Chapter 8 for
ethylene glycol recovery systems and VOC fugitive emissions control.
The installation cost factors used in each analysis and the annualized
cost factors used in all of the cost analysis are given in Tables 8-2
and 8-3, respectively.

E.2 FLARE DESIGN AND COST ESTIMATION PROCEDURE

Flares are open combustion devices that can be used to effectively
and inexpensively reduce VOC emissions. The polypropylene and polyethylene
industries commonly use flares to control large emergency releases and
some high VOC streams. Elevated flares were costed based upon state-of-
the-art industrial design (one-half of sonic velocity and a minimum of
115 Btu/scf). Flare height and diameter, which are the primary
determinants of capital cost, are dependent on flare flow rate, heating
value, and temperature. Associated piping and ducting from the process
sources to a header and from a header to the flare were conservatively
designed for costing purposes. Operating costs for utilities were
based on industry practice (1 fps purge of waste gas plus natural gas
for continuous flow flare; 80 scfh natural gas per pilot, number of
pilots based on flare tip diameter; 0.4 lb steam/lb hydrocarbon at
maximum smokeless rate).

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E.2.1 __Flare Design Procedure.__ Design of flare systems for the various combinations of waste streams was based on standard flare design equations for diameter and height presented by IT Enviroscience.[1] These equations were simplified to functions of the following waste gas characteristics: volumetric flow rate, lower heating value, temperature, and molecular weight. The diameter expression is based on the equation of flow rate with velocity times cross-sectional area. A minimum commercially available diameter of 2 inches was assumed. The height correlation premise is design of a flare that will not generate a lethal radiative heat level (1500 Btu/ft^2 hr, including solar radiation[2]) at the base of the flare (considering the effect of wind). Heights in 5-foot multiples with a minimum of 30 ft. were used.[3] Natural gas to increase the heating value to 115 Btu/scf was considered necessary to ensure combustion of streams containing no sulfur or toxic materials.[4] For flares with diameters of 24-inches or less, this natural gas was assumed to be premixed with the waste gas and to exit out the stack. For larger flares, a gas ring was assumed if large amounts of gas were required because separate piping to a ring injecting natural gas into the existing waste gas is more economical than increasing the flare stack size for large diameters. The flare height and diameter selection procedure is detailed in Table E-1.

Purge gas also may be required to prevent air intrusion and flashback. A purge velocity requirement of 1 fps was assumed during periods of continuous flow for standard systems without seals.[7] For flares handling only intermittent flows, purge gas requirements were assumed to be negligible according to the industry practice of not purging or perhaps purging before a planned intermittent release.[8] For combined streams with very large turndown ratios (intermittent flow : continuous flow), supplying purge gas to maintain an adequate continuous flow in a large flare (designed for the intermittent flow) can become more expensive than designing a second separate flare for the continuous flare. In such cases, a fluidic seal, which requires a greatly reduced purge rate, was used.

Table E-1. PROCEDURE TO DESIGN STATE-OF-THE-ART (0.5 MACH) ELEVATED STEAM-ASSISTED SMOKELESS FLARES

Item	Value
1. Waste gas flow rate, Q_{wg}, (scfm)[a]	from Chapter 6
2. Lower heating value of waste gas, LHV_{wg} (Btu/scf)	from Chapter 6
3. Temperature of waste gas, T_{wg} (°F)	from Chapter 6
4. Molecular weight of waste gas, MW_{wg} (lb/lb-mole)	from Chapter 6
5. Natural gas flow rate, Q_{ng} (scfm)	0, if $LHV_{wg} \geq 115$; $\dfrac{(115-LHV_{wg})Q_{wg}}{(815)}$, if $LHV_{wg} < 115$ Btu/scf
6. Total flare gas, $Q_{fl.g}$ (scfm)[b]	$Q_{wg} + Q_{ng}$
7. Lower heating value of flare gas, $LHV_{fl.g}$ (Btu/scf)	115, if $Q_{ng} > 0$; LHV_{wg}, if $Q_{ng} = 0$
8. Temperature of flare gas, $T_{fl.g}$ (°F)[c]	70, if $Q_{ng} = 0$ or $T_{wg} = 70$; $\dfrac{[Q_{wg} \times (MW_{wg}) \times T_{wg}] + [(Q_{ng}(1218)]}{(Q_{wg} \times MW_{wg}) + (17.4 \times Q_{ng})}$, if $Q_{ng} > 0$
9. Molecular weight of flare gas, $MW_{fl.g}$ (lb/lb-mole)	$MW_{w.g.}$, if $Q_{ng} = 0$; $= \dfrac{(Q_{wg} \times MW_{wg}) + (17.4 \times Q_{ng})}{Q_{wg} + Q_{ng}}$, if $Q_{ng} > 0$
10. Calculated flare diameter, D calc. (in.)[d]	$(1.582 \times 10^{-2}) \sqrt{(Q_{fl.g} \times MW_{fl.g})}$
11. Selected flare diameter, D (in.)[e]	Select standard size for which cost data available: 2, 3, 4, 6, 8, 10, 12, 14, 16, 18, 20, 24, or 60.
12. Actual exit velocity, V_e (fps)[f]	$(5.766 \times 10^{-3}) \times Q_{fl.g} \times (T_{fl.g.} + 46) / D^2$
13. Flare tip pressure drop, p (in. H_2O)[g]	$(1.818 \times 10^{-4}) \times (V_e)^2$
14. Flame angle, θ[h]	$TAN^{-1}(88.2/V_e)$
15. Calculated flare height, $H_{calc.}$ (ft)[i]	$(2.185 \times 10^{-2})\sqrt{Q_{fl.g} \times LHV_{fl.g}}\ -[(6.05 \times 10^{-3}) \times D \times V_e \times COS\theta]$
16. Selected flare height, H (ft)[j]	Select next large multiple of 5 with a minimum of 30.
17. Safe pipe length, l (ft)[j]	$\sqrt{[(4.095 \times 10^{-3}) \times Q_{fl.g} \times LHV_{fl.g}] - H^2}$

[a]Standard conditions for flare design and cost calculations: 70°F and 1 atm.

[b]Auxiliary natural gas requirement assumes 115 Btu/scf minimum lower heating value necessary to ensure combustion (Reference 4) and lower heating value of natural gas of 930 Btu/scf at 70°F (1000 Btu/scf at 32°F). From an energy balance

$$(Q_{ng} \times LHV_{ng}) + (Q_{wg} \times LHV_{wg}) = (Q_{ng} + Q_{wg}) \times (115 \text{ Btu/scf})$$

$$Q_{ng} \text{ (scfm)} = \left[\frac{115 - LHV_{wg} \text{ (Btu/scf)}}{LHV_{ng} \text{ (Btu/scf)} - 115} \right] \times Q_{wg} \text{ (scfm)}$$

[c]Temperature of mixture approximated by assuming uniform specific heats per unit mass.

[d]From Enviroscience (Reference 1, Appendix A)

$$D(in) = \left[\frac{(2.72 \times 10^{-3}) \times Q_{fl.g} \text{ (lb/hr)} \times \sqrt{(T_{wg} \text{ (°F)} + 460)/MW_{fl.g}}}{\sqrt{\Delta p \text{ (in. } H_2O)}} \right]^{\frac{1}{2}}$$

Using an approximation for sonic velocity, c, given by Straitz (Reference 5) that by comparison with the relationship given in Reference 6 implies that the ratio of specific heats, c_p/c_v is about 1.15, which is reasonable for the compounds and temperatures of interest, and using an exit velocity of one-half sonic velocity.

$$V_e \text{(fps)} = 0.5 \times c = 125 \times \sqrt{\frac{(T_{wg} \text{ (°F)} + 460)}{MW_{fl.g}}}$$

From the Environscience equation for flame angle,

$$p \text{ (in. } H_2O) = 55 [V_e \text{ (fps)}/550]^2 = (1.818 \times 10^{-4}) V_e^2$$

Substituting,

$$D \text{ (in.)} = \left[\frac{(2.72 \times 10^{-3}) \times Q_{fl.g} \text{ (scf/min)} \times MW_{fl.g} \text{ (lb/lb mole)} \times (60 \text{ min/hr}) \times (V_e/125)}{(387 \text{ scf at 70°F/lb-mole}) \times \sqrt{(1.818 \times 10^{-4}) V_e^2}} \right]^{\frac{1}{2}}$$

Which simplifies to the relationship given.

[e]Select next larger standard size than calculated diameter unless calculated diameter is within 10% of interval between next smaller and next larger size, if so, select next smaller standard size.

[f]Actual exit velocity V (fps) = Q (acfs)/A(ft^2)

$$= Q \text{ (scfm)} \times \frac{(T_{fl.g} \text{ (°F)} + 460) \text{ acf}}{(70°F + 460) \text{ scf}} \times \frac{\text{min}}{60 \text{ sec}} \div \frac{\pi}{4} \times [D(in.)]^2 \times \frac{ft^2}{144 \text{ in}^2}$$

Footnotes for Table E-1

gFrom p (in. H_2O) = 35 $[V_e$ (fps)/550$]^2$, (as in c.)

hFrom Enviroscience (Reference 1, Appendix A) assuming wind velocity, V_w, of 60 mph

$$\theta = TAN^{-1}\left[\frac{1.47 \text{ (fps/mph)} \times V_w \text{ (mph)}}{550\sqrt{\Delta p/35}}\right] = TAN^{-1}\left[\frac{1.47 \text{ (60)}}{V_e}\right]$$

iFrom Enviroscience (Reference 1, Appendix A), assuming flame emissivity, ϵ, of 0.12, flame radiation intensity, I, of 1200 Btu/hr ft^2

$$H = \sqrt{\frac{Q_{fl.g} \text{ (lb/hr)} \times LHV_{fl.g} \text{ (Btu/lb)} \times \epsilon}{(12.56) I}} \quad -3.33 \text{ D } (p/35) \text{ COS } \theta$$

$$= \sqrt{\frac{[Q_{fl.g} \text{(scf/min)} \times LHV_{fl.g} \text{ (Btu/scf)} \times 60 \text{ min/hr} \times 0.12]}{4\pi (1200 \text{ Btu/hr} - ft^2)}} \quad -3.33 \times D \times (V_e/550) \text{ COS } \theta$$

jSafe pipe length is the pipe length necessary to reach the horizontal distance from the flare where the flame radiation intensity, I, is reduced to 440 Btu/hr-ft^2 including solar radiation (300 Btu/hr-ft^2), a safe working level, (Reference 12):,

$$L = \sqrt{(r^2 - H^2)}$$

$$= \sqrt{\frac{[Q_{fl.g} \text{ (scfm)} \times LHV_{fl.g} \text{ (Btu/scf)} \times 60 \text{ min/hr} \times 0.12]}{12.56 (140 \text{ Btu/hr-}ft^2)} - H^2}$$

Natural gas was also assumed at a rate of 80 scfh per pilot flame to ensure ignition and combustion. The number of pilots was based on diameter according to available commercial equipment.[9]

Steam was added to produce smokeless combustion through a combined mixing and quenching effect. A steam ring at the flare tip was used to add steam at a rate of 0.4 lb steam/lb of hydrocarbons (VOC plus methane and ethane) in the continuous stream (or the intermittent stream if no continuous flow was present).[10] Availability and deliverability of this quantity of steam was assumed.

Piping (for flows less than 700 scfm) or ducting (for flows equal to or greater than 700 scfm) was designed from the process sources to a header combining the streams and from the header to the base of the flare. Since it is usual industry practice, adequate pressure (approximately 3 to 4 psig) was assumed available to transport all waste gas streams without use of a compressor or fan. The source legs from the various sources to the flare header were assumed to be 70 feet in length,[11] while the length of pipelines to the flare was based on the horizontal distance required to provide a tolerable and safe radiation level for continuous working (440 Btu/hr-ft^2, including solar radiation[9]). Piping and ducting were selected and costed as outlined in Section E.6.

E.2.2 Flare Cost Estimation Procedure. Flare purchase costs were based on costs for diameters from 2 to 24 inches and heights from 20 to 200 feet provided by National Air Oil Burner, Inc., (NAO) during November 1982 and presented in Table E-2.[9] A cost was also provided for one additional case of 60 inch diameter and 40 feet height.[10] These costs are October 1982 prices of self-supporting flares without ladders and platforms for heights of 40 feet and less and of guyed flares with ladders and platforms for heights of 50 feet and greater. Flare purchase costs were estimated for the various regulatory alternatives by either choosing the value provided for the required height and diameter or using two correlations developed from the NAO data for purchase cost as a function of height and diameter. (One correlation for heights of 40 feet and less, i.e., self-supporting flares and one for heights of 50 feet and greater, i.e., guyed flares.) An installation factor of 2.1 (see Table 8-2) was used to estimate installed flare costs. Installed costs were put on a June 1980 basis using the following

Table E-2. FLARE BUDGET PURCHASE COST ESTIMATES PROVIDED BY
NATIONAL AIR OIL BURNER, INC., IN OCTOBER 1982 DOLLARS[a]

Tip Diameter	Flow Rate (lb/hr)	Flare Height (ft)			Purchase Price		
		1000 Btu/scf	1500 Btu/scf	2000 Btu/scf	1000 Btu/scf	1500 Btu/scf	2000 Btu/scf
2	2,800	20	20	20	$6,693	$6,693	$6,693
3	6,300	20	20	25	7,560	7,560	8,208
4	11,200	20	25	35	8,097	8,692	9,956
6	25,200	30	40	50	11,565	13,590	23,862
8	44,800	40	55	65	13,962	25,564	29,007
10	70,000	50	70	85	27,786	34,029	38,049
12	100,800	60	80	100	31,653	41,061	47,277
14	137,200	70	95	115	35,680	49,643	57,458
16	179,200	80	110	130	44,325	57,845	64,777
18	226,800	90	120	150	58,362	71,944	86,064
20	280,000	100	135	165	64,726	80,755	95,704
24	403,200	120	160	200	84,228	105,082	125,827

[a]Price for heights of 40 feet or less are for self-supporting flares without ladders or platforms. Prices for heights of 50 feet and greater are for guyed flares with ladders and platforms.

Chemical Engineering Plant Cost Indices: the overall index for flares; the pipes, valves, and fittings index for piping; and the fabricated equipment index for ducting. Annualized costs were calculated using the factors presented in Table 8-3. The flare cost estimation procedure is presented in Table E-3.

E.3 THERMAL INCINERATOR DESIGN AND COST ESTIMATION PROCEDURE

Thermal incinerator designs for costing purposes were based on heat and mass balances for combustion of the waste gas and any required auxiliary fuel, considering requirements of total combustion air. Costs of associated piping, ducting, fans, and stacks were also estimated.

E.3.1 Thermal Incinerator Design Procedure

Designs of thermal incineration systems for the various combinations of waste gas streams were developed using a procedure based on heat and mass balances and the characteristics of the waste gas in conjunction with some engineering design assumptions. In order to ensure a 98 percent VOC destruction efficiency, thermal incinerators were designed to maintain a 0.75 second residence time at 870°C (1600°F).[12] The design procedure is outlined in this section.

Streams with low heat contents, which require auxiliary fuel to ensure combustion and sometimes require air dilution or fuel enrichment to prevent an explosive hazard, are often able to utilize recovered waste heat by preheating inlet air, fuel, and perhaps, waste gas. The design considerations for such streams are noted in the following discussion , but the combustion calculations, etc. are not detailed because all combined streams to thermal incinerators for polymers and resins regulatory alternatives had sufficient waste gas heating values to combust at 870°C (1600°F) without preheating the input streams. Therefore, only the design procedure for high heat content streams, independently able to sustain combustion at 870°C (1600°F), is detailed in this section.

The first step in the design procedure was to calculate the physical and chemical characteristics affecting combustion of the waste gas stream from the model plant characteristics given in Chapter 6, using Table E-4. In order to prevent an explosion hazard and satisfy insurance requirements, dilution air was added to any individual or combined waste stream with both a lower heating value between 13 and

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Item	Value
(1) Flare purchase cost, C'_{fl} (Oct. 1982$)	Select from Table E-2 if value given or use equations: $(3905.7) \pm 35.054)$ H x D + (900.36) D - (126.08)D^2, for 20 H 40 ft. $(6275.6) \pm (224.10)$ H + (12.782) H x D + (24.856)D^2, for 50 H 200 ft.
(2) Flare installed cost, C_{fl}[a] (Oct. 1982 $)	C'_{fl} x 2.1
(3) Total installed piping costs, C_p (Aug. 1978 $)	Method of Appendix E.6
(4) Total installed ducting costs, C_d (Dec. 1977 $)	Method of Appendix E.6
(5) June 1980 Installed costs	
(a) Piping[b]	C_p x 1.206
(b) Ducting[c]	C_d x 1.288
(c) Flare[d]	C_{fl} x 0.818
(d) Total flare system cost, C_{sys}	$(C_p + C_d + C_{fl})$

Annualized Costs[e]

Item	Value
(1) Operating labor, C_1	620 hr/yr x $18/hr = $11,160
(2) Maintenance, C_m	0.05 x C_{sys}
(3) Utilities	
(a) $(Q_{n.g.}^{scfh})$ pilot[f]	30 scfh, for $2 \leq D \leq 8$; 160 scfh, for $10 \leq D \leq 20$; 240 scfh, for D = 24; 320 scfh, for D = 60.
(b) $(Q_{n.g.}^{scfh})$ purge[g]	$[(0.3272)(D_{ft}(in))^2 - (Q_{fl.g}^{scfm})_{cont}] \times 60$
(c) Cost natural gas, $C_{n.g.}$[h]	$3,149 (Q_{n.g.}^{scfm})_{aux} + 53.45 [(Q_{n.g.}^{scfh})_{pilot} + (Q_{n.g.}^{scfh})$ purge)]
(d) Cost steam, C_{stm}[i]	$3.296 \left[Q \text{ scfm x MW x} \frac{wt.\% HC}{100} \right]_{cont. fl.g}$
(5) Taxes, admin. & insurance	C_{sys} x 0.04
(6) Total operating, C_{op}	$C_1 + C_m + C_{n.g.} + C_{stm} + C_{tax}$
(7) Total annualized, C_{tot}	$C_{op} + C_{cr}$

Footnotes for Table E-3

[a]For installation cost factor breakdown, see Table 8-2.

[b]Updated using Chemical Engineering Plant Cost pipes, valves and fittings index from August 1978 (273.1) to June 1980 (329.3).

[c]Updated using Chemical Engineering Plant Cost fabricated equipment index from December 1977 (226.2) to June 1980 (291.3).

[d]Adjusted using Chemical Engineering Plant Cost Index from October 1982 (317 estimated) to June 1980 (259.2).

[e]For annualized cost factors, see Table 8-3.

[f]Based on vendor information for pilots without energy conservation (Reference 9).

[g]Ensures continuous flow of at least 1 fps for flare with any continuous flow:

$$\frac{\pi}{4} \left\{ (D(in.))^2 \times \frac{ft^2}{144\ in^2} \times (1\ fps) \times (60\ sec/min) - [Q_{fl.g}\ (scfm)]_{cont} \right\} 60\ min/hr$$

[h]$\left\{ Q_{ng}(scfm) \times 60\ min/hr \times 8600\ hr/yr + [Q_{pilot}(scfh) + Q_{purge}(scfh)] \right\} \times 8760\ hr/y$

$$\times \frac{520°R}{530°F} \frac{scf\ at\ 60°F}{scf\ at\ 70°R} \times \frac{1,040\ Btu\ (HHV)}{scf\ at\ 60°F} \times \frac{\$5.98}{(10^6\ Btu\ (HHV))} \times \frac{(10^6\ Btu)}{10^6\ Btu}$$

[i]Assumes steam at 0.4 lb/lb of hydrocarbon at maximum continuous flaring rate for 8600 hr/yr:

$$Q_{cont}\ (scfm) \times MW_{cont} \times \left(\frac{wt.\%\ HC}{100\%}\right)_{cont.} \times 8600\ hr/yr \times 60\ min/hr$$

$$\times (lb\text{-}mole/387\ scf\ at\ 70°F) \times (0.4\ lb\ steam/lb\ HC) \times \frac{(1000\ lb\ steam)}{1000\ (lb\ steam)}$$

$$\times \$6.18/(1000\ lb\ steam)$$

Table E-1. WORKSHEET FOR CALCULATION OF WASTE GAS CHARACTERISTICS (molecular weight, molecular formula, lower heating value in Btu/scf)

emission factor: em.f. = _____ = kg VOC/Mg product; model plant capacity: CAP = _____ Gg/yr

Compound Name	Chemical Formula[a]	Molecular Weight (MW)	lb-moles/100 lb w.g. (= 100 lb ÷ MW w.g.)	lb-molesC/100 lb (= A × lb-moles/100 lb w.g.)	lb-molesH/100 lb (= B × lb-moles/100 lb w.g.)	lb-molesN/100 lb (= C × lb-moles/100 lb w.g.)	lb-molesO/100 lb (= D × lb-moles/100 lbw w.g.)	Lower Heating Value (LHV) in Btu/lb[b]	Btu/lb w.g.
hydrogen	H_2	~2.016			x2 =			51,571.4	
methane	CH_4	16.04		x1 =	x4 =			21,502	
nitrogen	N_2	28.01				x2 =		0	
ethylene (ethene)	C_2H_4	28.05		x2 =	x4 =			20,276	
air	$(0.79N_2, 0.21 O_2)$ 29.0 Air:					x1.58=	x0.42=		
ethane	C_2H_6	30.07		x2 =	x6 =			20,416	
methyl alcohol (methanol)	CH_3OH	32.04		x1 =	x4 =		x1 =	9,080c	
C_2–C_5 hydrocarbons	$(C_{2.5}H_5)$	35(ave)		x2.5=	x5 =			20,000d	
C_2–C_7 hydrocarbons	$(C_{2.7}H_{6.4})$	39(ave)		x2.7=	x6.4=			19,900d	
propylene (propene)	C_3H_6	42.08		x3 =	x6 =			19,683	
propane	C_3H_8	44.09		x3 =	x8 =			19,929	
C_2–C_5 hydrocarbons	$(C_{3.5}H_8)$	50(ave)		x3.5 =	x8 =			19,800d	
n-butene	C_4H_8	56.10		x4 =	x8 =			19,477	
n-butane	C_4H_{10}	58.12		x4 =	x10 =			19,665	
isobutane (2-methyl-propane)	C_4H_{10}	58.12		x4 =	x10 =			19,593	
isopropyl alcohol (2-propanol)	C_3H_7OH	60.11		x3 =	x8 =		x1 =	13,500c	
C_3–C_8 hydrocarbons	(C_5H_{12})	72(ave)		x5 =	x12 =			19,400d	

$$Btu/lb_{w.g.} \;(= lb/100\ lb_{w.g.}) \times \; LHV \times \frac{(100\ lb_{w.g.})}{100(lb_{w.g.})} \times$$

lb/100 lb w.g. (= wt. % × 100)

Table E-4. WORKSHEET FOR CALCULATION OF WASTE GAS CHARACTERISTICS (molecular weight, molecular formula, lower heating value in Btu/scf) (Continued)

I.D.: _____ emission factor: em.f. = _____ kg VOC/Mg product; model plant capacity: CAP = _____ Gg/yr

lb/100 lb$_{w.g.}$ ($= $ wt. % $\times \frac{1}{100}$) Compound Name	Chemical Formula[a]	Molecular Weight (MW)	lb-moles/ 100 lb w.g. ($\frac{lb/}{100\ lb} \div \frac{w.g.}{MW}$)	lb-molesC/ 100 lb ($= A$) w.g. lb-moles/ (100 lb$_{w.g.}$)	lb-molesH/ 100 lb ($= B$) w.g. lb-moles/ (100 lb$_{w.g.}$)	lb-molesN/ 100 lb ($= C$) w.g. lb-moles/ (100 lb$_{w.g.}$)	lb-moles0/ 100 lb ($= D$) w.g. lb-moles/ (100 lb$_{w.g.}$)	Lower Heating Value (LHV) in Btu/lb[b]	Btu/lb$_{w.g.}$ ($=$ lb/100 lb$_{w.g.}$) $\times$ LHV $\times$ $\frac{(100\ lb_{w.g.})}{100(lb\frac{w.g.}{w.g.})}$
cyclohexane C_6H_{12}		84.16		x6 =	x12 =			18,84b	
n-hexane C_6H_{14}		86.17		x6 =	x14 =			19,391	
$C_{10}-C_{12}$ ($C_{10.3}H_{20.4}$)		144(ave)		x10.3=	x20.4=			19,000e	
hydro-carbons									

wt. % VOK = __ (hydrocarbons except methane and ethane)

$\sum \frac{lb-moles}{100\ lb_{w.g.}}$: C: _____ H: _____ N: _____ O: $\sum = \frac{Btu}{lb\ w.g.} = LHV_{w.g.}$ $\frac{Btu}{lb}$

$M.W._{w.g.} = 100/\sum$

$Q_{w.g.} = \frac{(25.64) \times em.t. \times CAP}{wt. \%\ VOC\ nonmethane}$ $\frac{lb}{hr}$ or ethene

$Q^{scfm}_{w.g.} = Q_{w.g.} \times \frac{5.98}{MW_{w.g.}}$ scfm@32°F $T^{in}_{w.g.} =$ ___°F

$LHV^{scf}_{w.g.} = \sum \times \frac{MW_{w.g.}\ (lb/lb-mole)}{359scf/lb-mole} = \frac{Btu}{scf@32°F}$

[a] Parentheses indicate an approximate formula for a given molecular weight.
[b] Lower heating values taken from American Petroleum Institute Research Project 44, except as noted otherwise.
[c] Calculated from heats of formation.
[d] Estimated based on approximate heat of formation for assumed composition.
[e] Assumed from extrapolation to approximate composition.

50 Btu/scf at 0°C (32°F) (about 25 and 100 percent of the lower explosive limit) and an oxygen concentration of 12 percent or greater by volume. Dilution air was added to reduce the lower heating value of the stream to below 13 Btu/scf. (Adding dilution air is a more conservative assumption than the alternative of adding natural gas and is probably more realistic as other streams often have enough heat content to sustain the combustion of the combined stream for the regulatory alternative.)

The combustion products were then calculated using Table E-5 assuming 18 percent excess air for required combustion air, but 0 percent excess air for oxygen in the waste gas, i.e., oxygen thoroughly mixed with VOC in waste gas. The procedure would include a calculation of auxiliary fuel requirements for streams (usually with heating values less than 60 Btu/scf) unable to achieve stable combustion at 870°C (1600°F) or greater. Natural gas was assumed as the auxiliary fuel as it was noted by vendors as the primary fuel now being used by industry. Natural gas requirements would be calculated using a heat and mass balance assuming a 10 percent heat loss in the incinerator. Minimum auxiliary fuel requirements for low heating content streams would be set at 5 Btu/scf to ensure flame stability.[13]

The design procedure for streams able to maintain combustion at 870°C (1600°F) is presented in Table E-6. Fuel was added for flame stability in amounts that provided as much as 13 percent of the lower heating value of the waste gas for streams with heating values of 650 Btu/scf or less. For streams containing more than 650 Btu/scf, flame stability fuel requirements were assumed to be zero since coke oven gas, which sustains a stable flame, contains only about 590 Btu/scf. In order to prevent damage to incinerator construction materials, quench air was added to reduce the combustion temperature to below the incinerator design temperature of 980°C (1800°F) for the cost curve given by IT Enviroscience.[14]

The total flue gas was then calculated by summing the products of combustion of the waste gas and natural gas along with the dilution air. The required combustion chamber volume was then calculated for a residence time of 0.75 sec, conservatively oversizing by 5 percent according to standard industry practice.[15] The design procedure
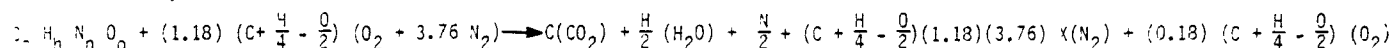
TABLE E-5. GENERALIZED WASTE GAS COMBUSTION CALCULATIONS

Basis: per 100 lb waste gas (w.g.)   Stream I.D.: _____

Assumptions: (1) 18 percent excess air (1 O_2:3.76 N_2 by volume or by mole);

              (2) known waste gas composition of C, H, N, O, (in lb-moles of atom per 100 lb w.g.);

              (3) known amount of AIR (in lb-moles/100 lb w.g.) in waste gas stream (also included in H & O compositions);

              (4) oxygen in waste gas (in air or hydrocarbons) thoroughly mixed with VOC so only stoichiometric requirement.

Reaction (in dry air):

$$C_c H_h N_n O_o + (1.18)(C + \frac{H}{4} - \frac{O}{2})(O_2 + 3.76 N_2) \longrightarrow C(CO_2) + \frac{H}{2}(H_2O) + \frac{N}{2} + (C + \frac{H}{4} - \frac{O}{2})(1.18)(3.76)(N_2) + (0.18)(C + \frac{H}{4} - \frac{O}{2})(O_2)$$

Products of Combustion in lb/100 lb w.g. (wet):

CO_2: $\frac{44.01 \text{ lb}}{\text{lb-mole}} \times C (\frac{\text{lb-mole } CO_2}{100 \text{ lb w.g.}}) = (44.01) C =$ _____

H_2O: if air required $\frac{18.02 \text{ lb}}{\text{lb-mole}} (\frac{H}{2} + [AIR + (C + \frac{H}{4} - \frac{O}{2})(4.76)(1.18)](0.013) \frac{\text{lb } H_2O}{\text{lb dry air}} a (29 \frac{\text{lb dry air}}{\text{lb-mole}})$
       i.e., $(C + \frac{H}{4}) > \frac{O}{2}$

                 $= (2.118) C + (9.539) H - (1.059) O + (0.377) AIR =$ _____

       if no air required: $\frac{18.02 \text{ lb}}{\text{lb-mole}} (\frac{H}{2}) + (AIR)(0.013) \frac{\text{lb } H_2O}{\text{lb dry air}} a (29 \frac{\text{lb dry air}}{\text{lb-mole}}) = (9.01) H + (0.377) AIR =$ _____
       i.e., $(C + \frac{H}{4}) \leq \frac{O}{2}$

N_2:   if air required $\frac{28.02 \text{ lb}}{\text{lb-mole}} [\frac{N}{2} + (C + \frac{H}{4} - \frac{O}{2})(3.76)(1.18)]$
       i.e., $(C - \frac{H}{4}) > \frac{O}{2}$:

                 $= (124.3) C + (31.08) H + (14.01) N - (62.16) O =$ _____

       if no air required: $\frac{28.02 \text{ lb}}{\text{lb-mole}} (\frac{N}{2}) = (14.01) N$
       i.e., $(C + \frac{H}{4}) \leq \frac{O}{2}$

O_2:   if air required, $\frac{32.00 \text{ lb}}{\text{lb-mole}} (0.18)(C + \frac{H}{4} - \frac{O}{2}) = (5.760) C + (1.44) H - (2.88) O =$ _____
       i.e., $C - \frac{H}{4}) > \frac{O}{4}$

       if no air required: $\frac{32.00 \text{ lb}}{\text{lb-mole}} [\frac{O}{2} - (C + \frac{H}{4})] = (-32.00)C - (8.00) H + (16.00) O =$ _____
       i.e., $(C + \frac{H}{4}) \leq \frac{O}{2}$

                                                                                                         _____

Total: $\frac{\text{prod.}}{\text{w.g.}} = \frac{\text{lb products (wet)}}{100 \text{ lb w.g.}} = \frac{\text{lb}(CO_2 + H_2O + N_2 + O_2)}{100 \text{ lb w.g.}} =$ _____

aPounds of water per pound of dry air at 80°F and 60% relative humidity.

TABLE E-6. PROCEDURE TO DESIGN
THERMAL INCINERATORS COMBUSTING STREAMS
WITH LOWER HEATING VALUES (LHV) GREATER THAN 60 Btu/scf[a]

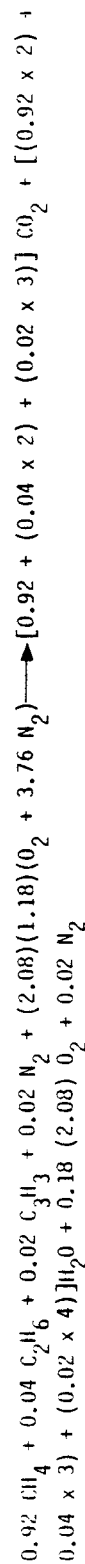| Item	Value
(1) Assign auxiliary fuel (natural gas) for flame stability (% aux)	For LHV ⎯⎯⎯ (% aux)[b] LHV ≤ 75 Btu/scf → 5 75 < LHV ≤ 100 → 0.32(LHV)-19 100 < LHV ≤ 200 → 16-0.03(LHV) 200 < LHV ≤ 375 → 10 375 < LHV ≤ 600 → 18.2-0.022(LHV) 600 < LHV ≤ 650 → 65-0.1(LHV) 650 < LHV → 0[c]
(2) Calculate dilution air to limit combustion temperature to 1300°F (basis of cost estimates)[d]	$W_{air}\left(\dfrac{lb\ air}{100\ lb\ w.g.}\right) = (0.198)\ \left(1+\dfrac{\%aux}{100}\right)LHV$ $-(1800 - T_{w.g.})\ [(5.95 \times 10^{-4})CO_2 + (1.14 \times 10^{-3})\ H_2O$ $+(5.92 \times 10^{-4})N_2 + (5.35 \times 10^{-4}\ O_2)]$ $+(1.65 \times 10^{-4})(\%aux) \times LHV$
(3) Calculate total flue gas volume: (a) waste gas products of combustion:	$Q_{w.g.\ prod.}\ (lb\text{-}moles/100\ lb\ w.g.) = \dfrac{CO_2}{44.01} + \dfrac{H_2O}{18.02}$ $+ \dfrac{N_2}{28.02} + \dfrac{O_2}{32.00}$
(b) auxiliary natural gas products of combustion :	$Q_{n.g.\ prod.}\ (lb\text{-}moles/100\ lb\ w.g.) = (1.66 \times 10^{-7})(\%aux)(LHV\ w.g.)$
(c) dilution air:	$Q_{dil.\ air}\ (lb\text{-}moles/100\ lb\ w.g.) = w_{air}\left(\dfrac{lb\ air}{100\ lb\ w.g.}\right)$ $\div\ 29\ lb\ air/lb\text{-}mole\ air$
(d) total flue gas:[f]	$Q_{f.g.}\ (scfm) = (5.98 \times 10^{-2})\ (Q_{w.g.\ prod.} + Q_{n.g.\ prod.} + Q_{dil.\ air}) \times Q_{w.g.}$
(4) Calculate combustion chamber volume:[g]	$V_{cc}(ft^3) = (6.029 \times 10^{-2})\ Q_{f.g.}\ (scfm)^h$

Footnotes for Table E-6

a Waste gas streams with lower heating values (LHV) of about 60 Btu/scf or greater can attain 1600°F combustion temperatures without additional fuel. Standard conditions for thermal incinerator design procedure are 32°F and 1 atm.
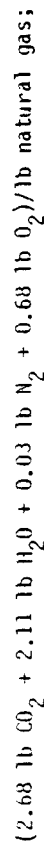
b (% aux) = percentage of LHV per standard cubic foot of waste gas (from Table E-4) to be supplied by auxiliary fuel. Amounts assigned based on a conservative approximation of curve in Figure III-2 of Reference 13.

c No additional fuel requirement for flare stability assumed for LHV 650 Btu/scf since 10% higher than LHV of coke oven gas, which is about 590 Btu/scf.

d $T_{w.g.}$ = inlet temperature of waste gas (from Ch. 6 and Table E-4); CO_2, H_2O, N_2, O_2 = lb. of waste gas combustion products per 100 lb of waste gas from Table E-5. Relationship based on following assumptions: (1) no preheat so inlet temperature of natural gas & dilution air assumed to be 60°F; (2) adequate unrequired oxygen in waste gas and in dilution air to combust small amounts of auxilliary natural gas for flame stability at 18 percent excess air; (3) mean specific heats(C_p) in Btu/lb °F between 60°F and 1800°F of 0.270 for CO_2, 0.517 for H_2O, 0.269 for N_2, 0.243 for O_2, and 0.261 for air; (4) natural gas composition (vol.%) of 92% CH_4, 4% C_2H_6, 2% C_3H_8, and 2 percent N_2 (MW = 17.40; LHV = 20,660 Btu/lb = 1000 Btu/scf). Therefore, for combustion of 1 mole of natural gas on a mole basis:

$0.92\ CH_4 + 0.04\ C_2H_6 + 0.02\ C_3H_3 + 0.02\ N_2 + (2.08)(1.18)(O_2 + 3.76\ N_2) \longrightarrow [0.92 + (0.04 \times 2) + (0.02 \times 3)]\ CO_2 + [(0.92 \times 2) + (0.04 \times 3) + (0.02 \times 4)]H_2O + 0.18\ (2.08)\ O_2 + 0.02\ N_2$

or simplifying and putting on a weight basis:

$(2.68\ lb\ CO_2 + 2.11\ lb\ H_2O + 0.03\ lb\ N_2 + 0.68\ lb\ O_2)/lb$ natural gas;

(5) 4.51 lb O_2 in waste gas and dilution air used for combustion of 1 lb of natural gas; (6) energy balance:

90 $LHV_{w.g.}$ [1+(% aux/100)]= $\sum (W_i c_{pi})$ i waste gas products $(1800-T_{w.g.}) + (1800-60)\{\sum (W_i c_{pi})_{ng}\} \times (\%aux\ LHV_{w.g.})\ LHV_{n.g.} + [Wc_p]air - 4.51(c_p)_0$], where $T_{w.g.}$ = inlet temperature of waste gas, W = weight, c_p = mean specific heats at constant pressure (assuming same for temperature change of $T_{w.g.}$ to 1800°F as for 60°F-100°F, and $LHV_{w.g.}$ generally about 60°F-100°F, and $LHV_{w.g.}$ and $LHV_{n.g.}$ = lower heating values of the waste gas and natural gas, respectively.

e [(1.06 + 2.04 + 0.02 + 0.37) lb-mole natural gas] x [(% aux)$LHV_{w.g.}$] Btu natural gas required/100 lb waste gas ÷ (20,660 Btu n.g./lb n.g. x 17.4 lb n.g./lb-mole n.g.)

f ($Q_{w.g.\ prod.} + Q_{n.g.\ prod.} + Q_{dil\ air}$) lb-moles/(100 lb w.g.) x (100 lb w.g.))/100 lb w.g. x $Q_{w.g.}$ (lb/hr) x (hr/60 min) x 359 scf at 32°F)/lb-mole.

g $q_{f.g.}$ (scfm) x (0.75 sec) x (min/60 sec) x (1300 + 460) °R/492°R x (1.05) design safety factor (Reference 15).

h If V_{cc} ÷ 35./ ft^3 (minimum commercially available size[16]), must add air and redesign.

If V_{cc} ÷ 7238 ft^3 (maximum shop-assembled unit[17]), larger units would require field fabrication; therefore, assume multiple units of equal sizes.

assumed a minimum commercially available size of 1.01 m^3 (35.7 ft^3) based on vendor information[16] and a maximum shop-assembled unit size of 205 m^3 (7,238 ft^3).[17]

The design procedure would allow for pretreating of combustion air, natural gas, and when permitted by insurance guidelines, waste gas using a recuperative heat exchanger in order to reduce the natural gas required to maintain a 870°C (1600°F) combustion temperature. If a plant had a use for it, heat could be recovered. (In fact, a waste heat boiler can be used to generate steam, generally with a net cost savings.)

E.3.2 Thermal Incinerator Cost Estimation Procedure

Thermal incinerator purchase costs for the calculated combustion chamber volume were taken directly from Figure E-1, (Figure A-1 in the IT Enviroscience document, Reference 14). An installation cost factor of 4.0 (see Table 8-2) was used based on the Enviroscience document.[18] The installed cost of one 150-ft. duct to the incinerator and its associated fan and stack were also taken directly from Figure E-2 (Figure IV-15, curve 3 in the IT Enviroscience study.[19] A minimum cost of $70,000 (in December 1979) was assumed for waste gas streams with flows below 500 scfm. The costs of piping or ducting from the process sources to the 150-ft. duct costed above were estimated as for flares. Installed costs were put on a June 1980 basis using the following Chemical Engineering Plant Cost Indices: the overall index for thermal incinerators; the pipes, valves, and fittings index for piping; and the fabricated equipment index for ducts, fans, and stacks. Annualized costs were calculated using the factors in Table 8-3. The electricity required was calculated assuming a 6-inch H_2O pressure drop across the system and a blower efficiency of 60 percent. The cost calculation procedure is given in Table E-7.

E.4 CATALYTIC INCINERATOR DESIGN AND COST ESTIMATION PROCEDURE

Catalytic incinerators are generally cost effective VOC control devices for low concentration streams. The catalyst increases the chemical rate of oxidation allowing the reaction to proceed at a lower energy level (temperature) and thus requiring a smaller oxidation
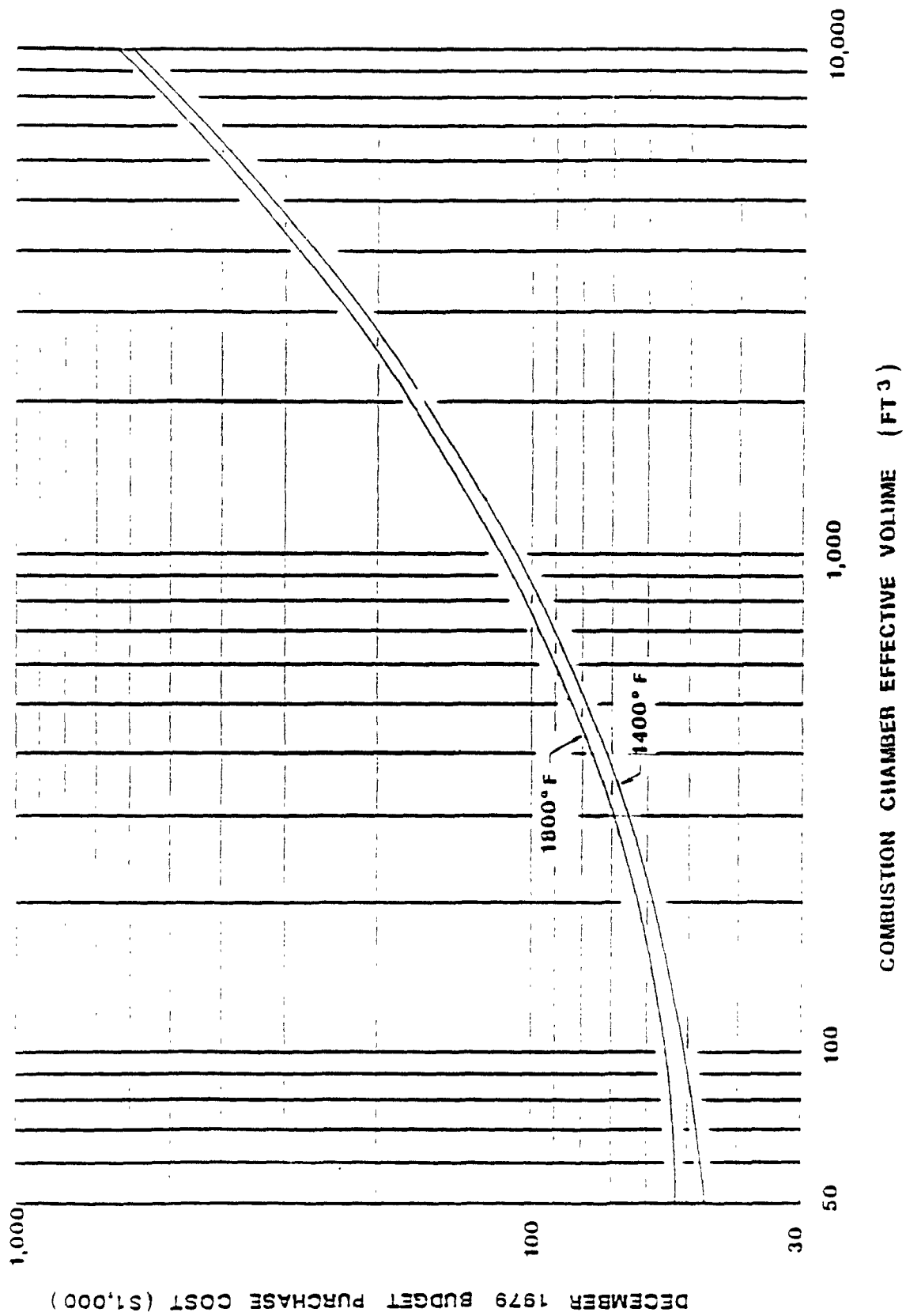
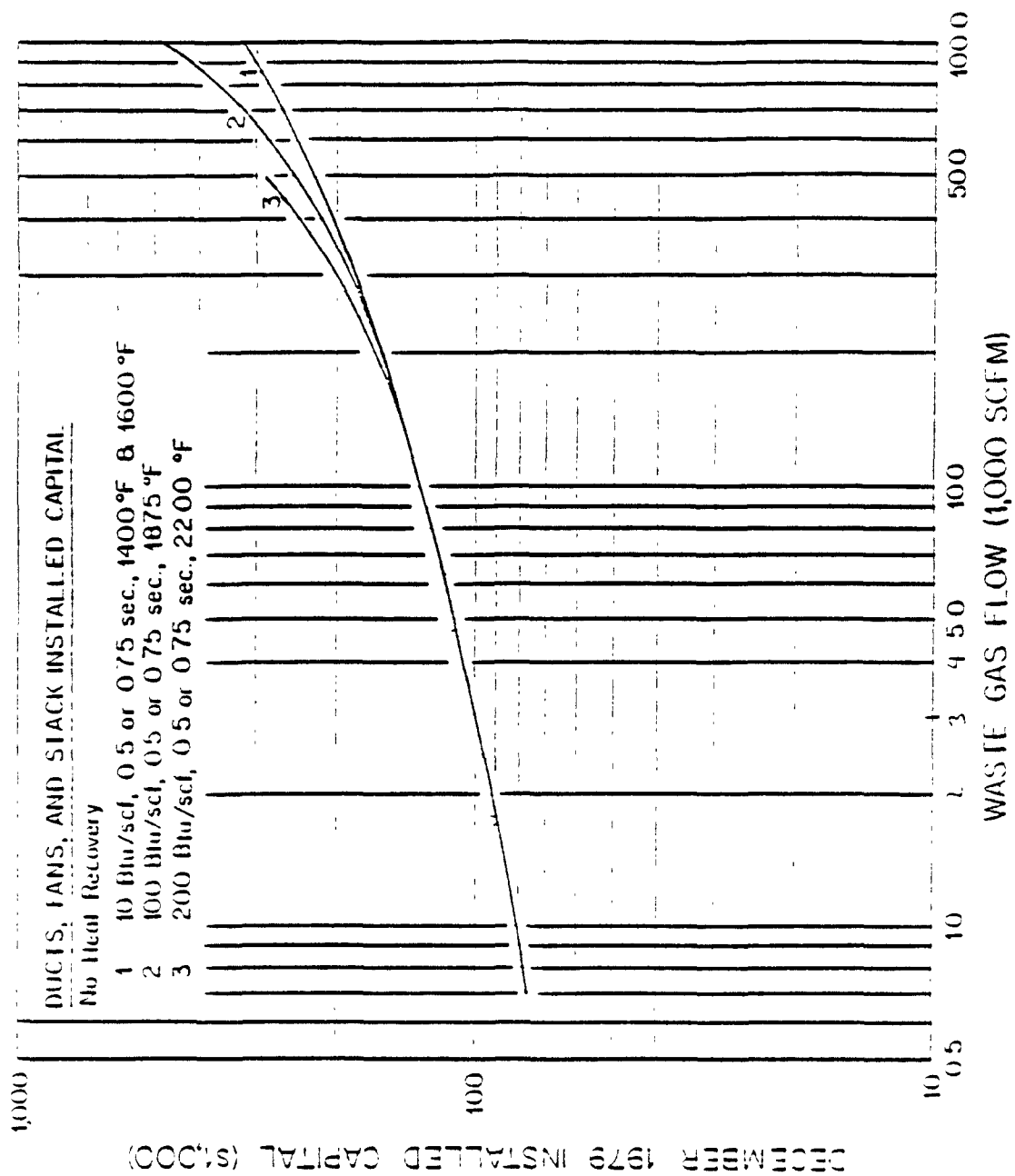Figure E-1. Purchase Costs for Thermal Incinerator Combustion Chambers

Figure E-2. Installed Capital Costs for Inlet Ducts, Waste Gas and Combustion Air Fans, and Stack for Thermal Incinerator Systems with no Heat Recovery

TABLE E-7. CAPITAL AND ANNUAL OPERATING COST ESTIMATES FOR
THERMAL INCINERATORS WITHOUT HEAT RECOVERY

Item	Value
Capital Costs	
Combustion Chamber	
Purchase cost	from Figure E-1 for V_{cc}
Installed cost	purchase cost x 4.0
Installed cost, June 1980[a]	installed cost x 1.047
Piping & Ducting (from sources to main incinerator duct)	
Installed cost	see Section E.7 for $Q_{w.g.}$ (scfm)
Installed cost, June 1980[b]	installed cost x 1.206 for piping installed cost x 1.288 for ducting
Ducts, Fans & Stacks (from main duct to incinerator and from incinerator to atmosphere)	
Installed cost[c]	from Figure E-2 for $Q_{w.g.}$; use $70,000 minimum
Installed cost, June 1980[d]	installed cost x 1.064
Total Installed Cost, June 1980	sum of combustion chamber, piping & ducting, and ducts, fans, & stacks
Annualized Costs[e]	
Operating labor	1200 hr/yr x $18/hr = $21,600
Maintenance material & labor	0.05 x total installed cost
Utilities	
natural gas[f]	(5.639×10^{-4}) (% aux) x $LHV_{w.g.}$
electricity[g]	(0.4955) x $Q_{f.g.}$ (scfm)
Capital recovery[h]	0.1627 x total installed cost
Taxes, administration & insurance	0.04 x total installed cost
Total Annualized Cost	operating labor + maintenance + utilities + capital recovery + taxes, administration & insurance

FOOTNOTES TO TABLE E-7

[a] Updated using <u>Chemical Engineering</u> Plant Cost Index from December 1979 (247.6) to June 1980 (259.2).

[b] Piping updated using <u>Chemical Engineering</u> Plant Cost pipes, valves, and fittings index from August 1978 (273.1) to June 1980 (329.3). Ducting updated using <u>Chemical Engineering</u> Plant Cost fabricated equipment index from December 1977 (226.2) to June 1980 (291.3).

[c] From Figures E-2 for no heat recovery from Enviroscience (Reference 19), which assumed 150-ft of round steel inlet ductwork with four ells, one expansion joint, and one damper with actuator; and costed according to the GARD Manual (Reference 20). Fans were assumed for both waste gas and combustion air using the ratios developed for a "typical hydrocarbon" and various estimated pressure drops and were costed using the Richardson Rapid System (Reference 21). Stack costs were estimated by Enviroscience based on cost data received from one thermal oxidizer vendor.

Although these Enviroscience estimates were developed for lower heating value waste gases using a "typical hydrocarbon" and no dilution to limit combustion temperature, the costs were used directly because Enviroscience found variations in duct, etc., design to cause only small variations in total system cost. Also, since the duct, fan, and stack costs are based on different flow rates (waste gas, combustion air and waste gas, and flue gas, respectively) the costs can not be separated to be adjusted individually.

[d] Updated using <u>Chemical Engineering</u> Plant Cost fabricated equipment index from December 1979 (273.7) to June 1980 (291.3).

[e] Cost factors presented in Table 8-3.

[f] $[(\% \text{ aux}) \times LHV_{w.g.}/LHV_{n.g.}]$ $(1b_{n.g.}/100 \ 1b_{w.g.}) \times Q_{w.g.}$ $(1b/hr) \times$ $(100 \ 1b_{w.g.})/100(1b_{w.g.}) \times (8600 \ hr/yr) \times (1b\text{-mole}/17.4 \ 1b_{n.g.}) \times$ $(379 \ scf \ at \ 60°F/1b\text{-mole}) \times (1040 \ Btu(HHV)/scf \ at \ 60°F) \times \$5.98/10^6$ Btu (HHV) $\times (10^6 \ Btu)/10^6$ (Btu).

[g] Electricty $= (6 \ in. \ H_2O \ pressure \ drop) \times Q_{f.g.}$ $(scfm) \times (8600 \ hrs/yr)$ $\times (0.7457 \ kW/h)p \times (5.204 \ 1b/ft^2/in. \ H_2O) \div [(60 \ sec/min) \times (550 \ ft\text{-}1b/$ $sec/hp) \times (0.6 \ kW \ blower/1 \ kW \ electric)$.

[h] 10 percent interest (before taxes) and 10 yr. life.

chamber, less expensive materials, and much less auxiliary fuel (especially for low concentration streams) than required by a thermal incinerator. The primary determinant of catalytic incinerator capital cost is volumetric flow rate. Annual operating costs are dependent on emission rates, molecular weights, VOC concentration, and temperature. Catalytic incineration in conjunction with a recuperative heat exchanger can reduce overall fuel requirements.

E.4.1 Catalytic Incinerator Design Procedure

The basic equipment components of a catalytic incinerator include a blower, burner, mixing chamber, catalyst bed, an optional heat exchanger, stack, controls, instrumentation, and control panels. The burner is used to preheat the gas to catalyst temperature. There is essentially no fume retention requirement. The preheat temperature is determined by the VOC content of gas, the VOC destruction efficiency, and the type and amount of catalyst required. A sufficient amount of air must be available in the gas or be supplied to the preheater for VOC combustion. (All the gas streams for which catalytic incinerator control system costs were developed are dilute enough in air and therefore require no additional combustion air.) The VOC components contained in the gas streams include ethylene, n-hexane, and other easily oxidizable components. These VOC components have catalytic ignition temperatures below 315°C (600°F). The catalyst bed outlet temperature is determined by gas VOC content. Catalysts can be operated up to a temperature of 700°C (1,300°F). However, continuous use of the catalyst at this high temperature may cause accelerated thermal aging due to recrystallization.

The catalyst bed size required depends upon the type of catalyst used and the VOC destruction efficiency desired. About 1.5 ft^3 of catalyst for 1,000 scfm is required for 90 percent control efficiency and 2.25 ft^3 is required for 98 percent control efficiency. As discussed earlier many factors influence the catalyst life. Typically the catalyst may loose its effectiveness gradually over a period of 2 to 10 years. In this report the catalyst is assumed to be replaced every 3 years.

Heat exchanger requirements are determined by gas inlet temperature and preheater temperature. A minimum practical heat exchanger efficiency

is about 30 percent. Gas temperature, preheater temperature, gas dew point temperature and gas VOC content determine the maximum possible heat exchanger efficiency. A heat exchanger efficiency of 65 percent was assumed for this analysis. The procedure used to calculate fuel requirements is presented in Table E-8. Estimated fuel requirements and costs are based on using natural gas, although either oil (No. 1 or 2) or gas can be used. Fuel requirements are drastically reduced when a heat exchanger is used. Total heat requirements are based on a preheat temperature of 600°F. A stack is used to vent flue gas to the atmosphere.

E.4.2 Catlaytic Incinerator Cost Estimation Procedure

The capital cost of a catalytic incinerator system is usually based on gas volume flow rate at standard conditions. For catalytic incineration, 70°F and 1 afm (0 psig) were taken as standard conditions. The operating costs are determined from the gas flow rate and other conditions such as gas VOC content and temperature. Table E-9 presents the basic gas parameters required for estimating system costs.

As noted earlier, equipment components of a catalytic incineration system include blower, preheater with a burner, mixing chamber, catalyst bed, an optional heat exchanger, stack, controls, and internal ducting including bypass. Calculations for capital cost estimates are based on equipment purchase costs obtained from vendors[22,23,24] and application of direct and indirect cost factors. Table E-10 presents third quarter 1982 purchase costs of catalyst incinerator systems with and without heat exchangers for sizes from 1,000 scfm to 50,000 scfm. The cost data are based on carbon steel for incinerator systems and stainless steel for heat exchangers. The heat exchanger costs are based on 65 percent heat recovery. Catalytic incinerator systems of gas volumes higher than 50,000 scfm can be estimated by considering two equal volume units in the system. The heat exchangers for small size systems would be costly and may not be practical. Table 3-2 presents the direct and indirect installation cost component factors used for estimating capital costs of catalytic incinerator systems. The geometric mean of the two vendor estimates for each flow rate was multiplied by the ratio of total installed costs to equipment purchase

Table E-8. OPERATING PARAMETERS AND FUEL REQUIREMENTS
OF CATALYTIC INCINERATOR SYSTEMS

Item	Source of information or calculation
Waste Gas Parameters	
(1) Flow rate (Q_2), scfm	From Table E-9
(2) Amount of air present in the gas, scfm	0 if the waste gas contains VOC and nitrogen or other inert gas; and [(1 - volume percent VOC) ÷ (volume percent VOC)] if the waste gas contains VOC and air
(3) Amount of air required for combustion at 20% excess, scfm	See footnote a.
(4) Net amount of additional air required (Q_3), scfm	Item (3) - Item (2); and 0 if [Item (3) - Item (2)] is negative
(5) Total amount of gas to be treated (Q_4), scfm	Item (1) + Item (4)
(6) Waste gas Temperature at the inlet of PHR^b, °F	From Table E-9
(7) Waste gas temperature at preheater outlet or catalyst bed inlet, °F	600°F
(8) Temperature rise in the catalyst bed, °F	(25°F/1% LEL) x (%LEL from Table E-9)
(9) Flue gas temperature at catalyst bed outlet °F	Item (7) + Item (8)
(10) Minimum possible temperature of flue gas at PHR outlet, °F	See footnote c.
(11) PHR efficiency at maximum possible heat recovery^d, %	[Item (1) x (Item (7) - 25°F - Item (6))]$_e$ ÷ [Item (5) x (Item (9) - Item (6))]$_e$
(12) PHR design efficiency^f, %	65

Table E-8. OPERATING PARAMETERS AND FUEL REQUIREMENTS
OF CATALYTIC INCINERATOR SYSTEM (concluded)

Item	Source of information or calculation
(13) Waste gas temperature at PHR outlet °F	0.65 [Item (9) - Item (6)] + Item (6)
(14) Amount of heat required by preheater at additional 10% for auxiliary, Btu/min	Item (5) x [Item (7) - Item (13)] x [Gas specific heat[g], Btu/scf, °F] x $(110\%)^h$
(15) Amount of heat required for preheater and auxiliary fuel, 10^6 Btu/h	[Item (14) x 60 minutes/hour] x $(10\%)^h$
(16) Amount of natural gas required per year, 10^3 cfm	[Item (14) x (8,600 x 60 minutes/year)] x 10^{-3} ÷ (1,040 Btu/cfm)

[a] On volume basis (scfm/scfm): 9.53 for methane, 16.68 for ethane, 23.82 for propane, 45.26 for hexane, and 14.3 for ethylene.

[b] Primary heat recovery unit.

[c] Heat exchanger should be designed for at least 50°F above the gas dew point.

[d] The heat exchanger will be designed for 25°F lower than the preheater temperature so as to not cause changes in catalyst bed outlet temperature.

[e] Though, the heat recovery to the temperature level of inlet gas is the maximum heat efficiency possible, in some cases this may not be possible due to gas dew point condition.

[f] Cost estimates are based on 65 percent heat recovery.

[g] Gas specific heat varies with composition and temperature. Used 0.019 Btu/ft^3 °F based on average specific heat of air for calculation purpose.

[h] Auxiliary fuel requirement is assumed to be 10 percent of total.

TABLE E-9. GAS PARAMETERS USED FOR ESTIMATING CAPITAL AND
OPERATING COSTS OF CATALYTIC INCINERATORS[a]

ITEM	VALUE
Stream identification	Identify the vent and the polymer industry from the BID
Stream conditions	
Temperature, $^\circ$F	
Pressure, psig	
VOC content:	
Emission factor, kg/Mg of product	
Weight % of total gas (W_1)	
Mass flow rate, kg/h	(Emission factor, E, kg/Mg) x (Plant production rate, P, Gg/yr) ÷ (8,600 h/yr)
lb/h	(kg/h) x (2.205 lb/kg)
Organic constituents, wt %	
Average mol. wt. (M_1), lbs	
Volume flow (Q_1), scfm	(VOC mass rate, lb/h) ÷ (60 min./h) ÷ (Molecular weight (M_1), lbs/lb mole) = 1.645 (EP/M_1)
Heat content (H_1), Btu/scf	$(174.273)(2.521N_c + N_H)^c$
Total gas:	
Constituents	VOC, air and others
Mass flow rate, lb/h	(VOC rate, lb/h) ÷ (wt% of VOC in gas, W_1)
Molecular weight (M_2)	
Volume flow (Q_2), scfm	Gas mass rate, lb/h) ÷ (60 min/h) ÷ (Gas molecular weight (M_2), lb/lb mole) x (385 ft³/lb mole) = 1.645 (EP/M_2W_1)
Air volume flow rate, scfm	(Total gas flow, scfm) x (volume percent air in total gas)
VOC concentration (A), % of LEL	(100)[(Volume flow of VOC, scfm)[d] ÷ (Volume flow of air, scfm] ÷ LEL[d]
Heat content (H_2), Btu/total scf[e]	

[a] Obtain gas parameters from Chapter 3 of the BID, except those to be calculated.

[b] Calculate using weight percent values of VOC components.

[c] If the VOC heating value is not available calculate it using on heat of combustion values of 14,093 Btu/lb from carbon converted to CO_2 and 51,623 Btu/lb from hydrogen converted to water. N_c and N_H denote number carbon and hydrogen atoms in VOC

[d] Lower explosion levels of ethylene, hexane, methanol, propane and butane are 3.1, 1.32, 7.3 and 2.5, and 1.9 respectively.

[e] Total gas heat content averages 50 Btu/scf at 100 percent LEL.

TABLE E-10 CATALYTIC INCINERATOR VENDOR
COST DATA[22,23]

| Flow (scfm) | Catalytic Incinerator Equipment Purchase Cost($)[a] | | | | | | | | |
| | Without heat recovery | | | With 65 % heat recovery | | |
	Vendor A	Vendor B	Geometric Mean	Vendor A	Vendor B	Geometric Mean
1,000	20,000	40,000	28,300	30,000	90,000	52,000
10,000	150-175,000	125,000	142,300	200-225,000	185,000	198,100
50,000	500,000	480,000	489,900	800-1,000,000	600,000	732,600

[a]Costs in third quarter 1982 dollars.

costs of 1.82 developed for a skid-mounted catalytic incinerator. Actual direct and indirect cost factors depend upon the plant specific conditions and may vary with system sizes.

Since the equipment purchase cost presented in Table E-10 represents the third quarter of 1982, the cost data was adjusted to represent June 1980 by using a cost index multiplying factor of 82.3 percent (based on Chemical Engineering plant cost indices of 259.2 for June 1980 and 315.1 for August 1982). The direct and indirect capital cost factors were applied to the adjusted purchase costs and the resultant estimates of catalytic incinerator installed capital costs as of June 1980 are presented in Figure E-3.

Installed costs of piping, ducts, fans, and stacks were estimated by the same procedure as for thermal incinerators. Installed costs were put on a June 1980 basis using the following Chemical Engineering Plant Cost indices: the overall index for catalytic incinerators; the pipes, valves, and fittings index for piping; and the fabricated equipment index for ducts, fans, and stacks.

Table 8-3 presents cost bases used for annualized cost estimates. The operating labor requirement value is based on conversations with vendors. The capital recovery factor is based on capital recovery period of 10 years and an interest rate (before taxes) of 10 percent. (Actually the current tax regulations allow the control system owners to depreciate the total capital expenditure in the first 5 years.) Fuel cost is the major direct cost item.

The total annual operating costs are calculated using the cost bases shown in Table 8-3 and the fuel requirements calculated in Table E-8. Table E-11 presents a procedure for calculating total annualized cost estimates of catalytic incinerators.

The amount of catalyst required usually depends upon the control efficiency. According to a vendor,[24] typical catalyst costs are about $3,000 per ft^3. Indirect additional costs involved in replacing the catalyst every 3 years are assumed to be 20 percent. Therefore, for 98 percent efficient systems, the annual catalyst replacement costs amount to $2.70/scfm.

Electricity cost calculations are based on pressure drops of 4 in. water for systems with no heat recovery and 10 in. water for
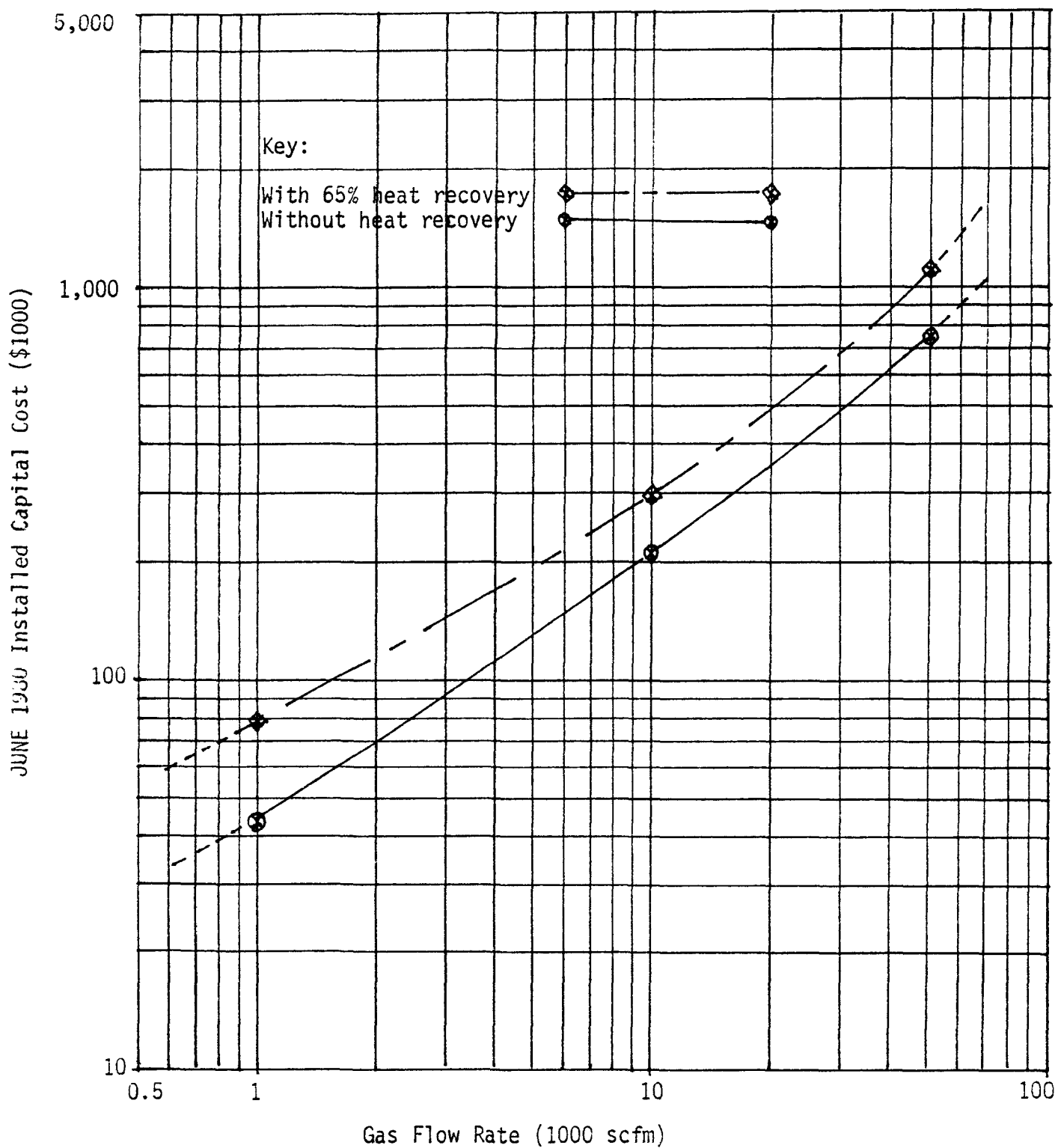
Figure E-3. Installed Capital Costs for Catalytic Incinerators
With and Without Heat Recovery

Table E-11. CALCULATION PROCEDURE FOR ESTIMATION OF ANNUALIZED
COSTS FOR CATALYTIC INCINERATOR SYSTEMS

Cost component	
Direct costs	
Operating labor	$11,200 for systems with no heat recovery; and $16,700 for systems with heat recovery
Maintenance material and labor	(0.05) x (Total installed capital cost, $ from Figure E-3)
Catalyst requirement	$2.7 x (Total gas volume flow$(Q_4)^a$,scfm, item 5 from Table E-8) = ($2.7 x Q_4)
Utilities:	
Fuel (natural gas)	($6.22/$10^3$ft^3) x (Amount of natural gas required, 10^3ft^3, Item 16 of Table E-8)^a
Electricity	($0.335/scfm) x (Total gas volume flow rate (Q_4), scfm, Item 5 from Table E-8) for units with no heat recovery; and ($0.838/scfm) x (Total gas volume flow rate (Q_4), scfm, Item 5 from Table E-8) for units with heat recover;
Indirect costs	
Capital recovery	(0.1627) x (Total installed capital cost, $ from Figure E-3)
Taxes, insurance and administrative charges	(0.04) x (Total installed capital cost, $ from Figure E-3)
Total annualized costs	Sum of total direct costs and total indirect costs

^{a}Total gas flow including waste gas and additional combustion air.

systems with heat recovery, and at 10 percent additional electricity required for instrumentation, controls, and miscellaneous. Therefore, at the conversion rate of 0.0001575 hp per inch of water pressure drop, 65 percent motor efficiency, and $0.049/kWh electricity unit cost, the total annual electricity costs amount to $0.335/scfm for units with no heat recovery (i.e., for 4 in. H_2O pressure drop) and $0.834/scfm for units with heat recovery (i.e., for 10 in. H_2O pressure drop).

E.5 SURFACE CONDENSER DESIGN AND COST ESTIMATION PROCEDURE

This section presents the details of the procedure used for sizing and estimating the costs of condenser systems applied to the gaseous streams from the continuous process polystyrene model plant. Two types of condensers are in use in the industry: surface condensers in which the coolant does not contact the gas or condensate; and contact condensers in which coolant, gas, and condensate are intimately mixed.

Surface condensers were evaluated for the following two streams from the polystyrene model plant: the styrene condenser vent and the styrene recovery unit condenser vent. These streams consist of styrene and steam which are immiscible. The nature of components present in the gas stream determines the method of condensation: isothermal or non-isothermal. The condensation method for streams containing either a pure component or a mixture of two immiscible components is isothermal. In the isothermal condensation of two immiscible components such as styrene and steam, the components condense at the saturation temperature and yield two immiscible liquid condensates. The saturation temperature is reached when the vapor pressure of the components equals the total pressure of the system. The entire amount of vapors can be condensed by isothermal condensation. Once the condensation temperature is determined, the total heat load is calculated and the corresponding heat exchanger system size is estimated. The following procedure and assumptions were used in evaluating the isothermal condensation systems for the two streams containing the immiscible styrene and steam from the continuous polystyrene model plant.

E.5.1 Surface Condenser Design

The condenser system evaluated consists of a shell and tube heat exchanger with the hot fluid in the shell side and the cold fluid in the tube side. The system condensation temperature is determined from the total pressure of the gas and vapor pressure data for styrene and steam. As the vapor pressure data are not readily available, the condensation temperature is estimated by trial-and-error using the Clausius Clapeyron equation which relates the stream pressures to the temperatures. The total pressure of the stream is equal to the vapor pressures of individual components at the condensation temperature. Once the condensation temperature is known, the total heat load of the condenser is determined from the latent heat contents of styrene and steam. The coolant is selected based on the condensation temperature. The condenser system is sized based on the total heat load and the overall heat transfer coefficient which is established from individual heat transfer coefficients of the gas stream and the coolant. An accurate estimate of individual coefficients can be made using such data as viscosity and thermal conductivity of the gas and coolant and the standard sizes of shell and tube systems to be used. For this study no detailed calculations were made to determine the individual and overall heat transfer coefficients. Since the streams under consideration contain low amounts of styrene, the overall heat transfer coefficient is estimated based on published data for steam. Then the total heat transfer area is calculated from the known values of total heat loads and overall heat transfer coefficient using Fourier's general equation. A tabular procedure for calculating heat exchanger size is presented in Table E-12.

E.5.2 Surface Condenser Cost Estimation Procedure

Since the gas volumes of the two streams are very low, the calculated heat transfer areas are also very low (about 10 ft^2). The heat exchanger costs for each stream were obtained for the minimum available size of 20 ft^2 from vendors.[25,26,27,28] An installation factor of 1.39 (See Table 8-2) was used to estimate installed condenser costs. No additional piping was costed since the condenser unit is so small (~1-2 ft. diameter) that it should be able to be installed

Table E-12. PROCEDURE TO CALCULATE HEAT TRANSFER AREA OF AN
ISOTHERMAL CONDENSER SYSTEM

Item	Value
Heat exchanger type	Cocurrent shell and tube heat exchanger with the hot fluid in the shell side and the cold fluid in the tube side
Gas stream condition (including temperature (T_1)°F pressure $(P.)$, psig, and composition)	Obtain from Chapters 3 and 6
Condensation temperature (T_2), °F	a
Total heat load (H), Btu/h	b
Coolant used[c]	water at 85°F, 25 gpm
Temperature rise of coolant, (ΔT), °F	H(Btu/h) ÷ [(25 gpm x 500 lb/h/gpm) x (1 Btu/lb°F)]
Coolant outlet temperature (T_3), °F	85°F + ΔT
Log mean temperature difference (LMTD), °F	$[(T_1-T_3) - (T_2 - 85)] \div \ln [(T_1-T_3)/(T_2-85)]$
Heat transfer coefficient[d] (U)	240 Btu/h ft^2°F
Heat transfer area (A), ft^2	(H)/U(LMTD)

[a] Determine from vapor pressures of styrene and steam and the total
gas pressure. Calculate the styrene vapor pressure using Clausius -
Clapeyron equation:

$$\ln P/P_0 = (\lambda/R) (1/T_0 - 1/T)$$

where P and P_0 are stream pressures, mm Hg; and T and T_0 are
corresponding temperatures, °K; λ is latent heat, cal/g-mole;
and R is universal gas constant = 1.99 cal/g mole°K.

The same equation can be rearranged to eliminate λ and R:

$$\frac{\ln(P/P_0)}{\ln(P_1/P_0)} = \frac{(1/T_0 - 1/T)}{(1/T_0 - 1/T_1)}$$

Using two known values of pressure and temperature, calculate
the pressure for an assumed temperature. Proceed by trial-and-error
until the temperature which gives a total value of styrene pressure
and steam pressure equals to the toal gas pressure.

[b] Total of latent heat of styrene and steam in stream per unit time
(Btu/hr): Calculate the latent heat of styrene from Clausius-Clapeyron
equation using pressure and temperature values of gas and condensation
condition, multiply by lb/hr styrene in stream and add to product of λ
(970.3 Btu/lb steam) and lb/hr steam in stream.

[c] Fixed amount of 25 gpm is used in order to maintain turbulent flow.

[d] Obtained using a clean overall heat transfer coefficient (U_c) of
366 Btu/h ft^2°F and a dirt factor (D.F.) of 0.003. The clean overall
heat transfer coefficient is obtained using weighted averages (86%
steam and 16% styrene) of pure fluid heat transfer coefficients,
1,000 Btu/h ft^2°F for steam and 35 Btu/h ft^2°F for styrene and the
following relationship:

$$\frac{1}{U_d} - \frac{1}{U_c} = D.F.$$

adjacent to the source. Table E-13 presents the estimated total capital and annual operating costs for the condenser system of 20 ft^2 heat transfer area.

E.6 ETHYLENE GLYCOL RECOVERY SYSTEMS DESIGN AND COST ESTIMATION PROCEDURE

This section outlines the basis and procedures used to design and estimate costs of the baseline and regulatory alternative ethylene glycol recovery systems. The resulting costs for the two systems are presented.

E.6.1 Ethylene Glycol Recovery System Design

The equipment selected to comprise the two ethylene glycol recovery systems, as well as the design and operating parameters, were based on information provided by industry sources. The baseline system recovers ethylene glycol (EG) from downstream of the cooling water tower. In this system, ethylene glycol emissions from the polymerization reactor and from the distillation column recovering EG from the esterifier emissions accumulate in the cooling water and are emitted from the cooling tower. The regulatory alternative system recovers ethylene glycol emitted from the polymerization reactor through use of an EG spray condenser and from the esterifier through use of a reflux condenser. The industry information (much of which was considered confidential) was used in conjunction with standard engineering references such as the Chemical Engineers' Handbook[29], and McCabe and Smith's Unit Operations of Chemical Engineering[30], and engineering judgement.[31] These design procedures are summarized in the footnotes to Tables E-14 and E-15 for the baseline and regulatory alternative systems, respectively.

E.6.2 Ethylene Glycol Recovery System Cost Estimation Procedure

The cost estimates and their bases are presented in Table E-14 for the baseline ethylene glycol recovery system and Table E-15 for the regulatory alternative system. The costs of the baseline system were estimated based on the design estimate developed and standard engineering procedures. The costs of the regulatory alternative ethylene glycol spray condenser and recovery system for new plants using the DMT and TPA process were derived from confidential cost data provided by an industry source for a similar system on a larger capacity

Table E-13. CAPITAL AND ANNUAL OPERATING COST ESTIMATES
FOR A 20 ft^2 CONDENSER SYSTEM FOR THE
STREAMS FROM THE CONTINUOUS POLYSTYRENE MODEL PLANT

Item	Value
Control system	Heat exchanger with a maximum capacity of 20 ft^2 heat transfer area
Capital Cost:	
Purchase cost	$2,000
Installed capital cost[a]	2,780
Annualized cost:	
Operating labor[b]	$1,080
Maintenance[c]	140
Utilities: Water[d]	$ 390
Electricity[e]	$ 150
Taxes, insurance, and administration[f]	$ 110
Capital recovery[g]	$ 450
Total annualized cost without recovery credit	$2,300
Total amount of styrene recovered from X lb/hr of styrene	(X lb/hr x 8,600 hr/yr x 99% heat exchanger efficiency x 90% recovery efficiency from the separator) ÷ 2,000 lb/ton = Y tons/year
Annual styrene recovery credit at $0.3575/lb	Y tons x 2,000 lb/ton x $0.3575/lb = $Z
Total annualized cost after credit	($2,300 - $Z)
Cost effectiveness of emission reduction ($/Mg)	($2,300 - $Z)/[X lb/hr x 8,600 hr/yr x 99% heat exchanger (VOC reduction) efficiency)/2,205 lb/Mg]

[a]Purchase cost times installation cost factor of 1.39 (see Table 3-2).

[b]Operating labor cost = 1 hr/wk x 52 wk/yr x 1.15 (with supervision/ without supervision) x $18/hr (including overtime).

[c]Maintenance cost = 0.05 x (installed capital cost).

[d]Water cost = 25 gpm x 60 min/hr x 8,600 hr/yr x 0.1 make-up/total x $0.30/(1,000 gal) x (1,000 gal)/1,000 gal.

[e]Electricity consumption (equations from Reference 29) and cost:
hydraulic horsepower = 50 ft x (1.0 specific gravity) x 25 gpm/3960 = 0.3157 hp

brake horsepower = 0.3157 hp x 745.7 W/hp x 8,600 hr/yr

x kW/1,000W ÷ 0.65 pump efficiency = 3110 kWh/yr

Cost = 3,110 kWh/yr x $0.049/kWh

[f]Taxes insurance, and administration cost = 0.04 x (installed capital cost).

[g]Capital recovery factor = 0.1627, for 10 percent interest (before taxes) and 10 year life.

Table E-14. EG RECOVERY COSTS FOR BASELINE SYSTEM

Item	Value
Capital Costs	
1 Distillation Column (CL-1)	24,000[a]
1 Distillation Column (CL-2)	24,000[b]
1 Pump	350[c]
1 Condenser	39,180[d]
Descalation factor to bring to 1980 dollars	0.82[e]
Installed Capital Cost Factor	3.26[f]
Total Installed Capital Cost	233,985
Annualized Costs	
Operating Labor	77,400[g]
Operating Materials	0
Maintenance Materials and Labor	10,350[h]
Electricity	940[i]
Steam	828,120[j]
Water	0
Taxes, Insurance and Administration	9,360[k]
Capital Recovery	38,140[l]
Recovery Credit	(407,100)[m]
Total (Annual Costs - Recovery Credit)	557,210

FOOTNOTES FOR TABLE E-14.

a Distillation column (CL-1) size of 4' diameter and 36' high, with 11 trays (Reference 32). Rough cost estimate of $24,000 obtained from Perry's (Reference 29), p. 18-55 in 1968 dollars adjusted to June 1980 dollars.

b Assumed to be same as CL-1.

c Determined to be 3 hp with cost of $350.

d See Reference 31, page 12.

e From Chemical Engineering indices.

f See Table 8-2, developed based on confidential industry information for alternative system.

g Confidential industry information on operating labor scaled down to assume 4 man-hours/shift, 3 shifts per day (= 4,300 hours per year) at $18 per hour.

h Confidential industry information on maintenance labor and material costs were scaled down to assume 520 hours per year for maintenance labor and $1,000 per year for material costs.

i 19,239 kWh per year for pump at $0.049/kWh

j Assumes steam required in distillation columns is equivalent to steam required in EG recovery steam in the regulatory alternative system (see footnote k for Table E-15).

k Based on 0.04 x Installed Capital Cost.

l Based on capital recovery factor of 0.163 for 10% interest and 10 year life.

m Ethylene glycol of 80 percent purity is recovered at a rate of 11.8 lbs/thousand lbs product (Reference 32). Thus, for a 230 x 10^6 capacity plant, total annual EG recovery would be 2,714,000 lbs. Raw material EG, which has a high purity, is worth about 27¢ per lb. The EG recovered from the baseline system is about 30 percent pure and was assumed to have a value of 15¢ per lb. (2,714,000 lbs x $0.15/lb = $407,100).

Table E-15. EG RECOVERY COSTS FOR REGULATORY ALTERNATIVE SYSTEM

Item	Value
Capital Costs	
7 Spray Condensers	148,400[a]
7 Reflux Condensers	148,000[b]
7 Pumps	2,450[c]
7 Heat Exchangers	37,800[d]
1 EG Recovery System	386,300[e]
Descalation Factor	0.82[f]
Installed Capital Cost Factor	4.24[g]
Total Installed Capital Cost	2,514,943
Annualized Costs	
Operating Labor	154,800[h]
Operating Materials	0
Maintenance Materials and Labor	21,700[i]
Electricity	70,550
Steam	828,120[k]
Water	4,650[l]
Taxes, Insurance and Administration	100,600[m]
Capital Recovery	409,940[n]
Recovery Credit	(1,292,000)[o]
Total (Annual Costs - Recovery Credit)	298,360

[a]Based on size estimate from Tennessee Eastman and cost estimate from Missouri Boiler. One spray condenser per line in model plant.

[b]Going to this system replaces feed lines from estifiers to distillation column (CL-2) with reflux condensers, one per line. The size and cost were assumed to be the same as for EG spray condensers on the reactors.

[c]See Reference 31, pp. 10 and 11, for pump sizing. A 3 hp pump is required per line at a cost of $350 per pump.

[d]See Reference 31, pp. 7-9, for heat exchanger sizing. Only sizing done for reactors in which industrial resins are produced. Assume size and costs for heat exchangers associated with reactors producing textile resins would be the same.

[e]Costs obtained from a industry source were considered confidential. A correction factor of 0.764 was obtained from this source to scale the costs down to our model plant capacity. Using the total equipment and steam jet ejector system cost yielded the EGRS cost given.

[f]From Chemical Engineering indices.

[g]See Table 8-2, factors based on sum of piping, insulation, painting, instruments, and electrical factors equaling 1.12 A for larger capacity system given in confidential industry information.

[h]Based on 8,600 man-hours per year at $18 per year. The number of man-hours has been scaled down from the number of man-hours provided by a confidential industry source.

[i]Based on maintenance labor requirements and maintenance materials cost given in confidential industry information for a larger system.

[j]This cost comes from 3 sources: pumps, recovery system, and heat exchanger to chill water. Pumps at 3 hp use 19,239 kWh per year. Recovery system was estimated to use 412,800 kWh per year based on confidential industry information for a larger system. Cost of electricity is $0.049/kWh. Chilled water is necessary for chilling the spent EG used in the EG spray condensers and is necessary for reactors producing high tenacity (high viscosity) industrial resins. The electricity requirement to maintain the chilled water in each heat exchanger is calculated to be 177,000 Btu/hr or $1,522 \times 10^9$ BTU per year. Using a conversion efficiency of 0.72 and converting to kWh, this is equal to 4.461×10^5 kWh per year. Assuming 2 of the 7 lines in a model plant produce high tenacity (high viscosity) industrial resins, total electricity usage in the model plant is $19,239 \times 7$ plus 412,800 plus $2 \times 4.461 \times 10^5$ equal to 1,439,673 kWh per year.

[k]Based on confidential industry information and scaling steam usage in the EG recovery system and its vacuum system proportionately according to plant capacity, steam usage was estimated to be 1.34×10^8 lbs/year. A cost of $6.18/1,000 lb of steam was used.

[l]Based on water consumption of 15.5×10^6 gallons per year (from confidential industry information scaled down by proportioning relative capacities).

[m]Based on 0.04 x Installed Capital Cost.

[n]Based on a capital recovery factor of 0.163.

[o]Based on a total EG recovery of 20.8 lbs of EG/1,000 lbs of product and a recovery credit of $0.27/lbs of EG (from Chemical Marketing Reporter). The 20.8 lbs of EG/1,000 lbs of product is based on 99 percent recovery of the 18.2 lbs of EG/1,000 lbs of product estimated entering the EG spray condenser (equal to 18 lbs/1000 lbs) plus 2.3 lbs of EG/1,000 lbs of product recycled from the estifiers that would have to be replaced with fresh feed if the baseline system was used.

E.7 PIPING AND DUCTING DESIGN AND COST ESTIMATION PROCEDURE

Control costs for flare and incinerator systems included costs of piping or ducting to convey the waste gases (vent streams) from the source to a pipeline via a source leg and through a pipeline to the control device. All vent streams were assumed to have sufficient pressure to reach the control device. (A fan is included on the duct, fan, and stack system of the incinerators.)

E.7.1 Piping and Ducting Design Procedure

The pipe or duct diameter for each waste gas stream (individual or combined) was determined by the procedure given in Table E-16. For flows less than 700 scfm, an economic pipe diameter was calculated based on an equation in the Chemical Engineer's Handbook[33] and simplified as suggested by Chontos.[34,35,36] The next larger size (inner diameter) of schedule 40 pipe was selected unless the calculated size was within 10 percent of the difference between the next smaller and next larger standard size. For flows of 700 scfm and greater, duct sizes were calculated assuming a velocity of 2,000 fpm for flows of 60,000 acfm or less and 5,000 fpm for flows greater than 60,000 acfm. Duct sizes that were multiples of 3-inches were used.

E.7.2 Piping and Ducting Cost Estimation Procedure

Piping costs were based on those given in the Richardson Engineering Services Rapid Construction Estimating Cost System[21] as combined for 70 ft. source legs and 500 ft. and 2,000 ft. pipelines for the cost analysis of the Distillation NSPS.[37] (see Tables E-17 and E-18) Ducting costs were calculated based on the installed cost equations given in the GARD Manual.[33] (see Table E-19)

Costs of source legs were taken or calculated directly from the tables. Costs of pipelines for flares were interpolated for the safe pipeline lengths differing by more than 10 percent from the standard lengths of 70, 500, and 2,000 ft.

Table E-16. PIPING AND DUCTING DESIGN PROCEDURE

Item	Value
(1) Pipe diameter, D	
(a) Piping[a]	D (in.) = 0.042 x Q (scfm) + 0.472, for Q < 40 scfm D (in.) = 0.009 x Q (scfm) + 2.85, for 40 ≥ Q < 700 scfm Select next larger inner diameter of schedule 40 pipe (given in Table E-18) unless calculated size with 10 percent of interval to next larger size.
(b) Ducting[b]	D (in) = $(0.3028)\sqrt{Q(\text{acfm})}$, for D > 12 in. or Q ≥ 700scfm and Q ≤ 60,000 acfm D (in.) = $(0.1915)\sqrt{Q(\text{acfm})}$, for D > 60,000 acfm Select size that is a multiple of 3 inches.
(2) Pipe length, L	
(a) Flares	Assumed 70-ft. source leg from each source to the pipeline. Assumed separate pipelines for large (≥40,000 scfm) intermittent streams and for all continuous streams together. Selected pipeline length of 70, 500 or 2,000 ft. if calculated safe pipeline length within 10 percent of standard length; if not selected calculated length between standard values.
(b) Incinerators	Assumed 70-ft. source legs from each source to the pipeline. Used duct, fan, and stack cost from Enviroscience, which assumes a 150-ft. duct cost based on the GARD Manual (Reference 38)

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[a] Economic pipe diameter equations from Reference 36 (which is based upon References 33 and 34).

[b] From continuity equation $Q = \frac{\pi}{4} D^2 V$; assumed velocity, V, of 2,000 fpm for lower flows and 5,000 fpm for higher flows.

Table E-17. PIPING COMPONENTS[a]

Equipment Type	Number of Equipment Type in Pipe Leg Type			
	Source	Compressor	Pipeline (500)	Pipeline (2,000)
Check Valves	1	1	1	1
Gate Valves	4	2	3	3
Control Valves	1	–	1	1
Strainers	1	1	1	1
Elbows	8	6	6	6
Tees	6	2	3	3
Flanges	15	10	20	35
Drip Leg Valves	1	1	1	1
Expansion Fittings	2	1	5	20
Bolt and Gasket Sets	15	12	21	38
Hangers	9	4	50	200
Field Welds	18	12	40	120
Pipe, Schedule 40 (ft)	70	20	500	2,000

[a]From Reference 36.

Table E-18. INSTALLED PIPING COSTS[a]

Pipe Diameter (ft)	Cost in August 1978 Dollars for Pipe Leg Type			
	(70') Source Leg	(20') Compressor Leg	500' Pipeline	2,000' Pipeline
0.0411	2,200	1,050	2,275	4,360
0.0518	2,262	1,085	2,555	4,955
0.0647	2,330	1,120	2,660	5,255
0.0874	2,470	1,170	2,870	5,800
0.1342	2,725	1,375	3,485	7,235
0.1722	3,340	1,490	3,990	8,190
0.2057	5,644	2,725	5,765	10,690
0.2557	6,045	2,900	6,640	12,500
0.2957	7,428	3,465	7,925	14,825
0.3355	7,786	3,830	9,000	16,870
0.4206	9,750	4,800	11,080	21,430
0.5054	11,415	5,570	13,220	26,120
0.6771	16,025	7,715	17,865	34,545
0.8542	23,925	11,440	26,500	48,920
1.021	33,133	18,280	36,320	65,000
1.111	45,750	20,900	48,680	83,400
1.281	58,680	26,670	61,554	96,850
1.448	72,800	32,670	77,450	123,200
1.604	92,581	41,140	97,010	149,420
1.937	121,200	54,205	124,075	186,385

[a]From SOCMI-Distillation NSPS Computer Program "DM Pipe" (Reference 37) based on component costs from Reference 21.

Table E-19. INSTALLED DUCTING COST EQUATIONS, DECEMBER 1977 DOLLARS[a]

Pipe Wall Thickness (in.)[b]	Pipe Leg Type as a Function of Diameter, D, (in.)				Permissable Diameters (in.)
	Source Leg	Compressor Leg	500 ft. Pipeline	2,000 ft. Pipeline	
1/4	$2074 + 396.8\,D + 2.40\,D^2$	$1054 + 202.5\,D + 1.600\,D^2$	$67.53 + 1058\,D + 1.68\,D^2$	$-3007 + 3713\,D + 1.68\,D^2$	$28 \le D \le 122$
3/16	$2083 + 352.2\,D + 2.050\,D^2$	$1052 + 183.5\,D + 1.367\,D^2$	$204.3 + 846.7\,D + 1.435\,D^2$	$-2436 + 2901\,D + 1.435\,D^2$	$30 \le D \le 90$
1/8	$2198 + 285.8\,D + 1.76\,D^2$	$1121 + 146.7\,D + 1.174\,D^2$	$420 + 673.9\,D + 1.232\,D^2$	$-1770 + 2309\,D + 1.232\,D^2$	$12 \le D \le 81$

[a]From equations for carbon steel pipe components (straight duct, elbows, tees, and transitions) given in the GARD Manual (Reference 38) combined for piping components in each pipe leg type as shown in Table E-17.

[b]After examining dependence of wall thickness, t, upon pressure, (psi) and diameter, D (in.):

$$t\ (in.) = \frac{p \times D}{[2[18,200 \times 0.50] + (0.4 \times p)]} + 0.16$$

for cases where pressure was known.

$$\begin{aligned} &= 1/4" \text{ for } p \ge 74.7 \text{ psia or } D \ge 11.5 \text{ ft.} \\ &= 3/16" \text{ for } Q \ge 30,000 \text{ acf}_m, \ (0\text{-}10 \text{ psig}) \\ &= 1/8" \text{ for } Q < 30,000 \text{ acf}_m \end{aligned}$$
assumed

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