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Office of Air Quality
Planning and Standards
Research Triangle Park NC 27711

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EPA

Polymer Manufacturing Industry - Background Information for Proposed Standards

Draft EIS

NSPS

ENVIRONMENTAL PROTECTION AGENCY

Background Information and Draft Environmental Impact Statement
for
Polymer Manufacturing Industry

Prepared by:

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9/23/87
Date

1. The proposed standards of performance would limit emissions of volatile organic compounds from new, modified, and reconstructed polymer manufacturing facilities. Section 111 of the Clean Air Act (42 U.S.C. 7411), as amended, directs the Administrator to establish standards of performance for any category of new stationary source of air pollution that ". . . causes or contributes significantly to air pollution which may reasonably be anticipated to endanger public health or welfare." Although all regions of the United States will be affected, the Southeast and Gulf States will be affected the most.

2. Copies of this document have been sent to the following Federal Departments: Labor, Health, and Human Services, Defense, Transportation, Agriculture, Commerce, Interior, and Energy; the National Science Foundation; the Council on Environmental Quality; State and Territorial Air Pollution Program Administrators; EPA Regional Administrators; Local Air Pollution Control Officials; Office of Management and Budget; and other interested parties.

3. The comment period for review of this document is 75 days from the date of publication of the proposed standard in the Federal Register. Mr. Gil Wood may be contacted at (919) 541-5625 regarding the date of the comment period.

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Polymer Manufacturing Industry - Background Information for Proposed Standards

Emission Standards and Engineering Division

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Air, Noise, and Radiation
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711

September 1985

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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, twelve 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 (SOCMI). Similarities between the polymers

and resins industry and SOCMI 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 either methanol emissions using a condenser or ethylene glycol emissions from the cooling tower by using a distillation column or making greater use of an existing distillation column. No additional regulatory alternatives were developed for the poly(ethylene terephthalate) model plant using the terephthalic acid process making a low viscosity product or a high viscosity product with a single end finisher per process line.

Additional regulatory alternatives for the eleven remaining model plants further reduce emissions through the additional control of process emissions. For the polypropylene, polyethylene, and expandable polystyrene 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 crystal or impact polystyrene model plant, additional control was obtained by applying additional recovery to the process emissions. For poly(ethylene terephthalate) plants using the dimethyl terephthalate process, additional control was obtained by greater recovery of the methanol stream from the methanol recovery system and by recovery of ethylene glycol emissions from the cooling tower with a distillation column. For poly(ethylene terephthalate) plants using the terephthalic acid process for producing a high viscosity product with multiple end finishers per process line, additional control was obtained by making greater use of the existing distillation column.

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	-1*	0	-1*
Regulatory Alternative III	+4**	+1**	-1*	-2*	-1**	-1*
Regulatory Alternative IV	+4**	+1**	-1*	-2*	-2**	-1*
Regulatory Alternative V	+4**	+1**	-1*	-2*	-2**	-1*
Regulatory Alternative VI	+4**	+1**	-1*	-3*	-2**	-2*
Regulatory Alternative VII	+4**	+1**	-1*	-3*	-2**	-2*

KEY: + Beneficial impact
- Adverse impact

0 No impact
1 Negligible impact
2 Small impact
3 Moderate impact
4 Large impact

* Short term impact
** Long-term impact
*** Irreversible impact

Alternative 2 impacts for each model plant are totaled as Regulatory Alternative II. Regulatory Alternative III represents Regulatory Alternative 3 impacts for ten model plants plus Regulatory Alternative 2 impacts for the polypropylene model plant using the gas phase process and the poly(ethylene terephthalate) model plant using the terephthalatic acid process producing a low viscosity product or a high viscosity product with a single end finisher per process line as no Regulatory Alternative 3 was developed for these model plants. Regulatory Alternative IV corresponds to Regulatory Alternative 4 for eight model plants plus Regulatory Alternative 3 impacts for two model plants plus Regulatory Alternative 2 impacts for two model plants. Regulatory Alternatives V, VI, and VII correspond to regulatory alternatives combined using the same procedure as outlined above. Regulatory VII, thus, represents the aggregate impact of the most stringent regulatory alternatives considered for each model plant.

1.2.1 Air Emissions Impacts

Total VOC emissions from new process lines in these industry segments in 1988 are projected to be approximately 7.33 gigagrams (Gg) under Regulatory Alternative I, compared to 5.84, 3.62, 3.18, 2.87, 2.80, and 2.77 Gg under Regulatory Alternatives II, III, IV, V, VI, and VII, respectively. The average percent emission reductions from the Regulatory Alternative I level achieved by Regulatory Alternatives II, III, IV, V, VI and VII are 20, 51, 57, 61, 61.8, and 62.2 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 through VII due to the use of catalytic incinerators. Spent catalyst from catalytic incinerator use may be generated at an annual rate of up to 4.21 m^3 (149 ft^3) under Regulatory Alternative VII.

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 nine of the twelve model plants and of additional methanol or ethylene glycol recovery in poly(ethylene terephthalate) plants result in a net increase in energy usage of 66 TJ (11,000 bbl oil) from what otherwise would occur under Regulatory Alternative 1. This increase is due solely to the energy required to implement Regulatory Alternative 2 for the poly(ethylene terephthalate) model plant using the terephthalic acid process producing a low viscosity product or a high viscosity product with a single end finisher per process line. Excluding this model plant, there would be a net decrease in energy usage of 65 TJ (10,000 bbl oil) under Regulatory Alternative II because the energy credit obtained through recovered VOC is greater than the energy expended to implement the controls.

Under Regulatory Alternatives III through VII, there is a net increase in energy usage as the energy needed to implement process VOC emission controls exceeds the energy credit obtained through recovered VOC. The net annual energy increase ranges from about 370 TJ (61,000 bbl oil) under Regulatory Alternative III (approximately a 37 percent increase over Regulatory Alternative I energy usage) up to about 940 TJ (154,000 bbl oil) under Regulatory Alternative VII, (approximately a 93 percent increase over Regulatory Alternative I 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 spend a net annual amount of around $1.9 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 cost of $600 for a poly(ethylene terephthalate) process line using the dimethyl terephthalate process up to $105,000 for

a poly(ethylene terephthalate) process line using the terephthalic acid process producing a low viscosity product or a high viscosity product with a single end finisher.

Under Regulatory Alternative III, the industry as a whole would spend an net annual amount of $7.7 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 an annual cost of $600 for a poly(ethylene terephthalate) process line using the dimethyl terephthalate process up to $404,000 for a high density polyethylene process line using the liquid phase solution process.

Under Regulatory Alternatives IV, V, and VI, fifth year annual costs of $10.2 million, $12.9 million, and $14.6 million, respectively, over and above what would be otherwise spent under Regulatory Alternative I would be realized by the industry as a whole. Annual costs for individual industry segments range from net annual costs of $600 for a poly(ethylene terephthalate) process line using the dimethyl terephthalate process up to a cost of $1.2 million (under Regulatory Alternative 6) for an expandable polystyrene process line using the post-impregnation suspension process.

Under the most costly combination of individual regulatory alternatives for each model plant, Regulatory Alternative VII, total additional annualized costs of controls in 1988 are estimated to be $15.4 million. The potential adverse economic impacts of these regulatory alternatives are expected to be minor in view of the generally 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

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 conducted 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

"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 segments 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 segments 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. (Docket Reference Number II-I-82.)

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 Plaquemine, LA	447 254
E.I. du Pont de Nemours & Co., Inc.	Orange, TX Victoria, TX	211 109
El Paso Polyolefins Co.	Odessa, TX Pasadena, TX	181 68
Exxon Chemical Co.	Baton Rouge, LA Mt. Belview, TX[d]	299 154[d]
Gulf Oil Chemical Co.	Cedar Bayou, TX Orange, TX Baytown, TX[b]	259 129 -
Mobil Chemical Co.	Beaumont, TX	136
National Distillers & Chemical Corp.	Deer Park, TX Tuscola, IL	249 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 Taft, LA Penuelas, P.R.	544[g] 272 141
United Foam Corp.[h]	Louisville, KY	73

[a]Source: SRI International, 1982 Directory of Chemical Producers, United States, unless otherwise indicated. (Docket Reference Number II-I-82.)

[b]Source: Texas Air Control Board communication. (Docket Reference Number II-D-62.)

[c]In a letter dated March 9, 1982, Cities Services Co. indicated that they were closing all their polymer and resins manufacturing plants. (Docket Reference Number II-D-46.)

[d]Source: Chemical Engineering. October 18, 1982. p. 26. (Docket Reference Number II-I-102.)

[e]Capacity obtained from April 14, 1982, letter from Northern Petrochemical Company. (Docket Reference Number II-D-52.)

[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. (Docket Reference Number II-I-69.)

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. (Docket Reference Number II-I-82.)

[b]Expansion from 191 Gg/yr to 261 Gg/yr completed in 1982. Chemical Engineering. October 4, 1982. p. 25. (Docket Reference Number II-I-101.)

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	3.6	-
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. (Docket Reference Number II-I-82.)

[b]Source: Industry communications.

[c]Source: Organic Chemical Producers Data Base. Product Data Report - Nationwide. February 17, 1981. p. 943. (Docket Reference Number II-I-68.)

[d]This plant is not currently in production. Letter from Gulf Oil to Texas Air Control Board. July 28, 1982. (Docket Reference Number II-I-93.)

[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. (Docket Reference Number II-E-33.)

Table 3-5. POLY(ETHYLENE 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	48	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 Kinston, NC Old Hickory, TN Wilmington, NC	DMT/TPA	736	Fiber
	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
Rohm and Haas Co.	Fayetteville, NC	-	91	Bottle Resins

[a]Source: SRI International, 1982 Directory of Chemical Producers, United States. (Docket Reference Number II-I-82.) 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 polyesters other than PET.

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

[c]Industry correspondence. (Docket Reference Number II-B-53.)

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 End-Uses of Polypropylene. 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).[5] 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).[6]

Like conventional LDPE, LLDPE is suitable for many end-uses. Specific applications may include housewares, lids, closures, blow molded parts (e.g., toys, bottles, and drums), wire and cable insulation, extruded pipe and tubing, and industrial and consumer films (e.g., food packaging, trash bags, and garment bags).[7]

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.[5] 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.[8] End-use sectors for HDPE include packaging (45 percent), consumer/institutional (11 percent), building and construction (9 percent), and other sectors (35 percent).[8]

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.[9]

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.[9]

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.[10] 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.[10]

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 that 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:

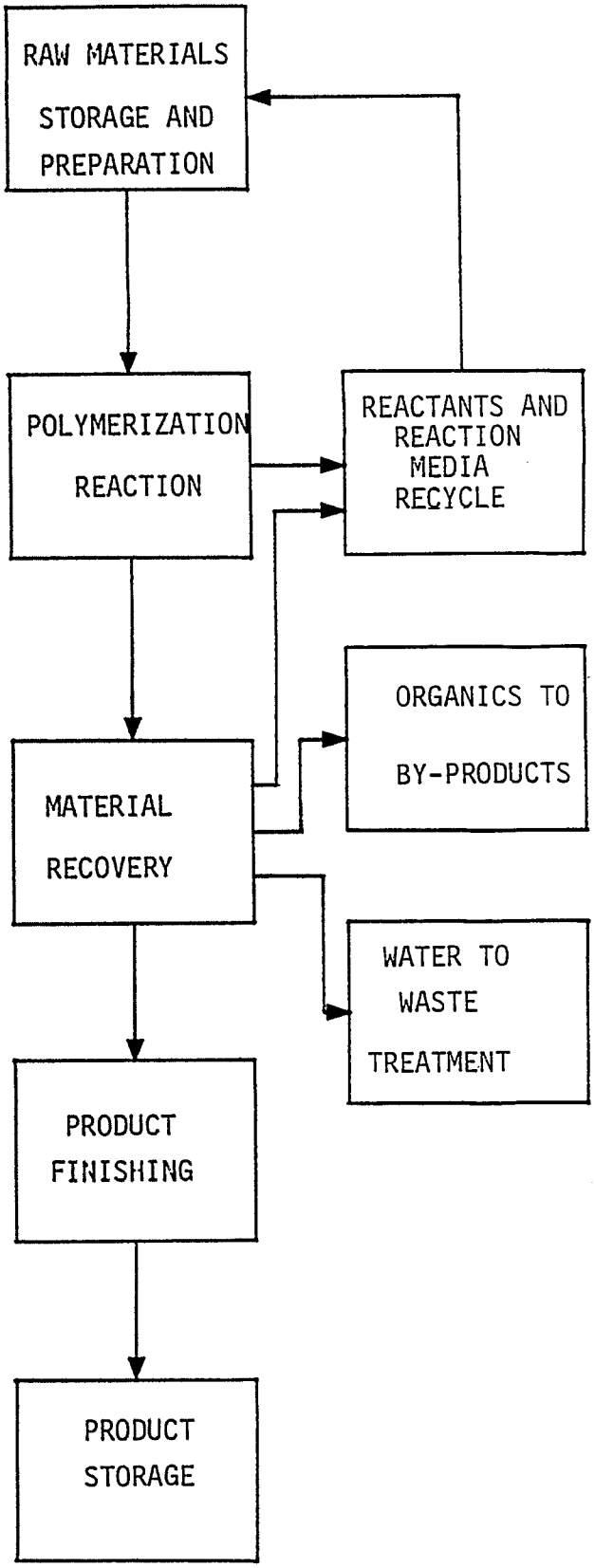
1. Raw materials storage and preparation
2. Polymerization reaction
3. Materials recovery
4. Product finishing
5. Product storage

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: (1) those that can be anticipated based on the process flow diagram and (2) those that can be identified only by sampling procedures (e.g., 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 materials preparation, polymer production,



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, and

(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

material recovery, polymer extrusion and pelletizing, and product cooling drying and storage. 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 process
4. LDPE - low-pressure process
5. HDPE - liquid phase slurry process
6. HDPE - liquid phase solution process
7. HDPE - gas phase process
8. Polystyrene - batch process
9. Polystyrene - continuous process
10. Expandable Polystyrene - post-impregnation suspension process
11. Expandable Polystyrene - in-situ suspension process
12. PET - Dimethyl terephthalate (DMT) process
13. 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 commercial thermoplastics. The general formula for polypropylene is:

$$\ldots - CH_2 - \underset{\displaystyle CH_3}{CH} - CH_2 - \underset{\displaystyle CH_3}{CH} - CH_2 - \underset{\displaystyle CH_3}{CH} - CH_2 - \ldots$$

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.[11]

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,[12] and can be a batch or continuous process. Batch polymerization is applicable particularly 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.[12] In conventional liquid phase processes, the catalyst provides a relatively low polymer yield on the order of 500 to 1,000 units per unit of catalyst.[13] 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 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 per unit of catalyst.[13] 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 required for catalyst removal and recovery. The elimination of this 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 in 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

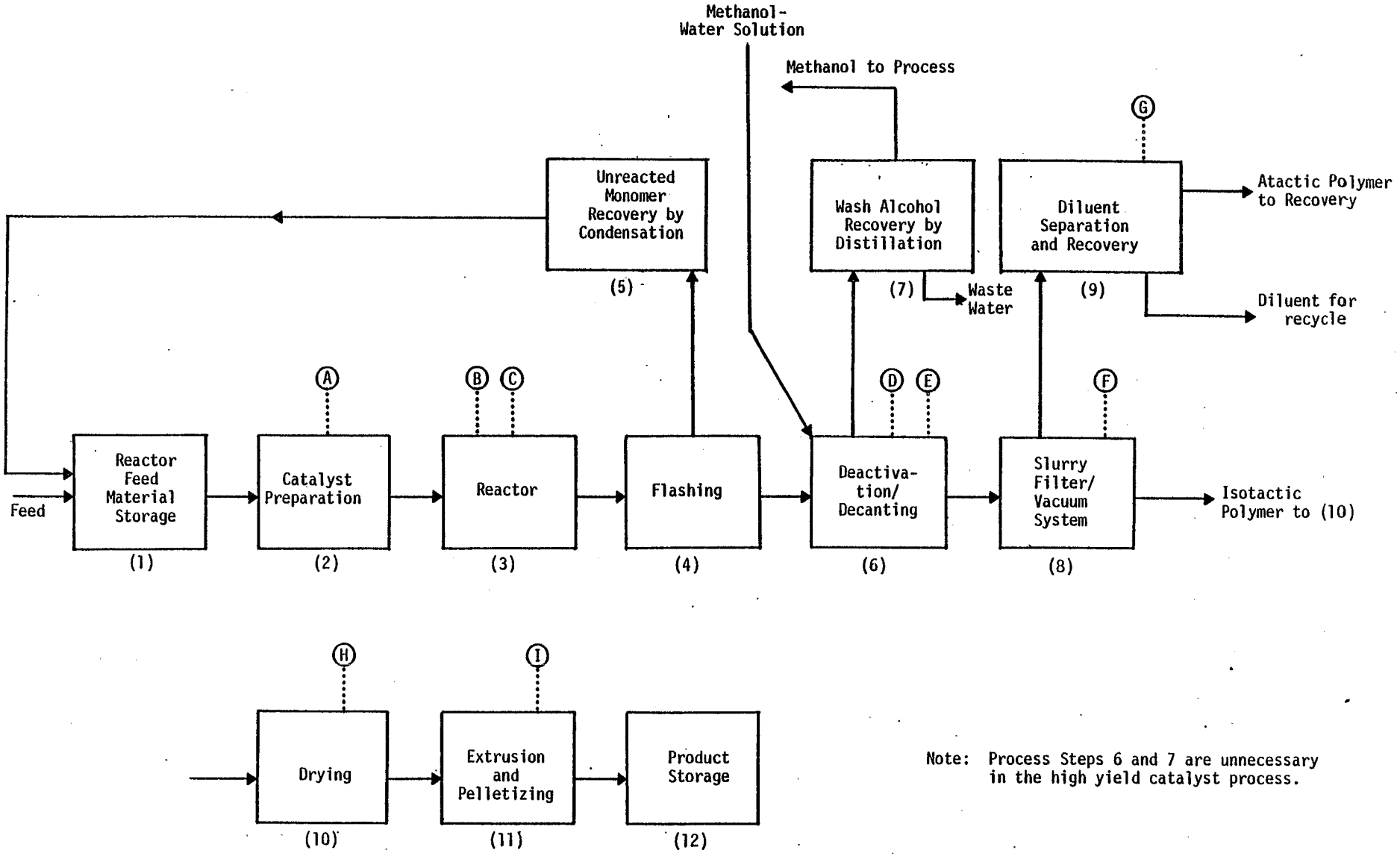


Figure 3-2. Simplified Process Block Diagram for the Polypropylene Continuous, Liquid Phase Slurry Process

may use either liquid propylene or a different organic compound as a diluent in which the polymer forms a slurry. The only difference between the conventional and high yield 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), a process diluent (which acts as a heat transfer agent), a polymer suspension medium, and a heterogeneous Ziegler-Natta type catalyst. Hexane is used often as a process diluent, although some processes use mixtures of other aliphatic hydrocarbons. The catalyst (2) sometimes is 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.[14] Reaction is carried out at temperatures of about 60°C (140°F) for approximately 8 hours.[15] A portion of the reaction effluent, which consists of polymer, monomer, and diluent, is drawn continuously 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 Emissions from the polypropylene continuous, liquid phase slurry process. The characteristics of vent streams from this process are shown in Table 3-6. Each vent stream is identified (circled letter) in Figure 3-2. The total process VOC emission rate for a conventional slurry process is approximately 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 23.8 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.

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, $C_{10}HC$
PR	C	Emergency Reactor Vents	Intermittent	0.38	66	-	C_2-C_6, N_2, H_2[f]
MR	D	Decanter vents	Continuous	11.49	38	0	C_3HC, $C_{10}HC$, IPA
MR	E	Neutralizer vents	Continuous	1.82	32-71	0	C_3HC, C_4HC, $C_{10}HC$, IPA
MR	F	Slurry vacuum/filter system vent	Continuous	7.93	32	0	$C_{10}HC$, IPA
MR	G	Diluent separation and recovery	Continuous	8.72	104	0	$C_{10}HC$, IPA
PF	H	Dryer vents	Continuous	0 to 0.69	85-104	0	Air and small amount of VOC
PF	I	Extrusion/pelletizing vent	Continuous	2.0			25% HC
		Total Emission Rate		37.1[h]			

[a]Source of information: Industry correspondence.

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

[c]See Figure 2-1 for stream identification.

[d]Streams are diluted in 10 to 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]Approximately 80 percent by weight C_2 - C_6; 10 percent by weight N_2; and 10 percent by weight H_2.

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

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

The vent from the catalyst preparation section, Stream A, releases diluent continuously. Reactor vent Stream B is continuous and releases monomer and diluent. Reactor vent Stream C is an intermittent stream that occurs during emergency dumps of the reactor. This stream is typically controlled by a flare.

The combined total of the decanter and neutralizer vents, Streams D and E, 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 F, 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 G, 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 H, 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 I. 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 since 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 Polypropylene Gas Phase Process. The gas phase process for producing polypropylene is relatively new and currently is 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 Process description. 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).

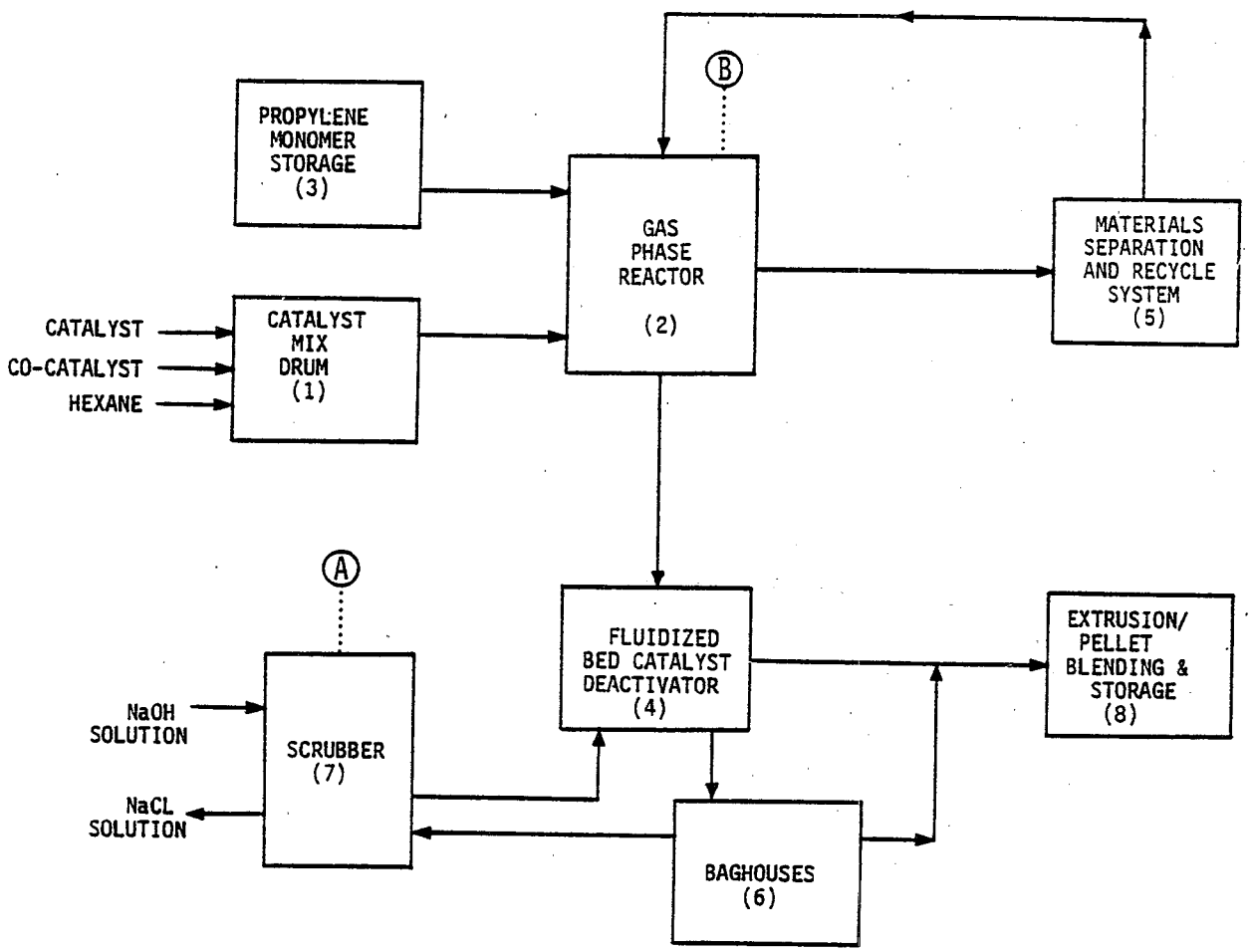
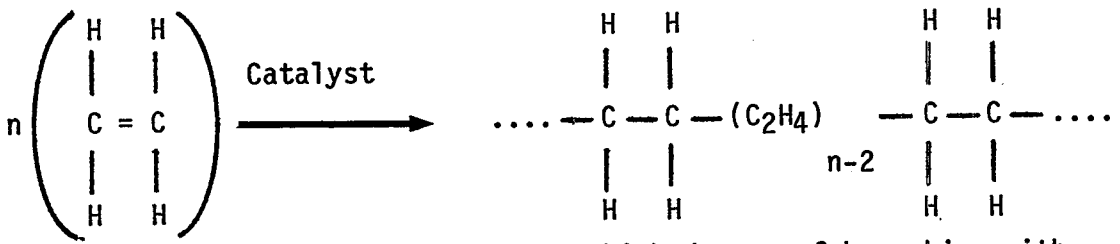


Figure 3-3. Simplified Process Block Diagram for the Polypropylene Gas Phase Process

3.2.1.2.2 Emissions from the polypropylene gas phase process. Two VOC offgas streams from the process, the scrubber vent, which is a continuous stream (A), and the reactor blowdown vent, which is an intermittent stream (B), are currently controlled by a flare system according to one industry source. The uncontrolled VOC emission rate for the entire process, based on the average expected flow rate 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.



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 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.[16] 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.[17] High-pressure processes, however, are not capable of polymerizing propylene or higher olefins to their corresponding polyolefin.[18] 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

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

Process Section[b]	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.

polymer in the reactor, the reactor configuration,[19] reaction temperature, or type and concentration of the catalyst.[20]

3.2.2.1.1 Process description. Figure 3-4 is a schematic representation of the 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.[21]

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).[21] 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.[22] 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

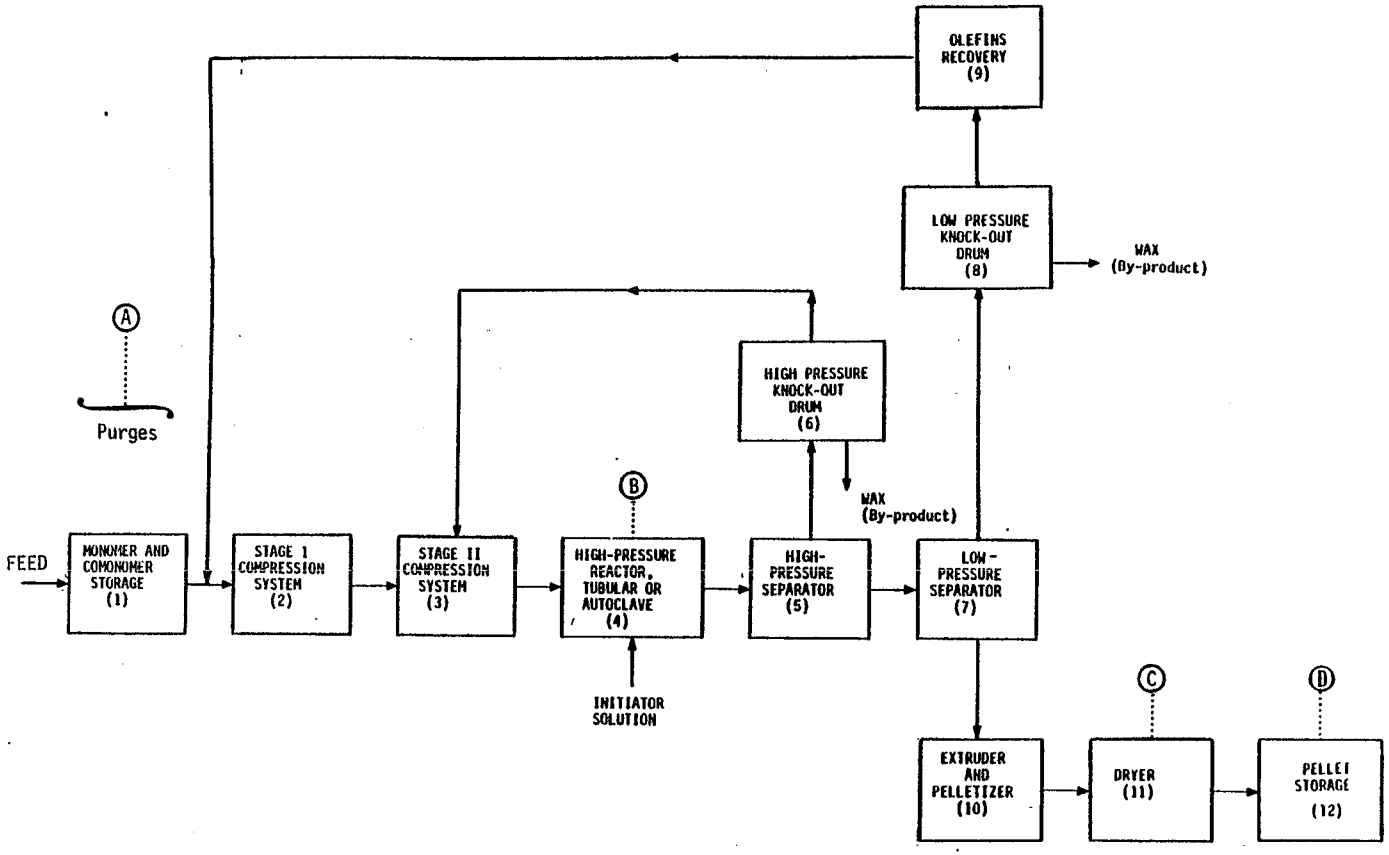


Figure 3-4. Simplified Process Block Diagram for the Low Density Polyethylene High-Pressure Process

homopolymer or copolymer product is extracted as the low-pressure separator bottoms which still contain residual ethylene monomer. These product 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 Emissions from the LDPE high-pressure process. The offgas stream characteristics for the high-pressure process are shown in Table 3-8. The total uncontrolled VOC emission rate for the entire process varies significantly, ranging from about 1 kg VOC/Mg product to about 29 kg VOC/Mg product or higher. 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 start-up, shutdown, and maintenance vents, Stream A, are intermittent and consist of VOC gases from a number of sources, such as compressors, within the plant. Included in this stream are the vents from the wax blowdown system, which can be a source of significant ethylene losses. The emission rate is highly dependent upon the design of the wax blowdown and discharge system.

The emergency reactor vent, Stream B, is intermittent. It is activated either manually or automatically as an emergency blowdown during a process upset. The total mass and instantaneous flow rate can vary dramatically depending on the nature of the upset. This VOC emission source 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 C and D, are usually combined. They are continuous streams consisting of mostly air with

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

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition Wt. %
Other	A	Start-up, shutdown, and maintenance purges	Intermittent	0.23-21.0	Varies	Varies	100 VOC[d]
PR	B	Emergency vents[e]	Intermittent	0.1-7.65	Varies	0-20	80-99 Ethylene 20-1 Other[f]
PF,PS	C,D	Combined dryer and storage bin vents	Continuous	0.71-6.01	47[g],35[h]	0	0.2 Ethylene 99.8 Air
		Total Emission Rate		0.71-29.25[i]			

[a]Source of information: Industry correspondences.

[b]PR = polymerization reaction; PF = product finishing; other = vents from all process sections.

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

[d]Typically at least 99 percent ethylene. One plant did report an 80-85% ethylene composition.

[e]Includes emergency releases from the low and high pressure separators.

[f]Such as polymer, propylene, water, CO, and CO_2.

[g]Dryer temperatures. Reported ranges: 38 to 60°C.

[h]Storage bin temperatures. Reported ranges: 27 to 43°C.

[i]There are actual reported total emission rates for total plants and do not represent the sum of the low or high estimates in the ranges reported above.

small amounts of ethylene. These streams are usually emitted to the atmosphere.

A considerable number of LDPE plants are located near or integrated with a plant which manufactures ethylene. As a result, the unreacted 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 Process.[23] Most, and probably all, new low density polyethylene plants will use the relatively new low-pressure 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 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 and/or reactor gas composition; reactor operating conditions remain the same.[24]

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 low-pressure 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).

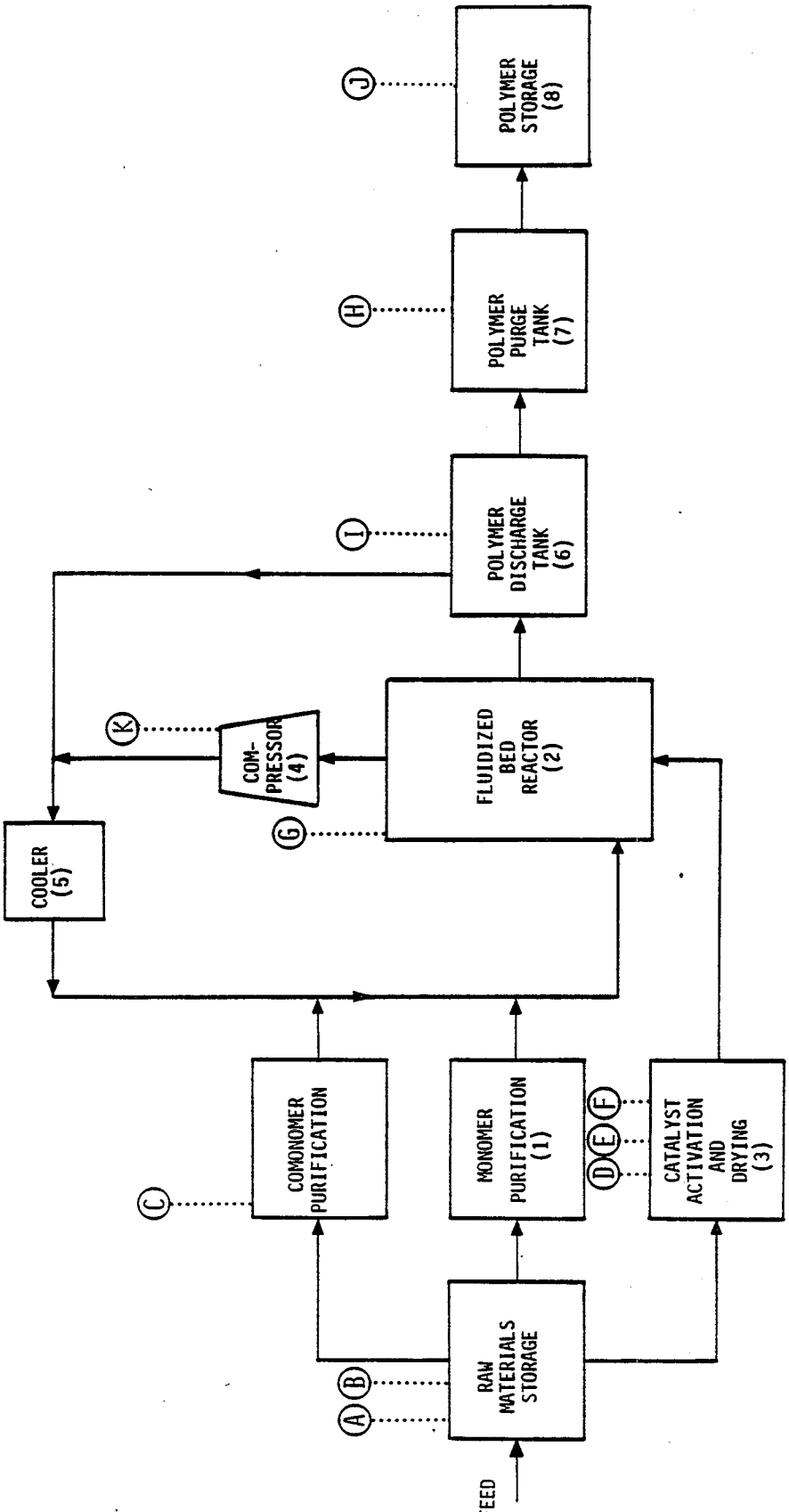


Figure 3-5. Simplified Process Block Diagram for the Low Density Polyethylene Low-Pressure and the High Density Polyethylene Gas Phase Processes

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 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 97 to 99 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.[20]

3.2.2.2.2 Emissions from the LDPE low-pressure process. The waste gas stream characteristics for the LDPE low-pressure process are shown in Table 3-9. The combined process VOC emission rate for this process is approximately 23.4 kg VOC/Mg product. One process vent, the product discharge vent, contributes slightly more than 95 percent (22.3 kg VOC/Mg product) of the total VOC discharged annually from the model plant for this process.

Table 3-9. CHARACTERISTICS OF VENT STREAMS FROM THE LOW DENSITY POLYETHYLENE LOW-PRESSURE PROCESS AND THE HIGH DENSITY POLYETHYLENE 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	0.10	-	-	92 VOC 8 N_2
RMP	D	Catalyst dehydrator handling vent	Intermittent	0.004	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	Reactor purge vent	Intermittent	0.71	-	-	1-83 VOC 99-17 N_2
PF	I	Product discharge vent	Continuous	22.3	85	1-20	35 VOC 65 N_2
PS	J	Bin vent	Continuous	0.05	50	0.2	0.02 VOC 99.98 Air
	K	Compressor seal (oil) vent[e]	Continuous	0.01	38	1.0	50 VOC 50 N_2
		Analyzer vents[f]	Continous	0.01	49	0.1	100 VOC
		Total Emission Rate		23.37			

[a]Source of information: Industry correspondence.

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

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

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

[e]This vent is a fugitive emission source.

[f]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.

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 827°C (100 to 1,520°F), and the pressure varies from 0.7 to 138 kPa (0.1 to 20 psig).

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 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.[25] 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

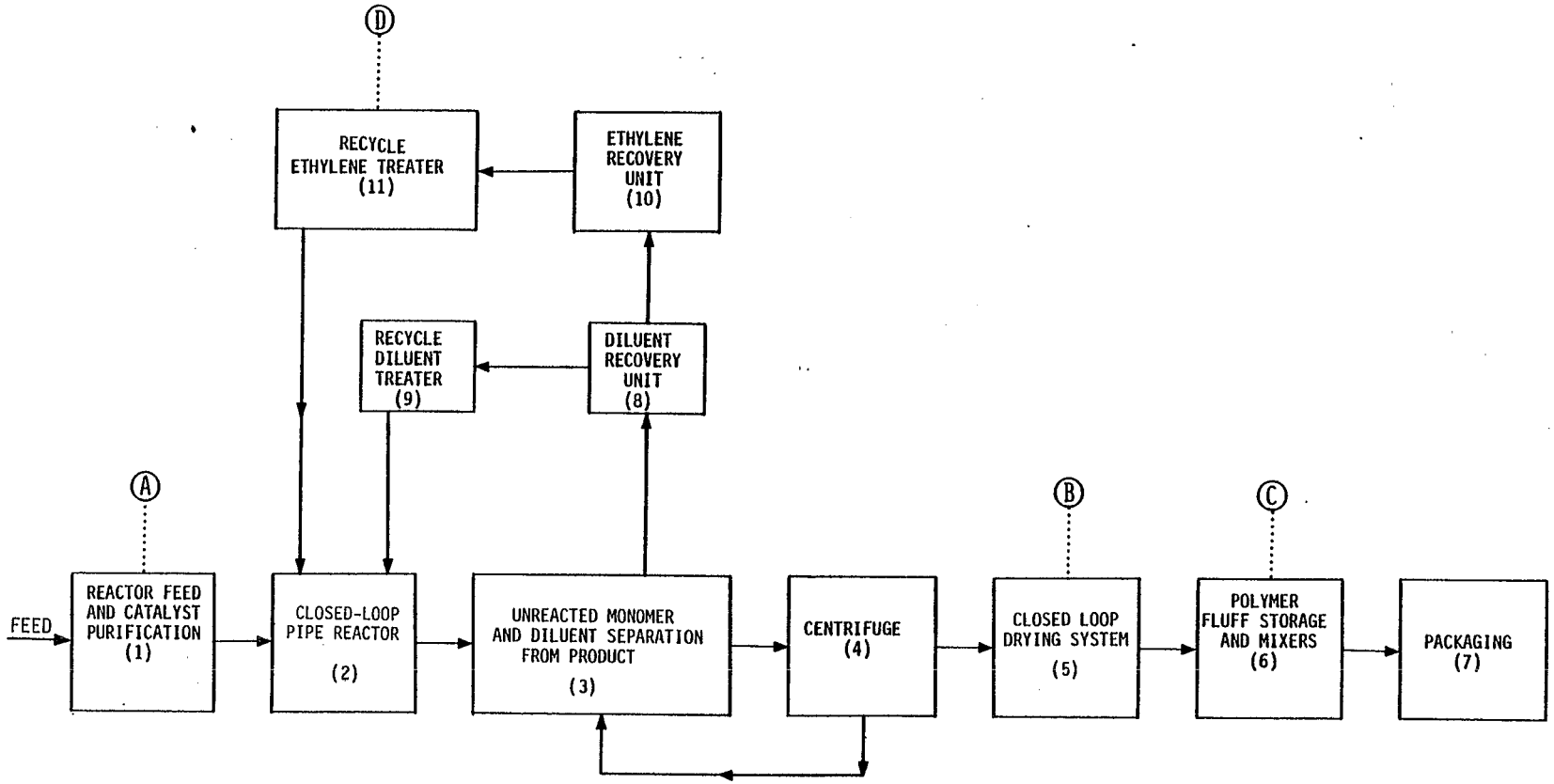


Figure 3-6. Simplified Process Block Diagram for the High Density Polyethylene Liquid Phase Slurry Process

screw conveyors are used to withdraw concentrated slurries containing 50 to 80 percent solids.[14] Unreacted monomer and diluent (3) are separated from the product by flashing. Final stripping of the gases from the 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 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 approximately 12.7 kg VOC/Mg product. The total emissions from the four streams varies from 13.0 to 13.3 kg VOC/Mg product. Table 3-10 shows the composition of these streams.

The emissions from the drying system (Stream B) are a dilute mixture of process solvent in nitrogen when a closed-loop nitrogen dryer is used or a mixture of process solvent and air when an air fluidized bed dryer is used. 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.

Table 3-10. CHARACTERISTICS OF VENT STREAMS FROM THE HIGH DENSITY POLYETHYLENE 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	0.06-0.4[f]	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		13.0-13.3		

[a]Source of information: Industry correspondence.

[b]Gas stream pressure data unknown.

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

[d]See Figure 3-6 for stream identification.

[e]One occurrence per month.

[f]Low end is for recycle nitrogen dryers; high end is for air fluidized bed dryers.

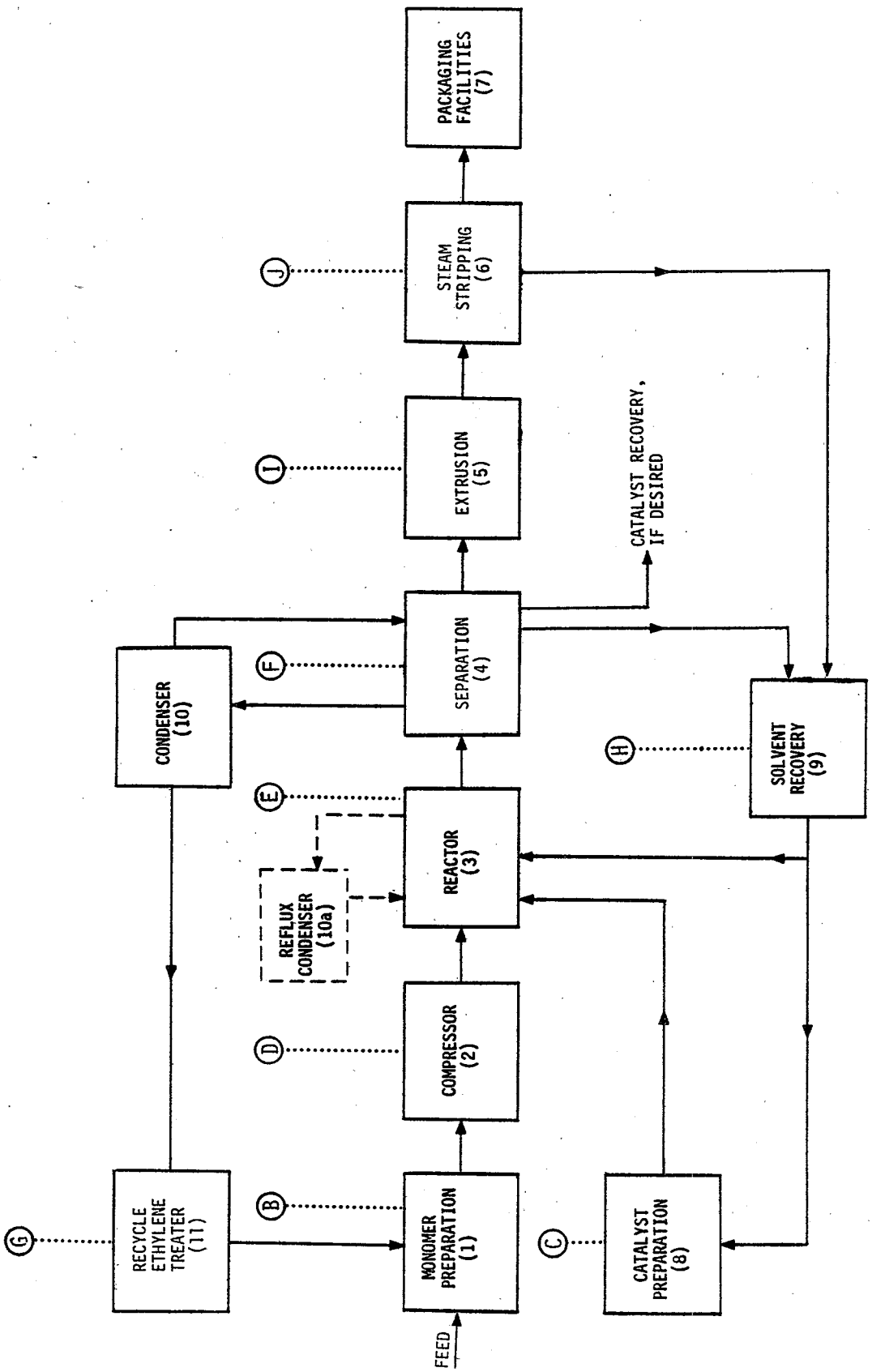
The intermittent offgas stream is the feed preparation stream (Stream A) which consists mostly of ethylene. Sources for this stream include drying/dehydrating and other feed purification operations. Its emission rate is 0.2 kg VOC/Mg product. Emissions occur when the treating vessels are purged prior 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 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 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.[26] 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.[27,28] 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.



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

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

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 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 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 16 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 55 percent of the emissions from those HDPE plants that do not have an adjacent ethylene plant. Table 3-11

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

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Composition, % by volume
RMP	A	Tank vents	Continuous; cyclic	0.024	-	-
RMP	B	Cyclohexane treater vents	Intermittent	(see stream Z)	-	-
RMP	C	Catalyst preparation vent	Generally continuous	(see stream Z)	-	-
PR	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[d]	Recycle ethylene treaters	Continuous	12.7	21	61.0 ethylene[e] 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[f] - 0.05 steam and air-99.959
PF	J	Stripper vents	Continuous[h]	0.85	40	VOC[f]- 0.01 air - 99.99
	Z	(All indicated streams to flare)		14.8	-	VOC[i] air
		Total Emission Rate		29.0		

[a]Source of information: Industry correspondences.

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

[c]See Figure 3-7 for stream identification.

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

[e]Percent by weight.

[f]Cyclohexane.

[g]Mostly steam.

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

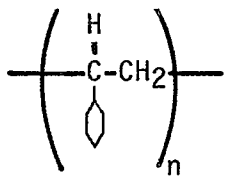
[i]Includes ethylene, cyclohexane, and other VOC comonomers. Ethylene is emitted at about 4.35 kg VOC/Mg product; cyclohexane at about 8.7 kg VOC/Mg product; and other VOC at about 1.74 kg VOC/Mg product.

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 Gas Phase Process. The HDPE gas phase process is similar to the LDPE low-pressure process described earlier (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 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 process serve also for production of HDPE when the same process is used.

3.2.4 Polystyrene

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:



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 impact 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.) Where a blowing (or expanding) agent is added to the polystyrene, the product is referred to as an expandable polystyrene. This may be done as part of the process (as in the production of expandable beads) or as part of the fabrication process (as in foamed polystyrene applications.)

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.[29] The suspension process is also commonly used, especially in the production of expandable beads.[29] The use of the emulsion process for producing homopolymer of styrene has decreased significantly since the mid-1940's.[30]

This section describes both the bulk (mass) batch and continuous processes for the production of crystal or impact polystyrene in Subsections 3.2.4.1 and 3.2.4.2, respectively, and the suspension process for the production of expandable polystyrene beads in Subsection 3.2.4.3.

3.2.4.1 Polystyrene Batch Process. 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 Process description. 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

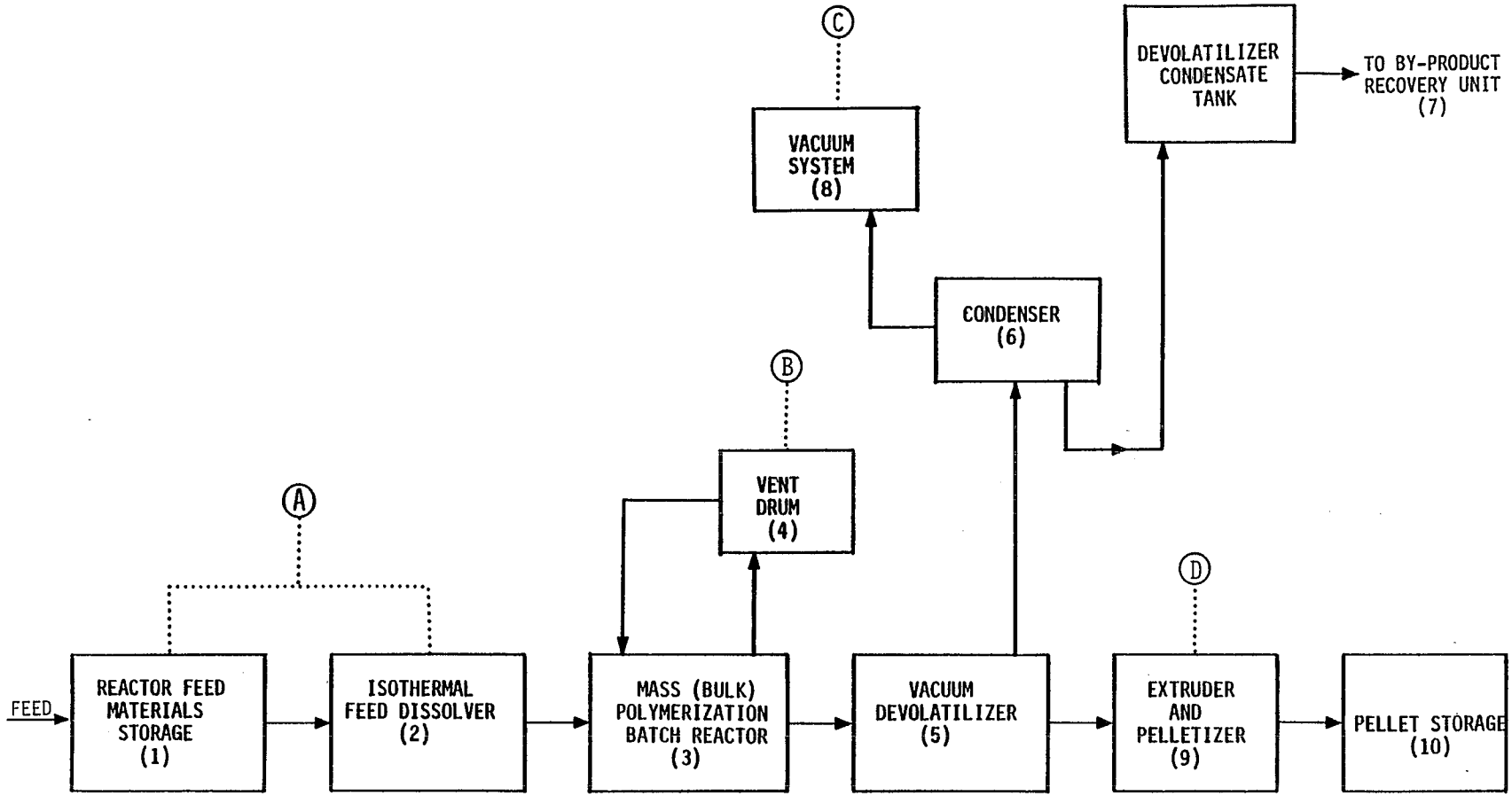


Figure 3-8. Simplified Process Block Diagram for the Polystyrene Batch Process

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).

3.2.4.1.2 Emissions from the polystyrene batch process. 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 0.61 to 2.5 kg VOC/Mg product. The major vent is the devolatilizer condenser vent (Stream C). 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 D) 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 D) 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 B). 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.

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.

3.2.4.2 Polystyrene Continuous Process. 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 Process description. The bulk (mass) continuous process is represented in Figure 3-9. The feed dissolver tank (1) is charged 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. Non-condensables (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

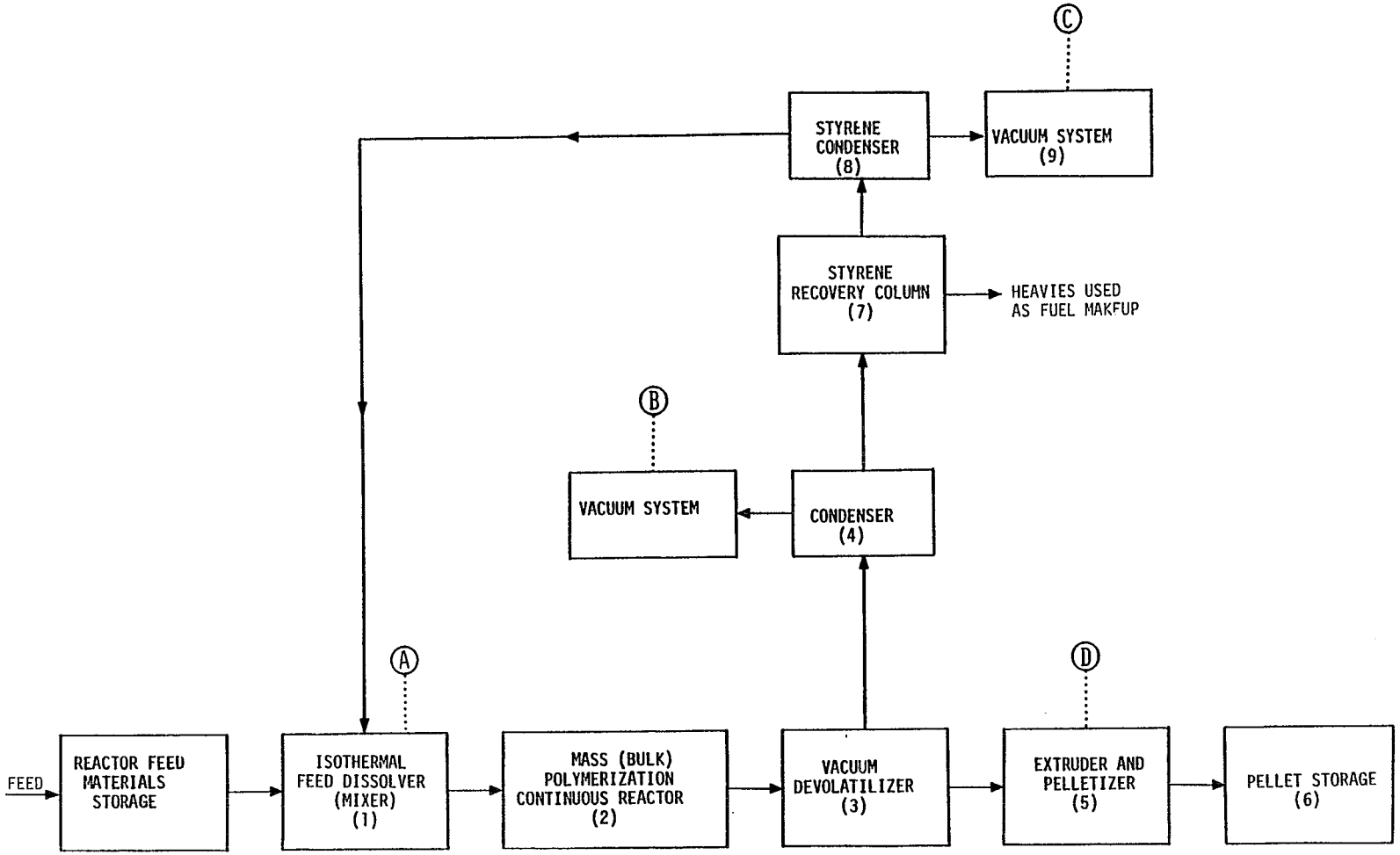


Figure 3-9. Simplified Process Block Diagram for the Polystyrene Continuous Process

(9). Column bottoms containing low molecular weight polymers are sometimes used as a fuel supplement.

3.2.4.2.2 Emissions from the polystyrene continuous process. 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 depending, in part, on the type of vacuum system used. 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. It is estimated that the typical total VOC emission rate for plants still using steam ejectors is about 3.25 kg VOC/Mg product. For plants that use vacuum pumps, it is estimated that the total VOC emission rate is about 0.26 kg VOC/Mg product. Table 3-13 presents the VOC vent stream characteristics.

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.

3.2.4.3 Expandable Polystyrene Suspension Process. The suspension process is a batch polymerization process that may be used to produce crystal, impact, or expandable polystyrene beads. An expandable polystyrene (EPS) bead consists typically of high-molecular weight crystal grade polystyrene and 5-8 percent low boiling aliphatic hydrocarbon blowing agent (typically pentane or isopentane although other blowing agents, such as esters, alcohols, and aldehydes, can be used).[31] When used to produce an EPS bead, the suspension process can be adapted in one of two ways for the impregnation of the bead with the blowing agent. One method is to add the blowing agent to a reactor after polymerization. The other method is to add the blowing agent to

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.

the monomer prior to polymerization. The former method (which shall be referred to as the "post-impregnation" suspension process) is more common than the latter method[32] (which shall be referred to as the in-situ suspension process). Both process are described below.

3.2.4.3.1 Process description. The post-impregnation suspension process is essentially a two-part process using two process trains in series. In the first process train, raw styrene monomer is polymerized and a finished polystyrene bead is produced. The second process train takes the finished bead from the first train, impregnates the bead with a blowing agent, and produces a finished EPS bead. Figure 3-10 is a schematic representation of this process.

In the first train, styrene monomer, water, initiator, and suspending agents form the basic charge to the suspension reactor (1).[33] The styrene to water ratio, about one-half to one-quarter monomer to water volume ratio,[33] varies depending on the type of polystyrene required. Initiators are commonly used because the reaction temperature is usually too low for adequate thermal initiation of polymerization. Suspending agents are typically protective colloids and insoluble inorganic salts. Protective colloids are added to increase the viscosity of the continuous water phase and insoluble inorganic salts, such as $MgCO_3$, are added to prevent coalescence of the drops upon collision.[33]

In the reactor, the styrene is dispersed as droplets throughout the water phase. The droplet size may range from about 0.1 to 1.0 mm.[34] The reactor is heated to start the polymerization process, which takes place within the droplets. An inert gas, such as nitrogen, is frequently used as a blanketing agent in order to maintain a positive pressure at all times during the cycle and prevent air leaks. Once polymerization starts, temperature control is typically maintained through a water-cooled jacket around the reactor and facilitated by the added heat capacity of the water in the reactor. The size of the product bead depends on the strength of agitation and on the nature of the monomer and suspending system.[33] Between 20 and 70 percent conversion, agitation becomes extremely critical.[35] If agitation weakens or stops between these limits, excessive agglomeration of the polymer particles may occur, followed by a runaway reaction. Polymerization typically occurs

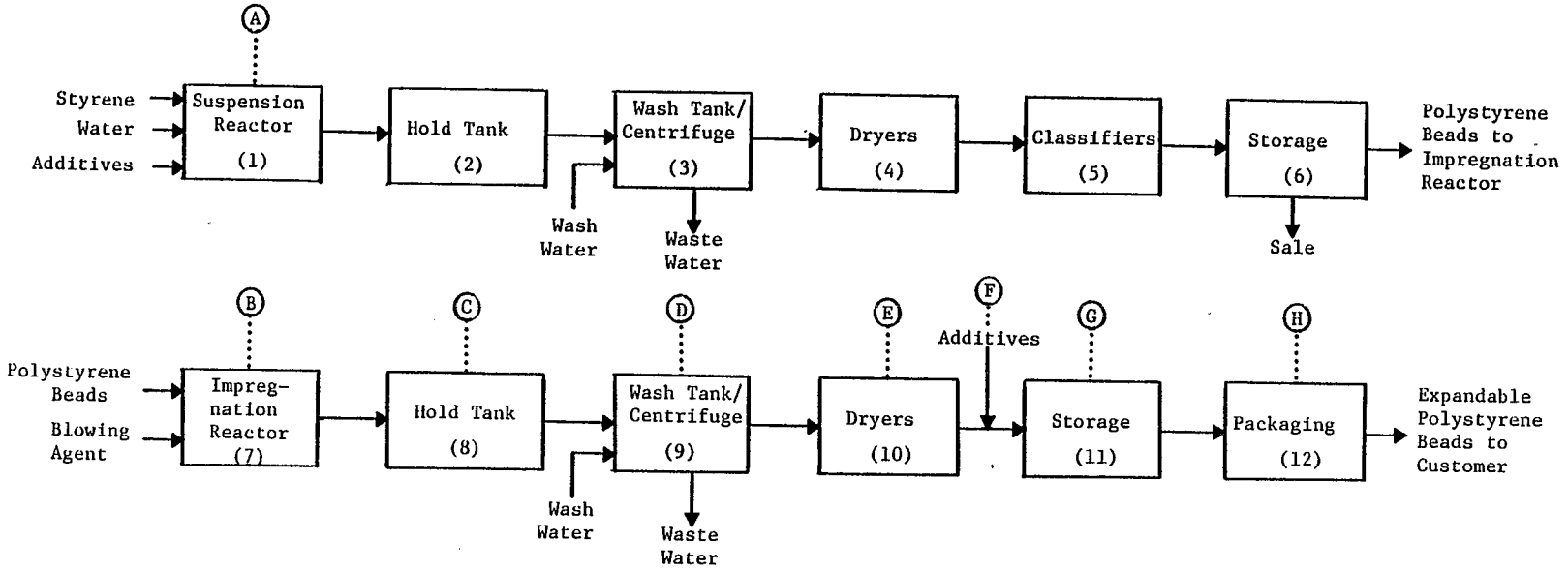


Figure 3-10. Simplified Process Block Diagram for the Expandable Polystyrene Post-Impregnation Suspension Process

within several hours, the actual time varying primarily with the temperature and the amount and type of initiator(s) used.[36]

Once the reaction has been completed, the polystryene-water slurry is normally pumped from the reactor to a hold tank (2), which has an agitator to maintain dispersion of the polymer particles. Hold tanks have at least three functions the polymer-water slurry is cooled to below the heat-distortion temperature of the polymer (generally 50 to 60°C (122 to 140°F), chemicals are added to promote solubilization of the suspension agents, and the tank serves as a storage tank until the slurry can be centrifuged.[37] From the hold tanks, the polymer-water slurry is fed to a centrifuge (3) where wash water is added and the slurry is then centrifuged to separate the wash water from the solids. The wash water is discarded and the polymer product, which may retain between 1 and 5 percent water,[37] are sent to dryers (4). From the dryers, the product beads may be sent to a classifer (5), which separates the beads according to size, and then to storage bins or tanks (6). The product beads do not always meet specified criteria for further processing into expandable beads. In these instances, the "off-spec" beads may be processed and sold as crystal (or possibly impact) polystyrene.

In the second train, the product bead (from the storage bins of the first train), water, blowing agent, and any desired additives are added to an impregnation reactor (7). The beads are impregnated with the blowing agent through the utilization of temperature and pressure. Upon completion of the impregnation process, the bead-water slurry is transferred to a hold tank (8). Acid may be fed to the hold tank and part of the water drained as wastewater. From the hold tanks, the slurry is washed and dewatered in centrifuges (9) and then dried in low temperature dryers (10). From the dryers, the EPS bead may undergo sizing, if not already done, before being transferred to storage silos (11) or directly to packaging (12) for shipment to the customer. In some instances, additives may be applied to the EPS bead before shipping to improve process characteristics.

The in-situ suspension process is shown schematically in Figure 3-11. The major difference between this process and the post-impregnation suspension process is that polymerization and impregnation takes place

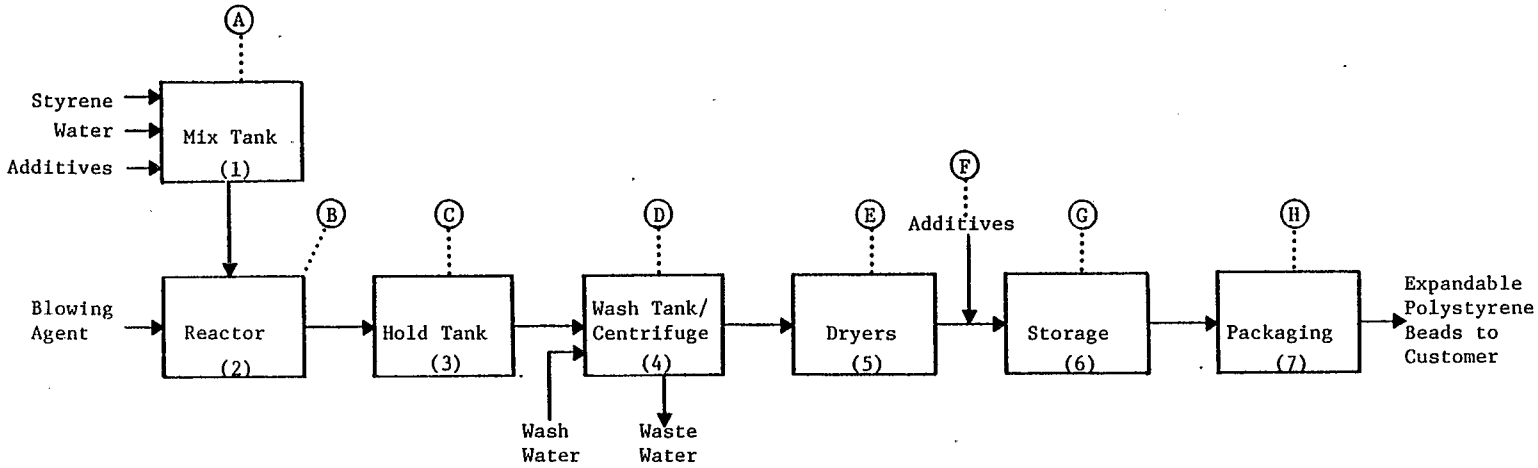


Figure 3-11. Simplified Process Block Diagram for the Expandable Polystyrene In-Situ Suspension Process

at the same time in a single reactor. The reaction mixture, composed of styrene monomer, water, polymerization catalysts, and additives, are charged to a reactor (1) to which a blowing agent is added.[38] The styrene monomer is polymerized at elevated temperatures and pressure in the presence of the blowing agent so that the blowing agent is entrapped in the polymerized bead, which may contain 5-7% blowing agent.[38] After polymerization and impregnation has taken place, the EPS bead-water slurry follows essentially the same steps as in the post-impregnation suspension process. These steps are repeated in Figure 3-11 from Figure 3-10.

3.4.4.3.2 Emissions from the expandable polystyrene suspension process. Table 3-14 summarizes the VOC emission sources and stream characteristics for the post-impregnation suspension process. Table 3-15 summarizes the VOC emission sources and stream characteristics for the in-situ suspension process.

Available information on the post-impregnation suspension process shows a rather large variation in emission rates for some of the emission points. For two plants for which complete information was available, overall emission rates of about 4.2 and 9.5 kg VOC per Mg of product were estimated. An emission rate of about 5 kg VOC per Mg of product was estimated for an in-situ suspension process. The blowing agent, which is continually diffusing out of the bead in the manufacturing process and during storage, constitutes almost all of the VOC emissions from both processes. A small amount of styrene is emitted from the suspension reactors in the post-impregnation process and from the mix tanks and reactors in the in-situ process.

3.2.5 Polyester Resin

Polyester resins used for fiber production may be classified chemically as either poly(ethylene terephthalate) [PET] resins or non-PET resins. Of these two types, PET resins are the most important polyester as a thermoplastic synthetic fiber. PET resins may, in turn, be classified as low viscosity or high viscosity resins. Low viscosity resins are used in such applications as fiber and bottle production. High viscosity resins are used in such applications as tire cord.

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

Table 3-14. CHARACTERISTICS OF VENT STREAMS FROM THE EXPANDABLE POLYSTYRENE POST-IMPREGNATION PROCESS[a]

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition, wt. %
PR	A	Suspension reactor vents	Intermittent	0.03 - 0.18	-	-	62 VOC[d] 38 Water Vapor
PF	B	Impregnation reactor vents	Intermittent	e - 0.21	-	-	VOC Air
PF	C	Holding tank vents	Intermittent	3.81	21	0	1.2 VOC[f] 98.8 Air
PF	D	Wash tank vents	Continuous	1.52 - 2.8	21	0	0.11 VOC[f] 99.89 Air
PF	E	Dryer vents	Continuous	0.64 - 1.48	21	0	0.02 VOC[f] 99.98 Air
PF	F	Product improvement vents	Intermittent	c			
PS	G	Storage vents	Continuous	0.4 - 2.27	21	0	0.17 VOC[f] 99.83 Air
PS	H	Packaging system losses	Intermittent	0.2 - 1.29	21	0	0.11 VOC[f] 99.89 Air
		Total Emission Rate		4.18[h] - 9.47[h]			

[a]Source of information: Industry correspondence.

[b]PR = polymerization reaction; PF = product finishing; PS = product storage.

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

[d]Styrene monomer.

[e]Negligible.

[f]Blowing agent.

[h]Sum of emission rates for one plant for which a complete set of emission rates were available.

Table 3-15. CHARACTERISTICS OF VENT STREAMS FROM THE EXPANDABLE POLYSTYRENE IN-SITU PROCESS[a]

Process Section[b]	Stream[c]	Name	Nature	Emission rate, kg VOC/Mg product	Temperature, °C	Pressure, psig	Composition, wt. %
RMP	A	Mix tank vents	Intermittent	0.123	21	0	0.75 Styrene 0.53 ACN 98.72 Air
PR	B	Reactor vents	Intermittent	1.14	35	0	58.8 VOC[d] 1.4 Styrene 39.8 Air
PF	C	Holding tank vents	Intermittent	0.055	24 - 35	0	0.66 VOC[d] 99.34 Air
PF	D	Wash tank vents	Continuous	0.024	27	0	0.35 VOC[d] 99.65 Air
PF	E	Dryer vents	Continuous	2.89	32 - 38	0	0.066 - 0.27 VOC[d] 99.934 - 99.73 Air
PF	F	Product improvement vents	Intermittent	e	27 - 38	0	0.002 VOC[d] 99.998 Air
PS	G	Storage vents	Continuous	0.746	21 - 38	0	0.002 VOC[d] 99.998 Air
		Total Emission Rate		4.98			

[a]Source of information: Industry correspondence.

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

[c]See Figure 3-11 for stream identification.

[d]Blowing agent.

[e]Negligible.

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. Most, if not all, high viscosity PET is produced in plants using the TPA process. 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.[39]

3.2.5.1.1 Process description. A schematic diagram of the continuous DMT process is presented in Figure 3-12. The primary reaction is:

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.

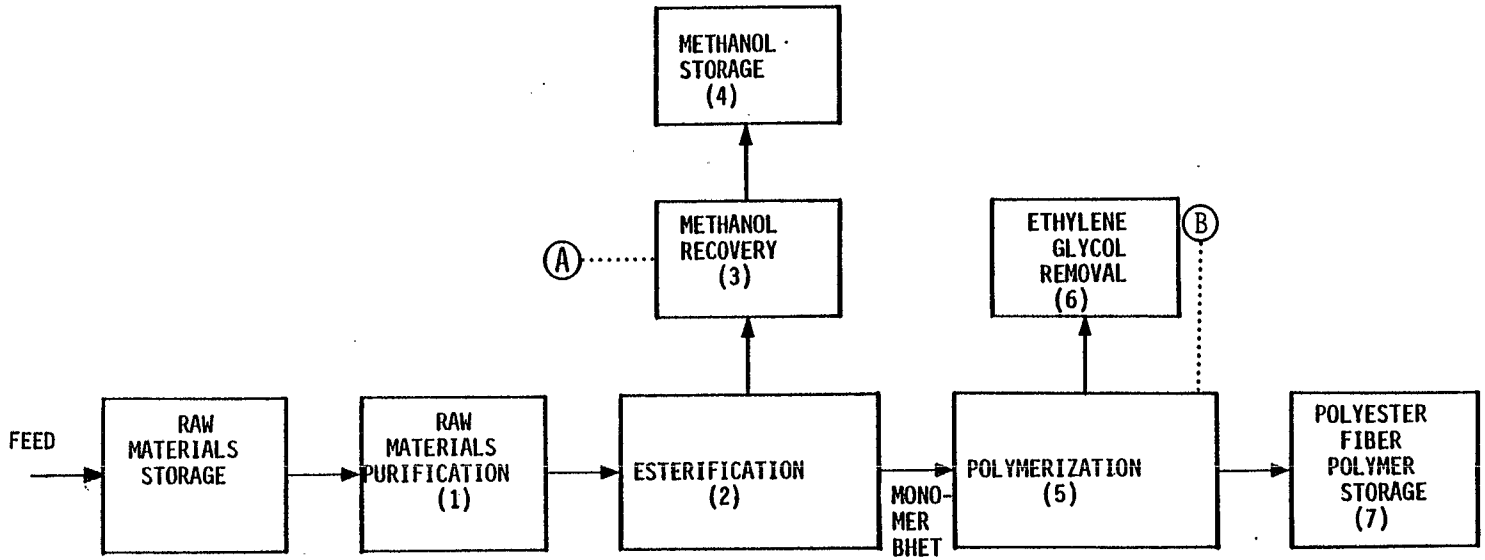


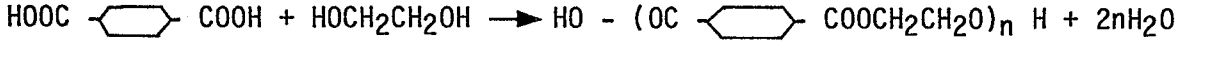
Figure 3-12. Simplified Process Block Diagram for the PET/DMT Process

3.2.5.1.2 Emissions from the PET/DMT process. 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 3-12. A summary of the vent stream characteristics is provided in Table 3-16. 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 streams from the reactors are 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 the number of polymerizers used and on the type of system used to recover the ethylene glycol. 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 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 3-13. The primary reaction is:



TPA EG PET

Table 3-16. CHARACTERISTICS OF VENT STREAMS FROM THE POLY(ETHLYENE TEREPHTHALATE) DMT PROCESS[a]

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 Reactors Cooling Tower Water	Continuous		25	0.2	VOC Air H_2O
		- with spent ethylene glycols spray condenser		0.355[d]			
		- without spray EG spray condenser		9.5[e]			
		Total Emission Rate					
		- with EG spray condenser		0.535			
		- without EG spray condenser		9.68			

[a]Source of information: Industry correspondence.

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

[c]See Figure 3-12 for stream identification.

[d]Assumed same for the total TPA emission rate with EG spray condenser.

[e]Estimated to be same as for the TPA process. All emissions from cooling water tower. The actual emission rate will vary depending upon the number of polymerization reactors per line at a plant.

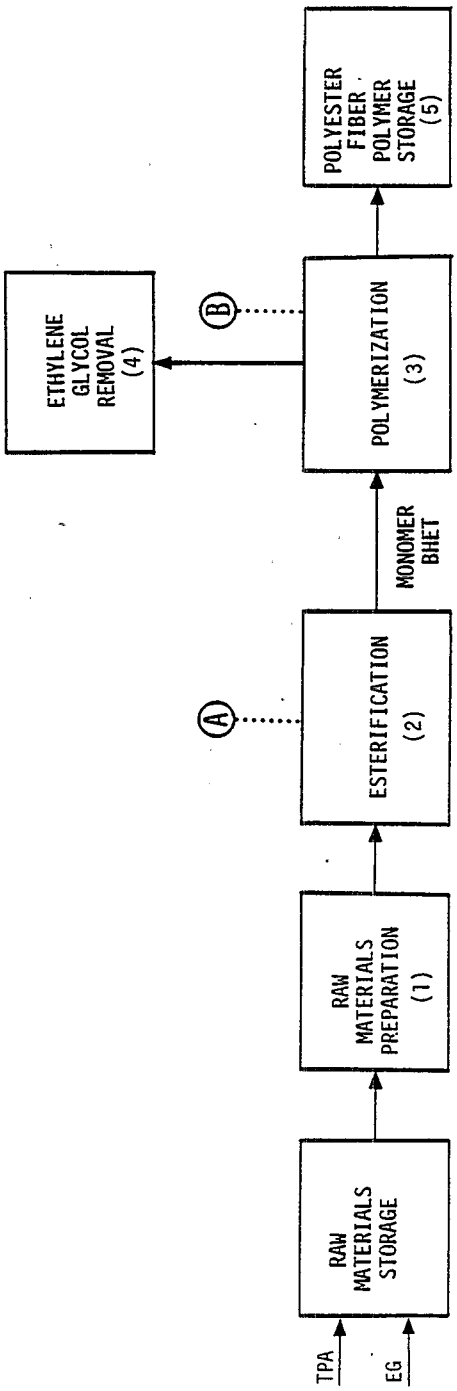


Figure 3-13. Simplified Process Block Diagram for the PET/TPA Process

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.

Polymerization is carried out as with the DMT process - in two or more reactors in series operated at increasing temperatures and successively lower pressures. In the production of high viscosity PET, there are two processes currently being used that are distinct from each other. One process can be described as essentially the low viscosity process with a single, additional end or final polymerizer. In this process, the first polymerization reactor, which may be called a prepolymizer, pre-polycondensation reactor, or low polymerizer, operates at temperatures between 220 and 290°C (430 and 555°F) and under vacuum [e.g., between 10 and 20 torr (10 mm Hg to 20 mm Hg)].[40,41] These temperature conditions are slightly higher and the vacuum level range is slightly smaller than for the production of low viscosity PET. The additional reactor, which may be called an end finisher or high polymerizer, may operate with a temperature of between 290 and 300°C (555 and 570°F) and a vacuum level of approximately 0.9 torr (0.9 mm Hg).[41,42]

The other high viscosity process contains a greater number of process vessels. The difference of primary importance is the number and operating conditions of the polymerization vessels. In this second process, the production of high viscosity PET takes place in two or more end finishers rather than in one as in the process discussed above. The first end finisher is typically operated with an intermediate vacuum level of about 2 torr (2 mm Hg). The polymer leaving this reactor then enters a second end finisher, which may have a vacuum level typically around 0.5 torr (0.5 mm Hg)[43] and as low as 0.25 torr (0.25 mm Hg).[44]

3.2.5.2.2 Emissions from the PET/TPA process. There are two sources of VOC emissions from this process as illustrated by Figure 3-13. These streams, which are continuous, are emitted from the esterifiers (Stream A), and from the polymerizers (Stream B). Offgas characteristics for these streams are summarized in Table 3-17.

Table 3-17. CHARACTERISTICS OF VENT STREAMS FROM THE POLY(ETHYLENE TEREPHTHALATE) TPA PROCESS[a]

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_2O
PR	B	Polymerizer Reactors Cooling Water Tower	Continous		25	0.2	VOC Air H_2O
		- with spent ethylene glycol spray condenser		0.315[e]			
		- without spent ethylene glycol spray condenser		9.5[f]			
		Total Emission Rate					
		- with EG spray condenser		0.355			
		- without EG spray condenser		9.5			

[a]Source of information: Industry correspondence.

[b]RMP = raw material preparation; PR = polymerization reaction.

[c]See Figure 3-13 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 0.295 kg VOC/Mg product from cooling water tower.

[f]Includes potential emissions from Stream A. The actual emission rate will vary depending upon the number of polymerization reactors per line at a plant.

The overall emission rate from the production of low viscosity PET is approximately 0.36 kg VOC per Mg of product. Of this amount, about 0.04 kg VOC per Mg of product is from the esterifiers and about 0.32 kg VOC per Mg of product is from the polymerization reactors via the cooling tower.

The overall emission rate from the production of high viscosity PET depends upon the particular process. In the first high viscosity PET process discussed above, the overall emission rate is about 0.36 kg VOC per Mg of product. Of this 0.36 kg VOC per Mg of product, about 0.04 kg VOC per Mg of product is from the esterifiers, which are controlled by reflux condensers, and 0.32 kg VOC per Mg of product is from the polymerization reactors via the cooling tower. In the second high viscosity PET process discussed above (i.e., where two or more end finishers are used), the overall emission rate has been estimated to be about 3.55 kg VOC per Mg of product. Of the overall emission rate, about 0.15 kg VOC per Mg product is from the esterifiers, which are controlled by distillation columns, and about 3.4 kg VOC per Mg product from the polymerizers via the cooling tower.

The difference in the emission rates from the polymerization reactors (via the cooling tower) between the two high viscosity processes is due to the ethylene glycol recovery system used with the two types of high viscosity PET processes. In the first process, ethylene glycol is recovered directly from the reactors by spray condensers and from the esterifiers by reflux condensers. In the second process, ethylene glycol emisions from the polymerizers are allowed to pass directly to the vacuum system and into the cooling tower. The ethylene glycol is then recovered from the water in the cooling tower. This arrangement allows for a higher ethylene glycol concentration in the cooling tower and a larger loss of ethylene glycol to the atmosphere from windage.

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 other industries. 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 45 through 48.

Table 3-18 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."

Data characterizing the uncontrolled levels of fugitive emissions specifically for 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 several other industries. 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 and because the chemicals processed in polymer and resin manufacturing are the same chemicals that form a large portion of the SOCMI data base, the SOCMI fugitive emission data are used to approximate the levels of fugitive emissions in the polymers and resins industry. The SOCMI emission factors, which are used for the purposes of this project, are presented in Table 3-19. (See Reference 48 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

Table 3-18. VAPOR PRESSURES OF MAJOR ORGANIC COMPOUNDS USED IN THE POLYMERS AND RESINS MANUFACTURING

Organic Compound	Vapor Pressure @20°C (kPa)	Process
Ethylene	gas	PP,PE
Ethane	gas	PE
Propylene	gas	PP,PE
Propane	gas	PP
Iso-butane	gas	PE
Butene	gas	PE
Methanol	12.3	PET (DMT)
Isopropyl Alcohol	4.4	PP,PE
Ethylbenzene	1.0	PS
Styrene	0.7	PS
Ethylene glycol	0.01	PET

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

Source: Weber, R.C., P.A. Parker, and M. Bowser. Vapor Pressure Distribution of Selected Organic Chemicals. EPA-600/2-81-021. IERL, U.S. EPA, Cincinnati, Ohio. February 1981. (Docket Reference Number II-A-38.)

Table 3-19. UNCONTROLLED FUGITIVE EMISSION RATES

Fugitive emission source	Uncontrolled emission rate,[a] kg/hr/source
Valves	
Gas	0.0056
Light liquid[b]	0.0071
Heavy liquid[c]	0.0023
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 48.

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

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

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

Low-pressure LDPE and gas phase HDPE 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.[49]

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 used to produce a crystal or impact grade polystyrene 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. Similarily, no routine control is applied in expandable polystyrene plants. There is some use of condensers or combustion devices, but control varies widely between plants. Flares are not usually installed in any polystyrene plant.

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 Current State VOC Regulations. 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 46 and 47.

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 Fugitive emissions regulations. 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 moves down into the stack.

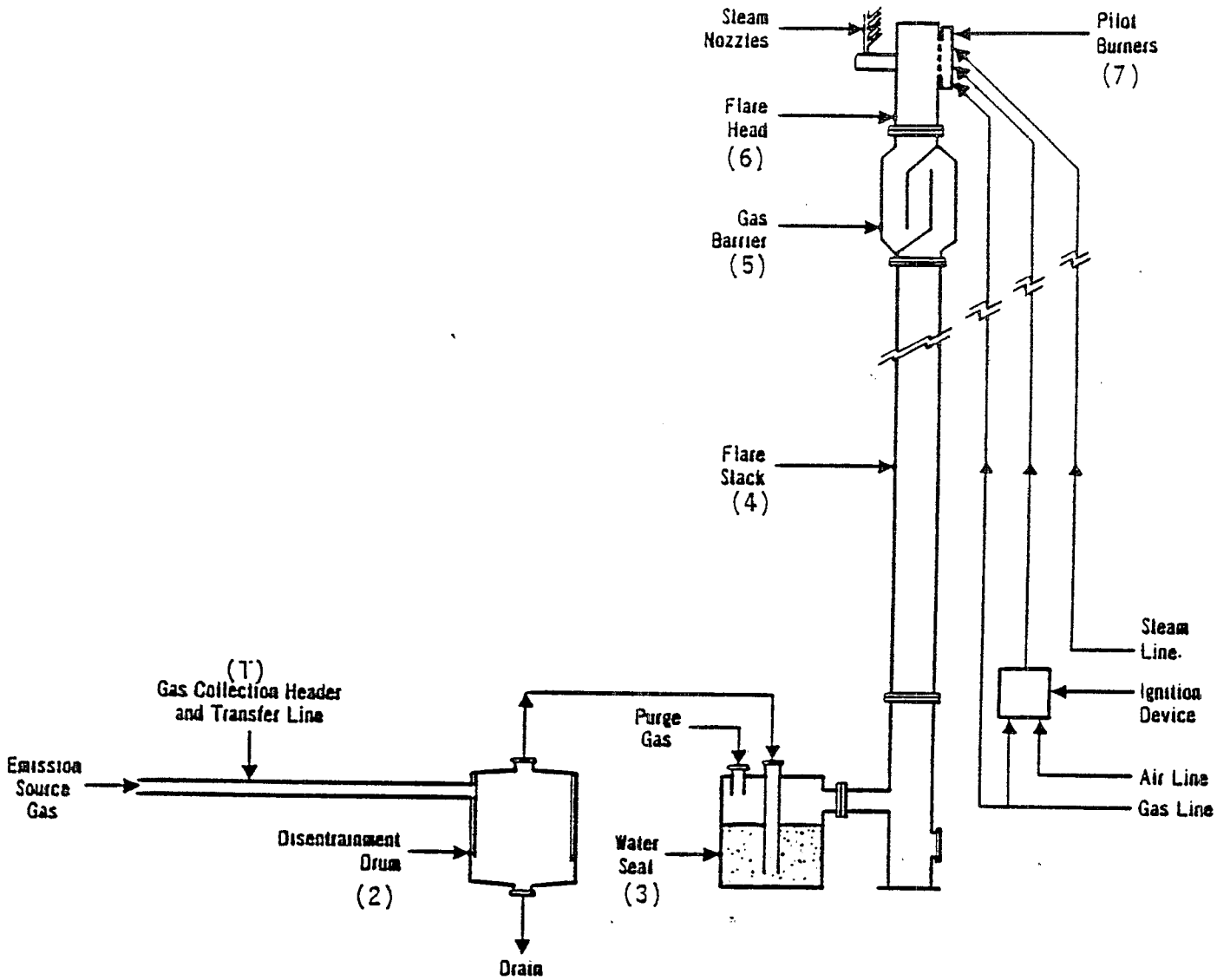

Figure 4-1. Steam Assisted Elevated Flare System

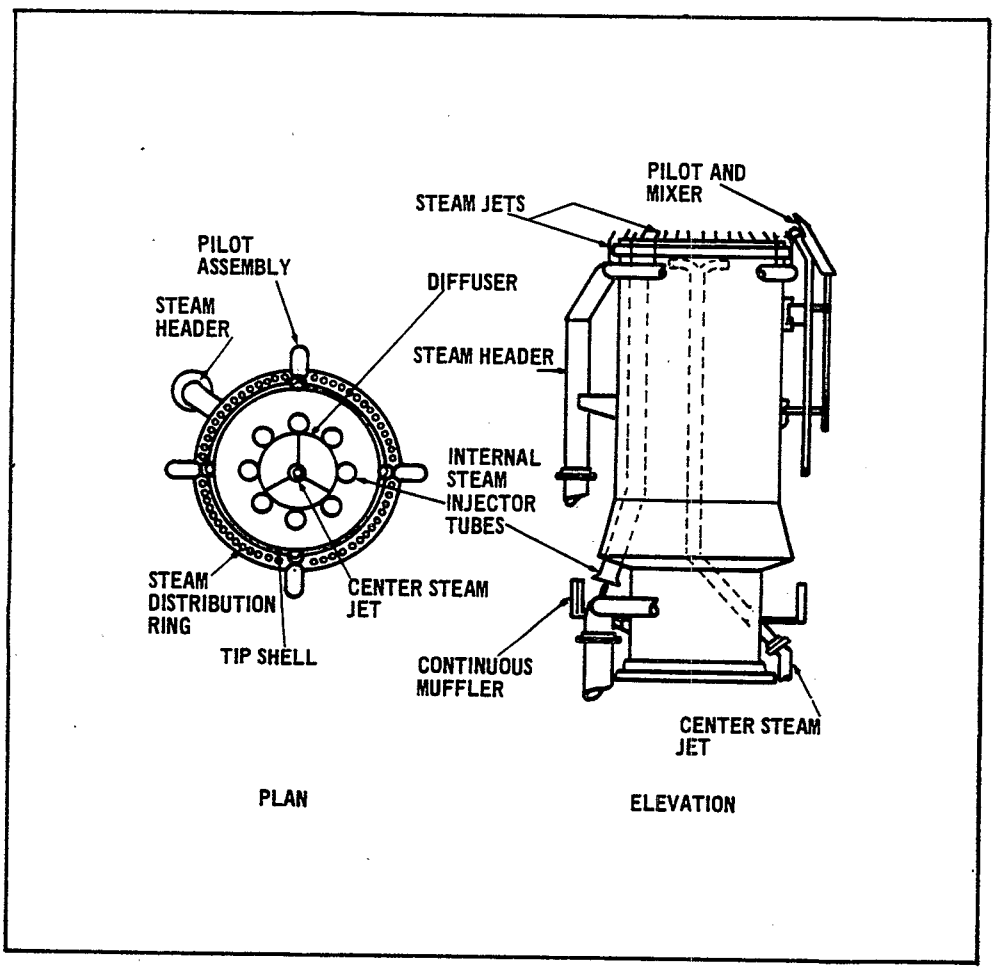

Figure 4-2. Steam Injection Flare Tip

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 hydrocarbon in the 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 x 10^6 J/sec (0.4-2.1 x 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 x 10^6 J/sec (235 x 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 x 10^6 J/sec (0.3 x 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 x 10^6 J/sec (44 x 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 x 10^6 J/sec (83.7 x 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 OMcDaniel, et al. (1982) o 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.

These results have been recently supplemented by tests conducted under EPA's flare testing program. A summary of these results is found

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 H.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	

in Docket Reference Number II-B-94. One result of these more recent tests is the determination of a broader range of exit velocity and heat content parameters under which steam-assisted and nonassisted flares can achieve 98 percent combustion efficiency.

After consideration of the results of these 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. In addition, steam-assisted and nonassisted flares can obtain 98 percent combustion efficiencies with exit velocities of up to 120 m/sec (400 ft/sec) provided the heat content of the gas stream is at least 37.3 MJ/scm (1,000 Btu/scf). 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:

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

where: v = maximum gas velocity in ft/sec, standard conditions.
HC = heat content of the combusted gas in Btu/scf.

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

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 4-3. (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 4-4. 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

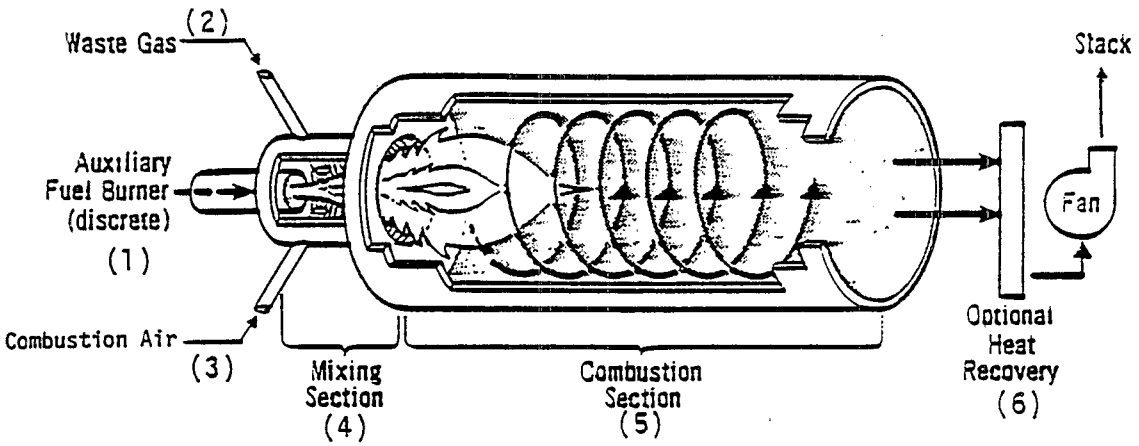

Figure 4-3. Discrete Burner Thermal Incinerator

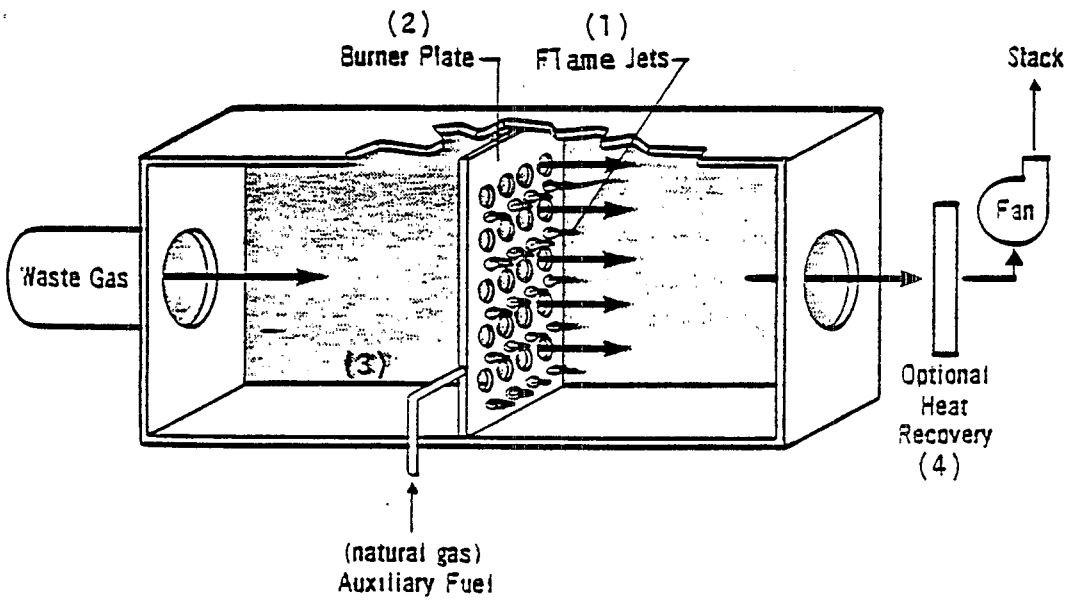

Figure 4-4. Distributed Burner Thermal Incinerator

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 Applicability. 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 Catalytic Incinerators. 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 4-5. 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

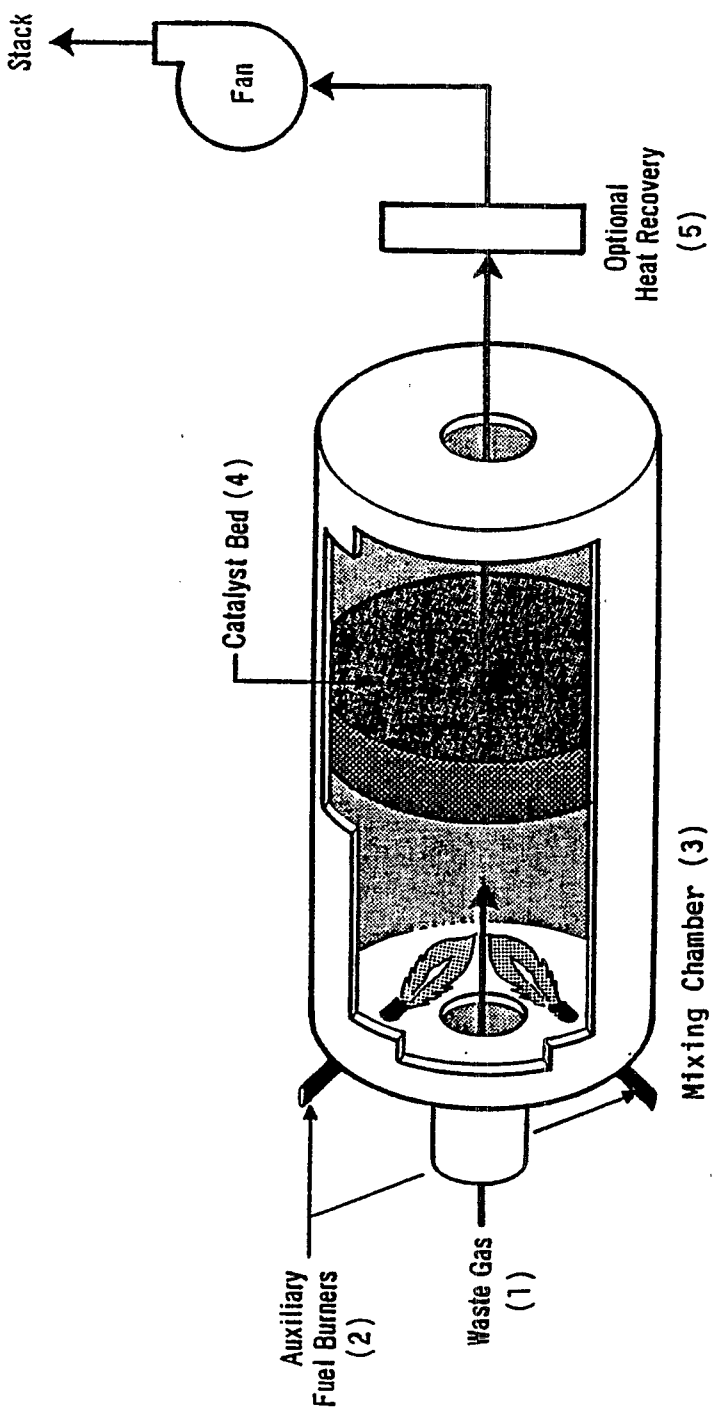

Figure 4-5. Catalytic Incinerator

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 Industrial Boilers. Fireboxes of boilers (and process 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.

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. A high intensity or vortex burner can be effective for vent streams with low heating values (i.e., streams where a conventional burner may not be applicable). Effective combustion of low heating value streams is accomplished in a high intensity burner by passing the combustion air through a series of spin vanes to generate a strong vortex.

4.1.1.4.1 <u>VOC destruction efficiency</u>. Furnace residence time and temperature profiles vary for industrial boilers depending on the furnace and burner configuration, fuel type, heat input, and excess air level.[31] A mathematical model has been developed which estimates the furnace residence time and temperature profiles for a variety of industrial boilers.[32] This model predicts mean furnace residence times of from 0.25 to 0.83 seconds for natural gas-fired water tube boilers in the size range from 15 to 150 x 10^6 Btu/hr. Furnace exit temperatures for this range of boiler sizes are at or above 1,200°C (2,200°F) with peak

furnace temperatures occurring in excess of 1,540°C (2,810°F). Residence times for oil-fired boilers and natural gas-fired boilers are similar. A boiler (or process heater) furnace can be compared to an incinerator where the average furnace temperature and residence time determine the combustion efficiency. However, when a vent gas is injected as a fuel into the flame zone of a boiler (or process heater), the required residence time is reduced due to the relatively high flame zone temperature. The following test data, which document the destruction efficiencies for industrial boilers and process heaters, are based on injecting the wastes identified into the flame zone of each combustion control device.

An EPA sponsored test was conducted in an effort to determine the destruction efficiency of an industrial boiler for polychlorinated biphenyls (PCB's).[33] The results of this test indicated that the PCB destruction efficiency of an oil-fired industrial boiler firing PCB-spiked oil was greater than 99.9 percent. This efficiency was determined based on the PCB content measured by a gas chromatagraph in the fuel feed and flue gas.

Firebox temperatures for process heaters show relatively wide variations depending on the application. Tests were conducted by EPA to determine the benzene destruction efficiency of five process heaters firing a benzene offgas and natural gas mixture.[34,35,36] The units tested are representative of process heaters with low temperature fireboxes (reboilers) and medium temperature fireboxes (superheaters). Sampling problems occurred while testing one of these heaters and, as a result, the data for that test may not be reliable and are not presented.[34] The reboiler and superheater units tested showed greater than a 98 percent overall destruction efficiency for C_1 to C_6 hydrocarbons.[35] Additional tests conducted on a second superheater and a hot oil heater showed that greater than 99 percent overall destruction of C_1 to C_6 hydrocarbons occurred for both units.[36] These efficiencies were determined based on the benzene content measured by a gas chromatagraph in the fuel feed and flue gas.

4.1.1.4.2 Applicability. Use of a boiler for VOC emission control in the polymers and resins industry is not common. Despite the potential problems, boilers are being used in at least two polypropylene plants[37]

and a high-density polyethylene plant.[38] 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 Condensers. 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 of 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 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,[39] 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.[40] 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).[41] 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."[42]

The major pieces of equipment used in a typical refrigerated surface condenser system are shown in Figure 4-6.[43] 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).[44] Temperatures as low as -62°C (-80°F) may be necessary to condense some VOC streams.[39]

4.1.2.1.1 Condenser control efficiency. 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 refrigerated condenser at an increased cost. Outlet concentrations for low boiling organics may be above 10,000 ppmv to 20,000 ppmv.[45]

4.1.2.1.2 Applicability of condensers. Water-cooled condensers are effective in reducing potential emissions of high boiling, easily

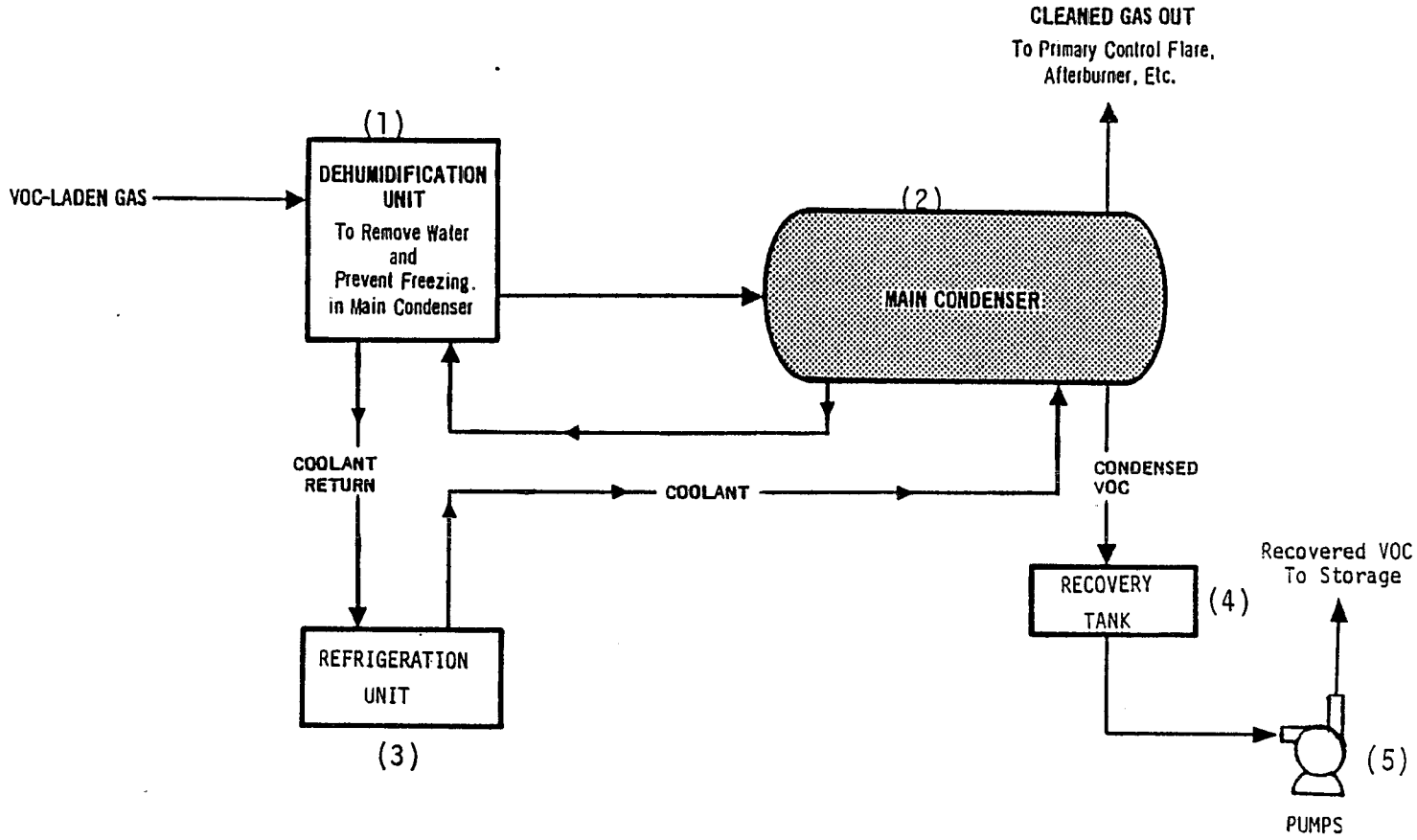

Figure 4-6. Condensation System

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 4-7 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 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

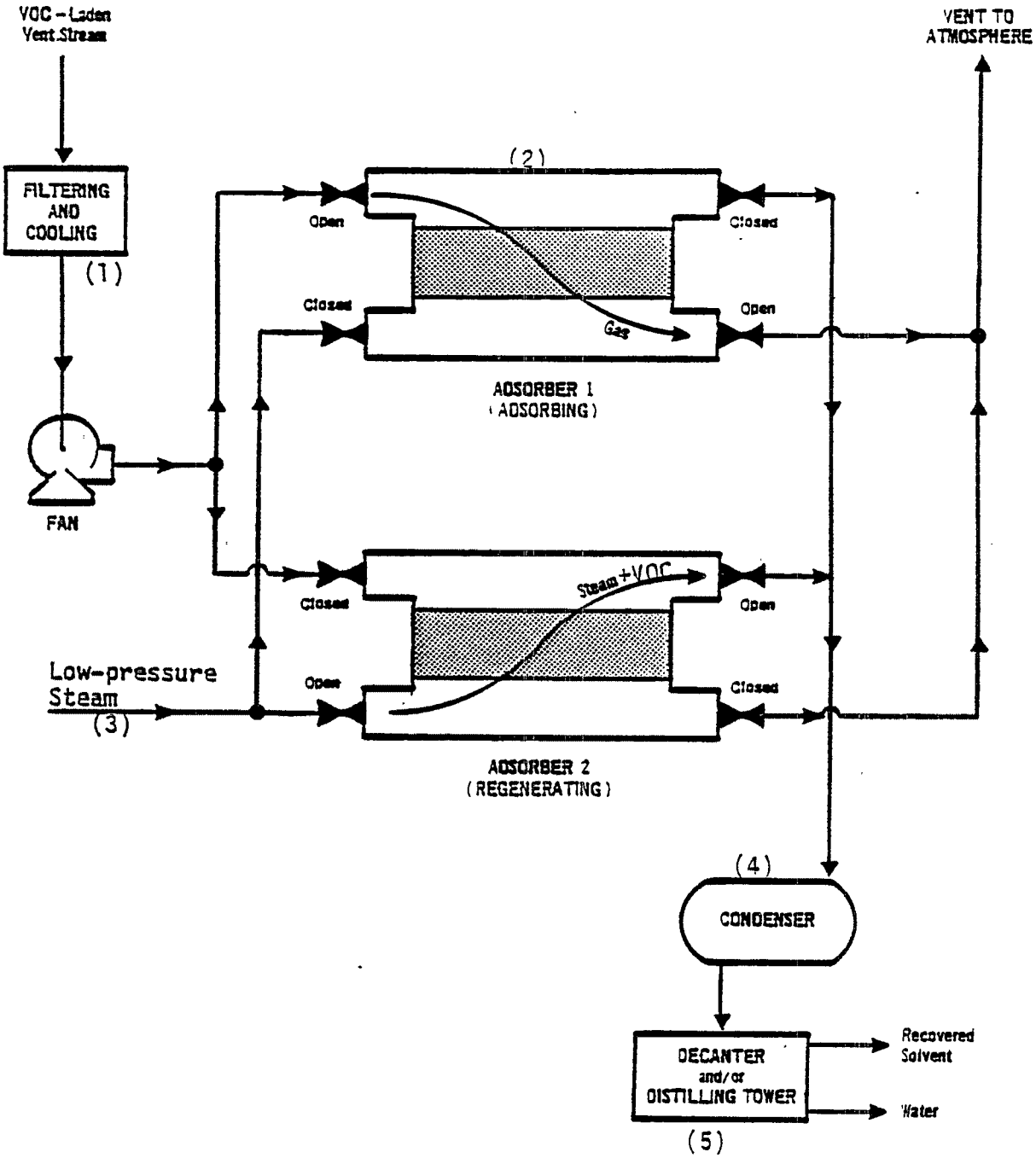

Figure 4-7. Two Stage Regeneration Adsorption System

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.[46] Removal efficiencies of 95 to 99 percent are achieved by well-designed and well-operated units.[47]

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 concentrations of organics are usually limited to about 0.5 to 1 percent.[44] The maximum practical inlet concentration is about 1 percent, or 10,000 ppmv.[48] 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) 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.[49]

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.[50] 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.[51]

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.[52] 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.[53]

A schematic of a packed tower is shown in Figure 4-8. 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.

4.1.2.3.1 Absorber control efficiency. 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.

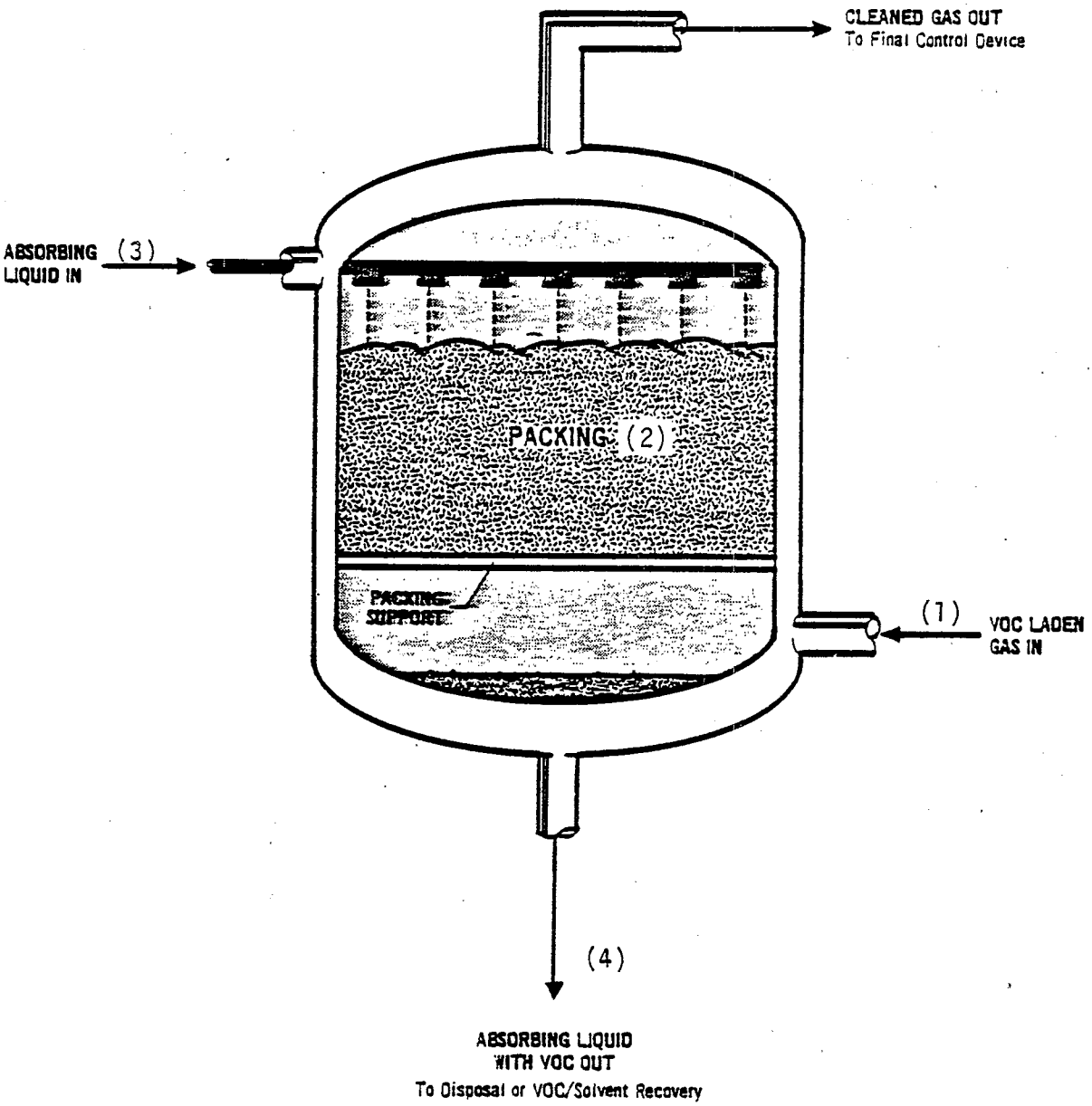

Figure 4-8. Packed Tower for Gas Absorption

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 that was 98 percent efficient.[54]

4.1.2.3.2 Applicability. The selection of absorption for VOC control depends on the availability of an appropriate solvent for the specific VOC. Absorption is sually not considered when the VOC concentration is below 200-300 ppmv.[55]

The use of absorbers is generally limtied 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. Abosrbers 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 appleid 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 leaking fugitive emissions sources are located and repaired at specified intervals. The second is a preventive approach whereby fugitive emissions are either captured and vented to a control device or eliminated through the installation of specified controls or leakless equipment. The technical application of these methods is briefly explained in the following subsections.

4.2.1 Leak Detection and Repair Program

4.2.1.1 Leak Detection. 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.[56]

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 packing gland. 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.[57] 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.

To implement a monitoring program, an instrument meter reading (organics concentration) which will indicate an equipment leak must be defined. The meter reading selected may vary from 1,000 ppmv to 100,000 ppmv.[58] 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 potential increase in emissions that can occur through attempts at repairing relatively minor leaks may dictate that these leaks be excluded from the leak definition. For the polymer and resins industry, a leak definition of 10,000 ppm was selected.

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 leakage for control. This approach is referred to as a preventive program. For example, in many cases, emissions 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.[59] A barrier fluid can be circulated between the mechanical seals. Degassing vents in the barrier fluid system allow the transport of leakage 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 reciprocating compressors can be controlled by creating a void in the packing and inserting one or more spacer rings into the packing gland or by enclosing the seal area and venting the leakage collected by either system to a control device.

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 second valves are devices that can essentially eliminate 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 180 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.

When an owner or operator replaces several components of an existing facility, that facility may become subject to applicable standards of performance under the provisions of Section 60.15. For example, if an owner or operator replaces several fugitive emission sources such as pumps, compressors, or sampling loops in an existing facility, and if the fixed capital costs for the new equipment exceeds 50 percent of the costs of all fugitive emissions sources in the unit, the Administrator may determine that reconstruction has occurred. Reconstructions may occur as a result of damage caused by fires, explosions, hurricanes, or other catastrophes. They might also result from feedstock changes, product changes, or other major process changes which would require additions or replacement of several fugitive emission sources.

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.

5.3 REFERENCES FOR CHAPTER 5

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*References can be located in Docket Number A-82-19 at the U.S. Environmental Protection Agency, Waterside Hall, Washington, D.C.

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 segments chosen for NSPS development. Section 6.1 presents the model plants chosen to represent the five polymers and resins segments, 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 segments, 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 polymers and resins segments. Therefore, twelve model plants were developed:

1. Polypropylene - continuous, liquid phase slurry process,
2. Polypropylene - gas phase process,
3. Low density polyethylene - high-pressure, process,
4. Low density polyethylene - low-pressure process and high density polyethylene - gas phase process,

5. High density polyethylene - liquid phase slurry process,
6. High density polyethylene - liquid phase solution process,
7. Polystyrene - continuous process,
8. Expandable polystyrene - post-impregnation suspension process,
9. Expandable polystyrene - in-situ suspension process,
10. Poly(ethylene terephthalate), (PET) - DMT process, and
11a. Poly(ethylene terephthalate), (PET) - TPA process, for low viscosity PET and high viscosity PET with a single end finisher,
11b. Poly(ethylene terephthalate), (PET) - TPA process for high viscosity PET with multiple end finishers.

The model plants were selected to be representative of basic manufacturing 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 for the production of crystal or impact polystyrene because available information indicates that new plants that produce crystal or impact polystyrene are unlikely to use this process.

Tables 6-1 through 6-11b 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 6-11b 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 6-12, was chosen to represent the fugitive emission characteristics of the polymers and resins industry. This model plant, however, was not applied to the poly(ethylene terephthalate) model plants because those PET plants using the DMT process are already covered by the

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

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Heating Value, Btu/scf	Composition, wt. %	Average VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	Cont.	0.07	10.5	3.5	0.2	0.07	29	4,320	100 VOC[a]	86
2. Polymerization Reaction	Cont.	4.1	615	205	25	8	54	2,260	100 VOC[b]	44
	Inter.	0.38	57	19	3,330	1,110	65	2,380	93 VOC[b] 6.5 N_2 0.5 H_2	48
3. Material Recovery	Cont.	30	4,500	1,500	140	46	56[f]	2,210	100 VOC[d]	54
4. Product Finishing	Cont.	2.6	390	130	230	71	29[f]	115	9.3 VOC[e] 90.7 Air	45
5. Product Storage	---	---			---				---	---
TOTALS		37.2	5,573	1,858						

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 dilutent, which, in turn, is assumed to be hexane.

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

[c]Assumes C_2 - C_6 with an average molecular weight of 50 and a heating value of 19,800 Btu/scf.

[d]Assumes 2.07 percent is C_3H_6; 2.07 percent, C_3H_8; 1.25 percent, C_4H_{10}; 62.04 percent hexane; and 33.52 percent, methanol.

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

[f]Assumed.

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

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Fstimated Flow (scfm)		Temp., °C	Heating Value, Btu/scf	Composition, wt. %	Average VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	---	---	---	---	---		---	---	---	---
2. Polymerization Reaction	Inter.	11.5	1,210	403	48,700[a]	15,600	38	2,100	99.8 VOC[b] 0.2 H_2	43
3. Material Recovery	Cont.	25	2,625	875	92	31	38	2,400	100 VOC[c]	47
4. Product Finishing	---	---	---	---	---		---	---	---	---
5. Product Storage	---	---	---	---	---		---	---	---	---
TOTALS		36.5	3,835	1,278						

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 86.8 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 LDPE HIGH PRESSURE PROCESS

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate,[a] kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Heating Value, Btu/scf	Composition, wt. %	Average VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	---	---	---	---	---		---	---	---	---
2. Polymerization Reaction	Inter.[b]	0.38	106	27	492,000[c]	250,000[c]	54	1,480	99.0 VOC[d] 1 Ethane	28
3. Material Recovery	---	---	---	---	---		---	---	---	---
4. Product Finishing	Cont.	0.2	56	14	1,600	400	54	3.0	0.2 VOC[e] 99.8 Air	28
5. Product Storage	Cont.	0.8	224	56	6,390	1,590	54	3.0	0.2 VOC[e] 99.8 Air	28
6. Other[f]	Inter.	8.5	2,380	595	24,135[g]	6,034[g]	270	1,470	100 VOC[e]	28
TOTALS		9.88	2,766	692						

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]Emission rates used for each process section reflect an estimate of what a newer plant is likely to emit.

[b]Emergency stream only. Also includes emergency releases from high and low pressure separators.

[c]Design flow based on peak flow for 2 seconds.

[d]Assumes 95.3 percent is ethylene; 1.5 percent, propylene; and 2.2 percent, isopropanol.

[e]Assumes all VOC is ethylene.

[f]Includes non-emergency intermittent releases, such as start-up, shutdown, and maintenance purges, from all process sections.

[g]Based on 100 occurrences per year and 30 minutes per occurrence.

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

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Heating Value, Btu/scf	Composition, wt. %	Average VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	Cont.	0.10	15	7.5	0.52	0.26	39[a]	2,430	92 VOC 8 N_2	56.1[b]
	Inter.	0.14	21.6	10.8	1,100[c]	550	298	98	6.7 VOC 58.8 N_2 34.5 Air	71.2[b]
2. Polymerization Reaction	Inter.[d]	0.76	114	57	32,700	16,350	79	1,359	77.9 VOC 22.1 N_2	36[b]
3. Material Recovery	---	---	---	---	---	---	---	---	---	---
4. Product Finishing	Cont.	22.3	3,345	1,673	510	255	85	560	35 VOC 62 N_2	39[b]
5. Product Storage	Cont.	0.05	7.5	3.8	2,140	1,070	50	0.3	0.02 VOC 99.98 Air	55[b]
TOTALS		23.4[f]	3,503	1,752						

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 for streams A, B, D, E, and F assumed to be 1000, 24, 40, 200, and 200 hours per year, respectively, based on industry data.

[d]Stream G emergency release and Stream H routine release considered separately for control.

[e]Total flow times for streams G and H assumed to be 0.13 and 26 hours per year, respectively, based on industry data.

[f]Stream H, reactor purge vent (the routine polymerization reaction release), and stream J, product discharge vent (product finishing), cannot be emitted simultaneously. Stream I, compressor seal oil vent, and the analyzer vents located throughout the plant, are considered to be fugitive emissions and, therefore, are not included in the model plant.

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

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Heating Value, Btu/scf	Composition, wt. %	Average VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	Inter.	0.2	42	14	1,710[a]	570	21	1,470	100 VOC[b]	28
2. Polymerization Reaction	---	---	---	---	---	---	---	---	---	---
3. Material Recovery	Cont.	12.7	2,667	889	205	68	21	1,450	79 VOC[c] 20 Ethane 1 H_2	32
4. Product Finishing	Cont.	0.406	85	28	690	230	21	10.3	0.7 VOC[d] 1.7 N_2 97.6 Air	58
5. Product Storage	---	---	---	---	---	---	---	---	---	---
TOTALS		13.3	2,794	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 all VOC is isobutane.

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

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Heating Value, Btu/scf	Composition, wt. %	Average VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	Cont. Inter.	0.21	19	6.3	1.2	0.4	167	1,390	91.9 VOC[b] 8.1 Air	30.2
2. Polymerization Reaction	Inter.	3.6	324	108	24	8	30	1,600	61.5 VOC[c] 38.5 Air	41.4
3. Material Recovery	Cont. Inter.	23.7	2,133	711	122	41	41	1,096	83.8 VOC[d] 11.3 Ethane 0.6 Hydrogen 4.3 Air	40
4. Product Finishing	Cont.	1.5	135	45	48,800[e]	16,240[e]	50	0. 2	0.2 VOC[f] 7.4 Steam 92.2 Air	84
5. Product Storage	---	---	---	---	---		---	---	---	---
TOTALS		29.0	2,611	870						

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]Emission characteristics for Raw Materials Preparation, Polymerization Reaction, and Material Recovery where estimated from industry information and engineering judgment.

[b]Assumes 87.5% ethylene and 4.4% cyclohexane.

[c]Assumes 30.8% ethylene, 25.6% cyclohexane, and 5.1% other VOC.

[d]Assumes 42.9% ethylene, 25.5% cyclohexane.

[e]Flow after condensation.

[f]Cyclohexane.

Table 6-7. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE CRYSTAL OR IMPACT POLYSTYRENE CONTINUOUS PROCESS[a]

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Pressure	Composition, wt. %	VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Material Recovery	Continuous	0.12	9	4.5	14	7	27	Atmospheric	1 Styrene 99 Air	104
2. Product Finishing	Continuous	0.15	11.3	5.6	10,280	5,140	21	-	Trace Styrene Steam	104
TOTALS		0.27	20.3	10.1						

Model Plant Capacity: 75 Gg/yr
No. of Process Trains: 2
Capacity per Process Train: 37.5 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 socpe of this project.

Table 6-8. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE EXPANDABLE POLYSTYRENE POST-IMPREGNATION SUSPENSION PROCESS

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Heating Value, Btu/scf	Composition, wt. %	Average VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	---	---	---	---	---	---	---	---	---	---
2. Polymerization Reaction	Inter.	0.3	2.25	0.75	6.9[a]	2.3	---	1,042[b]	62 VOC[c] 38 H_2O	104
3. Material Recovery	---	---	---	---	---	---	---	---	---	---
4. Product Finishing	Cont.	2.16	162	54	19,550	6,515	21	0.7	0.05 VOC[d] 99.95 Air	72.2
	Inter.	3.81	286	95	8,580[e]	2,860	21	17.7	1.2 VOC[d] 99.8 Air	72.2
5. Product Storage	Cont.	2.27	170	57	5,680	1,890	21	2.5	0.17 VOC[d] 99.83 Air	72.2
	Inter.	1.2	90	30	5,390[f]	1,800	21	1.6	0.11 VOC[d] 99.89 Air	72.2
TOTALS		9.47	710	237						

Model Plant Capacity: 75 Gg/yr
No. of Process Trains: 3
Capacity per Process Train: 25 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

[a]Assumes 200 hr/yr.

[b]In water vapor, before condensation.

[c]Assumes all styrene.

[d]Assumes all pentane.

[e]Assumes 1,350 hr/yr.

[f]Assumes 7,440 hr/yr.

Table 6-9. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE EXPANDABLE POLYSTYRENE IN-SITU SUSPENSION PROCESS

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Heating Value, Btu/scf	Composition, wt. %	Average VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	Inter.	0.12	9	1	102[a]	7.8	21	15.5	0.75 Styrene 0.53 ACN 98.72 Air	74.5
2. Polymerization Reaction	Inter.	1.14	82	9.1	595[c]	66.1	35	1,375	58.8 VOC[b] 1.4 Styrene 39.8 Air	72.7
3. Material Recovery	---	---	---	---	---	---	---	---	---	---
4. Product Finishing	Cont.	2.91	210	23	13,860	1,020	35	1.9	0.13 VOC[b] 99.87 Air	72.2
	Inter.	0.055	4	0.44	123.6[d]	13.7	30	9.7	0.66 VOC[b] 99.34 Air	72.2
5. Product Storage	Cont.	0.75	54	6	7,880	595	22	0.83	0.60 VOC[b] 99.94 Air	72.2
TOTALS		4.98	359	40						

Model Plant Capacity: 72 Gg/yr
No. of Process Trains: 9
Capacity per Process Train: 8 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

[a]Based on flow duration of 4,740 hrs/yr.

[b]Assumes all pentane.

[c]Based on flow durations of 71 hours for Stream B.

[d]Based on flow durations of 2,370 hours for Stream C.

Table 6-10. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE POLY(ETHYLENE TEREPHTHALATE) DMT PROCESS[a]

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Pressure, psig	Composition, wt. %	VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Material Recovery	Continuous	0.18	18.9	2.7	2.1	0.3	38	---	49 VOC[b] 51 Nitrogen	32
2. Polymerization Reaction	Continuous	0.355	37.3	5.33	---	---	25	0.2	---	62
TOTALS		0.535	56.2	8.03						

Model Plant Capacity: 105 Gg/yr
No. 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]This model plant assumes a production line configuration of two esterification vessels in series followed by a polymerization reactor.

[b]Methanol.

Table 6-11a. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE POLY(EHTYLENE TEREPHTHALATE) TPA PROCESS USING A SINGLE END FINISHER[a]

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Pressure, psig	Composition, wt. %	VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	Continuous	0.04	4	0.6	---	---	---	---	---	62
2. Polymerization Reaction	Continuous	0.315	33.1	4.7	---	---	25	0.2		
TOTALS		0.355	37.3	5.3						

Model Plant Capacity: 105 Gg/yr
No. 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]This model plant assumes a production line configuration of two esterification vessels in series followed by a single polymerization vessel for low viscosity PET or two polymerization reactors (a low polymerizer and an end finisher) for high viscosity PET also in series.

Table 6-11b. MODEL PLANT CHARACTERISTICS FOR PROCESS EMISSIONS FROM THE POLY(ETHYLENE TERPHTHALATE) TPA PROCESS USING MULTIPLE END FINISHERS[a]

Process Section	Nature of Stream(s)	Uncontrolled Emission Rate, kg VOC/ Mg Product	Uncontrolled VOC Emissions		Estimated Flow (scfm)		Temp., °C	Pressure, psig	Composition, wt. %	VOC Molecular Weight
			Total, Mg/yr	Process Train, Mg/yr	Total Plant	Process Train				
1. Raw Materials Preparation	Continuous	0.15	6	3	---	---	---	---	---	62
2. Polymerization Reaction	Continuous	3.4	136	68	---	---	25	0.2		
TOTALS		3.55	142	71						

Model Plant Capacity: 40 Gg/yr
No. of Process Trains: 2
Capacity per Process Train: 20 Gg/yr
Total Operating Hours per Year: 8,600
Standard Conditions: 21°C (70°F) and 0 psig

[a]This model plant assumes a production line configuration of two esterification vessels in series followed by a prepolymerizer, an initial end finisher, and one or more additional end finishers, also in series.

Table 6-12. 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
Pump 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	2,400	0.00083
Sampling connections	104	0.015
Open-ended lines	415	0.0017

TOTAL UNCONTROLLED FUGITIVE EMISSIONS: 151 Mg/yr

fugitive emission standards for the synthetic organic chemicals manufacturing industry (SOCMI) due to their methanol by-product production and those PET plants using the TPA process use only "heavy" liquids (i.e., ethylene glycol) and solids.

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. In addition, the regulatory alternatives were developed by applying baseline control and more stringent control to emissions from process sections within a single process line for each model plant.

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 or combustion technologies that achieve varying levels of VOC emission reductions. For the crystal or impact 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). As current industry practice in controlling emissions from expandable polystyrene plants varies considerably between plants, it was assumed for analyses purposes that there would be no baseline control. For the poly(ethylene terephthalate) segment, baseline control reflects the use of spent ethylene glycol condensers to recover the ethylene glycol from the polymerization reactors and reflux condensers on the esterifiers, if a low viscosity product is produced or if a high viscosity product is produced using a single end finisher. Baseline control for a plant producing high viscosity PET using multiple end finishers reflects the use of spent ethylene glycol spray condensers on the initial end finisher only and distillation columns on the esterifiers and cooling tower.

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 SOCMI.

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 poly(ethylene terephthalate) processes for VOC emission control. Some combustion of emissions from expandable polystyrene plants does take place.

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 poly(ethylene terephthalate).

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 6-13) was chosen as the regulatory alternative for controlling

Table 6-13. CONTROL SPECIFICATION FOR FUGITIVE EMISSIONS UNDER REGULATORY ALTERNATIVE 2

Source	Inspection Interval	Equipment Specification
Valves		
Vapor service	Monthly	None
Light liquid service	Monthly	None
Heavy liquid service	None	None
Pump seals		
Light liquid service	Monthly[a] Weekly visual	None
Heavy liquid service	None	None
Compressor seals	None	Controlled degassing vents
Safety relief valves		
Vapor service	None	Rupture disks on relief valves[b]
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.

[b]For safety relief valves, instrument monitoring would be necessary after an overpressure relief.

VOC fugitive emissions from the polymers and resins industry. This regulatory alternative was chosen for consistency with the fugitive emission regulations for SOCMI.

6.2.3 Regulatory Alternatives

The following sections present the regulatory alternatives, including baseline control, for a process line in 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 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 NSPS. 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 each remaining process section with potential uncontrolled emissions until all process sections within the process line were analyzed.

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. Combustion devices were also used to develop all but one of the regulatory alternatives for the expandable polystyrene model plants.

In developing the regulatory alternatives for the crystal or impact 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 crystal or impact polystyrene plants, condensers are the most likely recovery technique to be used and are used to develop the regulatory alternatives for this 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 6-14.) By applying the general criteria outlined in Section 6.2.1, the process emission streams identified under baseline control (Regulatory Alternative 1) are Streams B and C (reactor vents), Stream D (decanter vent), Stream E (neutralizer vent), Stream F (slurry filter/vacuum system vent) and Stream G (diluent separation and recovery). These streams correspond to the polymerization reaction section (Streams B and C) and the material recovery section (Streams D through G), 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 from a process line is equivalent to an annual process emission reduction of about 1,690 Mg. Fugitive emission baseline control, as discussed in Section 6.2.1, would reduce uncontrolled fugitive emissions from 50 Mg/yr to approximately 35 Mg/yr. Under baseline control, an annual emission reduction of 1,705 Mg is achieved or approximately 89 percent of the total uncontrolled emissions from a process line in 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 16 Mg per process line. Under this alternative, an annual emission reduction of 1,725 Mg per process line is achieved or approximately 90 percent.

Regulatory Alternative 3 includes the same control as in Regulatory Alternative 2 plus the combustion of the emissions from the product

Table 6-14. REGULATORY ALTERNATIVES FOR THE POLYPROPYLENE LIQUID PHASE PROCESS

Regulatory Alternative	Process Emissions		Fugitive Emissions	Annual Emission Reduction Per Process Line[b]	
	Process Section(s) Controlled[a]	Control Technique		Mg/yr	Percent[c]
1 (baseline)	PR + MR	Flare	d	1,705	89%
2	PR + MR	Flare	e	1,725	90%
3	PR, MR, plus PF	Combustion[f]	e	1,852	97.0%
4	PR, MR, PF, plus RMP	Combustion[f]	e	1,856	97.3%

[a]Process sections include the following:
RMP = raw materials 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 1,908 Mg/yr (1,858 Mg/yr process emissions plus 50 Mg/yr fugitive emissions).

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

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

finishing section. This alternative reduces annual emissions by 1,852 Mg per process line or approximately 97 percent.

Regulatory Alternative 4 includes the same control as in the third regulatory alternative plus the combustion of emissions from the raw materials preparation section. This alternative is somewhat more stringent than the third, reducing annual emissions from each process line by 1,856 Mg or 97.3 percent from uncontrolled levels in a process line.

6.2.3.2 Polypropylene - Gas Phase Process. (Table 6-15.) As stated in Table 6-2, the mass flowrate of emissions from the material recovery section is about 530 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. Based on information from the one company that uses this process, the emissions from the material recovery section are controlled by the same flare as the emissions from the polymerization reaction section. Thus, the baseline (Regulatory Alternative 1) includes control of both the intermittent and continuous streams by a flare. This baseline represents a 95 percent reduction (1,267 Mg/yr) of total uncontrolled VOC emissions from a process line in 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 16 Mg. This alternative achieves an annual emission reduction of about 1,290 Mg from each process line, or approximately 97 percent.

6.2.3.3 Low Density Polyethylene, High-Pressure Process. (Table 6-16.) The baseline control (Regulatory Alternative 1) for this model plant sends the non-emergency, purge-type intermittent emissions from all process sections to a flare. Under baseline control, process and fugitive emissions from the process line would be reduced by 594 Mg/yr or about 81 percent of the total uncontrolled emissions in a process line.

Regulatory Alternative 2, which represents fugitive emission controls plus baseline control, results in a total annual emission reduction of 609 Mg per process line or about 83 percent.

Regulatory Alternative 3 represents Regulatory Alternative 2 controls plus the combustion of emissions from the product storage section. This

Table 6-15. REGULATORY ALTERNATIVES FOR THE POLYPROPYLENE GAS PHASE PROCESS

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[a]	Process Emissions: Control Technique	Fugitive Emissions	Annual Emission Reduction Per Process Line[b]: Mg/yr	Annual Emission Reduction Per Process Line[b]: Percent[c]
1 (baseline)	Polymerization Reactor (PR) plus Material Recovery (MR)	Flare	d	1,267	95%
2	PR plus MR	Flare	e	1,287	97%

[a]Process sections include the following:
RMP = raw materials 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 1,328 Mg/yr (1,278 Mg/yr process emissions plus 50 Mg/yr fugitive emissions.)

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

Table 6-16. REGULATORY ALTERNATIVES FOR THE LDPE HIGH PRESSURE PROCESS

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[a]	Process Emissions: Control Technique	Fugitive Emissions	Annual Emission Reduction Per Process Line[b]: Mg/yr	Annual Emission Reduction Per Process Line[b]: Percent[c]
1 (baseline)	Purges	Flare	d	594	81%
2	Baseline (B)	Flare	e	609	83%
3	Baseline plus PS	Combustion[f]	e	664	91%
4	Baseline, PS plus PF	Combustion[f]	e	678	93%
5	Baseline, PS PF plus Emergency Vents from Reactor and Separators	Combustion[f]	e	704	96%

[a]Process sections include the following:
RMP = raw materials 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 730 Mg/yr (692 Mg/yr process emissions plus 38 Mg/yr fugitive emissions.)

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

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

alternative reduces total uncontrolled emissions from a process line by 664 Mg/yr or approximately 91 percent.

Regulatory Alternative 4 is the same as Regulatory Alternative 3 plus the combustion of emissions from the product finishing section. Total uncontrolled emissions from a process line under this alternative are reduced by 678 Mg/yr or about 93 percent.

Regulatory Alternative 5 is the same as Regulatory Alternative 4 plus the combustion of emissions from the emergency reactor vent in the polymerization reaction section. (Note: 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 available to most companies, Stream B is not likely to be controlled.) Total uncontrolled emissions under this alternative are reduced by 704 Mg/yr per process line or about 96 percent.

6.2.3.4 Low Density Polyethylene Low-Pressure and High Density Polyethylene Gas Phase Process. (Table 6-17.) The intermittent streams in the raw materials preparation, the polymerization reaction (except decompositions), and the product finishing sections and Stream J (product discharge vent) in the product finishing section were determined to be controlled by flaring under the baseline assumptions. Annual emission reductions from a process line under baseline control (Regulatory Alternative 1) are 1,725 Mg or about 95 percent of the total uncontrolled emissions from a process line in 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 23 Mg per process line. Under this alternative, an annual emission reduction of 1,762 Mg per process line is achieved, or approximately 96.1 percent.

Table 6-17. REGULATORY ALTERNATIVES FOR THE LDPE LOW-PRESSURE AND HDPE GAS PHASE PROCESS

Regulatory Alternative	Process Emissions		Fugitive Emissions	Annual Emission Reduction Per Process Line[b]	
	Process Section(s) Controlled[a]	Control Technique		Mg/yr	Percent[c]
1 (baseline)	RMP (Intermittent streams only), PR (Non-emergency intermittent streams only), and PF	Flare	d	1,725	94%
2	Baseline	Flare	e	1,755	96.1%
3	Baseline plus continuous streams from RMP	Combustion[f]	e	1,762	96.4
4	Baseline, continuous streams from RMP, plus PS	Combustion[f]	e	1,766	96.7%
5	Baseline, continuous streams from RMP and PS, plus PR emergency stream	Combustion[f]	e	1,770	96.9%

[a]Process sections include the following:
RMP = raw materials 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 1,827 Mg/yr (1,752 Mg/yr process emissions plus 75 Mg/yr fugitive emissions.)

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

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

Regulatory Alternative 3 represents Regulatory Alternatives 2 controls plus the combustion of continuous emissions from the raw materials preparation section. Under Regulatory Alternative 3, emissions are reduced by 1,762 Mg/yr per process line or about 96.4 percent.

Regulatory Alternative 4 represents, in addition to the control achieved in Regulatory Alternative 3, the combustion of continuous emissions in the product storage section. This alternative results in an annual emission reduction per process line of 1,766 Mg or 96.7 percent of the total uncontrolled emissions.

Regulatory Alternative 5 represents Regulatory Alternative 4 controls plus the control of emergency vents (i.e., decompositions) from the polymerization reaction section. This alternative results in an annual emission reduction 1,770 Mg per process line or about 96.9 percent of the total uncontrolled emissions from a single process line.

6.2.3.5 High Density Polyethylene Liquid Phase, Slurry Process. (Table 6-18.) Under baseline control (Regulatory Alternative 1) Stream A (feed preparation) and Stream D (recycle treaters) would be controlled. These streams correspond to the raw materials preparation and material recycle sections, respectively. Annual uncontrolled emissions from a process line are reduced by 900 Mg or about 92 percent.

The next level of control, Regulatory Alternative 2, represents baseline control plus the fugitive emission controls listed in Table 6-13. An annual emission reduction of 920 Mg per process line would be achieved, which is a 94 percent reduction from the uncontrolled level of emissions in a process line.

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 from a process line by 948 Mg/yr or about 97 percent.

6.2.3.6 High Density Polyethylene, Liquid Phase, Solution Process. (Table 6-19) 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

Table 6-18. REGULATORY ALTERNATIVES FOR THE HDPE LIQUID PHASE SLURRY PROCESS

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[a]	Process Emissions: Control Technique	Fugitive Emissions	Annual Emission Reduction Per Process Line[b]: Mg/yr	Annual Emission Reduction Per Process Line[b]: Percent[c]
1 (baseline)	RMP plus MR	Flare	d	900	92%
2	RMP plus MR	Flare	e	920	94%
3	RMP, MR plus PF	Combustion[f]	e	948	97%

[a]Process sections include the following:
RMP = raw materials 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 981 Mg/yr (931 Mg/yr process emissions plus 50 Mg/yr fugitive emissions.)

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

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

Table 6-19. REGULATORY ALTERNATIVES FOR THE HDPE LIQUID PHASE SOLUTION PROCESS

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[a]	Process Emissions: Control Technique	Fugitive Emissions	Annual Emission Reduction Per Process Line[b]: Mg/yr	Annual Emission Reduction Per Process Line[b]: Percent[c]
1 (baseline)	RMP, PR, MR	Flare	d	824	90%
2	RMP, PR, MR MR	Flare	e	844	92%
3	RMP, PR, MR plus PF	Combustion[f]	e	888	97%

[a]Process sections include the following:
RMP = raw materials 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 920 Mg/yr (870 Mg/yr process emissions plus 50 Mg/yr fugitive emissions.)

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

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

based. Under this assumption, all the streams in the raw materials preparation, polymerization reaction, and material recovery sections were assumed to be flared. Baseline control results in an annual emission reduction of 824 Mg per process line or about 90 percent of the total uncontrolled emissions from a process line.

Regulatory Alternative 2 requires the same control as baseline plus fugitive emission controls. This alternative reduces process and fugitive emissions from the process sections in a process line under baseline control by 844 Mg/yr or about 92 percent of the total uncontrolled emissions from a process line.

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 888 Mg per process or about 97 percent.

6.2.3.7 Polystyrene, Continuous Process. (Table 6-20.) 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 23 Mg per process line. This results in a 27 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 24 Mg. Under this alternative, an annual emission reduction of 53 Mg per process line is achieved, or approximately 62 percent of the total uncontrolled emissions from a process line.

Regulatory Alternatives 3 through 7 represent fugitive emission control under Regulatory Alternative 2 plus further recovery of emissions from the material recovery section by the use of refrigerated condensers to emission levels ranging from 0.012 to 0.0018 kg VOC/Mg product. Under these alternatives, emissions from a process line are reduced by about 57 Mg/yr, or about 67 percent of the total uncontrolled emissions from a process line.

Table 6-20. REGULATORY ALTERNATIVES FOR THE CRYSTAL OR IMPACT POLYSTYRENE CONTINUOUS PROCESS

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[a]	Process Emissions: Control Technique	Fugitive Emissions	Annual Emission Reduction Per Process Line[b]: Mg/yr	Annual Emission Reduction Per Process Line[b]: Percent[c]
1 (baseline)	-	-	c	23	27%
2	-	-	d	53	62%
3	Material recovery	Recovery[e]	d	57.05	67.1%
4	Material recovery	Recovery[f]	d	57.23	67.3%
5	Material recovery	Recovery[g]	d	57.37	67.4%
6	Material recovery	Recovery[h]	d	57.41	67.5%
7	Material recovery	Recovery[i]	d	57.43	67.6%

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

[b]Based on total uncontrolled emissions of 85 Mg/yr (10 Mg/yr process emissions plus 75 Mg/yr fugitive emissions.)

[c]Fugitive baseline control.

[d]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

[e]Control to an emission rate of 0.012 kg VOC/Mg product through use of refrigerated condensers.

[f]Control to an emission rate of 0.0072 kg VOC/Mg product through use of refrigerated condensers.

[g]Control to an emission rate of 0.0036 kg VOC/Mg product through use of refrigerated condensers.

[h]Control to an emission rate of 0.0024 kg VOC/Mg product through use of refrigerated condensers.

[i]Control to an emission rate of 0.0018 kg VOC/Mg product through use of refrigerated condensers.

6.2.3.8 Expandable Polystyrene, Post-Impregnation Suspension Process. (Table 6-21). As discussed earlier, no baseline control (Regulatory Alternative 1) is assumed for process emissions. Fugitive emission baseline control results in an annual emission reduction per process line of about 15 Mg, or approximatley 5 percent of total uncontrolled emissions from a process line.

Regulatory Alternative 2 reflects increased fugitive emission controls so that an annual emission per process line of 35 Mg, or about 12 percent of total uncontrolled emissions, is achieved.

Regulatory Alternative 3 reflects the fugitive emission control under Regulatory Alternative 2 plus the control of continuous emissions from the product storage section. This alternative results in an annual emission reduction of 91 Mg per process line, or about 32 percent of total uncontrolled emissions.

Regulatory Alternative 4 represents the emission controls of Regulatory Alternative 3 plus the control of intermittent emissions from the product finishing section. This alternative results in an annual emission reduction of 184 Mg, or about 64 percent of the total uncontrolled emissions, per process line.

Regulatory Alternative 5 represents the emission controls of Regulatory Alternative 4 plus the control of the intermittent streams from the product storage section. This alternative results in an annual emission reduction of 237 Mg, or about 83 percent of the total uncontrolled emissions, per process line.

Regulatory Alternative 6 represents the emission controls of Regulatory Alternative 5 plus the control of continuous emissions from the product finishing section. This alternative results in an annual emission reduction of 266 Mg, or slightly less than 93 percent of the total uncontrolled emissions, per process line.

Regulatory Alternative 7 represents the emission controls of Regulatory Alternative 6 plus the control of the emissions from the polymerization reaction section. This alternative results in an annual emission reduction of 267 Mg, or about 93 percent of the total uncontrolled emissions, per process line.

Table 6-21. REGULATORY ALTERNATIVES FOR THE EXPANDABLE POLYSTYRENE POST-IMPREGNATION SUSPENSION PROCESS

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[a]	Process Emissions: Control Technique	Fugitive Emissions	Annual Emission Reduction Per Process Line[b]: Mg/yr	Annual Emission Reduction Per Process Line[b]: Percent[c]
1 (baseline)	-	-	d	15	5%
2	-	-	e	35	12%
3	PS (continuous streams only)	Combustion[f]	e	91	32%
4	PS (continuous streams only plus PF (intermittent streams only)	Combustion[f]	e	184	64%
5	PS (continous streams only) plus PF (all streams)	Combustion[f]	e	237	83%
6	PS (all streams) plus PF (all streams	Combustion[f]	e	266	92.7%
7	PS, PF PR	Combustion[f] Recovery[g]	e	267	93.0%

[a]Process sections include the following:
PR = polymerization reaction
PF = product finishing
PS = product storage

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

[c]Based on total uncontrolled emissions of 287 Mg/yr (237 Mg/yr process emissions and 50 Mg/yr fugitive emissions).

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

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

[g]Recovery to 0.003 kg VOC/Mg product through use of condensers.

6.2.3.9 Expandable Polystyrene, In-situ Suspension Process. (Table 6-22). As discussed earlier, no baseline control (Regulatory Alternative 1) is assumed for process emissions. Fugitive emission baseline control results in an annual emission of 5 Mg per process line, or about 9 percent of the total uncontrolled emissions from a process line.

Regulatory Alternative 2 reflects increased fugitive emission control so that an annual emission reduction per process line of almost 12 Mg, or about 21 percent of the total uncontrolled emissions, is achieved.

Regulatory Alternative 3 represents the fugitive emission controls of Regulatory Alternative 2 plus the control of intermittent emissions from the polymerization reaction section. This alternative results in an annual emission reduction of 20.6 Mg, or about 36 percent of the total uncontrolled emission, per process line.

Regulatory Alternative 4 represents the emission controls of Regulatory Alternative 3 plus the control of continuous emissions from the product finishing section. This alternative results in an annual emission reduction of about 43.4 Mg, or about 76 percent of the total uncontrolled emissions, per process line.

Regulatory Alternative 5 represents the emission controls of Regulatory Alternative 4 plus the control of continuous streams from the product storage section. This alternative results in an annual emission reduction of 49.2 Mg, or about 86 percent of the total uncontrolled emissions, per process line.

Regulatory Alternative 6 represents the emission controls of Regulatory Alternative 5 plus the controls of intermittent emissions from the raw material preparation section. This alternative results in an annual emissions, per process line.

Regulatory Alternative 7 represents the emission controls of Regulatory Alternative 6 plus the control of intermittent emissions from the product finishing section. This alternative results in an annual emission reduction of 50.6 Mg, or about 89 percent of the total uncontrolled emissions, per process line.

Table 6-22. REGULATORY ALTERNATIVES FOR THE EXPANDABLE POLYSTYRENE IN-SITU SUSPENSION PROCESS

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[a]	Process Emissions: Control Technique	Fugitive Emissions	Annual Emission Reduction Per Process Line[b]: Mg/yr	Annual Emission Reduction Per Process Line[b]: Percent[c]
1 (baseline)	-	-	d	5	9%
2	-	-	e	11.7	21%
3	PR	Combustion[f]	e	20.6	36%
4	PR plus PF (continuous streams only)	Combustion[f]	e	43.4	76%
5	PR, PF (continuous streams) plus PS	Combustion[f]	e	49.2	86%
6	PR, PF (continuous streams), PS, plus RMP	Combustion[f]	e	50.2	88%
7	PR, PF (continuous streams), RMP, PS, plus PF (intermittent streams)	Combustion[f]	e	50.6	89%

[a]Process sections include the following:
RMP = raw materials preparation
PR = polymerization reaction
PF = product finishing
PS = product storage

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

[c]Based on total uncontrolled emissions of 57 Mg/yr (40 Mg/yr process emissions and 17 Mg/yr fugitive emissions).

[d]Fugitive baseline control.

[e]Control equivalent to Regulatory Alternative 2 (see Table 6-13).

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

6.2.3.10 Poly(ethylene terephthalate), DMT Process. (Table 6-23). As discussed earlier, baseline control (Regulatory Alternative 1) in Table 6-23 represents the use of spent ethylene glycol condensers to recover ethylene glycol from the polymerization reactors and the esterifiers. This corresponds to a recovery system currently in use in the industry; thus no emission reduction from uncontrolled emissions occurs under baseline control.

Regulatory Alternatives 2 through 6 reflect the use of refrigerated condensers to control the methanol from the material recovery section (i.e., methanol recovery) to emission levels of 0.018 to 0.0018 kg VOC/Mg of product. Under this alternative, annual emissions from a process line are reduced between 2.43 to 2.67 Mg, or from about 30 to 33 percent from uncontrolled levels.

Regulatory Alternative 7 reflects control under Regulatory Alternative 4 plus the additional recovery of ethylene glycol from the cooling tower (i.e., the polymerization reaction section) through the use of a distillation column. Under this alternative, annual emissions are reduced by 5.33 Mg, or 66 percent of uncontrolled emissions, from a process line.

6.2.3.11 Poly(ethylene terephthlate), TPA Process - low viscosity PET or high viscosity PET with a single end finisher. (Table 6-24a.) As discussed earlier, baseline control (Regulatory Alternative 1) is assumed to be equivalent to current industry practice in which ethylene glycol is recovered using spent ethylene glycol spray condensers. The resulting emissions are the same as the uncontrolled emissions and, thus, no emission reduction occurs under baseline control.

Regulatory Alternative 2 reflects the additional recovery of ethylene glycol from the cooling water tower using a distillation column that results in a 50 percent reduction in emissions to the atmosphere from the cooling water tower. This regulatory alternative would reduce emissions by 2.35 Mg/yr per process line or about 44 percent from uncontrolled emissions.

6.2.3.12 Poly(ethylene terephthalate), TPA Process - high viscosity PET with multiple end finishers. (Table 6-24b.) As discussed earlier, baseline control (Regulatory Alternative 1) is assumed to be the use of spent ethylene glycol spray condensers on the initial end finishers

Table 6-23. REGULATORY ALTERNATIVES FOR THE POLY(ETHYLENE TEREPHTHALATE) DMT PROCESS[a]

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[b]	Process Emissions: Control Technique	Annual Emission Reduction Per Process Line[c]: Mg/yr	Annual Emission Reduction Per Process Line[c]: Percent[d]
1 (baseline)	PR	Recovery[e]	0	0%
2	PR plus MR	Recovery[e] Recovery[f]	2.43	30%
3	PR plus MR	Recovery[e] Recovery[g]	2.48	31%
4	PR plus MR	Recovery[e] Recovery[h]	2.59	32.3%
5	PR plus MR	Recovery[e] Recovery[i]	2.66	33.1%
6	PR plus MR	Recovery[e] Recovery[j]	2.67	33.3%
7	PR plus MR	Recovery[k] Recovery[j]	5.33	66%

[a]These regulatory alternatives apply to a line producing a low viscosity or high viscosity product (with a single end finisher).

[b]PR = polymerization reaction
MR = material recovery

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

[d]Based on total uncontrolled emissions of 8.03 Mg/yr.

[e]Recovery of ethylene glycol to an emission rate of 0.355 kg VOC/Mg product through the use of spent ethylene glycol spray condensers recovering the ethylene glycol prior to the vacuum system servicing the polymerizers.

[f]Includes condenser on MR controlling emissions to 0.018 kg VOC/Mg of product.

[g]Includes condenser on MR controlling emissions to 0.0144 kg VOC/Mg product.

[h]Includes condenser on MR controlling emissions to 0.0072 kg VOC/Mg product.

[i]Includes condenser on MR controlling emissions to 0.0027 kg VOC/Mg product.

[j]Includes condenser on MR controlling emissions to 0.0018 kg VOC/Mg product.

[k]Additional recovery of ethylene glycol by recovery from the cooling water tower using a distillation column that results in a reduction in atmospheric emissions of 50 percent.

Table 6-24a. REGULATORY ALTERNATIVES FOR THE PQLY(ETHYLENE TEREPHTHALATE) TPA PROCESS PRODUCING LOW VISCOSITY PET OR HIGH VISCOSITY PET WITH A SINGLE END FINISHER

Regulatory Alternative	Process Emissions		Annual Emission Reduction Per Process Line[b]	
	Process Section(s) Controlled[a]	Control Technique	Mg/yr	Percent[c]
1 (baseline)	RMP, PR	Recovery[d]	0	0%
2	RMP, PR	Recovery[e]	2.35	44%

[a]RMP = raw materials recovery
PR = polymerization reaction

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

[c]Based on total uncontrolled emissions of 5.3 Mg/yr.

[d]Recovery of glycol from the raw materials preparation and polymerization reaction sections using spent ethylene glycol spray condensers.

[e]Control of emissions from the cooling water tower (i.e., polymerization reaction section) using a distillation column that results in a 50 percent reduction in atmospheric emissions from the cooling water tower.

Table 6-24b. REGULATORY ALTERNATIVES FOR THE PQLY(ETHYLENE TEREPHTHALATE) TPA PROCESS PRODUCING HIGH VISCOSITY PET WITH MULTIPLE END FINISHERS

Regulatory Alternative	Process Emissions: Process Section(s) Controlled[a]	Process Emissions: Control Technique	Annual Emission Reduction Per Process Line[b]: Mg/yr	Annual Emission Reduction Per Process Line[b]: Percent[c]
1 (baseline)	RMP, PR	Recovery[d]	0	0%
2	RMP, PR	Recovery[e]	7.2	10%
3	RMP, PR	Recovery[e]	14.7	21%
4	RMP, PR	Recovery[e]	22.3	31%
5	RMP, PR	Recovery[e]	26.1	37%
6	RMP, PR	Recovery[e]	29.8	42%
7	RMP, PR	Recovery[e]	37.3	53%

[a]RMP = raw materials recovery
PR = polymerization reaction

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

[c]Based on total uncontrolled emissions of 71 Mg/yr.

[d]Baseline recovery includes distillation columns on the esterifiers and cooling tower and spent ethylene glycol spray condensers on the initial end finishers.

[e]Additional control result of sending increasing amounts of cooling water from the cooling water tower to the distillation column. Feed rate from cooling water tower will determine resulting ethylene glycol concentration in the cooling water tower and the resulting level of ethylene glycol emissions to the atmosphere.

only and distillation columns on the esterifiers and the cooling tower. The resulting emissions are considered the same as the uncontrolled emissions and, thus, no emission reduction occurs under baseline control.

Regulatory Alternatives 2 through 7 reflect the increased flow of cooling water to the distillation column to recover additional ethylene glycol from the cooling water tower. By varying the feed rate of the cooling water to the distillation column, the ethylene glycol concentration in the cooling water will vary and emissions to the atmosphere will also vary. These alternatives reflect increased flow rate to the distillation column that results in effective emission reductions from the polymerization reaction section via the cooling water tower ranging from 7.2 Mg/yr or 10 percent reduction of the uncontrolled emissions, under Regulatory Alternative 2, up to 37.3 Mg/yr or about 53 percent reduction, under Regulatory Alternative 7, per process line.

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 6-25.

Table 6-25. SUMMARY OF UNCONTROLLED EMISSIONS AND EMISSION REDUCTIONS PER PROCESS LINE 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	Total Percent
PP, L	1,858	50	1,908	1	1,690	15	1,705	89%
				2	1,690	35	1,725	90%
				3	1,817	35	1,852	97%
				4	1,821	35	1,856	97.3%
PP, G	1,278	50	1,328	1	1,252	15	1,267	95%
				2	1,252	35	1,287	97%
LDPE, HP	692	38	730	1	583	11	594	81%
				2	583	26	609	83%
				3	638	26	664	91%
				4	652	26	678	93%
				5	678	26	704	96%
LDPE, LP HDPE, G	1,752	75	1,827	1	1,703	22	1,725	94%
				2	1,703	52	1,755	96.1%
				3	1,710	52	1,762	96.4%
				4	1,714	52	1,766	96.7%
				5	1,718	52	1,770	96.9%
HDPE, SL	931	50	981	1	885	15	900	92%
				2	885	35	920	94%
				3	913	35	948	97%
HDPE, SO	870	50	920	1	809	15	824	90%
				2	809	35	844	92%
				3	853	35	888	97%
PS, C	10	75	85	1	0	23	23	27%
				2	0	53	53	62%
				3	4.05	53	57.05	67.1%
				4	4.23	53	57.23	67.3%
				5	4.37	53	57.37	67.4%
				6	4.41	53	57.41	67.5%
				7	4.43	53	57.43	67.6%

Table 6-25. SUMMARY OF UNCONTROLLED EMISSIONS AND EMISSIONS REDUCTIONS PER PROCESS LINE FOR REGULATORY ALTERNATIVES BY MODEL PLANT (*concluded*)

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	Total Percent
EPS, PI	237	50	287	1	0	15	15	5%
				2	0	35	35	12%
				3	56	35	91	32%
				4	149	35	184	64%
				5	202	35	237	83%
				6	231	35	266	92.7%
				7	232	35	267	93.0%
EPS, IS	40	17	57	1	0	5	5	9%
				2	0	11.7	11.7	21%
				3	8.9	11.7	20.6	36%
				4	31.7	11.7	43.4	76%
				5	37.5	11.7	49.2	86%
				6	38.5	11.7	50.2	88%
				7	38.9	11.7	50.6	89%
PET/DMT	8	0	8	1	0	0	0	0%
				2	2.43	0	2.43	30%
				3	2.48	0	2.48	31%
				4	2.59	0	2.59	32%
				5	2.66	0	2.66	33%
				6	2.67	0	2.67	33.3%
				7	5.33	0	5.33	66%
PET/TPA[b]	5.3	0	5.3	1	0	0	0	0%
				2	2.4	0	2.4	44%
PET/TPA[c]	71	0	71	1	0	0	0	0%
				2	7.2	0	7.2	10%
				3	14.7	0	14.7	21%
				4	22.3	0	22.3	31%
				5	26.1	0	26.1	37%
				6	29.8	0	29.8	42%
				7	37.3	0	37.3	53%

[a]From uncontrolled levels.

[b]Low viscosity, and high viscosity with single end finisher.

[c]High viscosity with multiple end finishers.

KEY: PP = Polypropylene
LDPE = Low Density Polyethylene
HDPE = High Density Polyethylene
PS = Polystyrene
L = Liquid Phase Process
G = Gas Phase Process
SL = Slurry Process
SO = Solution Process
HP = High-Pressure Process
LP = Low-Pressure Process
PI = Post-Impregnation Process
IS = In-situ 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 93 percent less than uncontrolled emissions and about 20 percent less than estimated baseline emissions (Regulatory Alternative I). VOC emissions under the most stringent combination of regulatory alternatives, Regulatory Alternative VII, are almost 97 percent of uncontrolled emission levels and about 62 percent of baseline emission levels.

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 (primarily fugitive VOC control) is estimated to be 66 terajoules (TJ) (11,000 barrels of oil) per year over energy demands in the absence of any VOC control. Most of this increase energy usage is atrributable to

*In this chapter, arabic numerals are used to describe regulatory alternatives when used in the context of a single model plant, or model plants of a single type; Roman numerals are 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 are collectively described as Regulatory Alternative I.

one model plant [poly(ethylene terephthalate), terephthalic acid process using a single end finisher]. Except for the poly(ethylene terephthalate) plants, Regulatory Alternative 2 for all other model plants show a decrease in energy demand because the energy cost for fugitive VOC control under Regulatory Alternative 2 is less than the energy credit resulting from the reduced loss of VOC under these control measures. The most stringent combination of regulatory alternatives, Regulatory Alternative VII, requires about 1,950 TJ (318,000 barrels of oil) per year more than energy demands without VOC controls, and about 940 TJ (154,000 barrels of oil) per year more than the projected requirement under current VOC control practices (Regulatory Alternative I).

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 VII 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 VOC emission rates from process lines in each model plant are determined and used to project industrywide air quality impacts of new polymers amd resins process lines.

7.1.1 Average Annual Model Plant VOC Emissions

Annual VOC emission rates for a process line in 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 a process line in each 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, 1, 2, 3, 4, 5, 6, or 7);

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

F_j = annual VOC fugitive emission rate for a process line in each model plant under Regulatory Alternative $\underline{j}$.

The annual 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 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 a process line in model plant $\underline{i}$;

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

The uncontrolled process emission rates for each process section in a process line in each model plant are presented in Tables 6-1 through 6-11b. The control techniques employed on appropriate model plant process sections and their emission control efficiencies are presented in Tables 6-14 through 6-24b.

The quantity of fugitive VOC emissions are assumed to be the same for each model plant. The uncontrolled fugitive VOC emission rate is based on equipment component counts and uncontrolled component emission rates presented in Table 6-12. The effectiveness of fugitive VOC emission control through the use of leak detection and repair programs 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 6-13.

Table 7-1 presents the primary or VOC related air quality impacts of the regulatory alternatives for a process line in each model plant.

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

Model Plant	Regulatory Alternative	VOC Emissions per Process Line[a]			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined	Process	Fugitive	Combined
PP/Liquid	U[b]	1,858	50	1,908	-	-	-
	1	168	35	203	-	-	-
	2	168	15	183	0	57	10
	3	41	15	56	76	57	72
	4	37	15	52	78	57	74
PP/Gas	U[b]	1,278	50	1,328	-	-	-
	1	26	35	61	-	-	-
	2	26	15	41	0	57	33
LDPE/High Pressure	U[b]	692	38	730	-	-	-
	1	109	27	136	-	-	-
	2	109	12	121	0	56	11
	3	54	12	66	51	56	52
	4	40	12	52	63	56	62
	5	14	12	26	87	564	81
LDPE/Low Pressure HDPE/Gas Phase	U[b]	1,752	75	1,827	-	-	-
	1	49	53	102	-	-	-
	2	49	23	72	0	57	29
	3	42	23	65	14	57	36
	4	38	23	61	22	57	40
	5	34	23	57	31	57	44
HDPE/Slurry	U[b]	931	50	981	-	-	-
	1	46	35	81	-	-	-
	2	46	15	61	0	57	25
	3	18	15	33	61	57	59
HDPE/Solution	U[b]	870	50	920	-	-	-
	1	61	35	96	-	-	-
	2	61	15	76	0	57	21
	3	17	15	32	72	57	67
PS/Continuous	U[b]	10	75	85	-	-	-
	1	10	53	63	-	-	-
	2	10	23	33	0	57	48
	3	6	23	29	41	57	54
	4	5.8	23	28.8	42	57	54.3
	5	5.63	23	28.63	43.7	57	54.6
	6	5.59	23	28.59	44.1	57	54.6
	7	5.57	23	28.57	44.3	57	54.7
EPS/Post-impregnation	U[b]	237	50	287	-	-	-
	1	237	35	272	-	-	-
	2	237	35	252	0	57	7
	3	181	15	196	24	57	28
	4	88	15	103	63	57	62
	5	35	15	50	85	57	82
	6	6	15	21	97.5	57	92.3
	7	5	15	20	97.9	57	92.6

Table 7-1a. PRIMARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMERS AND RESINS PLANTS (Mg/yr) (concluded)

Model Plant	Regulatory Alternative	VOC Emissions per Process Line[a]			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined	Process	Fugitive	Combined
EPS/In-situ	U[b]	40	17	57	-	-	-
	1	40	12	52	-	-	-
	2	40	5.3	45.3	0	56	13
	3	31	5.3	36.3	23	56	30
	4	8.3	5.3	13.6	79	56	74
	5	2.5	5.3	7.8	94	56	85
	6	1.5	5.3	6.8	96	56	87
	7	1.1	5.3	6.4	97	56	88
PET/DMT[c]	U[b]	8	0	8	-	-	-
	1	8	0	8	-	-	-
	2	5.6	0	5.6	30	0	30
	3	5.55	0	5.55	31	0	31
	4	5.44	0	5.44	32	0	32
	5	5.37	0	5.37	33.1	0	33.1
	6	5.36	0	5.36	33.3	0	33.3
	7	2.7	0	2.7	66	0	66
PET/TPA[c]	U[b]	5.3	0	5.3	-	-	-
	1	5.3	0	5.3	-	-	-
	2	2.9	0	2.9	44	0	44
PET/TPA[d]	U[b]	71	0	71	-	-	-
	1	71	0	71	-	-	-
	2	64	0	64	10	0	10
	3	56	0	56	21	0	21
	4	49	0	49	31	0	31
	5	45	0	45	37	0	37
	6	41	0	41	42	0	42
	7	34	0	34	53	0	53

[a] From Table 6-25.

[b] U = uncontrolled.

[c] These regulatory alternatives apply to a process line producing PET with a single end finisher.

[d] These regulatory alternatives apply to a process line producing PET with multiple end finishers.

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 Process line[a]			Percent VOC Emission Reduction from Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined	Process	Fugitive	Combined
PP/Liquid	U[b]	2,048	55	2,103	-	-	-
	1	185	39	224	-	-	-
	2	185	17	202	0	57	10
	3	45	17	62	76	57	72
	4	41	17	57	78	57	74
PP/Gas	U[b]	1,408	55	1,463	-	-	-
	1	29	39	67	-	-	-
	2	29	17	45	0	57	33
LDPE/High Pressure	U[b]	763	42	804	-	-	-
	1	120	30	150	-	-	-
	2	120	13	133	0	56	11
	3	60	13	73	51	56	52
	4	44	13	57	63	56	62
	5	15	13	29	87	56	81
LDPE/Low Pressure HDPE/Gas Phase	U[b]	1,931	83	2,013	-	-	-
	1	54	58	112	-	-	-
	2	54	25	79	0	57	29
	3	46	25	72	14	57	36
	4	42	25	67	22	57	40
	5	37	25	63	31	57	44
HDPE/Slurry	U[b]	1,026	55	1,081	-	-	-
	1	51	39	89	-	-	-
	2	51	17	67	0	57	25
	3	20	17	36	61	57	59
HDPE/Solution	U[b]	959	55	1,014	-	-	-
	1	67	39	106	-	-	-
	2	67	17	84	0	57	21
	3	19	17	35	72	57	67
PS/Continuous	U[b]	11	83	94	-	-	-
	1	11	58	69	-	-	-
	2	11	25	36	0	57	48
	3	6.6	25	31.6	41	57	54
	4	6.4	25	31.4	42	57	54.3
	5	6.2	25	31.2	43.7	57	54.6
	6	6.16	25	31.16	44.1	57	54.6
	7	6.14	25	31.14	44.3	57	54.7
EPS/Post-impregnation	U[b]	261	55	316	-	-	-
	1	261	39	300	-	-	-
	2	261	17	278	0	57	7
	3	199	17	216	24	57	28
	4	97	17	114	63	57	62
	5	39	17	55	85	57	82
	6	6.6	17	23	97.5	57	92.3
	7	5.5	17	22	97.9	57	92.6

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

Model Plant	Regulatory Alternative	VOC Emissions per Process line[a]			Percent VOC Emission Reduction from Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined	Process	Fugitive	Combined
EPS/In-situ	U[b]	44	19	63	-	-	-
	1	44	13	57	-	-	-
	2	44	6	50	0	56	13
	3	34	6	40	23	56	30
	4	9.1	6	15.1	79	56	74
	5	2.8	6	8.8	94	56	85
	6	1.7	6	7.7	96	56	87
	7	1.2	6	7.2	97	56	88
PET/DMT[c]	U[b]	9	0	9	-	-	-
	1	9	0	9	-	-	-
	2	6.2	0	6.2	30	0	30
	3	6.1	0	6.1	31	0	31
	4	6.0	0	6.0	32	0	32
	5	5.92	0	5.92	33.1	0	33.1
	6	5.91	0	5.91	33.3	0	33.3
	7	2.98	0	2.98	66	0	66
PET/TPA[c]	U[b]	6	0	6	-	-	-
	1	6	0	6	-	-	-
	2	3	0	3	44	0	44
PET/TPA[d]	U[b]	78	0	78	-	-	-
	1	78	0	78	-	-	-
	2	70	0	70	10	0	10
	3	62	0	62	21	0	21
	4	54	0	54	31	0	31
	5	49	0	49	37	0	37
	6	45	0	45	42	0	42
	7	37	0	37	53	0	53

[a] From Table 6-25.

[b] U = uncontrolled.

[c] These regulatory alternatives apply to a process line producing PET with a single end finisher.

[d] These regulatory alternatives apply to a process line producing PET with multiple end finishers.

(Table 7-1 and other tables, as appropriate, in this chapter have been separated into 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 98 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. 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 57 percent in fugitive emissions of VOC from a process line in each model plant except for the PET process lines. Regulatory Alternative 2 for the poly(ethylene terepthalate) process lines reduces baseline emissions either through the use of a distillation column or a refrigerated condenser. Under Regulatory Alternative II, overall VOC emission reductions from baseline control (processes and fugitive emission reductions) range from 7 to 48 percent. The differences in emission reduction efficiency over baseline among process lines in 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 a process line in the model plant.

Under Regulatory Alternatives III through VII, more stringent controls are applied to process emissions. Implementation of Regulatory Alternative VII, the most stringent combination of regulatory alternatives, would result in a range of overall VOC emission reduction from baseline control of about 33 to 93 percent.

7.1.2 Industrywide VOC Emission Impacts of New Plants

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

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

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

E_{ij} = annual emission rate from a process line in model plant i under Regulatory Altertative j (Mg/yr-from Table 7-1);

n_i = number of new process lines of model plant type i projected to be built by 1988 (from Table 8-43).

The impact of the regulatory alternatives on VOC emissions from the 85 new process lines expected to be built are presented in Table 7-2. Baseline emissions from new process lines are projected to be 7,332 Mg/yr (8,080 tons/yr), or a 91 percent overall reduction from uncontrolled levels. The additional increment of emission control under Regulatory Alternative II results in overall VOC emissions of about 5,837 Mg/yr (6,432 tons/yr): a 93 percent decrease from the uncontrolled emission rate, or a 20 percent incremental reduction in emissions over Regulatory Alternative I. The additional process emission controls required under Regulatory Alternative VII, the most stringent combination of regulatory alternatives, are expected to reduce uncontrolled emissions by almost 97 percent to about 2,770 Mg/yr (3,056 tons/yr): a 62 percent incremental reduction in emissions over Regulatory Alternative I.

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

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 Process line[a]			Number of New Process Lines[b]	Industrywide VOC Emissions For New Process Lines			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined		Process	Fugitive	Combined	Process	Fugitive	Combined
PP/Liquid	U[c]	1,858	50	1,908		16,722	450	17,172	-	-	-
	1	168	35	203		1,512	315	1,827	-	-	-
	2	168	15	183	9	1,512	135	1,647	0	57	10
	3	41	15	56		369	135	504	76	57	72
	4	37	15	52		333	135	468	78	57	74
PP/Gas	U[c]	1,278	50	1,328		11,502	450	11,952	-	-	-
	1	26	35	61	9	234	315	549	-	-	-
	2	26	15	41		234	135	369	0	57	33
LDPE/High Pressure	U[c]	692	38	730		2,768	152	2,920	-	-	-
	1	109	27	136		436	108	544	-	-	-
	2	109	12	121	4	436	48	484	0	56	11
	3	54	12	66		216	48	264	51	56	52
	4	40	12	52		160	48	208	63	56	62
	5	14	12	26		56	48	104	87	56	81
LDPE/Low Pressure HDPE/Gas Phase	U[c]	1,752	75	1,827		35,040	1,500	36,540	-	-	-
	1	49	53	102		980	1,060	2,040	-	-	-
	2	49	23	72	20	980	460	1,440	0	57	29
	3	42	23	65		840	460	1,300	14	57	36
	4	38	23	61		760	460	1,220	22	57	40
	5	34	23	57		680	460	1,140	31	57	44
HDPE/Slurry	U[c]	931	50	981		5,586	300	5,886	-	-	-
	1	46	35	81	6	276	210	486	-	-	-
	2	46	15	61		276	90	366	0	57	25
	3	18	15	33		108	90	198	61	57	59
HDPE/Solution	U[c]	870	50	920		7,830	450	8,280	-	-	-
	1	61	35	96		549	315	864	-	-	-
	2	61	15	76	9	549	135	684	0	57	21
	3	17	15	32		153	135	288	72	57	67
PS/Continuous	U[c]	10	75	85		20	150	170	-	-	-
	1	10	53	63		20	106	126	-	-	-
	2	10	23	33	2	20	46	66	0	57	48
	3	6	23	29		12	46	58	41	57	54
	4	5.8	23	28.8		11.5	46	57.5	42	57	54.3
	5	5.63	23	28.63		11.3	46	57.3	43.7	57	54.6
	6	5.59	23	28.59		11.2	46	57.2	44.1	57	54.6
	7	5.57	23	28.57		11.1	46	57.1	44.3	57	54.7
EPS/Post-impregnation	U[c]	237	50	287		474	100	574	-	-	-
	1	237	35	272		474	70	544	-	-	-
	2	237	35	252	2	474	30	504	0	57	7
	3	181	15	196		362	30	392	24	57	28
	4	88	15	103		176	30	206	63	57	62
	5	35	15	50		70	30	100	85	57	82
	6	6	15	21		12	30	42	97.5	57	92.3
	7	5	15	20		10	30	40	97.9	57	92.6

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 Process line[a]			Number of New Process Lines[b]	Industrywide VOC Emissions For New Process Lines			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined		Process	Fugitive	Combined	Process	Fugitive	Combined
EPS/In-situ	U[c]	40	17	57		120	51	171	-	-	-
	1	40	12	52		120	36	156	-	-	-
	2	40	5.3	45.3	3	120	16	136	0	56	13
	3	31	5.3	36.3		93	16	109	23	56	30
	4	8.3	5.3	13.6		25	16	41	79	56	74
	5	2.5	5.3	7.8		7.5	16	23.5	94	56	85
	6	1.5	5.3	6.8		4.5	16	20.5	96	56	87
	7	1.1	5.3	6.4		3.3	16	19.3	97	56	88
PET/DMT[d]	U[c]	8	0	8		56	0	56	-	-	-
	1	8	0	8		56	0	56	-	-	-
	2	5.6	0	5.6	7	39.2	0	39.2	30	0	30
	3	5.55	0	5.55		38.9	0	38.9	31	0	31
	4	5.44	0	5.44		38.1	0	38.1	32	0	32
	5	5.37	0	5.37		37.6	0	37.6	33.1	0	33.1
	6	5.36	0	5.36		37.5	0	37.5	33.3	0	33.3
	7	2.7	0	2.7		18.9	0	18.9	66	0	66
PET/TPA[d]	U[c]	5.3	0	5.3		69	0	69	-	-	-
	1	5.3	0	5.3	13	69	0	69	-	-	-
	2	2.9	0	2.9		38	0	38	44	0	44
PET/TPA[e]	U[c]	71	0	71		71	0	71	-	-	-
	1	71	0	71	1	71	0	71	-	-	-
	2	64	0	64		64	0	64	10	0	10
	3	56	0	56		56	0	56	21	0	21
	4	49	0	49		49	0	49	31	0	31
	5	45	0	45		45	0	45	37	0	37
	6	41	0	41		41	0	41	42	0	42
	7	34	0	34		34	0	34	53	0	53
Total[f]	U[c]					80,258	3,603	83,861	-	-	-
	I					4,797	2,535	7,332	-	-	-
	II					4,742	1,095	5,837	1	57	20
	III				85	2,520	1,095	3,615	48	57	51
	IV					2,085	1,095	3,180	57	57	57
	V					1,773	1,095	2,868	63	57	61
	VI					1,708	1,095	2,803	64	57	61.8
	VII					1,678	1,095	2,773	65	57	62.2

[a] From Table 7-1.

[b] From Table 8-43.

[c] U = uncontrolled.

[d] These regulatory alternatives apply to a process line producing PET with a single end finisher.

[e] These regulatory alternatives apply to a process line producing PET with multiple end finishers.

[f] Where there is no emission 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 7-2b. INDUSTRYWIDE PRIMARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMERS AND RESINS PLANTS (tons/yr)

Model Plant	Regulatory Alternative	VOC Emissions per Process line[a]			Number of New Process Lines[b]	Industrywide VOC Emissions For New Process Lines			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined		Process	Fugitive	Combined	Process	Fugitive	Combined
PP/Liquid	UC	2,048	55	2,103		18,428	496	18,924	-	-	-
	1	185	39	224	9	1,666	347	2,013	-	-	-
	2	185	17	202		1,666	149	1,815	0	57	10
	3	45	17	62		407	149	555	76	57	72
	4	41	17	57		367	149	516	78	57	74
PP/Gas	UC	1,408	55	1,463		12,675	496	13,171	-	-	-
	1	29	39	67	9	258	347	605	-	-	-
	2	29	17	46		258	149	407	0	57	33
LDPE/High Pressure	UC	763	42	804		3,050	168	3,218	-	-	-
	1	120	30	150		480	119	599	-	-	-
	2	120	13	133	4	480	53	533	0	56	11
	3	60	13	73		238	53	291	51	56	52
	4	44	13	57		176	53	229	63	56	62
	5	15	13	29		62	53	115	87	56	81
LDPE/Low Pressure HDPE/Gas Phase	UC	1,931	83	2,013		38,614	1,653	40,267	-	-	-
	1	54	58	112		1,080	1,168	2,248	-	-	-
	2	54	25	79	20	1,080	507	1,587	0	57	29
	3	46	25	72		926	507	1,433	14	57	36
	4	42	25	67		838	507	1,344	22	57	40
	5	37	25	63		749	507	1,256	31	57	44
HDPE/Slurry	UC	1,026	55	1,081		6,156	331	6,486	-	-	-
	1	51	39	89		304	231	536	-	-	-
	2	51	17	67	6	304	99	403	0	57	25
	3	20	17	36		119	99	218	61	57	59
HDPE/Solution	UC	959	55	1,014		8,629	496	9,125	-	-	-
	1	67	39	106	9	605	347	952	-	-	-
	2	67	17	84		605	149	754	0	57	21
	3	19	17	35		169	149	317	72	57	67
PS/Continuous	UC	11	83	94		22	165	187	-	-	-
	1	11	58	69		22	117	139	-	-	-
	2	11	25	36	2	22	51	73	0	57	48
	3	6.6	25	31.6		13	51	64	41	57	54
	4	6.4	25	31.4		13	51	63	42	57	54.3
	5	6.2	25	31.2		12	51	63	43.7	57	54.6
	6	6.16	25	31.16		12	51	63	44.1	57	54.6
	7	6.14	25	31.14		12	51	63	44.3	57	54.7
EPS/Post-impregnation	UC	261	55	316		522	110	633	-	-	-
	1	261	39	300		522	77	599	-	-	-
	2	261	17	278		522	33	555	0	57	7
	3	199	17	216	2	399	33	432	24	57	28
	4	97	17	114		194	33	227	63	57	62
	5	39	17	55		77	33	110	85	57	82
	6	6.6	17	23		13	33	46	97.5	57	92.3
	7	5.5	17	22		11	33	44	97.9	57	92.6

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 Process line[a]			Number of New Process Lines[b]	Industrywide VOC Emissions For New Process Lines			Percent VOC Emission Reduction From Baseline (Regulatory Alternative 1)		
		Process	Fugitive	Combined		Process	Fugitive	Combined	Process	Fugitive	Combined
EPS/In-situ	U[c]	44	19	63		132	56	188	-	-	-
	1	44	13	57		132	40	172	-	-	-
	2	44	6	50	3	132	18	150	0	56	13
	3	34	6	40		102	18	120	23	56	30
	4	9.1	6	15.1		27	18	45	79	56	74
	5	2.8	6	8.8		8	18	26	94	56	85
	6	1.7	6	7.7		5	18	22	96	56	87
	7	1.2	6	7.2		4	18	21	97	56	88
PET/DMT[d]	U[c]	9	0	9		62	0	62	-	-	-
	1	9	0	9	7	62	0	62	-	-	-
	2	6.2	0	6.2		43	0	43	30	0	30
	3	6.1	0	6.1		43	0	43	31	0	31
	4	6.0	0	6.0		42	0	42	32	0	32
	5	5.92	0	5.92		41	0	41	33.1	0	33.1
	6	5.91	0	5.91		41	0	41	33.3	0	33.3
	7	2.98	0	2.98		21	0	21	66	0	66
PET/TPA[d]	U[c]	6	0	6		76	0	76	-	-	-
	1	6	0	6	13	76	0	76	-	-	-
	2	3	0	3		42	0	42	44	0	44
PET/TPA[e]	U[b]	78	0	78	1	78	0	78	-	-	-
	1	78	0	78		78	0	78	-	-	-
	2	70	0	70		70	0	70	10	0	10
	3	62	0	62		62	0	62	21	0	21
	4	54	0	54		54	0	54	31	0	31
	5	49	0	49		49	0	49	37	0	37
	6	45	0	45		45	0	45	42	0	42
	7	37	0	37		37	0	37	53	0	53
Total[f]	U[c]					88,444	3,971	92,415	-	-	-
	I					5,286	2,794	8,080	-	-	-
	II					5,225	1,207	6,432	1	57	20
	III				85	2,777	1,207	3,983	48	57	51
	IV					2,298	1,207	3,504	57	57	57
	V					1,954	1,207	3,160	63	57	61
	VI					1,882	1,207	3,089	64	57	61.8
	VII					1,849	1,207	3,056	65	57	62.2

[a] From Table 7-1.

[b] From Table 8-43.

[c] U = uncontrolled.

[d] These regulatory alternatives apply to a process line producing PET with a single end finisher.

[e] These regulatory alternatives apply to a process line producing PET with multiple end finishers.

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

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:

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

where S_{xj} = new process line 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); 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 a process line in model plant $\underline{i}$ under Regulatory Alternative $\underline{j}$ (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 a process line 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 process lines 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 implementaion 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 process lines are presented in Table 7-3.

The 85 new process lines that are expected to come on line by 1988 are projected to emit approximately 19.3 Mg/yr (21.3 tons/yr) of particulate, 822 Mg/yr (907 tons/yr) of SO_x, and 173 Mg/yr (190 tons/yr) of NO_x, if these plants operate under current industry practices (Regulatory

Table 7-3a. SECONDARY AIR QUALITY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMERS AND RESINS PLANTS

Model Plant	Regulatory Alternatives	Process Line Emissions, kg/yr			Number of New Process Lines[b]	Industrywide Emissions, Mg/yr		
		Particulate	SO_x	NO_x		Particulate	SO_x	NO_x
PP/Liquid	1	73	2,660	720		0.66	23.9	6.49
	2	73	2,660	720	9	0.66	23.9	6.49
	3	78	2,690	790		0.70	24.1	7.08
	4	83	2,690	860		0.74	24.2	7.73
PP/Gas	1	84	3,130	810		0.75	28.1	7.29
	2	84	3,130	810	9	0.75	28.1	7.29
LDPE/High Pressure	1	23	840	230		0.09	3.36	0.93
	2	23	840	230	4	0.09	3.36	0.93
	3	35	1,090	370		0.14	4.38	1.50
	4	38	1,160	410		0.15	4.61	1.64
	5	185	6,870	1,790		0.74	27.4	7.14
LDPE/Low Pressure HDPE/Gas Phase	1	100	3,275	1,050		2.01	65.4	21.0
	2	100	3,275	1,050		2.01	65.4	21.0
	3	105	3,280	1,130	20	2.11	65.5	22.5
	4	120	3,450	1,340		2.42	68.9	26.7
	5	160	4,690	1,720		3.16	93.7	34.3
HDPE/Slurry	1	49	1,780	480		0.29	10.7	2.89
	2	49	1,780	480	6	0.29	10.7	2.89
	3	50	1,820	500		0.30	10.9	2.98
HDPE/Solution	1	41	1,210	440		0.37	10.8	3.99
	2	41	1,210	440	9	0.37	10.8	3.99
	3	230	1,470	3,300		2.06	13.2	29.7
PS/Continuous	1	c	c	c		c	c	c
	2	c	c	c	2	c	c	c
	3	1	36	7		0.002	0.07	0.01
	4	1	39	7		0.002	0.08	0.01
	5	1	54	10		0.002	0.11	0.02
	6	1	56	11		0.002	0.11	0.02
	7	2	70	13		0.004	0.14	0.03
EPS/Post-impregnation	1	c	c	c		c	c	c
	2	c	c	c		c	c	c
	3	21	300	275	2	0.04	0.60	0.55
	4	225	970	3,280		0.44	1.94	6.55
	5	320	2,000	4,630		0.64	4.00	9.24
	6	1,070	2,140	16,100		2.13	4.27	32.0
	7	1,070	2,140	16,100		2.13	4.27	32.0

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

Model Plant	Regulatory Alternatives	Process Line Emissions, kg/yr			Number of New Process Lines[b]	Industrywide Emissions, Mg/yr		
		Particulate	SO_x	NO_x		Particulate	SO_x	NO_x
EPS/In-situ	1	c	c	c		c	c	c
	2	c	c	c		c	c	c
	3	12	320	135	3	0.04	0.95	0.40
	4	24	480	290		0.07	1.44	0.87
	5	32	575	410		0.10	1.72	1.22
	6	46	880	565		0.14	2.64	1.69
	7	59	1,190	720		0.18	3.56	2.16
PET/DMT[d]	1	756	33,910	6,480		5.28	237	38.0
	2	756	33,930	6,480	7	5.28	237	45.2
	3	756	33,930	6,480		5.28	237	45.3
	4	756	33,930	6,480		5.28	237	45.7
	5	757	33,940	6,490		5.28	237	45.3
	6	757	33,940	6,490		5.29	237	45.3
	7	1,000	44,990	8,600		7.02	315	60.2
PET/TPA[d]	1	756	33,910	6,480		9.81	440	84.1
	2	975	43,750	8,360	13	12.7	569	109
PET/TPA[e]	1	74	3,330	635		0.07	3.32	0.63
	2	84	3,770	720	1	0.08	3.76	0.72
	3	98	4,380	840		0.10	4.37	0.84
	4	115	5,230	1,000		0.12	5.22	1.00
	5	130	5,790	1,110		0.13	5.77	1.10
	6	145	6,460	1,240		0.14	6.45	1.23
	7	185	8,410	1,610		0.19	8.39	1.60
TOTAL[f]	I	-	-	-		19.3[c]	822[c]	173[c]
	II	-	-	-		22.2[c]	952[c]	197[c]
	III	-	-	-	85	24.2	958	227
	IV	-	-	-		25.0	964	238
	V	-	-	-		26.6	1,015	255
	VI	-	-	-		28.1	1,017	278
	VII	-	-	-		30.0	1,098	294

[d]Emission factors from References 2 and 3. Fuel consumption under each regulatory alternative is derived from Tables 8-21 and 8-31b. Factors used are for combustion of No. 4 fuel oil having 2 percent sulfur content by weight. (Source: Reference 4.)

[b]From Table 8-43.

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

[d]These regulatory alternatives apply to a process line producing PET with a single end finisher.

[e]These regulatory alternatives apply to a process line producing PET with multiple end finishers.

[f]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.

Table 7-3b. SECONDARY AIR QUALITY IMPACTS FOR THE REGULATORY ALTERNATIVES FOR POLYMERS AND RESINS PLANTS

Model Plant	Regulatory Alternatives	Process Line Emissions,[a] lbs/yr			Number of New Process Lines[b]	Industrywide Emissions, tons/yr		
		Particulate	SO_x	NO_x		Particulate	SO_x	NO_x
PP/Liquid	1	160	5,860	1,590		0.73	26.4	7.15
	2	160	5,860	1,590	9	0.73	26.4	7.15
	3	170	5,910	1,735		0.77	26.6	7.80
	4	180	5,930	1,890		0.82	26.7	8.52
PP/Gas	1	185	6,890	1,790		0.83	31.0	8.04
	2	185	6,890	1,790	9	0.83	31.0	8.04
LDPE/High Pressure	1	50	1,850	510		0.10	3.70	1.02
	2	50	1,850	510		0.10	3.70	1.02
	3	75	2,400	825	4	0.15	4.81	1.65
	4	85	2,540	905		0.17	5.08	1.81
	5	405	15,100	3,930		0.81	30.2	7.87
LDPE/Low Pressure HDPE/Gas Phase	1	220	7,200	2,320		2.22	72.1	23.2
	2	220	7,200	2,320	20	2.22	72.1	23.2
	3	235	7,220	2,480		2.33	72.2	24.8
	4	265	7,590	2,940		2.67	75.9	29.4
	5	350	10,330	3,780		3.48	103	37.8
HDPE/Slurry	1	108	3,910	1,060		0.32	11.7	3.18
	2	108	3,910	1,060	6	0.32	11.7	3.18
	3	111	3,990	1,100		0.33	12.0	3.28
HDPE/Solution	1	90	2,650	975		0.40	11.9	4.39
	2	90	2,650	975	9	0.40	11.9	4.39
	3	505	3,230	7,260		2.27	14.6	32.7
PS/Continuous	1	c	c	c		c	c	c
	2	c	c	c	2	c	c	c
	3	2	80	15		0.002	0.08	0.02
	4	2	85	16		0.002	0.09	0.02
	5	3	120	23		0.003	0.12	0.02
	6	3	125	24		0.003	0.12	0.02
	7	3	155	29		0.003	0.15	0.03
EPS/Post-impregnation	1	c	c	c		c	c	c
	2	c	c	c		c	c	c
	3	46	660	600		0.05	0.66	0.60
	4	490	2,140	7,220	2	0.49	2.13	7.22
	5	705	4,410	10,200		0.71	4.41	10.2
	6	2,350	4,700	35,300		2.35	4.70	35.3
	7	2,350	4,700	35,300		2.35	4.70	35.3

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

Model Plant	Regulatory Alternatives	Process Line Emissions,[a] lbs/yr			Number of New Process Lines[b]	Industrywide Emissions, tons/yr		
		Particulate	SO_x	NO_x		Particulate	SO_x	NO_x
EPS/In-situ	1	c	c	c		c	c	c
	2	c	c	c		c	c	c
	3	26	700	290	3	0.04	1.05	0.44
	4	52	1,060	640		0.08	1.58	0.96
	5	71	1,260	900		0.11	1.89	1.35
	6	100	1,940	1,240		0.15	2.91	1.87
	7	130	2,620	1,590		0.19	3.93	2.38
PET/DMT[d]	1	1,660	74,610	14,257		5.82	261	49.9
	2	1,660	74,645	14,263	7	5.82	261	49.9
	3	1,660	74,646	14,264		5.82	261	49.9
	4	1,660	74,652	14,265		5.82	261	49.9
	5	1,660	74,659	14,266		5.83	261	49.9
	6	1,660	74,667	14,268		5.83	261	49.9
	7	2,210	99,190	18,954		7.74	347	66.3
PET/TPA[d]	1	1,660	74,600	14,300		10.8	485	92.7
	2	2,150	96,450	18,430	13	14.0	627	120
PET/TPA[e]	1	165	7,320	1,400		0.08	3.66	0.70
	2	185	8,290	1,580	1	0.09	4.15	0.79
	3	215	9,640	1,840		0.11	4.82	0.92
	4	255	11,500	2,200		0.13	5.75	1.10
	5	285	12,700	2,430		0.14	6.37	1.22
	6	315	14,200	2,720		0.16	7.11	1.36
	7	410	18,500	3,540		0.21	9.25	1.77
TOTAL[f]	I	-	-	-		21.3[c]	907[c]	190[c]
	II	-	-	-		24.5[c]	1,049[c]	218[c]
	III	-	-	-		26.7	1,056	250
	IV	-	-	-		27.6	1,063	263
	V	-	-	-		29.3	1,119	281
	VI	-	-	-		31.0	1,121	307
	VII	-	-	-		33.0	1,210	324

[a]Emission factors from References 2 and 3. Fuel consumption under each regulatory alternative is derived from Tables 8-21 and 8-31b. Factors used are for combustion of No. 4 fuel oil having 2 percent sulfur content by weight. (Source: Reference 4.)

[b]From Table 8-43.

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

[d]These regulatory alternatives apply to a process line producing PET with a single end finisher.

[e]These regulatory alternatives apply to a process line producing PET with multiple end finishers.

[f]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.

Alternative I). Under Regulatory Alternative II, secondary air pollutant emissions will not increase appreciably for all model plants except one because the new process lines will employ fugitive emission control programs, which do not require the use of add-on pollution control systems. The new PET process lines using a single end finisher under Regulatory Alternative 2 are estimated, on a per process line basis, to emit an additional 0.2 Mg/yr (0.2 tons/yr) of particulate, 9.8 Mg/yr (12.3 tons/yr) of SO_x, and 1.9 Mg/yr (2.3 tons/yr) of NO_x above baseline control.

Particulate emissions generated by new process lines during control of VOC containing process streams under Regulatory Alternative III are about 24.2 Mg/yr (26.7 tons/yr), or 25 percent greater than secondary particulate emissions under baseline conditions. SO_x emissions under Regulatory Alternative III are projected at 958 Mg/yr (1,056 tons/yr): a 16 percent increase in SO_x emissions over Regulatory Alternative I. NO_x emissions under Regulatory Alternative III are estimated to be about 227 Mg/yr (250 tons/yr), which represents a 31 percent increase over baseline secondary NO_x emissions.

Projected particulate, SO_x and NO_x emissions from new process lines operating under Regulatory Alternative VII, the most stringent combination of regulatory alternatives, are 30.0 Mg/yr (33.0 tons/yr), 1,098 Mg/yr (1,210 tons/yr) and 294 Mg/yr (324 tons/yr), respectively. These emission levels are equivalent to increases over baseline levels of about 55 percent for particulates, 34 percent for SO_x, and 41 percent for NO_x. 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

Under the regulatory alternatives analyzed, increased water use will occur only in PET process lines producing a high viscosity product. Water usage in a PET plant comprised of seven process lines is estimated to increase by 58,674 m^3 (15.5 x 10^6 gal/year), or about 7 percent of the consumptive water requirements of a continuous polyester resin plant with an annual capacity of 105 Gg (231 x 10^6 lbs).[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 a number of regulatory alternatives for several different process lines.

The quantity of spent catalyst expected to be generated annually by new process lines 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 reaction, material recovery, product finishing, or product storage) in a process line in model plant $\underline{i}$ (i = liquid phase polypropylene and other process types) under Regulatory Alternative $\underline{j}$ (refer to Tables 6-1 to 6-11);

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

N_i = number of new process lines of model plant type $\underline{i}$ expected to be built by 1988 (Table 8-43).

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 44 new process lines expected to employ catalytic incinerators is projected to be 3.35 m^3/yr (119 ft^3/yr) under Regulatory Alternative III, 3.90 m^3/yr (138 ft^3/yr) under Regulatory Alternative IV, and 4.21 m^3/yr (149 ft^3/yr) under Regulatory Alternatives V through VII. The volume of solid waste generated by

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

Model Plant	Regulatory Alternative	Flow,[b] scfm	Solid Waste Generated,[c] $m^3(ft^3)/yr$[b]	Number of New Process Lines[d]	Solid Waste Generated, $m^3(ft^3)/yr$
LDPE, High Pressure	1	0	0		0
	2	0	0		0
	3	1,590	0.034 (1.2)	4	0.14 (4.8)
	4	1,990	0.042 (1.5)		0.17 (6.0)
	5	1,990	0.042 (1.5)		0.17 (6.0)
LDPE, Low Pressure and HDPE, Gas Phase	1	0	0		0
	2	0	0		0
	3	0	0	20	0
	4	1,070	0.023 (0.8)		0.45 (16)
	5	1,070	0.023 (0.8)		0.45 (16)
HDPE, Slurry	1	0	0		0
	2	0	0	6	0
	3	230	0.005 (0.2)		0.03 (1.2)
HDPE, Solution	1	0	0		0
	2	0	0	9	0
	3	16,240	0.345 (12.2)		3.1 (110)
EPS/Post-impregnation	1	0	0		0
	2	0	0		0
	3	1,890	0.04 (1.4)		0.08 (2.8)
	4	1,890	0.04 (1.4)	2	0.08 (2.8)
	5	8,405	0.18 (6.3)		0.36 (12.6)
	6	8,405	0.18 (6.3)		0.36 (12.6)
	7	8,405	0.18 (6.3)		0.36 (12.6)
EPS/In-situ	1	0	0		0
	2	0	0		0
	3	0	0		0
	4	1,020	0.022 (0.77)	3	0.07 (2.3)
	5	1,615	0.034 (1.2)		0.10 (3.6)
	6	1,615	0.034 (1.2)		0.10 (3.6)
	7	1,615	0.034 (1.2)		0.10 (3.6)
TOTAL[e]	I	-	-		0
	II	-	-		0
	III	-	-		3.35 (119)
	IV	-	-		3.90 (138)
	V	-	-		4.21 (149)
	VI	-	-		4.21 (149)
	VII	-	-		4.21 (149)

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

[b]Per process line. See Tables 6-1 through 6-11b and Tables 8-21 through 8-31b, as appropriate.

[c]Based on 2.25 ft^3 of catalyst/1,000 scfm of catalytic incinerator throughput and a three year catalyst life.

[d]See Table 8-43.

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

the use of spray towers or liquid jet scrubbers employed upstream of catalytic incineraters to treat corrosive exhaust gases under Regulatory Alternative VII represents about 0.01 percent of the volume of biological sludge generated annually as solid waste by polyethylene process operations.[7] Volumes of biological sludge generated by applicable process operations in new process lines 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 process lines is projected to be about 1,325 m^3/yr (46,750 ft^3yr) under Regulatory Alternative VII, the alternative having the greatest potential for secondary pollutant emissions. This fly ash generation rate represents 1.2 to 3.4 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]

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

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

Model Plant	Annual Production Rate Per Process Line,[b] Gg/yr	(lb/yr)	Number of New Process Lines[c]	Industrywide Biological Sludge Production Rate,[d] m^3/yr (ft^3/yr)
PP/Liquid	50	(110 x 10^6)	9	4,850-12,000(171,000-431,000)
PP/Gas	35	(77 x 10^6)	9	3,400-8,520(120,000-301,000)
LDPE/High	70	(154 x 10^6)	4	3,020-7,580(107,000-268,000)
LDPE/Low Pressure and HDPE/ Gas Phase	75	(165 x 10^6)	20	16,200-40,600(572,000-1,430,000)
HDPE/Slurry	70	(154 x 10^6)	6	4,530-11,400(160,000-403,000)
HDPE/Solution	30	(66 x 10^6)	9	2,910-7,310(103,000-258,000)
PS/Continuous	37.5	(83 x 10^6)	2	813-2,040(28,700-72,000)
EPS/Post-impregnation	25	(55 x 10^6)	2	539-1,350(19,000-47,700)
EPS/In-situ	8	(18 x 10^6)	3	264-664(9,320-23,400)
PET/DMT	15	(33 x 10^6)	7	2,840-11,400(100,000-403,000)
PET/TPA	15[e]	(33 x 10^6)	13	5,280-21,150(186,000-747,000)
	20[f]	(44 x 10^6	1	541-2,170(19,100-76,600)
TOTAL	-	-	85	39,400-103,000(1.39 x 10^6-3.64 x 10^6)[g] 27,500-68,900(970,000-2.43 x 10^6)[g,h]

[a]These devices were not a part of any regulatory alternative analyzed for PET/TPA process lines. For illustration purposes, however, this model plant is included in this table.

[b]From Tables 6-1 through 6-11b.

[c]From Table 8-43.

[d]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.)

[e]Single end finisher.

[f]Multiple end finishers.

[g]Does not include PET/TPA numbers.

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

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 process lines in each model plant are calculated as the sum of energy expended to control VOC emissions from process operations and the energy credit for reduced VOC losses by implementing fugitive VOC emission control practices and process VOC emission recovery 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, electricty 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 VOC emissions from a process line in each model plant under each regulatory alternative are presented in Table 7-6.

Energy impacts associated with process VOC emission control 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. 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 VII.

Table 7-6 shows that the energy impact of process VOC emission reduction in a model plant 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.

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

Model Plant	Regulatory Alternatives	Process Emissions Control Energy Impact (GJ/yr)				Fugitive Emission Control Energy Impact		Combined Emission Control Energy Impacts
		Natural Gas[b]	Electricity[c]	Steam[d]	Process Total[e]	Mg VOC Reduced	GJ/yr[f]	GJ/yr
PP/Liquid	1	2,260	-	2,720	4,990	0	0	4,990
	2	2,260	-	2,720	4,990	20	-915	4,080
	3	2,910	23	2,720	5,660	20	-915	4,750
	4	3,670	23	2,730	6,430	20	-915	5,510
PP/Gas	1	2,260	-	3,210	5,470	0	0	5,470
	2	2,260	-	3,210	5,470	20	-915	4,560
LDPE/High Pressure	1	754	-	862	1,620	0	0	1,620
	2	754	-	862	1,620	15	-706	910
	3	1,770	257	862	2,890	15	-706	2,180
	4	2,020	321	862	3,200	15	-706	2,500
	5	5,040	321	6,720	12,100	15	-706	11,400
LDPE/Low Pressure HDPE/Gas Phase	1	4,540	-	3,360	7,890	0	0	7,890
	2	4,540	-	3,360	7,890	30	-1,410	6,480
	3	5,290	-	3,360	8,650	30	-1,410	7,240
	4	7,200	172	3,360	10,700	30	-1,410	9,320
	5	8,700	172	4,640	13,500	30	-1,410	12,100
HDPE/Slurry	1	1,510	-	1,820	3,330	0	0	3,330
	2	1,510	-	1,820	3,330	20	-952	2,380
	3	1,600	37	1,820	3,460	20	-952	2,510
HDPE/Solution	1	2,260	-	1,240	3,500	0	0	3,500
	2	2,260	-	1,240	3,500	20	-952	2,550
	3	32,000	263	1,240	33,500	20	-952	32,570
PS/Continuous[g]	1	-	-	-	0	0	0	0
	2	-	-	-	0	30	-1,230	-1,230
	3	-	37	-	37	30	-1,230	-1,358[h]
	4	-	40	-	40	30	-1,230	-1,360[h]
	5	-	55	-	55	30	-1,230	-1,351[h]
	6	-	57	-	57	30	-1,230	-1,349[h]
	7	-	72	-	72	30	-1,230	-1,341[h]
EPS/Post-impregnation	1	-	-	-	-	0	0	0
	2	-	-	-	-	12	-516	-516
	3	2,290	307	0	2,600	12	-516	2,080
	4	32,850	307	679	33,800	12	-516	33,300
	5	45,000	1,360	679	47,000	12	-516	46,500
	6	165,900	1,360	785	168,000	12	-516	167,500
	7	165,900	1,360	785	168,000	12	-516	167,500

Table 7-6a. ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR POLYMER AND RESIN PLANTS[a] (concluded)

Model Plant	Regulatory Alternatives	Process Emissions Control Energy Impact (GJ/yr)				Fugitive Emission Control Energy Impact		Combined Emission Control Energy Impacts
		Natural Gas[b]	Electricity[c]	Steam[d]	Process Total[e]	Mg VOC Reduced	GJ/yr[f]	GJ/yr
EPS/In-situ	1	-	-	-	-	0	0	0
	2	-	-	-	-	6.7	-290	-290
	3	760	0	326	1,090	6.7	-290	800
	4	2,100	164	326	2,590	6.7	-290	2,300
	5	3,165	261	326	3,750	6.7	-290	3,460
	6	4,210	261	642	5,110	6.7	-290	4,820
	7	5,233	261	958	6,450	6.7	-290	6,160
PET/DMT[i]	1	0	2,400	32,350	34,750	0	0	34,750
	2	0	2,410	32,350	34,760	0	0	34,720[j]
	3	0	2,420	32,350	34,770	0	0	34,730[k]
	4	0	2,420	32,350	34,770	0	0	34,720[j]
	5	0	2,420	32,350	34,770	0	0	34,720[j]
	6	0	2,425	32,350	34,775	0	0	34,730[j]
	7	0	2,425	43,775	46,200	0	0	46,100[j]
PET/TPA[i]	1	0	2,400	32,350	34,750	0	0	34,750
	2	0	2,400	42,525	44,925	0	0	44,875[j]
PET/TPA[k]	1	0	2,090	1,320	3,410	0	0	3,410
	2	0	2,090	1,770	3,860	0	0	3,730[j]
	3	0	2,090	2,400	4,490	0	0	4,230[j]
	4	0	2,090	3,270	5,360	0	0	4,950[j]
	5	0	2,090	3,840	5,930	0	0	5,460[j]
	6	0	2,090	4,530	6,620	0	0	6,080[j]
	7	0	2,090	6,520	8,610	0	0	7,940[j]

[a]Impacts are reported on a per process line basis.

[b]10^6 Btu = 1.0551 GJ. This factor includes the energy conversion efficiency; it is not an energy equivalent. (Reference 12.)

[c]1 kWh = 9.476×10^{-3} GJ. This factor includes the energy conversion efficiency; it is not an energy equivalent. (Reference 12.)

[d]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.]

[e]1 GJ = 0.1628 bbl crude oil. (Reference 13.)

[f]Based on the following: 17,598 Btu/lb of styrene monomer; 19,683 Btu/lb of propylene; 20,276 Btu/lb of ethylene; 7,310 Btu/lb of ethylene glycol; and 19,499 Btu/lb of pentane. (Reference 14.)

[g]Estimates of energy use under baseline control are not available.

[h]Includes energy credit of 360 Gg for recovered styrene monomer.

[i]These regulatory alternatives apply to a process line producing PET with a single end finisher.

[j]Includes energy credit for recovered ethylene glycol.

[k]These regulatory alternatives apply to a process line producing PET with multiple end finishers.

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 PLANTS[a]

Model Plant	Regulatory Alternatives	Process Emissions Control Energy Impact: Natural Gas,[b] 10^6 Btu/yr	Electricity,[c] kWh/yr	Steam,[d] 1,000 lb/yr	Process Total,[e] bbl oil/yr	Fugitive Emission Control Energy Impact: Mg VOC reduced	bbl oil/yr[f]	Combined Emission Control Energy Impact: bbl/oil/yr
PP/Liquid	1	2,145	-	1,789	813	0	0	813
	2	2,145	-	1,789	813	22	-149	664
	3	2,760	2,450	1,789	922	22	-149	773
	4	3,475	2,450	1,794	1,046	22	-149	897
PP/Gas	1	2,145	-	2,105	891	0	0	891
	2	2,145	-	2,105	891	22	-149	742
LDPE/High Pressure	1	715	-	565	263	0	0	263
	2	715	-	565	263	17	-115	148
	3	1,675	27,140	565	470	17	-115	355
	4	1,915	33,880	565	521	17	-115	406
	5	4,775	33,880	4,405	1,965	17	-115	1,850
LDPE/Low Pressure HDPE/Gas	1	4,300	-	2,200	1,285	0	0	1,285
	2	4,300	-	2,200	1,285	33	-230	1,055
	3	5,015	-	2,205	1,410	33	-230	1,180
	4	6,820	18,160	2,205	1,750	33	-230	1,520
	5	8,250	18,160	3,040	2,200			1,970
HDPE/Slurry	1	1,430	-	1,195	542	0	0	542
	2	1,430	-	1,195	542	22	-155	387
	3	1,515	3,875	1,195	563	22	-155	408
HDPE/Solution	1	2,145	-	810	570	0	0	570
	2	2,145	-	810	570	22	-155	415
	3	30,350	27,780	810	5,460	22	-155	5,305
PS/Continuous[g]	1	-	-	-	0	0	0	0
	2	-	-	-	0	33	-200	-200
	3	-	3,878	-	6	33	-200	-221[h]
	4	-	4,224	-	7	33	-200	-221[h]
	5	-	5,837	-	9	33	-200	-220[h]
	6	-	6,061	-	9	33	-200	-220[h]
	7	-	7,571	-	12	33	-200	-218[h]
EPS/Post-impregnation	1	-	-	-	-	0	0	0
	2	-	-	-	-	13	-84	-84
	3	2,170	32,450	0	423	13	-84	339
	4	31,130	32,450	445	5,510	13	-84	5,420
	5	42,650	143,880	445	7,660	13	-84	7,570
	6	157,230	143,880	515	27,400	13	-84	27,300
	7	157,230	143,880	515	27,400	13	-84	27,300

Table 7-6b. ENERGY IMPACTS OF THE REGUALTORY ALTERNATIVES FOR POLYMER AND RESIN PLANTS[a] (concluded)

Model Plant	Regulatory Alternatives	Process Emissions Control Energy Impact: Natural Gas,[b] 10^6 Btu/yr	Electricity,[c] kWh/yr	Steam,[d] 1,000 lb/yr	Process Total,[e] bbl oil/yr	Fugitive Emission Control Energy Impact: Mg VOC reduced	bbl oil/yr[f]	Combined Emission Control Energy Impact: bbl/oil/yr
EPS/In-situ[g]	1	-	-	-	-	0	0	0
	2	-	-	-	-	9	-47	-47
	3	720	0	214	177	9	-47	130
	4	1,990	17,350	214	422	9	-47	375
	5	3,000	27,550	214	611	9	-47	564
	6	3,990	27,550	421	832	9	-47	785
	7	4,960	27,550	628	1,050	9	-47	1,000
PET/DMT[i]	1	0	253,060	21,215	5,657	0	0	5,657
	2	0	254,774	21,215	5,660	0	0	5,653[j]
	3	0	254,836	21,215	5,660	0	0	5,653[j]
	4	0	255,101	21,215	5,661	0	0	5,653[j]
	5	0	255,448	21,215	5,661	0	0	5,653[j]
	6	0	255,856	21,215	5,662	0	0	5,654[j]
	7	0	255,856	28,705	7,521	0	0	7,505[j]
PET/TPA[i]	1	0	253,060	21,215	5,657	0	0	5,657
	2	0	253,060	27,885	7,313	0	0	7,306[j]
PET/TPA[k]	1	0	220,815	864	555	0	0	555
	2	0	220,815	1,160	629	0	0	608
	3	0	220,815	1,573	731	0	0	688
	4	0	220,815	2,141	872	0	0	806
	5	0	220,815	2,516	965	0	0	888
	6	0	220,815	2,969	1,080	0	0	990
	7	0	220,815	4,278	1,400	0	0	1,293

[a] Impacts are on a per process line basis.

[b] 10^6 Btu = 1.0551 GJ. This factor includes the energy conversion efficiency; it is not an energy equivalent. (Reference 12).

[c] 1 kWh = 9.476×10^{-3} GJ. This factor includes the energy conversion efficiency; it is not an energy equivalent. (Reference 12.)

[d] 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.]

[e] 1 GJ = 0.1628 bbl crude oil. (Reference 13.)

[f] 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; and 19,499 Btu/lb of pentane. (Reference 14.)

[g] Estimates of energy use under baseline control are not available.

[h] Includes energy credit for recovered styrene monomer.

[i] These regulatory alternatives apply to a process line producing PET with a single end finisher.

[j] Includes energy credit for recovered ethylene glycol.

[k] These regulatory alternatives apply to a process line producing PET with multiple end finishers.

Note: Values on some lines may not total exactly due to rounding. Negative values indicate energy credits.

7.4.2 Industrywide Energy Impacts

Process line energy impacts presented in Table 7-6 are used to project industrywide energy impacts of the regulatory alternatives on new polymers and resins process lines expected to be built by the year 1988. The industrywide energy impacts of process and fugitive VOC emissions control in new process lines operating under each regulatory alternative are presented in Table 7-7.

Under Regulatory Alternative I, industrywide energy demand is estimated to be about 1,010 TJ/yr (164,000 bbl crude oil/yr). Under Regulatory Alternative II, industrywide energy demand is estimated to be about 1,140 TJ/yr (186,000 bbl oil/yr). The total energy demand increases when shifting from baseline control (Regulatory Alternative I) to Regulatory Alternative II due primarily to the process controls associated with poly-(ethylene terephthalate) plants using a single end finisher in each process line. Implementation of fugitive VOC control practices and increased recovery from process emissions in poly(ethylene terephthalate) process lines saves about 67 TJ/yr (11,000 bbl oil/yr). Energy impacts under Regulatory Alternative VII, the most stringent combination of regulatory alternatives, are approximately 2,020 TJ/yr (329,000 bbl oil/yr); the increase is due to a greater energy demand in controlling process emissions.

In most instances, the incremental energy requirements of the most stringent regulatory alternatives are increases over the energy required under baseline (Regulatory Alternative 1) control. Excluding the polystyrene process lines because no estimate of energy usage under baseline control is available and the poly(ethylene terephthalate) process lines because no estimate of energy usage for making the polymer is available, the next largest incremental percent and absolute size increase associated with the most stringent regulatory alternative over baseline control is for a process line producing high density polyethylene using a solution process. The absolute size of this energy requirement compared to the total energy used in making the polymer is small. A HDPE, solution process model plant comprised of three process lines would use approximately 23.8×10^3 TJ/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 100 TJ/yr (16.4×10^6 bbl oil/yr), less than 0.5 percent of the energy required for the production of the polymer.

Table 7-7a. INDUSTRYWIDE ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR NEW POLYMER AND RESIN PLANTS (TJ/yr)

Model Plant	Regulatory Alternatives	Process Line Energy Emissions[a]			Number of New Process Lines[b]	Industrywide Energy Emissions		
		Process	Fugitive[c]	Combined[c]		Process	Fugitive[c]	Combined
PP/Liquid	1	4.99	0	4.99		44.9	0	44.9
	2	4.99	-0.92	4.08	9	44.9	-8.2	36.7
	3	5.66	-0.92	4.75		51.0	-8.2	42.7
	4	6.43	-0.92	5.51		57.8	-8.2	49.6
PP/Gas	1	5.47	0	5.47		49.3	0	49.3
	2	5.47	-0.92	4.56	9	49.3	-8.2	41.0
LDPE/High Pressure	1	1.62	0	1.62		6.5	0	6.5
	2	1.62	-0.71	0.91	4	6.5	-2.8	3.6
	3	2.89	-0.71	2.18		11.5	-2.8	8.7
	4	3.20	-0.71	2.50		12.8	-2.8	10.0
	5	12.1	-0.71	11.4		48.3	-2.8	45.5
LDPE/Low Pressure HDPE/Gas Phase	1	7.89	0	7.89		158	0	158
	2	7.89	-1.41	6.48		158	-28.3	130
	3	8.65	-1.41	7.24	20	173	-28.3	145
	4	10.7	-1.41	9.32		215	-28.3	186
	5	13.5	-1.41	12.1		270	-28.3	242
HDPE/Slurry	1	3.33	0	3.33		20.0	0	20.0
	2	3.33	-0.95	2.38	6	20.0	-5.7	14.3
	3	3.46	-0.95	2.51		20.7	-5.7	15.0
HDPE/Solution	1	3.50	0	3.50		31.5	0	31.5
	2	3.50	-0.95	2.55	9	31.5	-8.6	22.9
	3	33.5	-0.95	32.6		302	-8.6	293
PS/Continuous	1	-	0	0		0	0	0
	2	-	-1.23	-1.23	2	0	-2.46	-2.46
	3	0.037	-1.23	-1.36[d]		0.073	-2.46	-2.715[d]
	4	0.04	-1.23	-1.36[d]		0.08	-2.46	-2.721[d]
	5	0.055	-1.23	-1.35[d]		0.111	-2.46	-2.703[d]
	6	0.057	-1.23	-1.35[d]		0.115	-2.46	-2.698[d]
	7	0.072	-1.23	-1.34[d]		0.143	-2.46	-2.682[d]
EPS/Post-impregnation	1	-	0	0		-	0	0
	2	-	-0.52	-0.52		-	-1.0	-1.0
	3	2.60	-0.52	2.08		5.2	-1.0	4.2
	4	33.8	-0.52	33.3	2	67.7	-1.0	66.6
	5	47.0	-0.52	46.5		94.1	-1.0	93.1
	6	168	-0.52	167.5		336	-1.0	335
	7	168	-0.52	167.5		336	-1.0	335

Table 7-7a. INDUSTRYWIDE ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR NEW POLYMER AND RESIN PLANTS (TJ/yr) (concluded)

Model Plant	Regulatory Alternatives	Process Line Energy Emissions[a]			Number of New Process Lines[b]	Industrywide Energy Emissions		
		Process	Fugitive[c]	Combined[c]		Process	Fugitive[c]	Combined
EPS/In-situ	1	-	0	0		-	0	0
	2	-	-0.29	-0.29		-	-0.87	-0.87
	3	1.09	-0.29	0.80		3.34	-0.87	2.37
	4	2.59	-0.29	2.30	3	7.78	-0.87	6.9
	5	3.75	-0.29	3.46		11.3	-0.87	10.4
	6	5.11	-0.29	4.82		15.3	-0.87	14.5
	7	6.45		6.16		19.4	-0.87	18.5
PET/DMT[e]	1	34.8	0	34.8		243	0	243
	2	34.8	0	34.7[f]	7	243	0	243[f]
	3	34.8	0	34.7[f]		243	0	243[f]
	4	34.8	0	34.7[f]		243	0	243[f]
	5	34.8	0	34.7[f]		243	0	243[f]
	6	34.8	0	34.7[f]		243	0	243[f]
	7	46.2	0	46.2[f]		323	0	323[f]
PET/TPA[e]	1	34.8	0	34.8	13	452	0	452
	2	44.9	0	44.9[f]		584	0	583[f]
PET/TPA[g]	1	3.41	0	3.41	1	3.4	0	3.4
	2	3.86	0	3.73[f]		3.9	0	3.7
	3	4.49	0	4.23[f]		4.5	0	4.2
	4	5.36	0	4.95[f]		5.4	0	5.0
	5	5.93	0	5.46[f]		5.9	0	5.5
	6	6.62	0	6.08[f]		6.6	0	6.1
	7	8.61	0	7.94[f]		8.6	0	7.9
TOTAL[h]	I	-	-	-		1,008	0	1,008
	II	-	-	-		1,141	-66	1,074
	III	-	-	-		1,448	-66	1,380
	IV	-	-	-	85	1,565	-66	1,497
	V	-	-	-		1,687	-66	1,619
	VI	-	-	-		1,934	-66	1,866
	VII	-	-	-		2,020	-66	1,951

[a]On a per process line basis. From Table 7-6.

[b]From Table 8-43.

[c]Negative values indicate energy credits.

[d]Includes energy credit for recovered styrene monomer.

[e]These regulatory alternatives apply to a process line producing PET with a single end finisher.

[f]Includes energy credit for recovered ethylene glycol.

[g]These regulatory alternatives apply to a process line producing PET with multiple end finishers.

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

NOTES: Dashes indicate that the entry is not applicable.

Table 7-7b. INDUSTRYWIDE ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR NEW POLYMER AND RESIN PLANTS (1,000 bbl oil/yr)

Model Plant	Regulatory Alternatives	Process Line Energy Emissions[a]			Number of New Process Lines[b]	Industrywide Energy Emissions		
		Process	Fugitive[c]	Combined[c]		Process	Fugitive[c]	Combined[c]
PP/Liquid	1	0.813	0	0.812		7.31	0	7.31
	2	0.813	-0.149	0.663	9	7.31	-1.34	5.97
	3	0.922	-0.149	0.773		8.30	-1.34	6.96
	4	1.05	-0.149	0.901		9.41	-1.34	8.07
PP/Gas	1	0.89	0	0.89	9	8.02	0	8.02
	2	0.89	-0.149	0.741		8.02	-1.34	6.68
LDPE/High Pressure	1	0.263	0	0.263		1.05	0	1.05
	2	0.263	-0.115	0.148		1.05	-0.46	0.59
	3	0.470	-0.115	0.355	4	1.88	-0.46	1.42
	4	0.521	-0.115	0.406		2.09	-0.46	1.63
	5	1.97	-0.115	1.86		7.86	-0.46	7.40
LDPE/Low Pressure HDPE/Gas Phase	1	1.29	0	1.29		25.7	0	25.7
	2	1.29	-0.23	1.06		25.7	-4.6	21.1
	3	1.41	-0.23	1.18	20	28.2	-4.6	23.6
	4	1.75	-0.23	1.52		34.9	-4.6	30.3
	5	2.20	-0.23	1.97		44.0	-4.6	39.4
HDPE/Slurry	1	0.542	0	0.542		3.25	0	3.25
	2	0.542	-0.155	0.387	6	3.25	-0.93	2.32
	3	0.563	-0.155	0.408		3.38	-0.93	2.45
HDPE/Solution	1	0.57	0	0.57		5.13	0	5.13
	2	0.57	-0.155	0.415	9	5.13	-1.40	3.73
	3	5.46	-0.155	5.31		49.1	-1.40	47.7
PS/Continuous	1	-	0	0		0	0	0
	2	-	-0.20	-0.20	2	0	-0.40	-0.40
	3	0.006	-0.20	-0.221		0.012	-0.40	-0.442[d]
	4	0.007	-0.20	-0.221[d]		0.013	-0.40	-0.443[d]
	5	0.009	-0.20	-0.22[d]		0.018	-0.40	-0.440d
	6	0.009	-0.20	-0.22[d]		0.019	-0.40	-0.439d
	7	0.012	-0.20	-0.218[d]		0.023	-0.40	-0.437d
EPS/Post-impregnation	1	-	0	0		-	0	0
	2	-	-0.084	-0.084		-	-0.17	-0.17
	3	0.423	-0.084	0.339		0.85	-0.17	0.68
	4	5.51	-0.084	5.43	2	11.0	-0.17	10.9
	5	7.66	-0.084	7.58		15.3	-0.17	15.2
	6	27.4	-0.084	27.3		54.7	-0.17	54.6
	7	27.4	-0.084	27.3		54.7	-0.17	54.6

Table 7-7b. INDUSTRYWIDE ENERGY IMPACTS OF THE REGULATORY ALTERNATIVES FOR NEW POLYMER AND RESIN PLANTS (1,000 bbl oil/yr) (concluded)

Model Plant	Regulatory Alternatives	Process Line Energy Emissions[a]			Number of New Process Lines[b]	Industrywide Energy Emissions		
		Process	Fugitive[c]	Combined[c]		Process	Fugitive[c]	Combined[c]
EPS/In-situ	1	-	0	0		-	0	0
	2	-	-0.047	-0.047		-	-0.14	-0.14
	3	0.177	-0.047	0.129		0.53	-0.14	0.39
	4	0.422	-0.047	0.374	3	1.27	-0.14	1.12
	5	0.611	-0.047	0.564		1.83	-0.14	1.69
	6	0.832	-0.047	0.785		2.50	-0.14	2.36
	7	1.05	-0.047	1.0		3.15	-0.14	3.01
PET/OMT[e]	1	5.66	0	5.66		39.6	0	39.6
	2	5.66	0	5.65[f]	7	39.6	0	39.6[f]
	3	5.66	0	5.65[f]		39.6	0	39.6[f]
	4	5.66	0	5.65[f]		39.6	0	39.6[f]
	5	5.66	0	5.65[f]		39.6	0	39.6[f]
	6	5.66	0	5.65[f]		39.6	0	39.6[f]
	7	7.52	0	7.51[f]		52.6	0	52.6[f]
PET/TPA[e]	1	5.66	0	5.66	13	73.6	0	73.6
	2	7.31	0	7.31[f]		95.0	0	95.0[f]
PET/TPA[g]	1	0.56	0	0.56		0.56	0	0.56
	2	0.63	0	0.61[f]		0.63	0	0.61[f]
	3	0.73	0	0.69[f]		0.73	0	0.69[f]
	4	0.87	0	0.81[f]	1	0.87	0	0.81[f]
	5	0.97	0	0.89[f]		0.97	0	0.89[f]
	6	1.08	0	0.99[f]		1.08	0	0.99[f]
	7	1.40	0	1.29[f]		1.40	0	1.29[f]
TOTAL[h]	I	-	-	-	85	164	0	164
	II	-	-	-		186	-10.8	175
	III	-	-	-		236	-10.8	225
	IV	-	-	-		255	-10.8	244
	V	-	-	-		275	-10.8	264
	VI	-	-	-		315	-10.8	304
	VII	-	-	-		329	-10.8	318

[a]On a per process line basis. From Table 7-6.

[b]From Table 8-43.

[c]Negative values indicate energy credits.

[d]Includes energy credit for recovered styrene monomer.

[e]These regulatory alternatives apply to a process line producing PET with a single end finisher.

[f]Includes energy credit for recovered ethylene glycol.

[g]These regulatory alternatives apply to a process line producing PET with multiple end finishers.

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

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 for most plants. 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 through VII 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 new process lines to meet Regulatory Alternatives III through VII 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 requirements of Regulatory Alternatives III through VII.

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 process lines 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 1,495 Mg/yr (1,648 tons/yr) in the absence of Regulatory Alternative II, and as much as 4,560 mg/yr (5,024 tons/yr) in the absence of Regulatory Alternative VII. 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 about 4 to 14 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 energy savings of 66 TJ/yr (11,000 bbl crude oil/yr). Delay in implementing Regulatory Alternatives III through VII may result in energy savings from 370 up to 940 TJ/yr (61,000 to 154,000 bbl oil/yr) when compared to baseline control.

7.6 REFERENCES FOR CHAPTER 7

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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 the next less stringent regulatory alternative, for each 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, the cost analysis is based on SOCMI Model Plant B.[1,2]

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 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-

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	MR	PF	PS	Other		Mg/yr	Percent
Polypropylene, liquid phase									
1 (Baseline)		C[4]	C[5]				No	1,705	89
2		C[4]	C[5]				Yes	1,725	90
3		C[4]	C[5]	C[5]			Yes	1,852	97.0
4	C[5]	C[4]	C[5]	C[5]			Yes	1,856	97.3
Polypropylene, gas phase									
1 (Baseline)		F[6]	F[5]				No	1,270	95
2		F[6]	F[5]				Yes	1,290	97
Low density, polyethylene high pressure									
1 (Baseline)						C[6]	No	594	81
2						C[6]	Yes	609	83
3					C[5]	C[6]	Yes	664	91
4				C[5]	C[5]	C[6]	Yes	678	93
5		C[6,7]		C[5]	C[5]	C[6]	Yes	704	96
Low density polyethylene, low pressure and high density polyethylene, gas phase									
1 (Baseline)	C[6,7]	C[6]		C[5]			No	1,725	94
2	C[6]	C[6]		C[5]			Yes	1,755	96.1
3	C[4]	C[6]		C[5]			Yes	1,760	96.4
4	C[4]	C[6]		C[5]	C[5]		Yes	1,765	96.7
5	C[4]	C[6,7]		C[5]	C[5]		Yes	1,770	96.9
High density polyethylene, slurry process									
1 (Baseline)	C[6]		C[5]				No	900	92
2	C[6]		C[5]				Yes	920	94
3	C[6]		C[5]	C[5]			Yes	950	97
High density polyethylene, solution process									
1 (Baseline)	C[4]	C[6]	C[4]				No	825	90
2	C[4]	C[6]	C[4]				Yes	845	92
3	C[4]	C[6]	C[4]	C[5]			Yes	890	97

Table 8-1. SUMMARY OF REGULATORY ALTERNATIVES FOR THE MODEL PLANTS (concluded)

Model Plant/ Regulatory Alternative	Process Sections Controlled[1,2]						Fugitive Emission Regulatory Control	Annual Emission Reduction[3]	
	RMP	PR	MR	PF	PS	Other		Mg/yr	Percent
Polystyrene, continuous process									
1 (Baseline)[8]		R					No	23	27
2		R					Yes	53	62
3		R	RR[9]				Yes	57.05	67.1
4		R	RR[10]				Yes	57.23	67.3
5		R	RR[11]				Yes	57.37	67.4
6		R	RR[12]				Yes	57.41	67.5
7		R	RR[13]				Yes	57.43	67.6
Expandable Polystyrene, post-impregnation suspension process									
1 (Baseline)							No	15	5
2							Yes	35	12
3					C[5]		Yes	91	32
4				C[6]	C[5]		Yes	184	64
5				C[4]	C[5]		Yes	237	83
6				C[4]	C[4]		Yes	266	92.7
7		R[5]		C[4]	C[4]		Yes	267	93.0
Expandable Polystyrene, in-situ suspension process									
1 (Baseline)							No	5	9
2							Yes	12	21
3				C[6]			Yes	21	36
4		C[6]		C[6]			Yes	43	76
5		C[6]		C[6]	C[6]		Yes	49	86
6	C[6]	C[6]		C[6]	C[6]		Yes	50.2	88
7	C[6]	C[6]		C[6]	C[4]		Yes	50.6	89
Poly(ethylene) terephthalate), dimethyl terephthalate									
1(Baseline)[14]	R	R					N.A.[15]	0	0
2	R	R	R[16]				N.A.	2.43	30
3	R	R	R[17]				N.A.	2.48	31
4	R	R	R[18]				N.A.	2.59	32
5	R	R	R[19]				N.A.	2.66	33
6	R	R	R[20]				N.A.	2.67	33.3
7	R	RR[21]	R[20]				N.A.	5.33	66
Poly(ethylene) terephthalate),[22] terephthalic acid									
1 (Baseline)[23]	R	R					N.A.	0	0
2	R	RR[24]					N.A.	2.4	44
Poly(ethylene) terephthalate),[25] terephthalic acid									
1 (Baseline)[26]	R	R					N.A.	0	0
2	R	RR[27]					N.A.	7.2	10
3	R	RR[27]					N.A.	14.7	21
4	R	RR[27]					N.A.	22.3	31
5	R	RR[27]					N.A.	26.1	37
6	R	RR[27]					N.A.	29.8	42
7	R	RR[27]					N.A.	37.3	53

Footnotes to Table 8-1

[1] Process Sections:

RMP - raw materials preparation (including esterification in poly(ethylene terephthalate) plants)
PR - polymerization reaction
Other - all emergency vents other than from the reactor
MR - material recovery
PF - product finishing
PS - product storage

[2] Control Devices:

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 on a per line basis.

[4] Control of intermittent and continuous streams.

[5] Control of continuous streams.

[6] Control of non-emergency intermittent streams.

[7] Control of emergency intermittent streams.

[8] Emissions from the polymerization reaction section are indirectly controlled as part of normal process operation by condensers. The control of the emissions from this material recovery is indicated in Regulatory Alternatives 3 and 4.

[9] The additional control is the application of a refrigerated condenser controlling emissions of 0.12 kg VOC per Mg of product from the material recovery section down to 0.012 kg VOC per Mg of product.

[10] The additional control is the application of a refrigerated condenser controlling emissions to 0.0072 kg VOC per Mg of product.

[11] Control by refrigerated condensers to 0.0036 kg VOC per Mg product.

[12] Control by refrigerated condensers to 0.0024 kg VOC per Mg product.

[13] Control by refrigerated condensers to 0.0018 kg VOC per Mg product.

[14] As part of normal process operations, the methanol emissions are recovered by condensers. The emissions from this methanol recovery are controlled as indicated in Regulatory Alternatives 2 through 6. In addition, as part of normal process operation, ethylene glycol emissions from the polymerizers are recovered by spray condensers prior to the vacuum system evacuating the polymerizers. Additional control was evaluated for this section in Regulatory Alternative 7.

[15] N.A. = Not Applicable.

[16] Control by refrigerated condensers to 0.018 kg VOC per Mg product.

[17] Control by refrigerated condensers to 0.0144 kg VOC per Mg product.

[18] Control by refrigerated condensers to 0.0072 kg VOC per Mg product.

[19] Control by refrigerated condensers to 0.0027 kg VOC per Mg product.

[20] Control by refrigerated condensers to 0.0018 kg VOC per Mg product.

[21] Control by distillation column on cooling water tower resulting in a 50 percent reduction in emissions from polymerization reaction section via the cooling water tower.

[22] These regulatory alternatives are for a low viscosity PET line or a high viscosity PET line with a single end finisher.

[23] As part of normal process operations, ethylene glycol is recovered from the process vessels using a spent ethylene glycol spray condensers. Additional control was evaluated under Regulatory Alternative 2.

[24] Additional control reflects the use of a distillation column on the cooling water tower that results in a 50 percent reduction of ethylene glycol emissions to the atmosphere from the polymerization reaction section via the cooling water tower.

[25] These regulatory alternatives are for a high viscosity PET line with multiple end finishers.

[26] As part of normal process operations, ethylene glycol is recovered from the process sections. Baseline control includes distillation columns on the esterifiers and cooling water tower, and spent ethylene glycol spray condensers on the first end finisher.

[27] Additional control reflects increased flow of cooling water to the distillation column on the cooling water tower that results in emission reductions ranging from 15 percent, under Regulatory Alternative 2, up to 90 percent, under Regulatory Alternative 7.

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.

The fugitive control capital cost includes costs for monitoring instruments, caps for open-ended lines, piping to vent compressor seals to an existing enclosed combustion device or vapor recovery header, rupture disk assemblies, closed-purge sampling connections, and initial leak repair. 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.

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

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	0.0	0.2	0.0	0.0	0.0	0.0
Total Equipment (A)	1.0	1.2	1.0	1.0	1.0	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	0.15	-	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[b]
= 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]

R-12 Refrigerant: $2.74/kg ($1.24/lb)[o]

R-502 Refrigerant: $5.13/kg ($2.33/lb)[o]

Indirect Cost Factors

Interest rates:

10 percent (in the absence of taxes)

Equipment life, N:[p]

15 years for flare
10 years for thermal incinerator, catalytic incinerator, condenser, ethylene glycol recovery system, piping, initial labor for fugitive emissions control, other fugitive emission equipment except rupture disks and monitoring instruments
6 years for monitoring instruments
2 years for rupture disks

TABLE 8-3. ANNUALIZED COST FACTORS FOR POLYMERS AND RESINS NSPS
(June 1980 Dollars) (Continued)

Indirect Cost Factors (Con't)

Capital recovery charge factor = $\frac{i(1 + i)^N}{(1 + i)^N - 1}$

0.131 for flare
0.163 for thermal incinerator, catalytic incinerator, condenser, ethylene glycol recovery system, piping

Taxes, insurance, and administration: 0.04 x total installed capital cost[q]

Maintenance cost: 0.5 x Total installed capital cost[r]
$3,000 for monitoring instrument, calibration, and maintenance

Operating hours: 8600 hours/yr

[a] Includes wages plus 40 percent for labor-related adminstrative 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. LEnvironmental 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 GTGD 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.

Footnotes to Table 8-3 Concluded

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. pp. 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] Telecon: K. Meardon, PES to ARC Supply Co. May 1984 prices of $1.31/lb for Freon-12 and $3.11/lb for Freon-502 deflated to June 1980 dolalrs using Chemical Engineering Producer Price Industrial Chemicals Index (June 1980 = 327.3; May 1984 = 344.8).

[p] Average equipment lives given by Neverill in reference cited in j., pp. 3-16.

[q] 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. pp. 5-16.

[r] Per reference cited in footnote q:

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).

to functions of the following waste gas characteristics: volumetric flow rate, lower heating value, temperature, and molecular weight for a state-of-the-art exit velocity of one-half sonic velocity (0.5 Mach). 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. For flares handling very small flows in a minimum-available-size flare, the cost of supplying sufficient purge gas can be greater than the cost of a fluidic seal on the flare tip to prevent air intrusion and subsequent flashback. 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. For flares handling large intermittent flows, steam requirements were calculated for a small continuous "bleed" rate that is necessary to prevent shortened steam nozzle equipment life (due to steam condensate contacting the high temperature metal).[11,12]

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,[13] 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[9]). For flows less than 700 scfm, an economic pipe diameter was calculated based on an equation in the Chemical Engineer's Handbook[14] and simplified as suggested by Chontos.[15,16,17] 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[18] as combined for 70 ft. source legs and 500 ft. and 2,000 ft. pipelines for the cost analysis of the Distillation NSPS.[19] Ducting costs were calculated based on the installed cost equations given in the GARD Manual.[20] 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 Thermal Incineration Design. 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).[21] The design procedure is outlined in this section.

In order to prevent an explosion hazard and satisfy insurance requirements, dilution air would be 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)[22,23] and an oxygen concentration of 12 percent or greater by volume[24] that required preheating to maintain a combustion temperature of 870°C (1600°F). Dilution air would be 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 only stream (Stream G in the polypropylene, liquid phase plant) for which a thermal incinerator was costed in this heat content range, however, did not require preheating, because the combined streams for the section had a heat content greater than 50 Btu/scf.

The combustion products were then calculated assuming 18 percent excess air for required additional combustion air,[25,26] 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.[27,28] Minimum auxiliary fuel requirements for low heating value streams were set at 5 Btu/scf to ensure flame stability.[29]

For streams able to maintain combustion at 870°C (1600°F) such as the combined streams costed for the Polymers and Resins industry, 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.[30]

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.[31] The design procedure assumed a minimum commercially available size of 1.01 m^3 (35.7 ft^3) based on vendor information[32] and a maximum shop-assembled unit size of 205 m^3 (7,238 ft^3).[33]

The design procedure could 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 greater than 870°C (1600 °F), and even at greater than 980°C (1800 °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 Thermal Incinerator Costing. Thermal incinerator purchase costs were taken directly from the IT Enviroscience graph for the calculated combustion chamber volume.[30] (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).[34] The installed cost used for the minimum commercially available size of 1.01m^3 (35.7 ft^3) was $217,800 (June 1980) ($52,000 December 1979 purchase cost).

The installed cost of one 150-ft. duct to the incinerator and its associated fan and an 80-ft. stack were also taken directly from the IT Enviroscience study.[35] A minimum installed 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 noted 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 1.5 kPa (6-inch H_2O) pressure drop across the system and a blower efficiency of 60 percent.

8.1.3 Catalytic Incinerator Design and Cost Basis

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 Catalytic Incinerator Design. 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 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. Heat exchanger requirements are determined by gas inlet temperature and preheater temperature. A minimum practical heat exchanger efficiency is about 30 percent; a maximum of 65 percent was assumed for this analysis. Gas temperature, preheater temperature, gas dew point temperature, and gas VOC content determine the maximum feasible 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.[36] Catalyst replacement every three years was assumed.

8.1.3.2 Catalytic Incinerator Costing. 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.[36,37,38] 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. A minimum available unit size of 500 scfm was assumed[39,40]; the installed cost of this minimum size unit, which can be used without addition of gas or air for stream flows greater than about 150 scfm[40], was estimated to be $53,000 (June 1980). 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 Condenser Design and Cost Basis

This section outlines the general procedures used for sizing and estimating the costs of surface condenser systems applied to the gaseous streams from the continuous process polystyrene model plant and the poly(ethylene terephthalate), dimethyl terephthate process model plant. For illustrative purposes, the following paragraphs describe the condenser design and costs for the polystyrene model plant. Details of the condenser

design and costs for both model plants are found in Docket Reference Number II-B-93.

8.1.4.1 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 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.

Although some older polystyrene processes emit styrene in steam, which is more readily condensed and thus less costly to control, saturated styrene in air at 27°C (80°F) was assumed for costing design purposes since new plants under baseline conditions were assumed to use vacuum pumps. 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 in air. As the vapor pressure data are not readily available, the condensation temperature was estimated by a regression equation of available data points[41] 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 content of the condensed styrene and the sensible heat changes of styrene and air. The coolant is selected based on the condensation temperature.

The styrene-in-air refrigerated condenser systems were designed according to procedures for calculating shell-side[42], tube-side[43], and condensation[44] heat transfer coefficients, mass transfer coefficients,[45] and, finally, an overall heat transfer coefficient for condensation in the presence of a non-condensible using a multiple section analysis.[46] Heat exchanger[47,48] and refrigeration unit[49] characteristics were developed from vendor information in conjunction with information given primarily in the Chemical Engineers' Handbook.[42,50] Refrigerant characteristics were taken primarily from the Chemical Engineers' Handbook[51,52] and publications of the American Society of Heating, Refrigerating, and Air-conditioning Engineers (ASHRAE).[53] The total required heat transfer area and refrigeration capacity then were calculated from the total heat load, temperature difference, and overall heat transfer coefficient and commercially available sizes were selected.

8.1.4.2 Surface Condenser Costing. Since the gas volumes of the two streams are low, the calculated required (selected) heat transfer areas are also low - about 1.7 (2.3) and 4.5 (7.2), respectively, for 90 and 98 percent reduction of styrene emissions from a single process line). The purchase costs of the heat exchanger[47,48] and refrigeration systems were estimated from data provided by vendors.[49] An installation factor of 1.39 (see Table 8-2) was used to estimate installed condenser costs. The corresponding required refrigeration capacities for 90 and 98 percent styrene reduction were only 0.056 and 0.080 tons (0.20 and 0.28 kW or 670 and 960 Btu/hr), respectively. For 90 percent reduction, the required capacity is much smaller than the 0.117 tons (0.412 kW or 1405 Btu/hr) available capacity for the minimum available size refrigeration unit of 1/4 compressor horsepower (0.186 kW compressor) and a coolant temperature of 0°F (-17.8°C). For 98 percent reduction, however, 1/2 compressor horsepower (0.373 kW compressor) was required to provide sufficient available refrigeration capacity for a coolant temperature of -30°F (-34.4°C). Installed costs were put on a June 1980 basis using Chemical Engineering Cost Indexes. No additional piping was costed since the condenser unit is so small ($\leq$ 5 in. diameter) that it should be able to be installed adjacent to the source. Details of the design and cost estimates are given in Appendix E (Section E.5).

8.1.5 Ethylene Glycol Recovery System Design and Cost Basis

This section briefly describes the procedures used to estimate the cost of the baseline systems and of using a distillation column to recover ethylene glycol from a cooling water tower.

8.1.5.1 Ethylene Glycol Recovery System Design. The baseline system for a PET plant recovers ethylene glycol through spent ethylene glycol spray condensers, reflux condensers, and distillation columns. The equipment selected to comprise baseline recovery systems was obtained from information provided by industry sources, which also provided basic operating and maintenance parameters. 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 TPA process can be as much as 35 percent less than that from a DMT 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.)

Industry sources also provided the basic equipment requirements and design considerations for installing a distillation column on a cooling water tower to recover ethylene glycol from the cooling water. The information provided was for a PET/TPA plant of larger capacity than the model plant.

8.1.5.2 EG Recovery System Costing. The costs of the baseline system were estimated based on the preliminary design using industry information in conjunction with standard engineering procedures. 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. The costs of the distillation column were derived from information 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 the SOCMI fugitive VOC emission regulations. The following sections outline the specific requirements of the regulatory alternative and the procedures used to estimate its capital and annual costs.

8.1.6.1 Design of Fugitive VOC Regulatory Alternative. The following equipment in the polymers and resins industry were considered for regulation: 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 were described in Chapter 6.

The technical parameters for the polymers and resins model plant for fugitive emissions are given in Table 8-5.

Table 8-4. FUGITIVE VOC REGULATORY ALTERNATIVE CONTROL SPECIFICATIONS[a]

Emission source	Control Specification: Inspection/monitoring interval[b]	Control Specification: Equipment
Valves		
Gas	Monthly	None
Light liquid	Monthly	None
Heavy liquid	None	None
Pumps seals		
Liqht 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]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 MODEL PLANTS [a,b]

Emission source	Number of sources[c]	Uncontrolled emission factor, kg/hr/ source	Annual uncontrolled emissions,[d] Mg/yr	Control method[e]	Control efficiency	Annual controlled emissions,[d] Mg/yr	Annual emission reductions due to control Mg/yr
Valves							
Gas	402	0.0056	19.72	Monthly leak	0.73	5.32	14.40
Light liquid	524	0.0070	32.59	detection and	0.59	13.36	19.23
Heavy liquid	524	0.00023	1.06	repair[f]	--	1.06	0
Pump seals							
Light liquid	29	0.0494	12.55	Monthly leak	0.61		
Heavy liquid	30	0.0214	5.62	detection and repair[f]	--	4.92 5.62	7.63 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
Compressor seals	2	0.228	3.99	Vented seal	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. EPA-450/3-82-010. April 1982. p. B-2 and B-3. (Reference 37.)

[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.

[f] Only for equipment in light liquid service; no control method for equipment in heavy liquid service.

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 NSPS analysis.[1,2] 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 describes the cost estimation procedures and assumptions used to derive these costs 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.

(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,[54] 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 AID.[55]

Table 8-6. 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	20,800
b) Maintenance	5,700
c) Miscellaneous[a]	2,600
d) Annualized capital costs	14,100
Total	43,200
3. VOC recovery credits[b]	(31,000)[c]
4. Net annualized cost	12,200

[a] Taxes, insurance, and administration.

[b] Based on 58.69 Mg of VOC recovered annually at a credit of $528 per Mg.

[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	Compressor seals	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[b]	6,600	1,000	2,200	2,100	13,500
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	4,100	1,900					6,000
b) Maintenance			1,700	300	700	3,000	5,700
c) Miscellaneous			1,300	300	600	400	2,600
4. Total annual cost	14,600	8,400	9,600	1,600	3,500	5,500	43,200
5. VOC recovery credit	(17,800)[a]	(4,000)	(5,300)	(2,100)	(1,800)	0	(31,000)
6. Net annualized cost	(3,200)	4,400	4,300	(500)	1,700	5,500	12,200
7. Cost effectiveness per Mg of VOC	(95)	575	430	(125)	520	-	208[c]

[a] Values in parentheses denote credit.

[b] Replacement seals.

[c] Based on an emission reduction of 58.69 Mg/yr.

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 $18/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)

Service	No. of valves per model	Type of monitoring	Monitoring time, man-min[a]	Fraction screened	Monitoring labor required, man-hours	No. of leaks per year[b]	Repair time, man-hours	Leak repair labor-hours required, man-hours
Gas	402	Instrument	2	11.79	158	76.4	1.13	86.3
Light-liquid	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 LDAR model, based on fraction of sources operated on equal to 19 percent. (Reference 2)

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.

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 AID.[55]

Safety/Relief Valves

The fugitive VOC emission control specifications for safety/relief valves include installation of rupture disk systems on the relief valves. 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.[56] 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

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. (Reference 2)

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, $	1,650
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.

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

Although all open-ended lines were assumed controlled under baseline and, thus, no fugitive emissions would occur (see Table 8-5), cost data were generated for open-ended lines and are presented in Table 8-18. 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.

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[56] 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

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
(2) Rupture disk systems with 3-way valve.	Capital costs
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).

Table 8-18. CAPITAL AND NET ANNUALIZED COSTS FOR CONTROL OF EMISSIONS FROM OPEN-ENDED LINES
(May 1980 dollars)

Costs	Caps
Installed capital cost, $	53
Annualized cost at 16.3 percent of capital cost, $	8.6
Annualized operating costs, $	
Maintenance at 5 percent of capital cost, $	2.7
Miscellaneous at 4 percent of capital cost, $	2.1
Total annual costs before credit, $	13.4
Recovery credit[a], $	7,870
Net annualized costs, $	(7,857)[b]
Cost effectiveness, $/Mg	(527)

[a]Recovery credit based on uncontrolled VOC emission factor of 0.0408 kg/day. Based on 100 percent control efficiency for caps and $528/Mg VOC emission reduction.

Emission Reduction:

0.0408 kg/day/open-ended line x 365 day/yr x 1 Mg/1,000 kg = 14.9 Mg/yr.

Recovery Credit:

14.9 Mg/yr x $528/Mg VOC = $7,870/yr/open-ended line.

[b]Values within parentheses denote credit.

The above costs include connection of the degassing reservoir to an existing 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 8-19.

Sampling Systems

Equipment costs were computed for closed loop sampling connections. The cost estimates were based on information from the Hydroscience (now ITE) report[56] and was for the following items per sampling system. Table 8-20 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 8-21 to 8-31b, for the twelve model plants. Tables 8-32 to 8-42b show the costs and associated emission reductions of regulatory alternatives for the eleven model plants for the implementation of an alternative both from the baseline control level and from the next less stringent alternative.

National control costs of the most stringent regulatory alternatives over baseline controls are estimated, as shown in Table 8-43, based on the projected number of new process lines, which are described in Chapter 9 "Economic Impact". The total nationwide fifth-year net annualized cost in these new process lines is estimated to range from $11.2 million for baseline control to about $26.7 million for the maximum achievable control level. The maximum achievable control level from baseline is equivalent to an annual VOC emission reduction of about 4,700 Mg.

Table 8-19. 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
Annualized operating costs, $	
Maintenance at 5 percent of capital cost, $	320
Miscellaneous at 4 percent of capital cost, $	260
Total annual costs before credit, $	1,620
Annual product recovery credit[d]	(2,110)[e]
Net annualized costs, $	(490)[e]
Cost effectiveness, $/Mg	(120)[e]

[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.) (Reference 2)

[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 reduction of 3.99 Mg/yr.

[e]Values within parentheses denote credit.

Table 8-20. CAPITAL AND NET ANNUALIZED COSTS FOR CONTROL OF EMISSIONS FROM SAMPLING SYSTEMS FOR THE MODEL UNIT
(May 1980 dollars)

Costs	Closed purge sampling connections
Installed capital cost,[a] $	13,800
Annualized cost at 16.3 percent of capital cost,[b] $	2,250
Annualized operating costs, $	
Maintenance at 5 percent of capital cost, $	690
Miscellaneous at 4 percent of capital cost, $	550
Total annual costs, $	3,490
Annual product recovery credit[c], $	(1,810)[d]
Net annualized costs, $	1,680
Cost effectiveness, $/Mg	490

[a]Based on $530 per sampling system and 26 sampling systems in the model plant.

[b]Capital recovery factor, 0.163, is based on the equipment amortized over 10 years at 10 percent interest.

[c]Product credit is calculated based on $528/Mg and 3.42 Mg/yr of VOC recovered annually (see Table 8-5).

[d]Values within parentheses denote credit.

Table 8-21. POLYPROPYLENE, LIQUID PHASE MODEL PLANT REGULATORY ALTERNATIVES COSTS (June 1980 dollars)[a,b]

	Regulatory Alternative 1	2	3	4
INSTALLED CAPITAL COST ($)				
Flare(s)	45,600	45,600	45,600	58,500
Flare Ducting	40,400	40,400	40,400	45,800
Thermal Incinerator			217,800	217,800
Thermal Incinerator Ducts, Fans & Stack			86,700	86,700
Catalytic Incinerator				
Catalytic Incinerator Ducts, Fans & Stack				
Condenser				
Fugitive Leak Detection and Repair (LDAR)		28,400	28,400	28,400
Total	85,900	114,300	418,800	437,100
ANNUALIZED COST ($/yr)				
Direct				
Operating Labor	33,500	40,500	62,100	73,200
Operating Materials (e.g., catalyst)				
Maintenance Materials and Labor	4,300	8,200	23,500	24,400
Utilities				
Natural Gas	12,800	12,800	16,400	20,700
Electricity			120	120
Steam	11,100	11,100	11,100	11,120
Subtotal	61,700	72,500	113,200	129,500
Indirect				
Taxes, Insurance, & Administration[c]	3,440	4,570	16,800	17,500
Capital Recovery	12,400	18,500	22,100	70,600
Subtotal	15,800	23,000	84,700	88,100
Recovery Credit		10,300	10,300	10,300
Total (Direct + Indirect - Credit)	77,500	85,200	187,600	207,300

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] Includes miscellaneous operating costs for fugitive leak detection and repair program.

Table 8-22. POLYPROPYLENE, GAS PHASE MODEL PLANT REGULATORY ALTERNATIVES COSTS (June 1980 dollars)[a,b]

	Regulatory Alternative 1	Regulatory Alternative 2
INSTALLED CAPITAL COST ($)		
Flare(s)	104,400	104,400
Flare Ducting	59,300	59,300
Thermal Incinerator		
Thermal Incinerator Ducts, Fans & Stack		
Catalytic Incinerator		
Catalytic Incinerator Ducts, Fans & Stack		
Condenser		
Fugitive Leak Detection and Repair (LDAR)		28,400
Total	163,700	192,100
ANNUALIZED COST ($/yr)		
Direct		
Operating Labor	22,300	29,300
Operating Materials (e.g., catalyst)		
Maintenance Materials and Labor	8,200	12,100
Utilities		
Natural Gas	12,800	12,800
Electricity		
Steam	13,000	13,000
Subtotal	56,400	67,200
Indirect		
Taxes, Insurance, & Administration[c]	6,540	7,670
Capital Recovery	23,400	29,500
Subtotal	29,900	37,200
Recovery Credit		10,300
Total (Direct + Indirect - Credit)	86,300	94,000

[a] Costs are on a per basis line basis.

[b] Some totals do not add up exactly due to rounding.

[c] Includes miscellaneous operating costs for fugitive leak detection and repair program.

Table 8-23. LOW DENSITY POLYETHYLENE, HIGH PRESSURE MODEL PLANT REGULATORY ALTERNATIVES COSTS (June 1980 dollars)[a,b]

	Regulatory Alternative				
	1	2	3	4	5
INSTALLED CAPITAL COST ($)					
Flare(s)	24,500	24,500	24,500	24,500	526,900
Flare Ducting	71,700	71,700	71,700	71,700	477,700
Thermal Incinerator					
Thermal Incinerator Ducts, Fans & Stack					
Catalytic Incinerator			112,000	165,000	165,000
Catalytic Incinerator Ducts, Fans & Stack			101,500	195,300	195,300
Particulate Removal					
Fugitive Leak Detection and Repair (LDAR)		23,600	23,600	23,600	23,600
Total	96,200	119,800	333,300	480,100	1,388,500
ANNUALIZED COST ($/yr)					
Direct					
Operating Labor	11,200	16,400	33,100	49,800	61,000
Operating Materials (e.g., catalyst)			4,290	5,360	5,360
Maintenance Materials and Labor	4,800	8,400	19,100	26,400	71,800
Utilities					
Natural Gas	4,280	4,280	10,000	11,500	28,600
Electricity			1,330	1,660	1,660
Steam	3,500	3,500	3,500	3,500	27,200
Subtotal	23,700	32,600	71,300	98,200	195,600
Indirect					
Taxes, Insurance, & Administration[c]	3,850	4,750	13,300	19,200	55,500
Capital Recovery	14,900	20,000	54,700	78,600	210,700
Subtotal	18,700	24,800	68,100	97,800	266,300
Recovery Credit		7,750	7,750	7,750	7,750
Total (Direct + Indirect - Credit)	42,500	49,700	131,700	188,300	454,200

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] Includes miscellaneous operating costs for fugitive leak detection and repair program.

Table 8-24. LOW DENSITY POLYETHYLENE, LOW PRESSURE AND HIGH DENSITY POLYETHYLENE, GAS PHASE MODEL PLANT REGULATORY ALTERNATIVES COSTS[a,b]
(June 1980 dollars)

	Regulatory Alternative				
	1	2	3	4	5
INSTALLED CAPITAL COST ($)					
Flare(s)	45,000	45,000	57,900	57,900	138,000
Flare Ducting	88,900	88,900	94,300	94,300	169,200
Thermal Incinerator					
Thermal Incinerator Ducts, Fans & Stack					
Catalytic Incinerator				81,600	81,600
Catalytic Incinerator Ducts, Fans & Stack				94,000	94,000
Condenser					
Fugitive Leak Detection and Repair (LDAR)		38,000	38,000	38,000	38,000
Total	133,900	171,900	190,100	365,700	520,800
ANNUALIZED COST ($/yr)					
Direct					
Operating Labor	33,500	43,900	55,000	71,700	82,900
Operating Materials (e.g., catalyst)				2,870	2,870
Maintenance Materials and Labor	6,690	11,000	12,000	20,700	28,500
Utilities					
Natural Gas	25,700	25,700	30,000	40,800	49,300
Electricity				890	890
Steam	13,590	13,590	13,630	13,630	18,800
Subtotal	79,500	94,200	110,600	150,700	183,300
Indirect					
Taxes, Insurance, & Administration[c]	5,360	6,860	7,590	14,600	20,800
Capital Recovery	20,400	28,500	31,100	59,600	82,400
Subtotal	25,700	35,300	38,600	74,200	103,200
Recovery Credit		15,500	15,500	15,500	15,500
Total (Direct + Indirect - Credit)	105,200	114,100	133,800	209,400	270,900

[a] Costs are on a per process line basis.

Table 8-25. HIGH DENSITY POLYETHYLENE, LIQUID PHASE-SLURRY MODEL PLANT REGULATORY ALTERNATIVES COSTS[a,b] (June 1980 dollars)

	Regulatory Alternative		
	1	2	3
INSTALLED CAPITAL COST ($)			
Flare(s)	25,100	25,100	25,100
Flare Ducting	45,300	45,300	45,300
Thermal Incinerator			
Thermal Incinerator Ducts, Fans & Stack			
Catalytic Incinerator			53,000
Catalytic Incinerator Ducts, Fans & Stack			89,100
Condenser			
Fugitive Leak Detection and Repair (LDAR)		28,400	28,400
Total	70,400	98,800	240,900
ANNUALIZED COST ($/yr)			
Direct			
Operating Labor	22,300	29,300	46,000
Operating Materials (e.g., catalyst)			620
Maintenance Materials and Labor	3,520	7,420	14,500
Utilities			
Natural Gas	8,560	8,560	9,070
Electricity			190
Steam	7,390	7,390	7,390
Subtotal	41,800	52,600	77,700
Indirect			
Taxes, Insurance, & Administration[c]	2,820	3,950	9,630
Capital Recovery	10,700	16,800	39,900
Subtotal			
Recovery Credit		10,300	10,300
Total (Direct + Indirect - Credit)	55,300	63,000	116,900

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] Includes miscellaneous operating costs for fugitive leak detection and repair program.

Table 8-26. HIGH DENSITY POLYETHYLENE, LIQUID PHASE-SOLUTION MODEL PLANT REGUALTORY ALTERNATIVES COSTS[a,b]

(June 1980 dollars)

	Regulatory Alternative		
	1	2	3
INSTALLED CAPITAL COST ($)			
Flare(s)	38,000	38,000	38,000
Flare Ducting	29,400	29,400	29,400
Thermal Incinerator			
Thermal Incinerator Ducts, Fans & Stack			
Catalytic Incinerator			425,100
Catalytic Incinerator Ducts, Fans & Stack			181,600
Condenser			
Fugitive Leak Detection and Repair (LDAR)		28,400	28,400
Total	67,400	95,800	702,500
ANNUALIZED COST ($/yr)			
Direct			
Operating Labor	33,500	40,400	57,100
Operating Materials (e.g., catalyst)			43,800
Maintenance Materials and Labor	3,370	7,270	37,600
Utilities			
Natural Gas	12,800	12,800	181,500
Electricity			13,600
Steam	5,020	5,020	5,020
Subtotal	54,700	65,500	338,700
Indirect			
Taxes, Insurance, & Administration[c]	2,700	3,830	28,100
Capital Recovery	9,780	15,900	114,600
Subtotal	12,500	19,700	142,700
Recovery Credit		10,300	10,300
Total (Direct + Indirect - Credit)	67,200	74,900	471,000

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] Includes miscellaneous operating costs for fugitive leak detection and repair program.

Table 8-27. POLYSTYRENE CONTINUOUS MODEL PLANT REGULATORY ALTERNATIVES COSTS (June 1980 dollars)[a,b]

	Regulatory Alternative						
	1	2	3	4	5	6	7
INSTALLED CAPITAL COST ($)							
Condenser			1,550	1,550	1,717	2,409	2,890
Fugitive Leak Detection and Repair (LDAR)		38,000	38,000	38,000	38,000	38,000	38,000
Total	0	38,000	39,550	39,550	39,717	40,409	40,890
ANNUALIZED COST ($/yr)							
Direct							
Operating Labor		10,400	10,897	10,897	10,897	10,897	10,897
Operating Materials (e.g., catalyst)							
Maintenance Materials and Labor		4,350	4,428	4,428	4,436	4,470	4,495
Utilities							
Natural Gas							
Electricity			192	207	286	297	371
Steam							
Freon			2	2	2	6	8
Subtotal	0	14,800	15,519	15,534	15,621	15,670	15,771
Indirect							
Taxes, Insurance, & Administration[c]		1,500	1,562	1,562	1,569	1,596	1,616
Capital Recovery		8,100	8,352	8,352	8,379	8,492	8,570
Subtotal	0	9,600	9,914	9,914	9,948	10,088	10,186
Recovery Credit		15,500	18,373	18,500	18,596	18,628	18,644
Total (Direct + Indirect - Credit)	0	8,850	7,060	6,948	6,973	7,130	7,313

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] Includes miscellaneous operating costs for fugitive leak detection and repair program.

Table 8-28. EXPANDABLE POLYSTYRENE POST-IMPREGNATION SUSPENSION PROCESS MODEL PLANT REGULATORY ALTERNATIVES COSTS (June 1980 dollars)[a,b]

	Regulatory Alternative						
	1	2	3	4	5	6	7
INSTALLED CAPITAL COST ($)							
Flare(s)				19,900	19,900	36,500	36,500
Flare Ducting				20,400	20,400	38,100	38,100
Thermal Incinerator							
Thermal Incinerator Ducts, Fans & Stack							
Catalytic Incinerator			114,800	114,800	337,700	337,700	337,700
Catalytic Incinerator Ducts, Fans & Stack			105,900	105,900	271,300	271,300	271,300
Condenser							26,300
Fugitive Leak Detection and Repair (LDAR)		28,400	28,400	28,400	28,400	28,400	28,400
Total	0		249,100	289,400	677,600	712,000	738,300
ANNUALIZED COST ($/yr)							
Direct							
Operating Labor		6,930	23,600	34,800	51,500	62,700	63,700
Operating Materials (e.g., catalyst)			5,110	5,110	22,700	22,700	22,700
Maintenance Materials and Labor		3,900	14,900	16,900	36,350	38,100	39,400
Utilities							
Natural Gas			13,000	186,100	255,000	940,300	940,300
Electricity			1,590	1,590	7,050	7,050	7,050
Steam				2,760	2,760	3,200	3,200
Subtotal	0	10,800	58,200	247,300	375,400	1,073,900	1,076,300
Indirect							
Taxes, Insurance, & Administration[c]		1,130	9,950	11,600	27,100	28,500	29,500
Capital Recovery		6,100	42,000	47,900	111,100	116,200	120,400
Subtotal	0	7,230	52,000	59,500	138,200	144,600	150,000
Recovery Credit	0	10,300	10,300	10,300	10,300	10,300	10,900
Total (Direct + Indirect - Credit)	0	7,730	99,900	296,500	503,200	1,208,200	1,215,400

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] Includes miscellaneous operating costs for fugitive leak detection and repair program.

Table 8-29. EXPANDABLE POLYSTYRENE IN-SITU SUSPENSION PROCESS MODEL PLANT REGULATORY ALTERNATIVES COSTS (June 1980 dollars)[a,b]

	Regulatory Alternative						
	1	2	3	4	5	6	7
INSTALLED CAPITAL COST ($)							
Flare(s)			12,600	12,600	12,600	25,100	37,700
Flare Ducting			12,240	12,240	12,240	18,200	24,200
Thermal Incinerator							
Thermal Incinerator Ducts, Fans & Stack							
Catalytic Incinerator				79,200	137,400	137,400	137,400
Catalytic Incinerator Ducts, Fans & Stack				115,900	213,000	213,000	213,000
Condenser							
Fugitive Leak Detection and Repair (LDAR)		15,600	15,600	15,600	15,600	15,600	15,600
Total	0	15,600	40,400	235,500	390,800	409,300	427,900
ANNUALIZED COST ($/yr)							
Direct							
Operating Labor		2,310	13,510	30,200	46,900	58,000	69,200
Operating Materials (e.g., catalyst)				2,740	4,340	4,340	4,340
Maintenance Materials and Labor		3,300	4,500	14,300	22,100	23,000	23,900
Utilities							
Natural Gas			4,290	11,900	17,900	23,900	29,700
Electricity				850	1,350	1,350	1,350
Steam			1,320	1,320	1,320	2,600	3,880
Subtotal	0	5,610	23,600	61,300	93,900	113,200	132,400
Indirect							
Taxes, Insurance, & Administration[d]		640	1,630	9,430	15,600	16,400	17,100
Capital Recovery		3,430	7,100	38,800	64,100	66,700	69,300
Subtotal	0	4,070	8,660	48,200	79,700	83,100	86,400
Recovery Credit	0	3,440	3,440	3,440	3,440	3,440	3,440
Total (Direct + Indirect - Credit)	0	6,240	28,900	106,100	170,200	192,900	215,400

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] Includes miscellaneous operating costs for fugitive leak detection and repair program.

Table 8-30. POLY(ETHYLENE TEREPHTHALATE) (PET) - DMT PROCESS MODEL PLANT REGULATORY ALTERNATIVES COSTS (June 1980 dollars)[a,b]

	Regulatory Alternative						
	1	2	3	4	5	6	7
INSTALLED CAPITAL COST ($)							
Condenser		1,520	1,520	1,520	1,520	1,520	1,520
EGRS[c] (Baseline)[d]	1,628,400	1,628,400	1,628,400	1,628,400	1,628,400	1,628,400	1,628,400
Distillation Column							239,300
Total	1,628,400	1,629,920	1,629,920	1,629,920	1,629,920	1,629,920	1,869,220
ANNUALIZED COST ($/yr)							
Direct							
Operating Labor	22,100	22,600	22,600	22,600	22,600	22,600	37,350
Operating Materials (e.g., catalyst)							
Maintenance Materials and Labor	3,100	3,176	3,176	3,176	3,176	3,176	5,146
Utilities							
Natural Gas							
Electricity	12,400	12,484	12,487	12,500	12,517	12,537	12,537
Steam	131,100	131,100	131,100	131,100	131,100	131,100	177,400
Water	664	665	665	665	665	665	665
Subtotal	169,364	170,025	170,028	170,041	170,058	170,078	233,100
Indirect							
Taxes, Insurance, & Administration	65,100	65,161	65,161	65,161	65,161	65,161	74,733
Capital Recovery	264,940	265,190	265,190	265,190	265,190	265,190	304,100
Subtotal	330,040	330,351	330,351	330,351	330,351	330,351	378,860
Recovery Credit	174,000	174,328	174,342	174,350	174,359	174,361	175,950
Total (Direct + Indirect - Credit)	325,400	326,047	326,037	326,041	326,050	326,068	436,000

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] EGRS - ethylene glycol recovery system.

[d] Ethylene glycol is recovered from the polymerization reactors through use of EG spray condensers and additional recovery equipment and from the esterifiers through use of reflux condensers.

Table 8-31a. POLY(ETHYLENE TEREPHTHALATE) (PET) - TPA PROCESS MODEL PLANT REGULATORY ALTERNATIVES COSTS, LOW VISCOSITY OR HIGH VISCOSITY WITH SINGLE END FINISHER (June 1980 dollars)[a,b]

	Regulatory Alternative	
	1	2
INSTALLED CAPITAL COST ($)		
EGRS[c] (Baseline)[d]	1,628,400	1,628,400
Distillation Column		239,300
Total	1,628,400	1,867,700
ANNUALIZED COST ($/yr)		
Direct		
Operating Labor	22,100	36,850
Operating Materials (e.g., catalyst)		
Maintenance Materials and Labor	3,100	5,070
Utilities		
Natural Gas		
Electricity	12,400	12,400
Steam	131,100	172,300
Water	664	664
Subtotal	169,400	227,300
Indirect		
Taxes, Insurance, & Administration	65,100	74,710
Capital Recovery	265,400	303,900
Subtotal	330,500	378,600
Recovery Credit	174,000	175,400
Total (Direct + Indirect - Credit)	325,900	430,500

[a] Costs are on a per process line basis.

[b] Some totals do not add up exactly due to rounding.

[c] EGRS - ethylene glycol recovery system.

[d] Ethylene glycol is recovered from the polymerization reactors through use of EG spray condensers and additional recovery equipment and from the esterifiers through use of reflux condensers.

Table 8-31b. POLY(ETHYLENE TEREPHTHALATE) (PET) - TPA PROCESS MODEL
PLANT REGULATORY ALTERNATIVES COSTS, HIGH VISCOSITY
WITH MULTIPLE END FINISHERS
(June 1980 dollars)[a,b]

	Regulatory Alternative						
	1	2	3	4	5	6	7
INSTALLED CAPITAL COST ($)							
EGRS[c] (Baseline)[d]	298,500	298,500	298,500	298,500	298,500	298,500	298,500
Distillation Column	568,000	580,500	597,000	618,900	633,000	649,800	696,300
Total	866,500	879,000	895,500	917,400	931,500	948,300	994,800
ANNUALIZED COST ($/yr)							
Direct							
Operating Labor	14,750	14,750	14,750	14,750	14,750	14,750	14,750
Operating Materials (e.g., catalyst)							
Maintenance Materials and Labor	1,970	1,970	1,970	1,970	1,970	1,970	1,970
Utilities							
Natural Gas							
Electricity	10,820	10,820	10,820	10,820	10,820	10,820	10,820
Steam	5,340	7,170	9,720	13,230	15,550	18,350	26,440
Water							
Subtotal	32,880	34,710	37,260	40,770	43,090	45,890	53,980
Indirect							
Taxes, Insurance, & Administration	34,660	35,160	35,820	36,700	37,260	37,930	39,790
Capital Recovery	140,980	143,010	145,700	149,260	151,560	154,290	161,850
Subtotal	175,640	178,170	181,520	185,960	188,820	192,220	201,640
Recovery Credit	123,390	125,750	128,250	130,740	131,970	133,220	135,680
Total (Direct + Indirect - Credit)	85,130	87,130	90,530	95,990	99,940	104,890	119,940

[a] Costs are on a per line basis.

[b] Some totals do not add up exactly due to rounding.

[c] EGRS - ethylene glycol recovery system.

[d] Ethylene glycol is recovered from the esterifiers and cooling tower using distillation columns and from the initial end finishers using spent ethylene glycol spray condensers.

Table 8-32. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR POLYPROPYLENE, LIQUID PHASE PROCESS[a]

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	7,700	20	385	7,700	20	385
3	110,100	147	750	102,400	127	805
4	129,800	151	860	19,700	4	4,925[b]

[a] Costs and emission reductions are on a per process line basis.

[b] It is unlikely that this incremental cost will be incurred. Due to the small size of the emission stream being controlled in this alternative, the stream can be easily ducted with other streams to a common control device rather than to its own control device. The cost of reduction for this alternative decreases to $805 per Mg if these additional emissions are sent to a common control device serving a process line and to $280 per Mg if serving a whole plant.

Table 8-33. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR POLYPROPYLENE, GAS PHASE PROCESS[a]

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	7,700	20	385	7,700	20	385

[a]Costs and emission reductions are on a per process line basis.

Table 8-34. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR LDPE, HIGH PRESSURE PROCESS[a]

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	7,200	15	480	7,200	15	480
3	89,200	70	1,270	82,000	55	1,490
4	145,800	84	1,740	56,600	14	4,040
5	411,700	110	3,740	265,900	26	10,200

[a]Costs and emission reductions are on a per process line basis.

Table 8-35. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR LDPE, LOW PRESSURE AND HDPE, GAS PHASE PROCESS[a]

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	8,900	30	295	8,900	30	295
3	28,600	37	770	19,700	7	2,800[b]
4	104,200	41	2,540	75,600	4	18,900
5	165,700	45	3,680	61,500	4	15,400

[a] Costs and emission reductions are on a per process line basis.

[b] It is unlikely that this incremental cost will be incurred. Due to the small size of the emission stream being controlled in this alternative, the stream can be easily ducted with other streams to a common control device rather than to its own control device. The cost of reduction for this alternative decreases to $100 per Mg if these additional emissions are sent to a common control device serving a process line and $97 per Mg if serving a whole plant.

Table 8-36. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR HDPE, LIQUID PHASE SLURRY PROCESS[a]

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	7,700	20	385	7,700	20	385
3	61,600	48	1,280	53,900	28	1,925[b]

[a] Costs and emission reductions are on a per process line basis.

[b] It is unlikely that this incremental cost will be incurred. Modification is more likely to occur at all product finishing sections in a plant rather than just one. When product finishing emissions from a typical plant are combined and sent to a common control device, the cost of reduction decreases to about $960 per Mg. Furthermore, the cost of reduction drops to $670/Mg when these emissions are combined with others in a process line and sent to a common control device and to $60/Mg when combined with others in a whole plant and sent to a common control device.

Table 8-37. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR HDPE, LIQUID PHASE SOLUTION PROCESS[a]

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	7,700	20	385	7,700	20	385
3	403,800	64	6,310	396,100	44	9,000

[a] Costs and emission reductions are on a per process line basis.

Table 8-38. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR POLYSTYRENE, CONTINUOUS PROCESS[a]

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	8,850	30	295	8,850	30	295
3	7,060	34.05	207	(1,790)[b]	4.05	(442)[b]
4	6,948	34.23	203	(112)[b]	0.18	(622)[b]
5	6,973	34.37	203	25	0.14	179
6	7,130	34.41	207	157	0.04	3,925
7	7,313	34.43	212	183	0.02	9,150

[a] Costs and emission reductions are on a per process line basis.

[b] Numbers in parentheses represent savings or negative costs.

Table 8-39. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR EPS, POST-IMPREGNATION SUSPENSION PROCESS[a]

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, $/yr	Annualized Cost, $	VOC Reduction, Mg/yr	Cost of Reduction, $/Mg
1 (Baseline)	-	-	-	-	-	-
2	7,700	20	385	7,700	20	385
3	99,900	76	1,315	92,200	56	1,645
4	296,500	169	1,750	196,600	93	2,110
5	503,200	222	2,270	206,700	53	3,900
6	1,208,200	251	4,810	705,000	29	24,310
7	1,215,400	252	4,820	7,200	1	7,200

[a] Costs and emission reductions are on a per process line basis.

Table 8-40. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR EPS, IN-SITU SUSPENSION PROCESS[a]

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	6,240	6.7	930	6,240	6.7	930
3	28,900	15.6	1,855	22,700	8.9	2,550
4	106,100	38.4	2,760	77,160	22.8	3,355
5	170,200	44.2	3,850	64,100	5.8	11,050
6	192,900	45.2	4,270	22,700	1.0	22,700
7	215,400	45.6	4,720	22,500	0.4	56,300

[a]Costs and emission reductions are on a per process line basis.

Table 8-41. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR PET/DMT PROCESS[a]

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	647	2.43	266	647	2.43	266
3	637	2.48	257	(10)[b]	0.05	(20)[b]
4	641	2.59	247	4	0.11	36
5	650	2.66	244	9	0.07	129
6	668	2.67	250	18	0.01	1,800
7	110,600	5.42	20,400	109,930	2.75	39,975

[a]Costs and emission reductions are on a per process line basis.

[b]Numbers in parentheses represent credits or negative dollar amounts.

Table 8-42a. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR PET/TPA PROCESS, LOW VISCOSITY, AND HIGH VISCOSITY WITH SINGLE END FINISHER[a]

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	104,600	2.4	43,600	104,600	2.4	43,600

[a]Costs and emission reductions are on a per process line basis.

Table 8-42b. COSTS AND ASSOCIATED EMISSION REDUCTIONS OF REGULATORY ALTERNATIVES FOR PET/TPA PROCESS, HIGH VISCOSITY WITH MULTIPLE END FINISHERS[a]

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	2,000	7.2	278	2,000	7.2	278
3	5,400	14.7	367	3,400	7.6	447
4	10,860	22.3	487	5,460	7.6	718
5	14,810	26.1	567	3,950	3.7	1,068
6	19,760	29.8	663	4,950	3.8	1,300
7	34,810	37.3	933	15,050	7.5	2,007

[a]Costs and emission reductions are on a per process line basis.

Table 8-43. 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 Costs Per Process Line[a] $/yr	Number of New Process Lines	Total Nationwide Fifth-Year Net Annualized Process and Fugitive Costs, $/yr
PP				($)
Liquid	4	129,800	9	1,168,200
Gas	2	7,700	9	69,300
LPDE				
High Pressure	5	411,700	4	1,646,800
Low Pressure	5	165,700	10	1,657,000
HDPE				
Slurry	3	61,600	6	369,600
Solution	3	403,800	9	3,634,200
Gas	5	165,700	10	1,657,000
PS				
Continuous Post-Impregnation Suspension Process	7	7,310	2	14,620
In-situ Suspension Process	7	1,215,300	2	2,430,600
	7	215,400	3	646,200
PET				
DMT	7	110,600	7	774,200
TPA[b]	2	104,600	13	1,359,800
TPA[c]	7	34,800	1	34,800
TOTAL			85	15,462,000

[a] From baseline annualized costs.

[b] Low viscosity PET line or high viscosity PET line with a single end finisher.

[c] High viscosity PET line with multiple end finishers.

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 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.[57]

8.2.1.2 Safe Drinking Water Act (SDWA). 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 contaiminant 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 Occupational Safety and Health Regulations

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 Superfund. 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 Clean Air Act

There are only three potential Federal air pollution standards that might impose costs on the facilities that will be impacted by this potential polymers and resins NSPS. All three are based on Section 111 of the Clean Air Act, and have been initiated by EPA since the Act's Amendments of 1977. First, of course, is this polymers and resins NSPS. Second is the fugitive emission NSPS for the synthetic organic chemical manufacturing industry (SOCMI). It will affect only PET/DMT facilities, because all other fugitive emissions from polymer and resin facilities will be covered by this polymer and resin NSPS. Third is the volatile organic liquid (VOL) storage NSPS. It will affect only polystyrene and PET/DMT facilities. Of these standards, only the SOCMI fugitive emissions NSPS has been promulgated.

The SOCMI fugitive emission NSPS will apply to PET/DMT because methanol is produced, and methanol is a chemical the production of which is expressly covered by the fugitive emission NSPS. (All other chemical products produced by the polymerization facilities of concern are not covered by the fugitive emission NSPS.) However, any costs incurred in compliance with the fugitive emission NSPS will be substantially offset by credits for the value of organics saved. Thus, no significant costs will accrue to PET/DMT facilities due to the fugitive emission NSPS.

The VOL storage NSPS will apply to the storage of inputs to polystyrene production facilities. In situations where styrene is produced on site, the costs of complying with the VOL storage NSPS will be shared between

the styrene and the polystyrene units. The VOL storage NSPS also will apply to the storage of methanol at PET/DMT facilities. In all other cases the inputs and outputs from polymerization facilities have vapor pressures under 3.4 kPa, and therefore are not covered, or over 76.6 kPa, and therefore would routinely be stored in pressure vessels meeting the VOL storage NSPS. In any event, costs incurred in compliance with the VOL storage NSPS will be substantially offset by credits for the value of inputs and outputs saved. Thus, no significant costs will accrue to polmerization facilities due to the VOL storage NSPS.

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9.0 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, bottles and film. Because of data limitations, only data for PET fiber are used in this analysis; however, all PET production is being regulated. 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

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 producer. 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, half 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

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	1	2				3
Fiber Industries					5	5
Firestone Tire & Rubber					1	1
Goodyear Tire & Rubber					2	2
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.

polymer 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 Capacity share. 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 accounted for about 40 percent of capacity.

Phillips Petroleum had the largest production capacity (about 25 percent) of HDPE. Oil companies, in general, accounted for approximately 60 percent of the HDPE, 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	Manufacturer	PP	LDPE	HDPE	PS	PET	Total
	Gg							Gg					
A&E Plastics				15		15	Hercules	600		5			605
Akzona					75	75	Huntsman-Goodson				10		10
Allied Corporation			300		40	340	ICI Americas					35	35
American Hoechst			135	285	295	715	InterNorth	90	275				365
American Petrofina				225		225	Kama				10		10
Atlantic Richfield	180	180	160	215		735	Minnesota Mining					35	35
Avtex Fibers					20	20	Mobil Chemical		135		90		225
BASF Wyandotte				215		215	Monsanto				315	90	405
Chemplex		190	120			310	National Distillers		325				325
Cities Service		330				330	National Petrochemical			285			285
Crest Container				20		20	Phillips Petroleum	90		680			770
Dart Industries				70		70	Polysar Group				125		125
Dow Chemical		700	205	425		1,330	Rohm and Haas					90	90
Du Pont		320	205		900	1,425	Shell Chemical	270			135		405
Eastman Chemicals	65	165			480	710	Soltex Polymer	90		340			430
El Paso Natural Gas	135	250				385	Standard Oil (Indiana)	235		170	195		600
Exxon	180	455				635	Texstyrene Plastics				25		25
Fiber Industries					680	680	Union Carbide		960	90	5		1,055
Firestone Tire & Rubber					25	25	United Foam		75				75
Goodyear Tire & Rubber					135	135	USS Chemicals	235			10		245
Carl Gordon				70		70	Vititek				5		5
Gulf Oil	180	390	260	160		990							
							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.

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 Vertical integration. 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.

Backward-to-suppliers integration. 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 involves 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.

Forward-to-final product market integration. 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 Horizontal integration--product diversification. 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. 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, PET plastics, 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 chloride, for purposes of this study 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; consequently, the total SIC 2821 and 3079 shipments in 1981 for PP, HDPE, LDPE, and PS is estimated to be $15.0 billion.

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 Plants. 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 Size. 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 first produced commercially in the U.S. in 1953.

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 1981 employment data for plants producing the five considered polymers and resins are included in the Census of Manufactures data by SIC code.

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
LDPE	10.0	40.0	35.0	15.0	100
HDPE	25.0	50.0	12.5	12.5	100
PS	75.7	21.6	2.7	-	100
PET	59.3	22.2	14.8	3.7	100

[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, 6]

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 28213; this is based on the proportion of PVC production to the total production.[3]
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.[1]
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.

9.1.1.3 Markets.[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 applications in both the auto industry and the fibers market.

9.1.1.3.1 Substitutes. 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 Imports-Exports. 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 of this 50 Gg was 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 Processes and uses. 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 End-use markets. 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 split 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 Performance

The initial commercial development of the major thermoplastics occurred in the period 1935-1945 with the introduction of such plastics as PS and LDPE. In the 1950's HDPE, PP, and PET 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. Although industry statistics may vary according to source, the data shown in Table 9-9 represent the general trends in the industry. 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. During the period PET fibers had 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 a high of 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
	Utilizlation (%)	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 consumer 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 Finances.[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 Sales and value of production. 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 economic expansion 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 model plants or process lines. Eighty-five new model process lines equivalent in size to 27 model plants (defined in Chapter 6) are projected. The numbers of model plants and process lines by type of polymer and process are:

Polymer/Process	Number of Plants	Number of Process Lines
PP		
Liquid phase	3	9
Gas phase	3	9
LPDE		
High pressure	1	4
Low pressure	5	10
HDPE		
Liquid phase-slurry	2	6
Liquid phase-solution	3	9
Gas phase	5	10
PS		
Continuous	1	2
EPS, PI	0.7	2
EPS, IS	0.3	3
PET		
DMT	1	7
TPA - single end finisher	1.9	13
- multiple end finishers	0.5	1

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) take into account the current geographic distribution (SIP and non SIP states) of the existing plants and assumes 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:

$$NC = FP \frac{1}{CU} - CC + RC - IC \quad (1)$$

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

$$NP = NC \div MP \quad (2)$$

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 is projected by applying the applicable growth rate as determined from published projections[23,24] to the 1981 domestic production and net exports and summing. The values so calculated are tabulated in Table 9-11.

The published growth rates are developed from historical trends and end uses data, from communications 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. 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 decline in demand in 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 growth rates are determined for both domestic markets and net exports. The annual growth rates assumed for each polymer and resin are discussed individually below and shown in Table 9-11.

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 is 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 is 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 is 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 is 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 is 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 is 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 are 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 is 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 to 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 is assumed to 1984. From 1984 on, a growth rate of 4 percent is 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 are assumed to remain constant in order to arrive at a conservative projection.

9.1.3.2 Projected Capacity. Projected capacity required in 1988 is calculated by applying a utilization rate to the projected 1988 production.

In order to project the capacity, a capacity utilization rate that will trigger additional capacity must be assumed. The capacity utilization rates used are estimated on the basis of published information[29] and discussions with an 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 are used.

The projected 1988 capacities for the five polymers and resins, calculated from the assumed utilization rates and the projected 1988 production, are shown in Table 9-11.

9.1.3.3 Retired Capacity. In calculating the retired capacity, the useful life of all plants is assumed to be 20 years. Therefore, any capacity built between January, 1964 and January, 1969, is 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 are assumed to be replaced. For the PET industry where there are no known plans to rebuild a portion of the retired capacity, it is 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 Chemical Engineering, are shown in Table 9-11.[31]

9.1.3.4 Interim Capacity. Interim capacities are the capacities of plants to be completed in 1982 and 1983. Values for PS, LDPE, HDPE and PS are taken from the Facts and Figures of the U.S. Plastics

Table 9-11. PROJECTED REQUIRED NEW CAPACITY FOR POLYMERS AND RESINS, 1984 THROUGH 1988

Product	1981 capacity[a]	1981 production[b]	1981 U.S. domestic consumption[c]	Net exports[d]	Growth rate domestic	Growth rate net exports	Projected 1988 production	1988 utilization rate	Projected 1988 capacity	Retired capacity 1984-88	Interim capacity 1982-83	Total required new capacity
	Gg/yr	Gg	Gg	Gg	%	%	Gg	%	Gg/yr	Gg/yr	Gg/yr	Gg/yr
PP	2,290	1,790	1,480	310	7.0	-4.7	2,600	88	2,950	115	0	775
LDPE	4,430	3,490	3,040	450	5.6	-10.0	4,660	88	5,300	520	390	1,000
HDPE	2,720	2,130	1,970	160	7.9	0	3,510	88	3,990	330	195	1,410
PS	2,350	1,640	1,570	70	4.0	4.0	2,160	85	2,540	120	160	150
PET	2,100	1,890	1,510	380	2.3[e]	0	2,150	88	2,440	180[f]	90	340

[a] Table 9.9
[b] Table 9.9
[c] 1981 production minus net exports
[d] Exports (Table 9.6) minus imports (Section 9.1.1.3.2)
[e] Zero through 1984; 4% 1985 through 1988
[f] Assumed only half the retired capacity replaced. (See text p. 9-30)

Industry.[9] Values for PET are 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) i.e., the 1981 capacity and the interim capacity are subtracted from the projected 1988 capacity, and the retired capacity is added to obtain the total required new 1988 capacity. A sensitivity analysis was not conducted in light of the low level of price effects as discussed below.

In projecting the additional capacity required in 1988 it should be noted that this 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 or process section
- 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 economic impact analysis it was assumed that all affected facilities will be new model size plants, i.e., no modifications and reconstructions are included. (See Chapter 5.)

9.1.3.6 New Plants. The projected capacity requires the equivalent of 27 new model size plants or 85 new model process lines. Industry sources indicate that some of this capacity may be in the form of new grass roots plants and some capacity will stem from adding new process trains to existing plants. Sufficient information is not available, however, to

allow the projection of the amount of new capacity that will be added by new plants and the amount that 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 that can make both LLDPE or HDPE. Thus, data on plants 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 reflect 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 the extent 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 economic impact analysis, therefore, the projected new capacities are expressed in terms of equivalent new model plants.

9.1.3.7 Projections by Process. 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 is tabulated in Table 9-12. (For ease of comparison with other information given in this document, capacities are also expressed in terms of model process lines.)

9.1.3.7.1 PP plants. 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

Table 9-12. PROJECTED NUMBER OF NEW POLYMERS AND RESINS PLANTS, BY PROCESS 1984 THROUGH 1988

Product	New Capacity Required[a] Gg/yr	Process	Model Plant Capacity[b] Gg/yr	Model Process Line Capacity[b] Gg/yr	Projected Number of New Plants 1984 Through 1988	Projected Number of New Process Lines 1984 Through 1988
PP	775	Liquid phase	150	50	3	9
		Gas phase	105	35	3	9
LDPE	1,000	High pressure	280	70	1	4
		Low pressure	150	75	5	10
HDPE	1,410	Liquid phase slurry	210	70	2	6
		Liquid phase solution	90	30	3	9
		Gas phase	150	75	5	10
PS[c]	150	Continuous	75	37.5	1	2
		EPS, PI	75	25	0.7	2
		EPS, IS	72	8	0.3	3
PET	340	DMT	105	15	1	7
		TPA - single end finisher	105	15	1.9	13
		- multiple end finishers	40	20	0.5	1

[a]See Table 9-11.

[b]See Tables 6-1 to 6-11b.

[c]The two processes for expandable polystyrene are post-impregnation (PI) and in-situ (IS).

three new liquid phase process plants are projected. The capacity of these six plants, 765 Gg/yr, is essentially equal to the projected required new capacity.

9.1.3.7.2 LDPE plants. Most new construction will employ gas-phase, low pressure, Unipol or similar proprietary process technology; however, the use of the high pressure liquid process cannot be ruled out. The consequent projections assume that one high pressure plant will be built and that the remainder of the capacity added will be low pressure plants. These assumptions result in one high-pressure plant and five low-pressure 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 such as expandable PS (EPS), then a batch plant is more suitable. Indications are that new capacity will probably be added in both forms; therefore, one continuous plant and EPS process lines constituting the equivalent of one EPS plant are projected. The projected process lines result in 345 Gg/yr; the projected required capacity is 340 Gg/yr.

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. Of the required 21 model process lines, 20 are projected to be for fiber resins and one for industrial resins. The industrial resins process line is expected to be a TPA line.

The combined capacity of the projected PET plants is 320 Gg/yr, slightly less than the 340 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. 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.

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 projected plants.

The additional annualized costs of controls in 1988 (Table 9-16), the fifth year of controls, for the 85 projected model process lines (27 projected polymers and resins manufacturing plants) are estimated to be but $15.5 million under the combination of the most stringent regulatory alternatives for each plant. Thus, a 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 most of the alternatives would be less than five-tenths of a percent. The maximum price increase required to fund any alternative would not exceed 3.3 percent. Because of the minor increases required by these plants, no significant economic impacts 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 Tax Equity and Fiscal Responsibility Act of 1982.
- 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:

$$NPV = \sum_{y=1}^{n} \frac{CF_y}{(1+d)^y} - I_o \qquad (1)$$

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.

$$CF_y = R_y - OM_y - T_y \quad (6a)$$

$$= R_y - OM_y - t(R_y - OM_y - D_y) + ITC_y \quad (6b)$$

$$= R_y - OM_y - tR_y + tOM_y + tD_y + ITC_y \quad (6c)$$

$$= (1-t)R_y - (1-t)OM_y + tD_y + ITC_y \quad (6d)$$

Further, by substitution of Equations (3) and (4) into (6d), operating cash flow, CF_y, may be expressed by:

$$CF_y = (1-t)R_o(1+inf)^y - (1-t)OM_o(1+inf)^y + tD_y + ITC_y \quad (7)$$

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),

$$NPV = -I_o + \sum_{y=1}^{n} \frac{(1-t)R_o(1+inf)^y - (1-t)OM_o(1+inf)^y + tD_y + ITC_y}{(1+d)^y} \quad (8)$$

Equivalently, each term under summation can be summed individually so that

$$NPV = -I_o + \sum_{y=1}^{n} \frac{(1-t)R_o(1+inf)^y}{(1+d)^y} - \sum_{y=1}^{n} \frac{(1-t)OM_o(1+inf)^y}{(1+d)^y} + \sum_{y=1}^{n} \frac{ITC_y + tD_y}{(1+d)^y} \quad (9)$$

By imposing the constraint that incremental NPV=0, it is possible to solve for the annual revenue requirement, R_o, (in base year dollars) that is equivalent to incremental emissions control investment, I_o, 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)

$$I_0 - \sum_{y=1}^{n} \frac{ITC_y + tD_y}{(1+d)^y} = I_0 \left[1 - \frac{(.1+.15t)}{(1+d)} - \frac{.22t}{(1+d)^2} - \frac{.21t}{(1+d)^3} - \frac{.21t}{(1+d)^4} - \frac{.21t}{(1+d)^5} \right]$$

$$= I_0 \{TAXF\} \qquad (12)$$

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

$$R_o = \frac{(1-t)OM_o \sum_{y=1}^{n} \frac{(1+inf)^y}{(1+d)^y} + I_o \{TAXF\}}{(1-t) \sum_{y=1}^{n} \frac{(1+inf)^y}{(1+d)^y}} \quad (13)$$

Simplifying Equation (13) results in

$$R_o = OM_o + I_o \frac{TAXF}{(1-t) \sum_{y=1}^{n} \frac{(1+inf)^y}{(1+d)^y}} \quad (14)$$

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:

$$R_o = OM_o + I_{15} \frac{TAXF}{(1-t) \sum_{y=1}^{15} \frac{(1+inf)^y}{(1+d)^y}} + I_{10} \frac{TAXF}{(1-t) \sum_{y=1}^{10} \frac{(1+inf)^y}{(1+d)^y}} \quad (15)$$

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

$$PI = \frac{R_o}{Q_o} \tag{16}$$

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

$$PPI = \frac{PI}{P_o} \times 100 \tag{17}$$

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 Required Revenue and Maximum Price Increases. 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 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 process lines and 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 process lines are summarized in Table 9-13.

The model plants represent 13 product-process combinations. Twenty-seven affected plants consisting of 85 process lines are projected for the first five years, i.e., 1984 through 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 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 Facts and Figures.[35] That table reflects both quantities of sales and net 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 prices were not available from the above source.

The PET price represents the price of fiber processed within the PET plants. Prices of PET used for bottles and film, which amounts to only a small portion of the total PET produced, are not available in published sources. Determining an average annual price for the PET fibers plant was more involved than determining that for the other plants, because the PET plant produces a mix of product fibers of varying prices. Chemical and Engineering News (C&EN)[36] reports 1980 prices for three polyester fibers:

Table 9-13. OPERATIONAL CHARACTERISTICS OF POLYMERS AND RESINS MODEL PROCESS LINES BUILT 1984 THROUGH 1988 (JUNE, 1980 DOLLARS)

Product	Process	Projected Number of Affected Lines	Annual Capacity per Line (Gg)	Utilization Rate[a] (%)	Annual[b] Production per Line (Gg)	Price[c] ($/Kg)	Annual Sales per Line[d] ($1000)
PP	Liquid phase	9	50	85	42.5	.70	29,750
	Gas phase	9	35	85	29.8	.70	20,825
LDPE	High pressure	4	70	85	59.5	.82	48,790
	Low pressure	10	75	85	63.8	.82	52,275
HDPE	Liquid phase slurry	6	70	85	59.5	.79	47,005
	Liquid phase solution	9	30	85	25.5	.79	20,145
	Gas phase	10	75	85	63.8	.79	50,363
PS[e]	Continuous	2	37.5	85	31.9	1.02	32,513
	EPS, PI	2	25	85	21.3	1.73	36,848
	EPS, IS	3	8	85	6.8	1.73	11,791
PET	DMT	7	15	85	12.8	1.80	22,950
	TPA - single end finisher	13	15	85	12.8	1.80	22,950
	- multiple end finishers	1	20	85	17.0	1.80	30,600

[a]Generally profitable 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; the price of EPS is estimated to be 1.7 times that of PS.

[d]Annual sales equals annual production times price.

[e]The two processes for expandable polystyrene (EPS) are post-impregnation (PI) and in-situ (IS).

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 increases for polymer and resin model process lines and percent required maximum price increases for process lines (and also model plants) to comply with potential NSPS regulatory alternatives are shown in Table 9-14. These measures are shown with the associated process line estimated investment, operating and maintenance costs, and recovery credits for each regulatory alternative. For each model, 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 PROCESS LINE CONTROL COSTS[a] AND MAXIMUM PRICE INCREASES FOR MODEL PLANT BY PRODUCT, PROCESS, AND REGULATORY ALTERNATIVE (JUNE, 1980 DOLLARS)

Model process lines by product and process	Regulatory alternative	Investment[b]	Recovery credit[c]	O&M[d]	Required revenue increase[e]	Maximum price increase[f]
		($1000)				(%)
PP						
o Liquid phase	2	28.4	10.3	12.0	11.5	0.04
	3	332.9	10.3	64.9	110.7	0.37
	4	351.2	10.3	81.9	130.5	0.44
o Gas phase	2	28.4	10.3	11.9	7.4	0.04
LDPE						
o High pressure	2	23.6	7.8	9.8	9.9	0.02
	3	237.1	7.8	57.1	89.6	0.18
	4	383.9	7.8	89.8	144.6	0.30
	5	1,292.3	7.8	223.6	418.0	0.86
o Low pressure	2	38.0	15.5	16.3	8.4	0.02
	3	56.2	15.5	33.4	27.8	0.05
	4	231.8	15.5	80.5	101.6	0.19
	5	386.9	15.5	119.2	160.9	0.31
HDPE						
o Liquid phase slurry	2	28.4	10.3	11.9	7.4	0.02
	3	170.5	10.3	42.8	59.9	0.13
o Liquid phase solution	2	28.4	10.3	12.0	7.5	0.04
	3	635.1	10.3	309.4	397.1	1.97
o Gas phase	2	38.0	15.5	16.3	8.4	0.02
	3	56.2	15.5	33.4	27.8	0.05
	4	231.8	15.5	80.5	101.6	0.19
	5	386.9	15.5	119.2	160.9	0.31
PS						
o Continuous	2	38.0	15.5	16.3	10.4	0.03
	3	39.6	18.4	17.1	8.5	0.03
	4	39.6	18.5	17.1	8.4	0.03
	5	39.7	18.6	17.2	8.4	0.03
	6	40.4	18.6	17.3	8.6	0.03
	7	40.9	18.6	17.4	8.8	0.03
o EPS, PI	2	28.4	10.3	12.0	7.5	0.02
	3	249.1	10.3	68.2	97.3	0.26
	4	289.4	10.3	258.4	292.8	0.79
	5	677.6	10.3	402.6	496.0	1.35
	6	712.0	10.3	1,102.5	1,200.6	3.26
	7	738.3	10.9	1,105.8	1,207.3	3.28
o EPS, IS	2	15.6	3.4	6.3	6.2	0.05
	3	40.4	3.4	25.2	29.0	0.25
	4	235.5	3.4	70.7	104.2	0.88
	5	390.8	3.4	109.5	166.6	1.41
	6	409.3	3.4	129.6	189.1	1.60
	7	427.9	3.4	149.5	211.3	1.79
PET						
o DMT	2	1.5	0.3	0.7	0.6	0.00
	3	1.5	0.3	0.7	0.6	0.00
	4	1.5	0.4	0.7	0.6	0.00
	5	1.5	0.4	0.8	0.6	0.00
	6	1.5	0.4	0.8	0.6	0.00
	7	240.8	2.0	73.4	39.2	0.17
o TPA						
- Single end finisher	2	239.3	1.4	67.5	38.5	0.17
- Multiple end finishers	2	12.5	2.4	2.3	1.9	0.01
	3	29.0	4.9	5.5	5.1	0.02
	4	50.9	7.4	9.9	10.3	0.03
	5	65.0	8.6	12.8	14.1	0.05
	6	81.8	9.8	16.3	18.9	0.06
	7	128.3	12.3	26.2	33.4	0.11

a Control costs and recovery credits are based on data contained in Tables 8-21 to 8-31b.

b Investment for each alternative represents the sum of the incremental capital costs (over baseline, Alternative I) for the various types of equipment as aggregated by economic life (costs were rounded to the nearest $100). For example, the investment for Regulatory Alternative 5, LDPE low pressure is $386,900 (shown above as 386.9). This is obtained from Table 8-24 and represents the sum of incremental costs of $1,250 for equipment with an economic life of 2 years, $9,200 for equipment with an economic life of 6 years, $283,000 for equipment with an economic life of 10 years, and $93,000 for equipment with an economic life of 15 years. The sum of the incremental costs indicated above also equals the difference of the total cost under baseline of $133,900 and the cost under Alternative 5 of $520,800.

c Except for PET, recovery credits are obtained directly from the tables in Chapter 8; incremental costs are not involved since there are no credits under baseline. However, for PET there are recovery credits under baseline; consequently, incremental credits are used.

d O&M represents the incremental annualized costs (excluding capital recovery which is incorporated in the formula shown below). In the example, the incremental O&M cost for LDPE low pressure is $119,200 which is the difference between the annualized cost of $84,800 under baseline (105,200 less 20,400) and $204,000 under Alternative 5 (286,400 less 82,400).

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

$$R_0 = OM_0 + .577I_1 + .230I_2 + .152I_3, + .115I_4$$

where

R_0 = Required revenue increase
OM_0 = O&M less recovery credit
I_1 = Incremental capital costs of equipment with economic life of 2 years.
I_2 = Incremental capital costs of equipment with economic life of 6 years.
I_3 = Incremental capital costs of equipment with economic life of 10 years.
I_4 = Incremental capital costs of equipment with economic life of 15 years.

For Alternative 5 of LDPE low pressure, the calculations are:

$$R_0 = (119.2 - 15.5) + .577(1.3) + .230(9.2) + .152(283.0) + .115(93.0)$$

$$= 160.9$$

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:

$$PPI = \frac{Ro}{PoQo} \times 100 \text{ or } \frac{\text{Required revenue increase}}{\text{dollar sales volume}} \times 100$$

The dollar sales volume for each model process line is shown in Table 9-13. Percentage price increases are the same for both the model process line and the multiple process line model plant.

Maximum price increases for all alternatives of the models are relatively insignificant except in the few cases indicated below. The maximum price increase for Alternative 2 of all model plants except PET is less than .05 percent; for PET plants the increase is .4 percent. Increases for the remaining alternatives are less than 1 percent for all model plants except the HDPE liquid phase solution model and the EPS models. The price increase under Alternative 3 for the HDPE model is 2.0 percent; for the EPS models the price increases range up to 3.3 percent under Alternatives 6 and 7 for the EPS, PI process line.

9.2.2.2 Expected Price and Profitability Impacts. 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 Other Economic Effects. 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 Fifth Year Costs and Benefits

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 PROCESS LINE PRODUCT AND PROCESS (JUNE, 1980 DOLLARS)

Model process lines by product and process	Regulatory alternative	Net annualized cost ($1000)	Number of process lines	Total annualized cost ($1000)
PP				
o Liquid phase	2	7.7	9	69.3
	3	110.1		990.9
	4	129.8		1,168.2
o Gas phase	2	7.7	9	69.3
LDPE				
o High pressure	2	7.2	4	28.8
	3	89.2		356.8
	4	145.8		583.2
	5	411.7		1,646.8
o Low pressure	2	8.9	10	89.0
	3	28.6		286.0
	4	104.2		1,042.0
	5	165.7		1,657.0
HDPE				
o Liquid phase slurry	2	7.7	6	46.2
	3	61.6		369.6
o Liquid phase solution	2	7.7	9	69.3
	3	403.8		3,634.2
o Gas phase	2	8.9	10	89.0
	3	28.6		286.0
	4	104.2		1,042.0
	5	165.7		1,657.0
PS				
o Continuous	2	8.9	2	17.8
	3	7.1		14.2
	4	7.0		14.0
	5	7.0		14.0
	6	7.2		14.4
	7	7.3		14.6
o EPS, PI	2	7.7	2	15.4
	3	99.8		199.6
	4	296.0		592.0
	5	503.4		1,006.8
	6	1,208.4		2,416.8
	7	1,215.3		2,430.6
o EPS, IS	2	6.2	3	18.6
	3	28.9		86.7
	4	106.1		318.3
	5	107.2		321.6
	6	192.9		578.7
	7	215.4		646.2
PET				
o DMT	2	0.6	7	4.2
	3	0.6		4.2
	4	0.6		4.2
	5	0.6		4.2
	6	0.7		4.9
	7	110.6		774.2
o TPA				
- Single end finisher	2	104.6	13	1,359.8
- Multiple end finishers	2	2.0	1	2.0
	3	5.4		5.4
	4	10.9		10.9
	5	14.8		14.8
	6	19.8		19.8
	7	34.8		34.8

[a] These costs represent the costs of the regulatory alternatives over and above the baseline as described in Chapter 8.

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 of aggregate cost to society, is $15.5 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 human cell damage known as chromosomal aberrations.
- Vegetation effects. Ozone reduces citrus, grape, cotton, and other crop yields by damaging leaves and plants. The reduction has been 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 process lines by product and process	Number of process lines	Regulatory alternative	Total annualized cost ($)
PP			
o Liquid Phase	9	4	1,168,200
o Gas Phase	9	2	69,300
LDPE			
o High Pressure	4	5	1,646,800
o Low Pressure	10	5	1,657,000
HDPE			
o Liquid Phase Slurry	6	3	369,600
o Liquid Phase Solution	9	3	3,634,200
o Gas Phase	10	5	1,657,000
PS			
o Continuous	2	7	14,600
o EPS, PI	2	7	2,430,600
o EPS, IS	3	7	646,200
PET			
o DMT	7	7	774,200
o TPA			
- Single end finishers	13	2	1,359,800
- Multiple end finishers	1	7	34,800
TOTAL			15,462,000

[a] Annualized fifth year costs are based on the highest alternative cost for each model process line as shown in Table 9-15.

- o 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.
- o 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) as those with 1,000 or fewer. 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 businesses constructing new plants would be small or, if small, would remain small. Furthermore, as the analysis in Section 9.2 explains, any potential adverse economic impacts on new plants, regardless of whether they are owned or operated by 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.

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*References can be located in Docket Number A-82-19 in the U.S. Environmental Protection Agency Library, Waterside Mall, Washington, D.C.

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 that have occurred in the development of background information for the new source performance standards for the polymers and resins industry.

Date	Activity
August 30, 1979	Pullman Kellogg Report, "Polymer Industry Ranking by VOC Emissions Reduction that would occur from New Source Performance Standards."
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.
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.

Date	Activity
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.

Date	Activity
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.
April 14, 1982	Tennessee Eastman's comments on draft BID Chapters 3-6.
April 15, 1982	Allied Fibers and Plastics' comments on draft BID Chapters 3-6.
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 20, 1982	Trip to Texas Air Control Board, Austin, Texas.
September 20, 1982	114 Letter to Fiber Industries requesting information on emissions and their control at Fiber Industries' Polyester resin plant, Salisbury, N.C.
September 22, 1982	Letter to Gulf Oil Chemical requesting additional information regarding Gulf's polystyrene plant, Marietta, Ohio.
September 29, 1982	Plant visit to Fiber Industries' polyester facilities at Salisbury, North Carolina.
November 2, 1982	Letter to Tennessee Eastman requesting information on ethylene glycol recovery and costs at polyester resins plants.

Date	Activity
November 2, 1982	Letter to Cosden Oil requesting information on expandable polystyrene industry.
November 10, 1982	Letter from Cosden Oil responding to 11/2/82 letter requesting information on expandable polystyrene industry.
November 11, 1982	Letter from Gulf Oil responding to 9/22/82 letter requesting additional information on polystyrene facility at Marietta, Ohio.
November 22, 1982	Letter from Tennessee Eastman responding to 11/2/82 letter requesting information on ethylene glycol recovery and costs at polyester resins plants.
November 23, 1982	Meeting with Fiber Industries to discuss poly(ethylene terephthalate) manufacture and emission control.
December 16, 1982	114 Letter sent to four manufacturers of expandable polystyrene requesting information on production, emissions, and emission control.
February 18, 1983	Letters from ARCO and BASF Wyandotte responding to 114 letter on expandable polystyrene.
March 16, 1983	Mailout of material for National Air Pollution Control Techniques Advisory Committee meeting.
March 30, 1983	Letter from Cosden Oil Company responding to a 114 letter requesting information on expandable polystyrene.
April 19, 1983	Letter from Union Carbide Corporation commenting on draft NSPS regulation and background document.
April 20, 1983	Letter from DuPont commenting on draft NSPS regulation and background document.
April 20 and 21, 1983	Trip to Union Carbide Corporation, South Charleston, W.V., to discuss control of emergency releases from high pressure, LDPE tubular reactors.
April 26, 1983	Presentation made to the National Air Pollution Control Techniques Advisory Committee.
April 27, 1983	Letter from DuPont commenting on draft NSPS regulation and background document.
May 4, 1983	Letter from American Hoechst responding to a 114 letter requesting information on expandable polystyrene.
May 12, 1983	Letter from Monsanto commenting on draft NSPS regulation and background document.

Date	Activity
May 19, 1983	Meeting with Allied Fibers to discuss their concerns with the draft NSPS regulation.
May 19, 1983	Letter from Hercules Inc., commenting on draft NSPS regulation and background document.
May 23, 1983	114 Letter to Gulf Oil Chemicals requesting information on emissions and emission control from high pressure, LDPE manufacturing.
May 31, 1983	Letter from Dow Chemical commenting on draft NSPS.
June 2, 1983	Meeting with Gulf Oil Chemicals to discuss control of emergency releases from high pressure LDPE autoclave reactors.
June 16, 1983	Letter to USS Chemicals requesting additional information concerning intermittent emissions from their plant.
June 21, 1983	Letter from Union Carbide Corporation transmitting safety report excerpts concerning high velocity vents for emergency releases from UNIPOL reactors.
June 24, 1983	Letter from Texas Chemical Council commenting on the draft NSPS.
June 30, 1983	Letter from Gulf Oil Chemicals transmitting TACB operating permits for Gulf's Cedar Bayou plant.

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 3-63.
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 4-37.

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 6-15 through 6-41.
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 7-24.
Energy impacts	The energy impacts of the various regulatory alternatives are discussed in Chapter 7, Section 7.4, pages 7-24 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 8-61.
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.

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 the first phase of an extended EPA study. More recent test results are found in Docket Reference Number II-B-94.

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

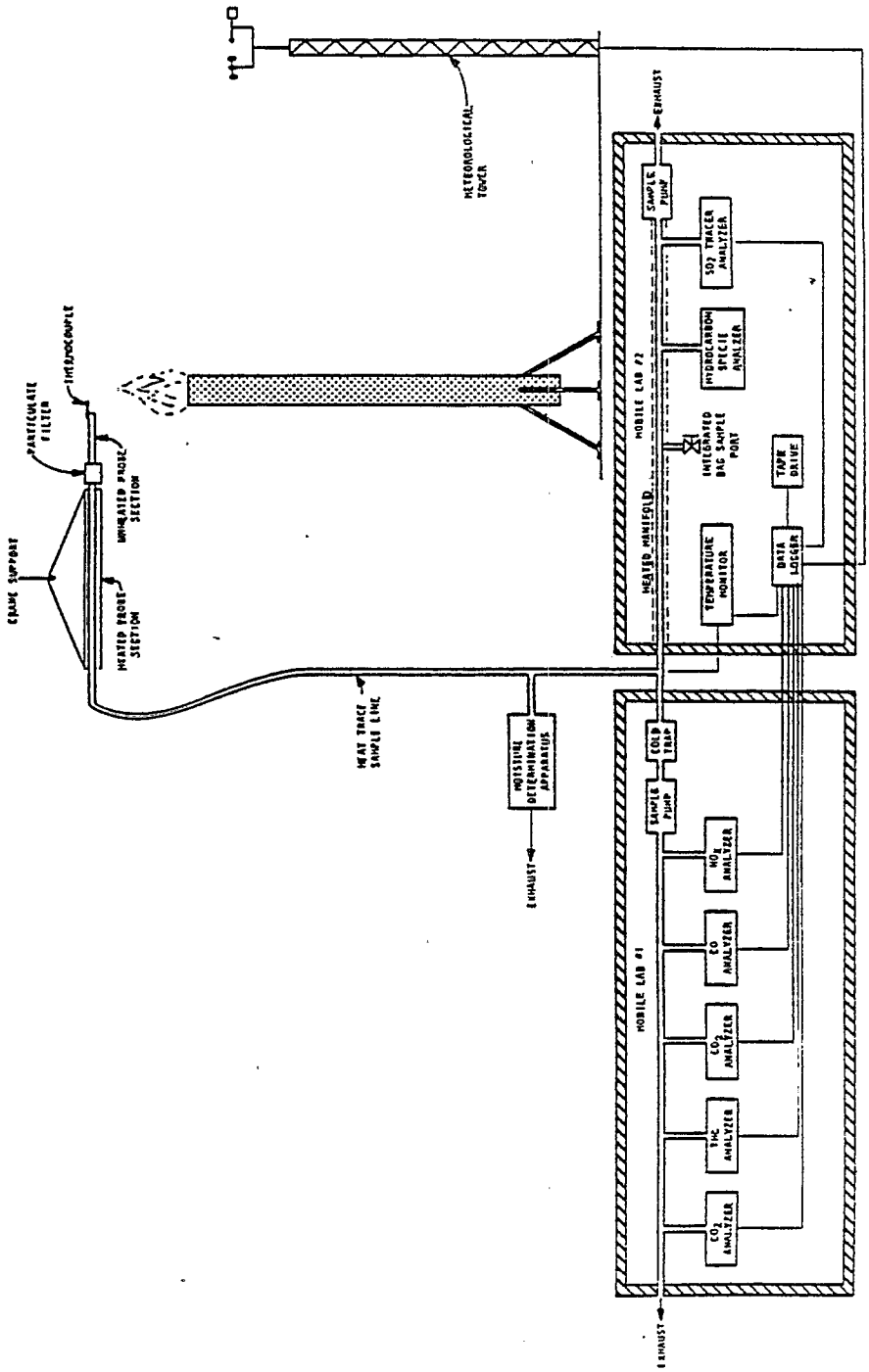


Figure C-1. Flare sampling and analysis system

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 Test Results

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/m^3 (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[b] (scfm)	Velocity (ft/min)	Lower Heating Value (Btu/scf)	Propylene Flow (lbs/hr)	Propylene Flow (scfm)	Nitrogen Flow (lbs/hr)	Nitrogen Flow (lbs/hr)	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	3,027	465	-	-	1,355	0.448	99.82	Incipient smoking flare
4	283	1,509	2,183	1,875	283	-	-	-	-	98.80	Smoking flare
8	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	232	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 473 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	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[d]	1545	1.0	99.7		1
Union Carbide Corp., Taft, LA	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.

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 181,000 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

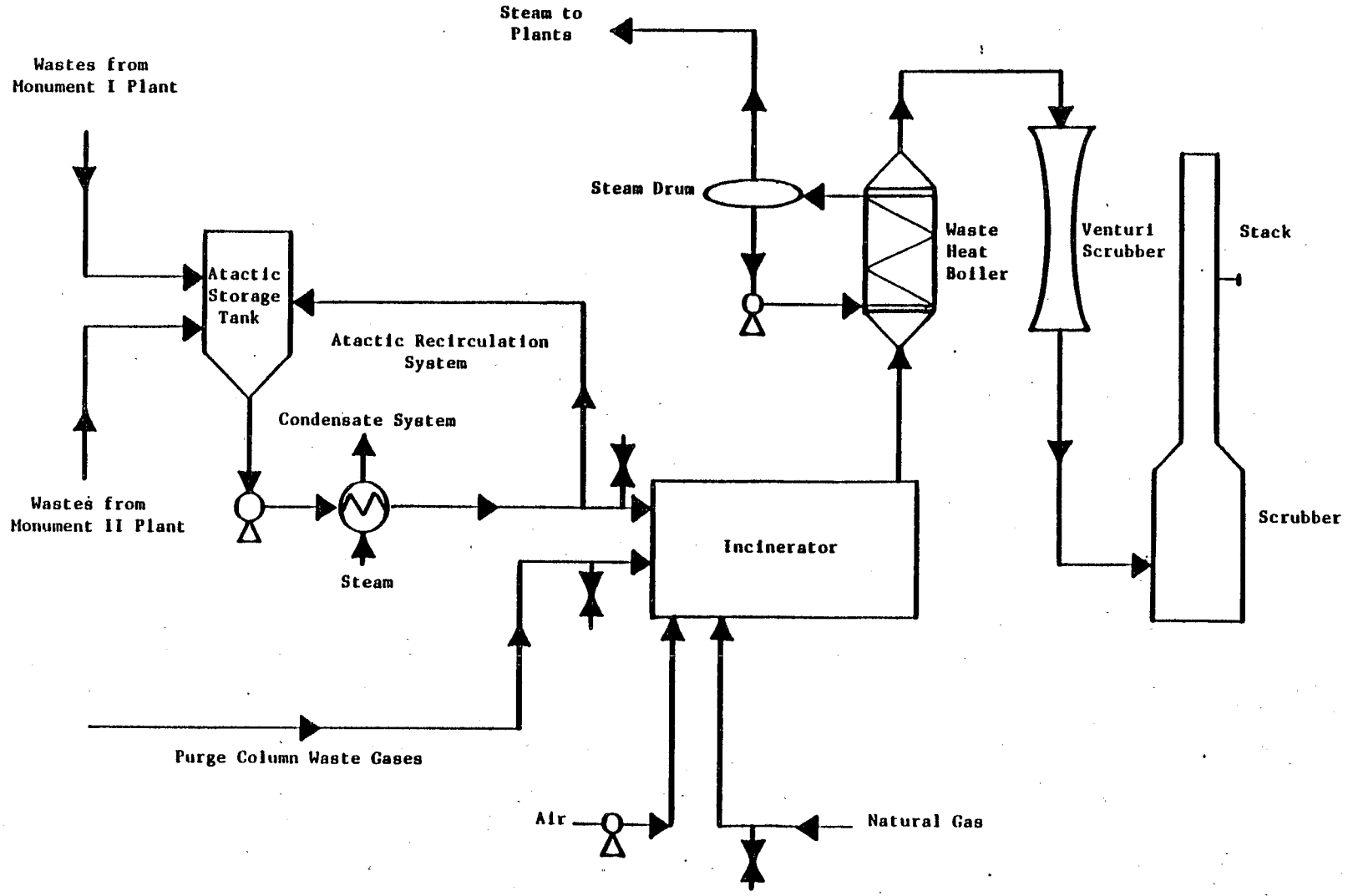


Figure C-2. Schematic of incineration system at ARCO polypropylene facility

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 ≈2.18 MJ/s (7.45×10^6 Btu/hr),

b) air supply ≈15.1 standard m^3/s at 0°C (33,900 scfm, at 60°F)

c) firebox temperature ≈980°C average and 1,200°C maximum (1,800°F average and 2,200°F maximum),

d) firebox residence time ≈1.5 seconds, and

e) pressure ≈19 kPa (78 in. H_2O).

C.2.1.2 Sampling and Analytical Techniques. 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	-	-	-	585 kPa(85 psig)
	Range	960-1,030 kPa(140-150 psig)	998-1,070 kPa(145-155 psig)	21-41 kPa(3-6 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(351b/hr)	-	-
	Range	-	-	11-13 g/sec(90-100 lb/hr) (propylene)	-

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_2O)[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_2O.

[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 1,300°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	99.8 ± .1	99.9 ± .1	99.6 ± .4	76 ± 20	
NG/WG 1,600°F	>99.76 ± .07	99.8 ± .10	99.88 ± .04	66 ± 10	99.88 ± .04
AW/NG 1,600°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 = $100 - \frac{\text{gC in Stack gas}}{(\text{gC in Atactic Waste} + \text{gC in Waste Gas})}$

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 sm^3/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. H_2O 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.

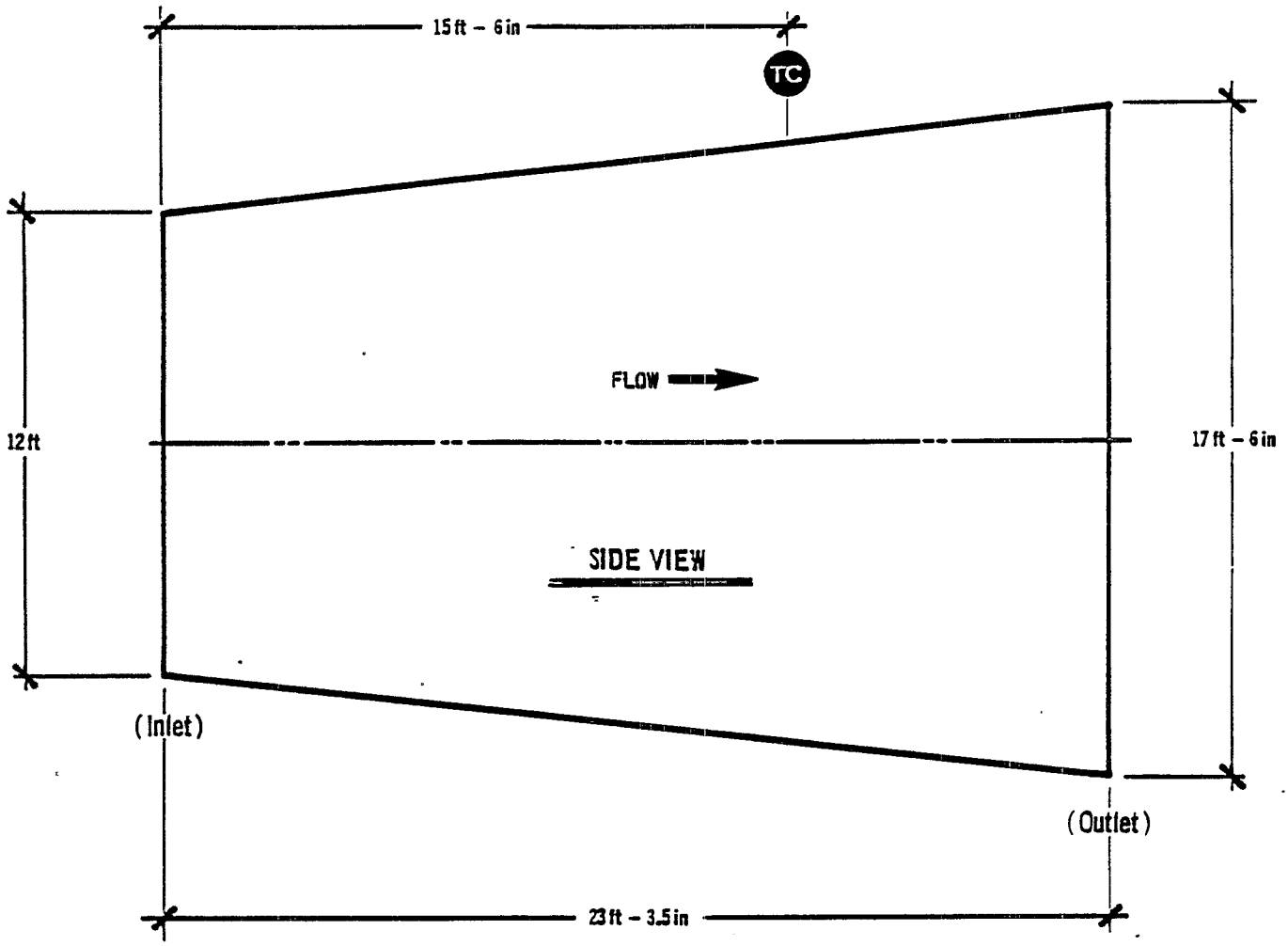
deviations for this measurement method. The accuracy and representativeness of these values obtained from Method 25 is, thus, questionable. If Method 25 results are disregarded, the DE's for all testing combinations are found to be consistently above 99 percent.

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. Sampling and Analytical Techniques. 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 cm^3 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 (Federal Register, 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[a] (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		NA[e]			47	99

[a] At production capacity.

[b] At 70 percent rather than 100 percent of total capacity.

[c] Total waste gas flow rate of 52,500 scfm comprised 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).

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.

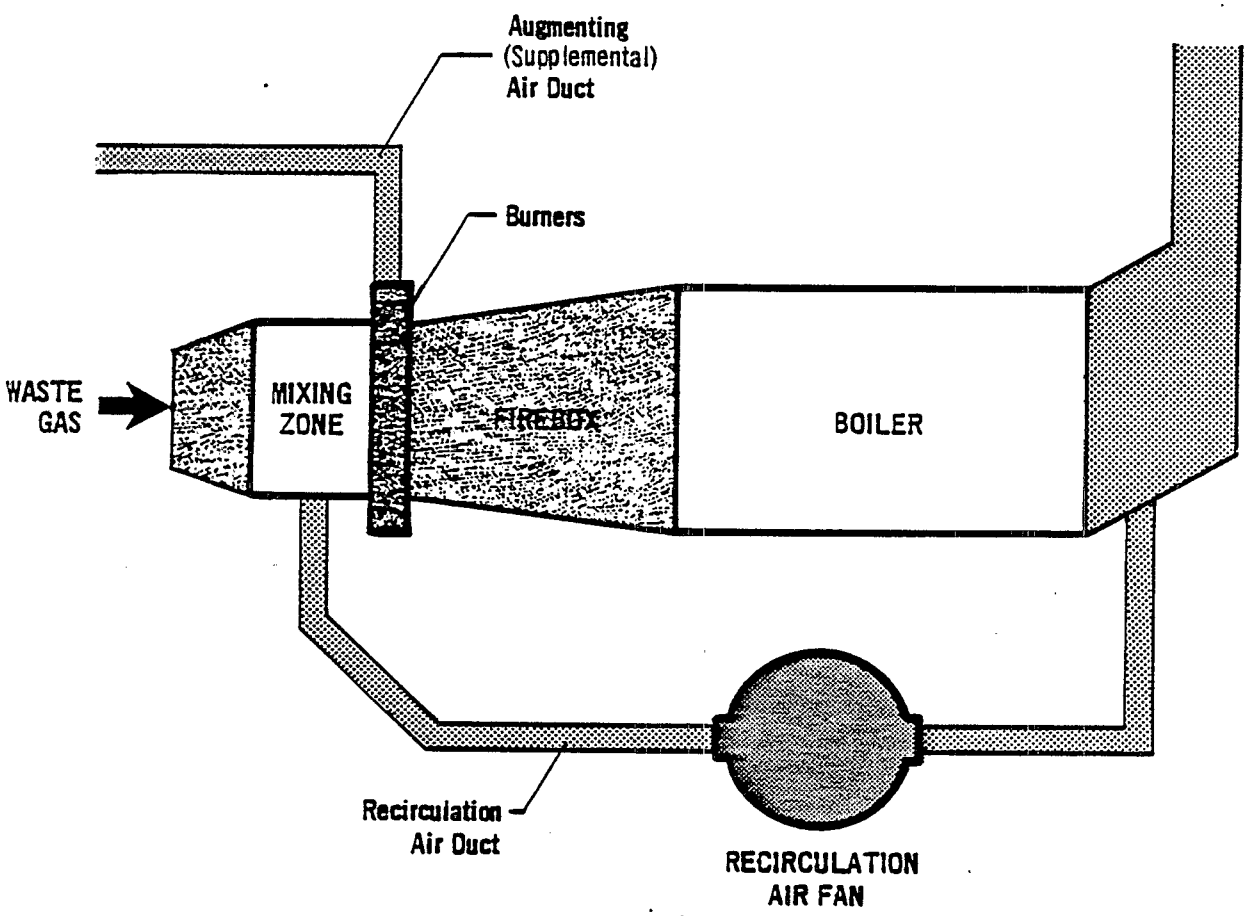


Figure C-4. Petro-Tex oxo unit incinerator.

2. Sampling and Analytical Techniques. 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. Test Results. 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.

November 1977

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 about 90 cm^3/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.

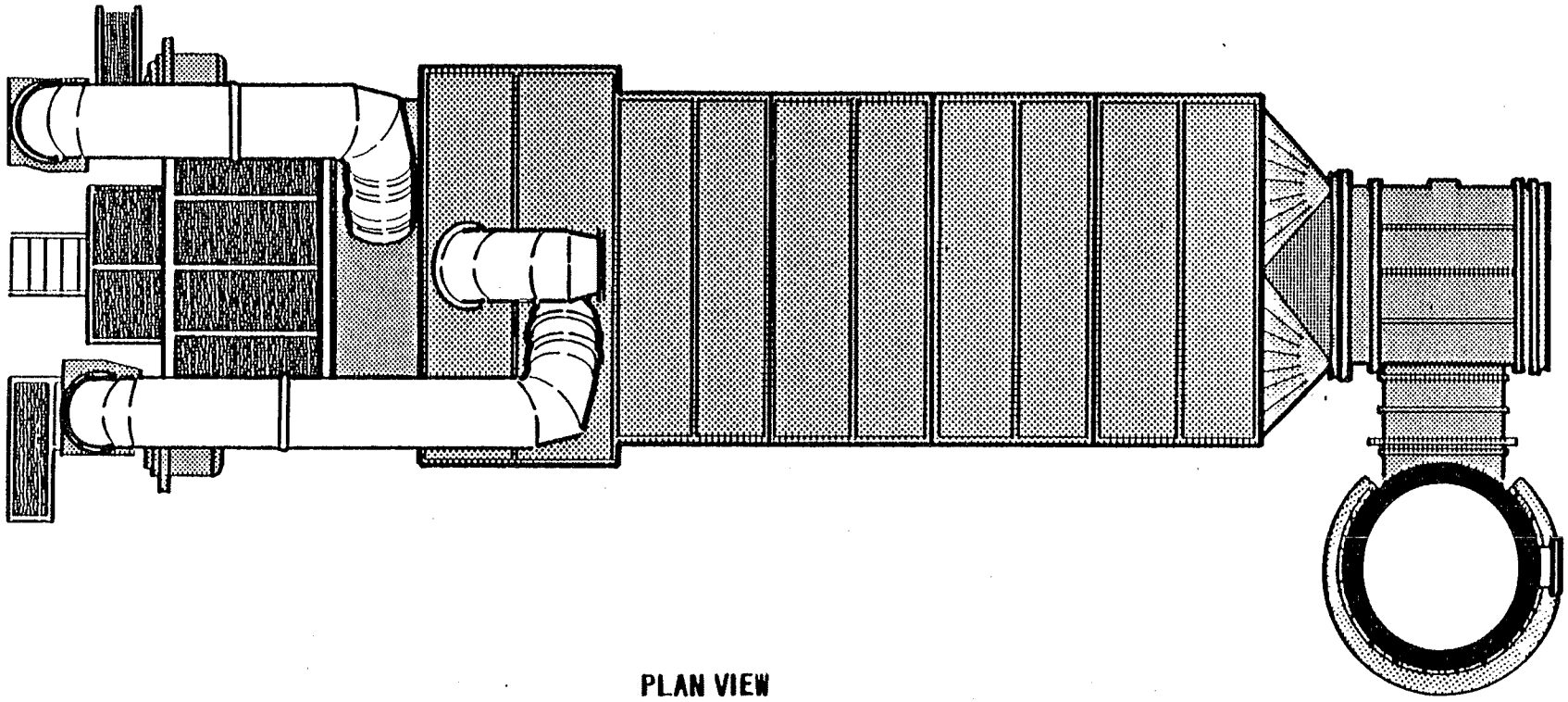
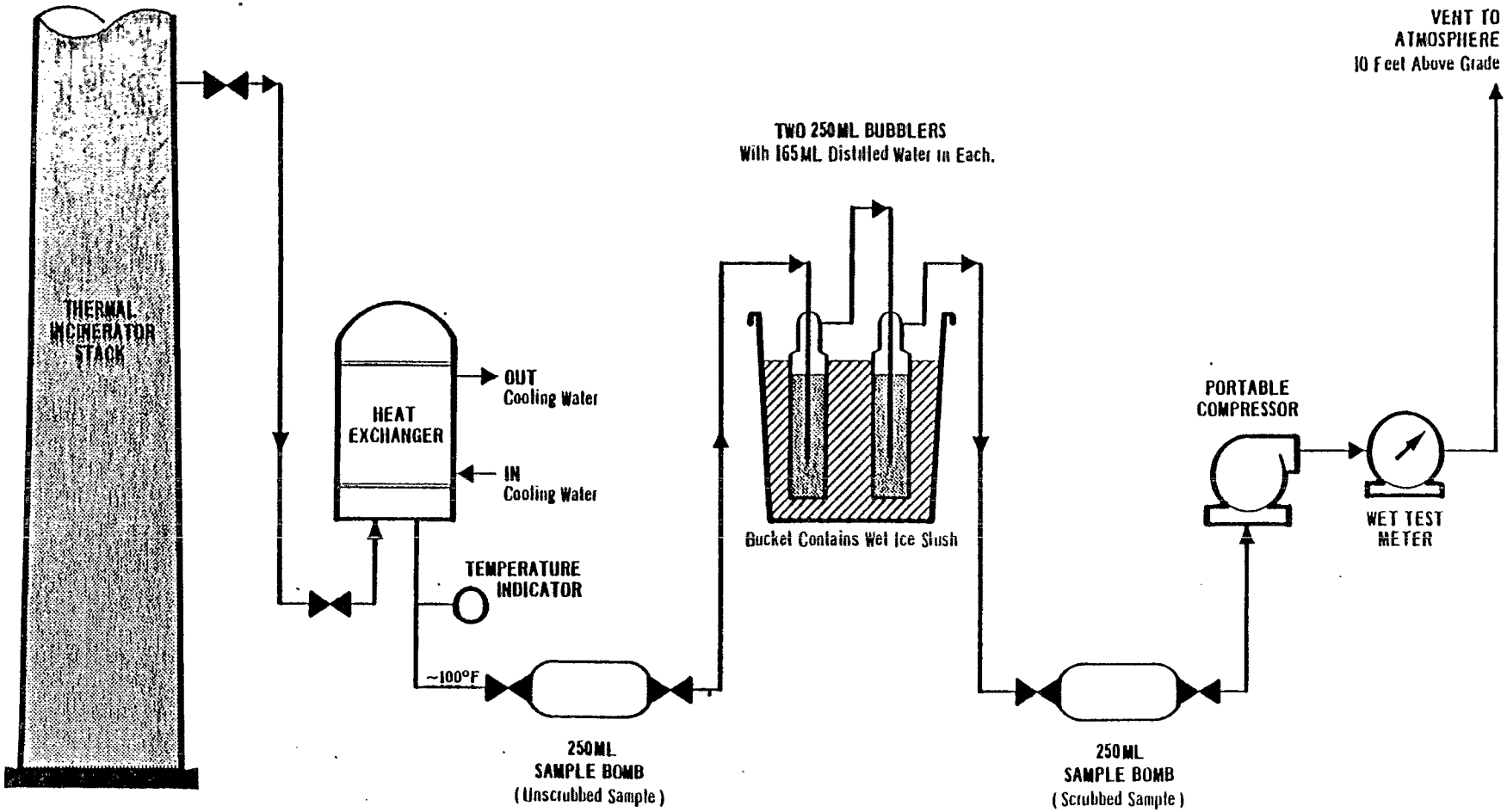


Figure C-5. Off-gas incinerator, Monsanto Co., Chocolate Bayou Plant.



NOTE: From Exchanger Process Outlet, All Lines are Vacuum Tubing.

Figure C-6. Thermal incinerator stack sampling system.

3. Test Results. 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 Union Carbide Lab-Scale Test Data[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/m^3 (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]			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	50	157	
Vapor Service	11[d]	42[d]	130[d]	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
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	53.8	

FOOTNOTES FOR TABLE C-9

[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, 31; B, 11; C, 33.

[e] All open-ended lines are 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

Equipment Component[a]	Polymers and Resins Plant Data by Type of Plant							
	A:PP-LP	B:PP-G	C:LDPE-HP	D:LDPE-LP	E:HDPE-GP	F:HDPE-GP	G:HDPE-LP	H:HDPE-LP
Pump Seals		24[b]			7	1[b]	2[b]	6[b]
Light Liquid Service								
Heavy Liquid Service								
Valves		3,000			490[c]	-	293	2,313[c]
Vapor Service							78	
Light Liquid Service							215	
Heavy Liquid Service								
Safety/relief valves					19[d,e]	-	7[d,e]	
Vapor Service							6[g]	
Light Liquid Service							1	
Heavy Liquid Service								
Open-ended lines					-	-		
Vapor Service								
Light Liquid Service								
Heavy Liquid Service								
Compressor seals		7			3	2	3	3
Sampling connections					-	-	3	
Flanges					485	-	695	
Total Estimated Emissions (kg/hr)[h]								
- baseline:	24.0[i]	19.6	22.4[j]	20.1[k]	5.1	0.5	3.5	15.9
- uncontrolled:	25.1	19.6	34.8	28.7	6.5	0.5	4.0	16.4

[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 11 safety/relief valves.

[f] All open-ended lines are 100 percent controlled at baseline.

FUGITIVE VOC EMISSION SOURCES IN POLYMERS AND RESINS PLANTS

Polymers and Resins Plant Data by Type of Plant								Assumed For Polymers And Resins Analysis
I:HDPE-LP	J:HDPE-LP	K:PS-Cont	L:EPS-PI	M:EPS-PI	N:EPS-IS	O:PET-DMT	P:PET-TPA	
66[b]	67[b]	22[b]	4[b]	3[b]	10[b]	100[b]	11[b]	
								29
								30
2,010	-	100[c]	8	13	175	1,500[c]	43[c]	1,450
								402
								524
								524
	270[d,e]	-	-	2[d]	-		42	42[e]
					unknown	250[f]	-	415[f]
8	3	0	0	0	0	0	0	2
		14[g]				100[g]	-	104[g]
	2,000	100				4,000	-	2,400
16.3	5.7	1.8	0.24	0.43	1.47	25.2	0.82	12.1
16.3	5.7	2.7	0.24	0.43	1.47	47.8	0.82	17.2

[g] Seventy-five percent of sampling connections are assumed to be controlled at baseline; therefore, the baseline emissions estimates are based on 26 sampling counts.

[h] Calculated using SOCMI emission factors for each equipment component given in Table C-10.

[i] Based on a total equipment count of about 9,400.

[j] Based on a total equipment count of about 8,200.

[k] Based on a total equipment count of about 10,200.

C.6 REFERENCES FOR APPENDIX C

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*References can be located in Docket Number A-82-19 at the U.S. Environmental Protection Agency Library, Waterside Mall, Washington, D.C.

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 disadvantage 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.

I-D.3 REFERENCES

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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. However, where soaping is possible, it can

be used as a preliminary screening technique. Soap solution would be applied to the potential leak surfaces and if no bubbles are observed, the source is presumed not to be leaking.

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.

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.

II-D.4 REFERENCES

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11. Blacksmith, J.R., et al. "Frequency of Leak Occurrence for Fittings in Synthetic Organic Chemical Plant Process Units" EPA Report No. 600/2-81-003. September 1980. EPA/IERL, RTP, NC. Docket Reference Number II-A-14.*

12. "Emission Test Report: Sun Petroleum Products Co., Toledo, OH," EMB Report No. 78-OCM-12B, October 1980. Docket Reference Number II-A-15.*

13. "Emission Test Report: Union Carbide Corp., Torrance, CA." EMB Report No. 78-OCM-12A, November 1980. Docket Reference Number II-A-17.*

*References can be located in Docket Number A-82-19 at the U.S. Environmental Protection Agency Library, Waterside Mall, Washington, D.C.

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, ethylene glycol recovery systems, 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; E.6, ethylene glycol recovery systems; and E.7, piping and ducting. Sufficient detail was presented in Chapter 8 for 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).

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 for a state-of-the-art exit velocity of one-half sonic velocity (0.5 Mach). 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. For flares handling very small flows in a minimum-available-size flare, the cost of supplying sufficient purge gas can be

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.	Auxiliary natural gas flowrate, Q_{ng}(scfm)	0, if $LHV_{wg} \geq 115$; $\frac{(115-LHV_{wg})Q_{wg}}{(815)}$, 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 $T_{wg} \approx 70$; $\frac{[Q_{wg} \times (MW_{wg}) \times T_{wg}] + [(Q_{ng}(1218)]}{(Q_{wg} \times MW_{wg}) + (17.4 \times Q_{ng})}$, if $Q_{ng} > 0$
9.	Molecular weight of flare gas, $MW_{fl.g}$ (lb/lb-mole)	$MW_{w.g.}$, if $Q_{ng} = 0$ $= \frac{(Q_{wg} \times MW_{wg}) + (17.4 \times Q_{ng})}{Q_{wg} + Q_{ng}}$, 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]	$(5.766 \times 10^{-3}) \times Q_{fl.g} \times (T_{fl.g.} + 460) / D^2$
13.	Flare tip pressure drop, Δp (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)	Select next larger multiple of 5 with a minimum of 30.
17.	Safe pipe length, L (ft)[j]	$\sqrt{[(4.095 \times 10^{-3}) \times Q_{fl.g} \times LHV_{fl.g}] - H^2}$

Footnotes for Table E-1

[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

$$(Q_{ng} \times LHV_{ng}) + (Q_{wg} \times LHV_{wg}) = (Q_{ng} + Q_{wg}) \times (115\ Btu/scf)$$

$$Q_{ng}\ (scfm) = \left[\frac{115 - LHV_{wg}\ (Btu/scf)}{LHV_{ng}\ (Btu/scf) - 115}\right] \times Q_{wg}\ (scfm)$$

[c] Temperature of mixture approximated by assuming uniform specific heats per unit mass.

[d] From Enviroscience (Reference 1, Appendix A)

$$D(in) = \left[\frac{(2.72 \times 10^{-3}) \times Q_{fl.g}\ (lb/hr) \times \sqrt{(T_{wg}\ (°F) + 460)/MW_{fl.g}}}{\sqrt{\Delta p\ (in.\ H_2O)}}\right]^{\frac{1}{2}}$$

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.

$$V_e(fps) = 0.5 \times c = 125 \times \sqrt{\frac{(T_{wg}\ (°F) + 460)}{MW_{fl.g}}}$$

From the Environscience equation for flame angle,

$$\Delta p\ (in.\ H_2O) = 55\ [V_e\ (fps)/550]^2 = (1.818 \times 10^{-4})\ V_e^{\ 2}$$

Substituting,

$$D\ (in.) = \left[\frac{(2.72 \times 10^{-3}) \times Q_{fl.g}\ (scf/min) \times MW_{fl.g}\ (lb/lb\ mole) \times (60\ min/hr) \times (V_e/125)}{(387\ scf\ at\ 70°F/lb\text{-}mole) \times \sqrt{(1.818 \times 10^{-4})\ V_e^{\ 2}}}\right]^{\frac{1}{2}}$$

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)

$$= Q\ (scfm) \times \frac{(T_{fl.g}\ (°F) + 460)\ acf}{(70°F + 460)\ \ scf} \times \frac{min}{60\ sec} \div \frac{\pi}{4} \times [D(in.)]^2 \times \frac{ft^2}{144\ in^2}$$

Footnotes for Table E-1 (concluded)

[g] From Δp (in. H_2O) = 55 $[V_e$ (fps)/550$]^2$, (as in c.)

[h] From Enviroscience (Reference 1, Appendix A) assuming wind velocity, V_w, of 60 mph

$$\theta = TAN^{-1}\left[\frac{1.47\ (fps/mph) \times V_w\ (mph)}{550\sqrt{\Delta p/55}}\right] = TAN^{-1}\left[\frac{1.47\ (60)}{V_e}\right]$$

[i] From Enviroscience (Reference 1, Appendix A), assuming flame emissivity, ϵ, of 0.12, flame radiation intensity, I, of 1200 Btu/hr ft^2

$$H = \sqrt{\frac{Q_{fl.g}\ (lb/hr) \times LHV_{fl.g}\ (Btu/lb) \times \epsilon}{(12.56)\ I}} \cdot \quad -3.33 \times D \times (\Delta p/55) \times \cos\theta$$

$$= \sqrt{\frac{[Q_{fl.g}(scf/min) \times LHV_{fl.g}\ (Btu/scf) \times 60\ min/hr \times 0.12]}{4\pi\ (1200\ Btu/hr - ft^2)}} \quad -3.33 \times D \times (V_e/550) \cos\theta$$

[j] Safe 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 9):,

$$L = \sqrt{(r^2 - H^2)}$$

$$= \sqrt{\frac{[Q_{fl.g}\ (scfm) \times LHV_{fl.g}\ (Btu/scf) \times 60\ min/hr \times 0.12]}{12.56\ (140\ Btu/hr\text{-}ft^2)} - H^2}$$

greater than the cost of a fluidic seal on the flare tip to prevent air intrusion and subsequent flash back. In such cases, a fluidic seal, which requires a greatly reduced purge rate, was used.

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 (i.e., the maximum continuous rate or the intermittent stream if no continuous flow was present).[10] Availability and deliverability of this quantity of steam was assumed. For flares handling large intermittent flows, steam requirements were approximated (see Figure E-1) for a small continuous "bleed" rate that is necessary to prevent shortened steam nozzle equipment life (due to steam condensate contacting the high temperature metal.)[11,12]

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,[13] 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.7.

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

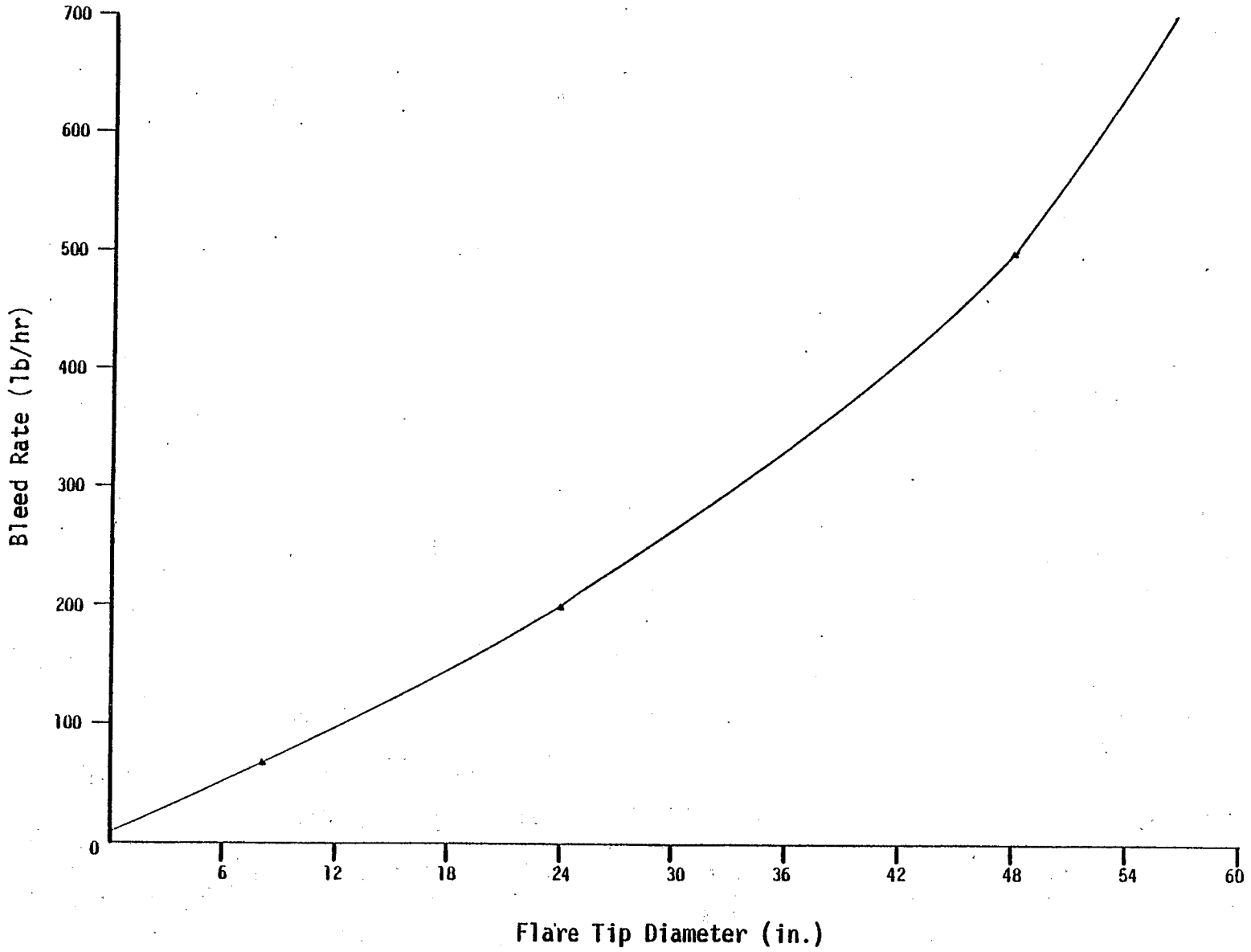

Figure E-1. Interpolation of Steam Bleed Rates for Intermittent Flares

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]Prices 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.

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.) Purchase costs of large diameter, 40-ft. high flares were approximated using a curve developed from the NAO data (see Figure E-2). Purchase costs for fluidic seals were approximated using a curve based on data provided by NAO[10] (See Figure E-3). 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 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).[14] 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

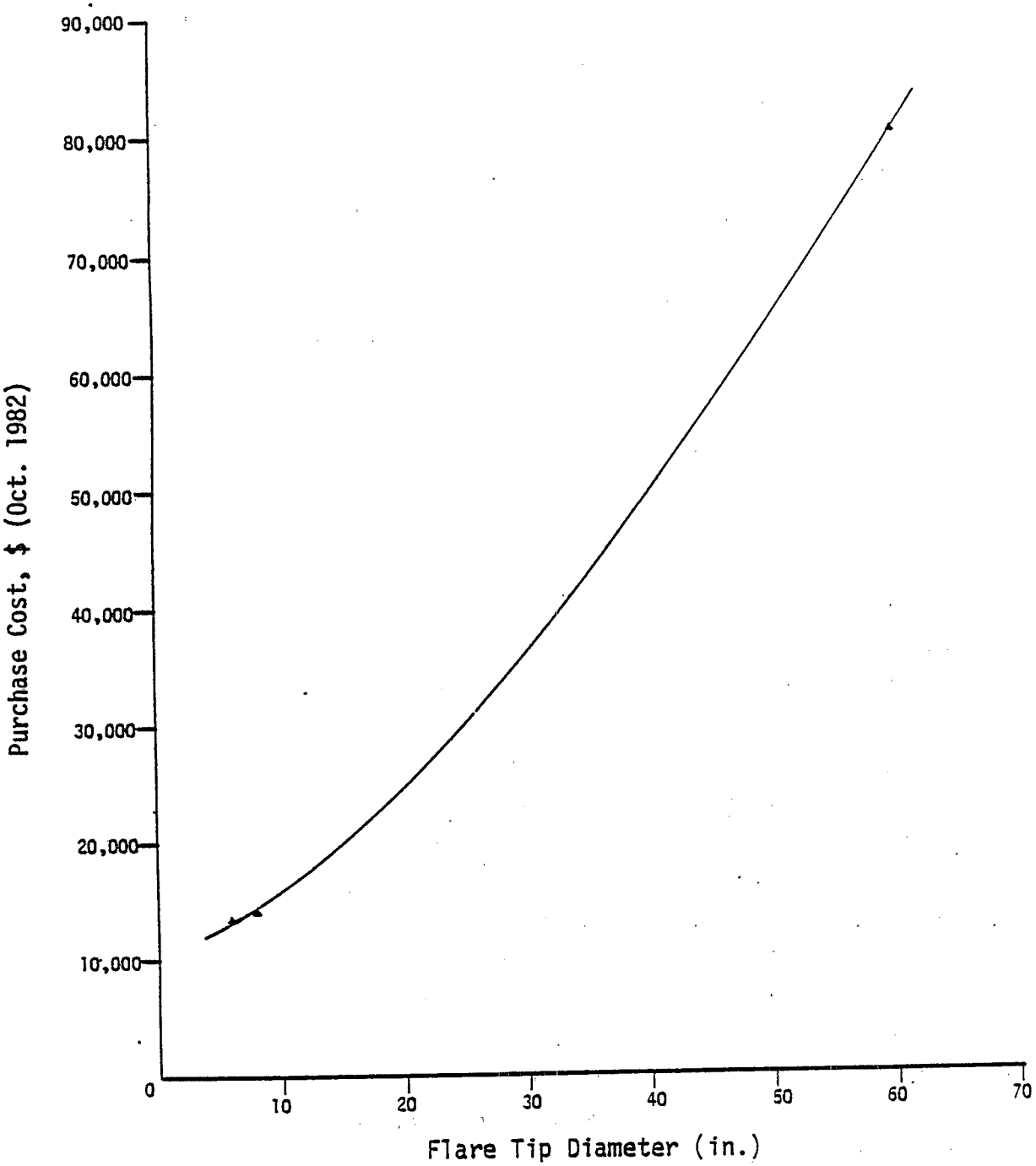

Figure E-2. Estimated Flare Purchase Cost for 40 ft Height

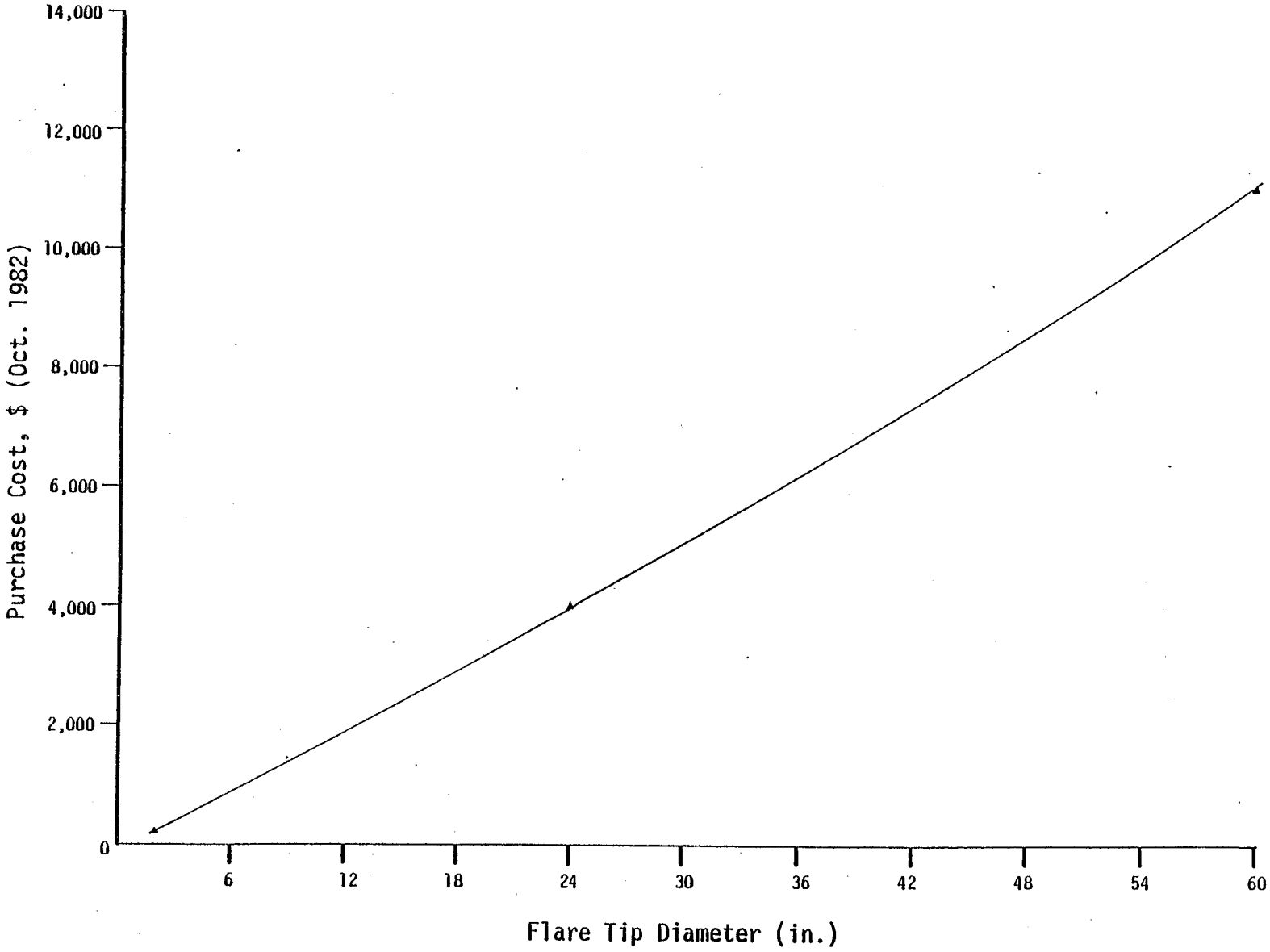

Figure E-3. Approximate Fluidic Seal Costs

Table E-3. CAPITAL AND ANNUAL OPERATING COST ESTIMATION PROCEDURE FOR STATE-OF-THE-ART STEAM-ASSISTED SMOKELESS FLARES

Item	Value
Capital Costs	
Flare purchase cost, C'''_{fl} (Oct. 1982 $)	Select from Table E-2 if value given or use equations: $(3905.7) + (35.054)\ H \times D + (900.36)\ D - (126.08)D^2$, for $20 \leq H \leq 40$ ft and $D \leq 8$ in. $(6275.6) + (224.10)\ H + (12.782)\ H \times D + (24.856)D^2$, for $50 \leq H \leq 200$ ft. or from Figure E-2 if H = 40 ft and D > 8 in.
Fluidic seal purchase cost, $C'''_{fl.s.}$ (Oct. 1982 $)	See footnote a.
Flare system purchase cost, C''_{fl}	$C'''_{fl} + C'''_{fl.s.}$
Flare installed cost, C'^{b}_{fl} (Oct. 1982 $)	$C''_{fl} \times 2.1$
Total installed piping costs, C'_p (Aug. 1978 $)	Method of Appendix E.7
Total installed ducting costs, C'_d (Dec. 1977 $)	Method of Appendix E.7
June 1980 Installed costs	
Piping[c], C_p	$C'_p \times 1.206$
Ducting[d], C_d	$C'_d \times 1.288$
Flare[e], C_{fl}	$C'_{fl} \times 0.818$
Total flare system cost, C_{sys}	$(C_p + C_d + C_{fl})$

Table E-3. CAPITAL AND ANNUAL OPERATING COST ESTIMATION PROCEDURE FOR STATE-OF-THE-ART STEAM-ASSISTED SMOKELESS FLARES (Concluded)

Item	Value
Annualized Costs[f]	
Operating labor, C_l	620 hr/yr x \$18/hr = \$11,160
Maintenance, C_m	0.05 x C_{sys}
Utilities	
$(Q_{n.g.}^{scfh})$ pilot[g]	80 scfh, for $2 \leq D \leq 8$; 160 scfh, for $10 \leq D \leq 20$; 240 scfh, for $D = 24$; 320 scfh, for $D = 60$.
$(Q_{n.g.}^{scfh})$purge[h]	$[(0.3272)(D \text{ in})^2 - (Q_{fl.g}^{scfm})_{cont}] \times 60$
Cost natural gas, $C_{n.g.}$[i]	$0.3661 \sum_i [(Q_{n.g.}^{scfm})_i \times t_i]$ aux. + $53.45 [(Q_{n.g.}^{scfh})_{pilot} +$ $(Q_{n.g.}^{scfh})_{purge})]$
Cost steam, C_{stm}	See footnote j
Taxes, admin. & insurance, C_{tax}	$C_{sys} \times 0.04$
Capital recovery, C_{cr}	$0.1315\ C_{fl} + 0.1627\ (C_p + C_d)$
Total annualized, C_{tot}	$C_l + C_m + C_{n.g.} + C_{stm} + C_{cr} + C_{tax}$

Footnotes for Table E-3

[a]Fluidic seal is costed only if cost of purge gas without seal is greater than the annualized cost of the seal plus any purge gas required with the seal, i.e., taking the October 1982 purchase cost of a seal, $C_{fl.s.}$ from Figure E-3 for D, if

$$53.45 \frac{\$/yr}{scfh} \times 60 \frac{min}{hr} \left\{ \left(0.372 \frac{scfm\ @\ 1\ fps}{in^2}\right) \left(D(in.)\right)^2 - \left(Q^{scfm}_{fl.g.}\right)_{cont.} \right\}$$

$$> \left[(0.1315 + 0.05 + 0.04) \frac{\$/yr}{\$\ capital} \right.$$

$$\left. \times\ 0.818 \frac{Jun.\ '80\$}{Oct.\ '80\$} \times 2.1 \frac{installed}{purchase} \times C_{fl.s.} \right]$$

$$+ 53.45 \frac{\$/yr}{scfh} \left\{ \left(0.45 \frac{scfh}{in.^2}\right) \times \left[D(in.)\right]^2 - \left(Q^{scfm}_{fl.g.}\right)_{cont.} \right\}$$

or, simplifying,

$$\text{if} \quad 1169\left[D(in.)\right]^2 - 3154 \left(Q^{scfm}_{fl.g.}\right)_{cont.} > (0.3805 \times C_{fl.s.})$$

$$\text{then} \quad C'''_{fl.s.} = C_{fl.s.}$$

$$\left\{ \text{Note: This condition will be in error to the degree that } \left[0.45\ D^2 - \left(Q^{scfm}_{fl.g.}\right)_{cont.}\right] < 0 \right\}$$

$$\text{Otherwise,} \quad C'''_{fl.s.} = 0$$

[b]For installation cost factor breakdown, see Table 8-2.

[c]Updated using Chemical Engineering Plant Cost pipes, valves and fittings index from August 1978 (273.1) to June 1980 (329.3).

[d]Updated using Chemical Engineering Plant Cost fabricated equipment index from December 1977 (226.2) to June 1980 (291.3).

[e]Adjusted using Chemical Engineering Plant Cost Index from October 1982 (317 estimated) to June 1980 (259.2).

[f]For annualized cost factors, see Table 8-3.

[g]Based on vendor information for pilots without energy conservation (Reference 9).

[h]Ensures continuous flow of at least 1 fps for flare with any continuous flow not using fluidic seal:

$$\frac{\pi}{4} \left\{ (D(in.))^2 \times \frac{ft^2}{144\ in^2} \times (1\ fps) \times (60\ sec/min) \right.$$

$$\left. - \left[Q_{fl.g.}\ (scfm)\right]_{cont.} \right\} 60\ min/hr$$

Footnotes for Table E-3 (Concluded)

Ensures sufficient continuous flow per vendor information (See Docket Item No. II-E-20) for flare with any continuous flow using a fluidic seal:

$$0.45 \ \frac{\text{scfh}}{\text{in}^2} \times [D(\text{in.})]^2$$

$$^{i}\left\{\sum_{i}[Q_{n.g}\ (\text{scfm})_i \times 60\ \text{min/hr} \times (t_i \text{ operating hours per year at stream combination } i) + [Q_{pilot}(\text{scfh}) + Q_{purge}(\text{scfh})] \times 8760\ \text{hr/yr}\right\}$$

$$\times \frac{520°\text{R scf at } 60°\text{F}}{530°\text{F scf at } 70°\text{R}} \times \frac{1{,}040\ \text{Btu (HHV)}}{\text{scf at } 60°\text{F}} \times \frac{\$5.98}{(10^6\ \text{Btu (HHV)}} \times \frac{(10^6\ \text{Btu})}{10^6\ \text{BTU}}$$

[j]For continuous streams only, assumes steam at 0.4 lb/lb of hydrocarbon at maximum continuous flaring rate for 8600 hr/yr:

$$Q_{cont.}\ (\text{scfm}) \times MW_{cont.} \times \left(\frac{\text{wt.\% HC}}{100}\right)_{cont.} \times 8600\ \text{hr/yr} \times 60\ \text{min/hr} \times (\text{lb-mole/387 scf at } 70°\text{F}) \times (0.4\ \text{lb steam/lb HC}) \times \frac{(1000\ \text{lb steam})}{1000\ (\text{lb steam})} \times \$6.18/(1000\ \text{lb steam})$$

or simplifying,

$$3.296\left[Q_{w.g.}(\text{scfm}) \times MW \times \frac{\text{wt.\% HC}}{100}\right]_{\substack{\text{cont.}\\ \text{fl.g.}}}$$

For intermittent streams only and for combinations of continuous and intermittent streams, determine approximate steam bleed rate (SBR) for flare diameter, D, using Figure E-1 (which extrapolates to about 800 lb/hr at 60 in. and 925 lb/hr at 66 in.) annual cost for intermittent flow duration, t_{int}:

$$\text{SBR, lb steam/hr} \times \frac{(1000\ \text{lb steam})}{1000\ (\text{lb steam})} \times \frac{\$6.18}{(1000\ \text{lb steam})} \times \left(8600 - \sum t_i\right)\text{hr/yr} + \frac{3.296}{8600}\sum_{i}\left[t_i \times MW \times \frac{\text{wt.\% HC}}{100} \times Q\right]_i$$

where i indicates those stream combinations for which the design hourly steam cost (based on waste gas only since natural gas does not require steam for smokeless combustion) exceeds the hourly steam bleed cost, i.e.,

$$\left[\frac{t_i}{8600} \times 3.296\ \left(MW \times \frac{\text{wt.\% HC}}{100} \times Q\right)_i\right] - (0.00618 \times \text{SBR}) > 0.$$

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 would be 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)[15,16] and an oxygen concentration of 12 percent or greater by volume[17] that required preheating to maintain a combustion temperature of 870°C (1600°F). Dilution air would be 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 additional combustion air,[18,19] 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.[20,21] Minimum auxiliary fuel requirements for low heating content streams would be set at 5 Btu/scf to ensure flame stability.[22]

The design procedure for streams, such as the combined streams costed for the polymers and resins industry, 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

Table E-4. WORKSHEET FOR CALCULATION OF WASTE GAS CHARACTERISTICS (molecular weight, molecular formula, lower heating value in Btu/scf)

I.D.: _____ emission factor: em.f. = _____ kg VOC/Mg product; model plant capacity: CAP = _____ Gg/yr

lb/100 $lb_{w.g.}$ (= wt. % x 100)	Compound Name	Chemical Formula[a]	Molecular Weight (MW)	lb-moles/100 lb (lb/100 lb / MW) $_{w.g.}$	lb-molesC/100 lb (= A x $_{w.g.}$ lb-moles/(100 $lb_{w.g.}$)	lb-molesH/100 lb (= B x $_{w.g.}$ lb-moles/100 $lb_{w.g.}$)	lb-molesN/100 lb (= C x $_{w.g.}$ lb-moles/100 $lb_{w.g.}$)	lb-molesO/100 lb (= D x $_{w.g.}$ lb-moles/100 $lb_{.g.}$)	Lower Heating Value (LHV) in Btu/lb[b]	Btu/$lb_{w.g.}$ (= lb/100 $lb_{w.g.}$) x LHV x (100 $lb_{w.g.}$ / 100($lb_{w.g.}$)) x
	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=	0	
	ethane	C_2H_6	30.07		x2 =	x6 =			20,416	
	methyl alcohol (methanol)	CH_3OH	32.04		x1 =	x4 =		x1 =	9,080[c]	
	C_2-C_5 hydrocarbons	$(C_{2.5}H_5)$	35(ave)			x2.5=	x5 =		20,000[d]	
	C_2-C_5 hydrocarbons	$(C_{2.7}H_{6.4})$	39(ave)			x2.7=	x6.4=		19,900[d]	
	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 =	13,500[c]	
	C_3-C_8 hydrocarbons	(C_5H_{12})	72(ave)			x5 =	x12 =		19,400[d]	

Table E-4. WORKSHEET FOR CALCULATION OF WASTE GAS CHARACTERISTICS (molecular weight, molecular formula, lower heating value in Btu/scf) (Concluded)

I.D.: _____ emission factor: em.f. = _____ kg VOC/Mg product; model plant capacity: CAP = _____ Gg/yr

lb/100 $lb_{w.g.}$ (= wt. % x 100)	Compound Name	Chemical Formula[a]	Molecular Weight (MW)	lb-moles/100 $lb_{w.g.}$ (lb/100 $lb_{w.g.}$ ÷ MW)	lb-molesC/100 $lb_{w.g.}$ (= A x lb-moles/100 $lb_{w.g.}$)	lb-molesH/100 $lb_{w.g.}$ (= B x lb-moles/100 $lb_{w.g.}$)	lb-molesN/100 $lb_{w.g.}$ (= C x lb-moles/100 $lb_{w.g.}$)	lb-molesO/100 $lb_{w.g.}$ (= D x lb-moles/100 $lb_{w.g.}$)	Lower Heating Value (LHV) in Btu/lb[b]	Btu/$lb_{w.g.}$ (= lb/100 $lb_{w.g.}$) x LHV x (100 $lb_{w.g.}$ / 100($lb_{w.g.}$))
	cyclohexane	C_6H_{12}	84.16		x6 =	x12 =			18,846	
	n-hexane	C_6H_{14}	86.17		x6 =	x14 =			19,391	
	C_{10}-C_{12} hydrocarbons	($C_{10.3}H_{20.4}$)	144(ave)		x10.3=	x20.4=			19,000[e]	

wt.% VOC = _____ (hydrocarbons except methane and ethane) $\sum$= lb-moles/100 $lb_{w.g.}$: _____ C: _____ H: _____ N: _____ O: _____ $\sum$= Btu/lb w.g. = $LHV_{w.g.}$ = _____ Btu/lb

$Q_{w.g.}$ = (25.64) x em.f. x CAP / wt % VOC nonmethane or eth ne = _____ lb/hr

$H.W._{w.g.}$ = 100/$\sum$ = _____

$Q^{scfm}_{w.g.}$ = $Q_{w.g.}$ x 5.98/$MW_{w.g.}$ = _____ scfm@32°F $T^{in}_{w.g.}$ = _____ °F

$LHV^{scf}_{w.g.}$ = $\sum$ x $MW_{w.g.}$ (lb/lb-mole / 359scf/lb-mole = _____ Btu/scf@32°F

[a] Parentheses indicate an approximate formula for a given moleuclar 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.

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 N & O compositions);

(4) oxygen in waste gas (in air or hydrocarbons) thoroughly mixed with VOC so only stoichiometric requirement.

Reaction (in dry air):

$$C_c H_h N_n O_o + (1.18)\left(C + \frac{H}{4} - \frac{O}{2}\right)(O_2 + 3.76\ N_2) \longrightarrow C(CO_2) + \frac{H}{2}(H_2O) + \frac{N}{2} + \left(C + \frac{H}{4} - \frac{O}{2}\right)(1.18)(3.76)\ X(N_2) + (0.18)\left(C + \frac{H}{4} - \frac{O}{2}\right)(O_2)$$

Products of Combustion in lb/100 lb w.g. (wet):

CO_2: $\frac{44.01\ lb}{lb\text{-mole}} \times C\ \left(\frac{lb\text{-mole}\ CO_2}{100\ lb\ w.g.}\right) = (44.01)\ C =$ _____

H_2O: if air required i.e., $(C + \frac{H}{4}) > \frac{O}{2}$: $\frac{18.02\ lb}{lb\text{-mole}}\left(\frac{H}{2} + [AIR + \left(C + \frac{H}{4} - \frac{O}{2}\right)(4.76)(1.18)](0.013)\ \frac{lb\ H_2O}{lb\ dry\ air}^a\left(29\ \frac{lb\ dry\ air}{lb\text{-mole}}\right)\right.$

$= (2.118)\ C + (9.539)\ H - (1.059)\ O + (0.377)\ AIR =$ _____

if no air required i.e., $(C + \frac{H}{4}) \leq \frac{O}{2}$: $\frac{18.02\ lb}{lb\text{-mole}}\left(\frac{H}{2}\right) + (AIR)\ (0.013)\ \frac{lb\ H_2O}{lb\ dry\ air}^a\left(29\ \frac{lb\ dry\ air}{lb\text{-mole}}\right) = (9.01)\ H + (0.377)\ AIR =$ _____

N_2: if air required i.e., $(C + \frac{H}{4}) > \frac{O}{2}$: $\frac{28.02\ lb}{lb\text{-mole}}\left[\frac{N}{2} + \left(C + \frac{H}{4} - \frac{O}{2}\right)(3.76)(1.18)\right]$

$= (124.3)\ C + (31.08)\ H + (14.01)\ N - (62.16)\ O =$ _____

if no air required i.e., $(C + \frac{H}{4}) \leq \frac{O}{2}$: $\frac{28.02\ lb}{lb\text{-mole}}\left(\frac{N}{2}\right) = (14.01)\ N$

O_2: if air required i.e., $(C + \frac{H}{4}) > \frac{O}{2}$: $\frac{32.00\ lb}{lb\text{-mole}}(0.18)\left(C + \frac{H}{4} - \frac{O}{2}\right) = (5.760)\ C + (1.44)\ H - (2.88)\ O =$ _____

if no air required i.e., $(C + \frac{H}{4}) \leq \frac{O}{2}$: $\frac{32.00\ lb}{lb\text{-mole}}\left[\frac{O}{2} - \left(C + \frac{H}{4}\right)\right] = (-32.00)C - (8.00)\ H + (16.00)\ O =$ _____

_____

Total: $\frac{prod.}{w.g.} = \frac{lb\ products\ (wet)}{100\ lb\ w.g.} = \frac{lb(CO_2 + H_2O + N_2 + O_2)}{100\ lb\ w.g.} =$ _____

[a]Pounds 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)[b]	For LHV(Btu/scf) / (% aux)[b] LHV ≤ 75 / 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 1800°F (basis of cost estimates)[d]	$W_{air}(\frac{\text{lb air}}{100\text{ lb w.g.}}) = (0.198)\ (1 + \frac{\%aux}{100})LHV\ (Btu/lb)$ $-(1800 - T_{w.g.})\ [(5.95 \times 10^{-4})CO_2 + (1.14 \times 10^{-3})\ H_2O$ $+(5.92 \times 10^{-4})N_2 + (5.35 \times 10^{-4}\ O_2)]$ $+(1.65 \times 10^{-4})(\%aux) \times LHV\ (Btu/lb)$
(3) Calculate total flue gas volume: (a) waste gas products of combustion:	$Q_{w.g.\ prod.}\ (\text{lb-moles/100 lb w.g.}) = \frac{CO_2}{44.01} + \frac{H_2O}{18.02} + \frac{N_2}{28.02} + \frac{O_2}{32.00}$
(b) auxiliary natural gas products of combustion[e]:	$Q_{n.g.\ prod.}\ (\text{lb-moles/100 lb w.g.}) = (1.66 \times 10^{-7})(\%aux)(LHV\ w.g.)$
(c) dilution air:	$Q_{dil.\ air}\ (\text{lb-moles/100 lb w.g.}) = W_{air}\ (\frac{\text{lb air}}{100\text{ lb w.g.}}) \div 29\ \text{lb air/lb-mole air}$
(d) total flue gas:[f]	$Q_{f.g.}\ (scfm) = (5.98 \times 10^{-2})\ (Q_{w.g.\ prod.} + Q_{n.g.\ prod.} + Q_{dil.\ air}) \times Q_{w.g.}\ (lb/hr)$
(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 21.

[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. 2 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:

$$90\ LHV_{w.g.}\ [1+(\%\ aux/100)] \leq \sum(W_i c_{pi})\ \text{waste gas products}\ (1800-T_{w.g.}) + (1800-60)\left\{\sum(W_i\ c_{pi})_{ng} \times (\%aux\ LHV_{w.g.})\ LHV_{n.g.} + [W(c_p)_{air} - 4.51(c_p)_0]\right\}$$

, 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 to 1800°F since $T_{w.g.}$ 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 (1800 + 460) °R/492°R x (1.05) design safety factor (Reference 24).

[h] If $V_{cc} < 35.7\ ft^3$ (minimum commercially available size[25]), must add air and redesign.

If $V_{cc} > 7238\ ft^3$ (maximum shop-assembled unit[26]), larger units would require field fabrication; therefore, assume multiple units of equal sizes.

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.[23]

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.[24] The design procedure assumed a minimum commercially available size of 1.01 m^3 (35.7 ft^3) based on vendor information[25] and a maximum shop-assembled unit size of 205 m^3 (7,238 ft^3).[26]

The design procedure could 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 heat contents to combust at greater than 870°C (1600°F), and even at greater than 980°C (1800°), 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.)

E.3.2 Thermal Incinerator Cost Estimation Procedure

Thermal incinerator purchase costs for the calculated combustion chamber volume were taken directly from Figure E-4, (Figure A-1 in the IT Enviroscience document, Reference 23). An installation cost factor of 4.0 (see Table 8-2) was used based on the Enviroscience document.[27] The installed cost used for the minimum commercially available size of 1.01 m^3 (35.7 ft^3) was $217,800 (June 1980; based on a December 1979 purchase cost of $52,000). The installed cost of one 150-ft. duct to the incinerator and its associated fan and stack were taken directly from Figure E-5 (Figure IV-15, curve 3 in the IT Enviroscience study[28]).

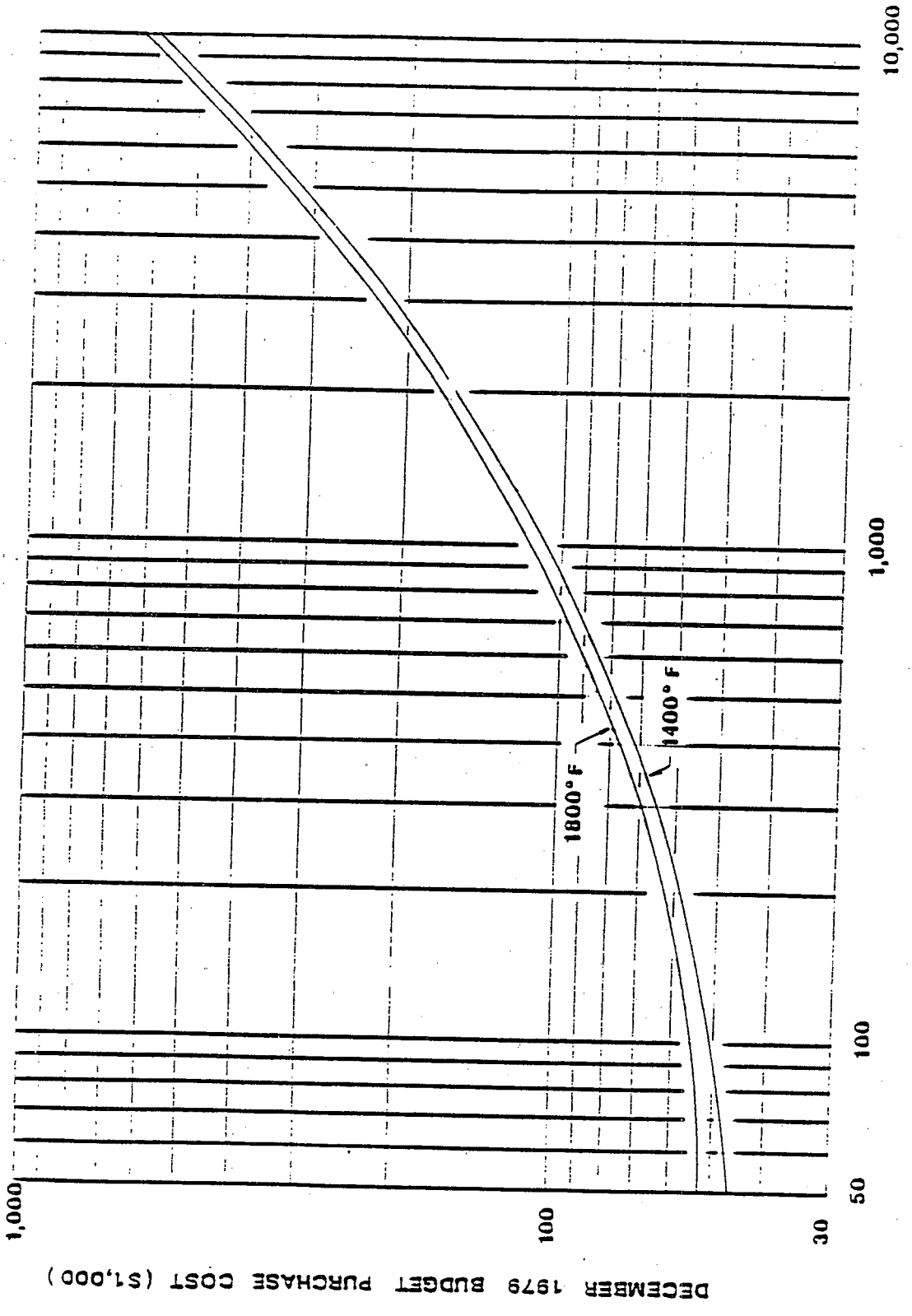

Figure E-4. Purchase Costs for Thermal Incinerator Combustion Chambers

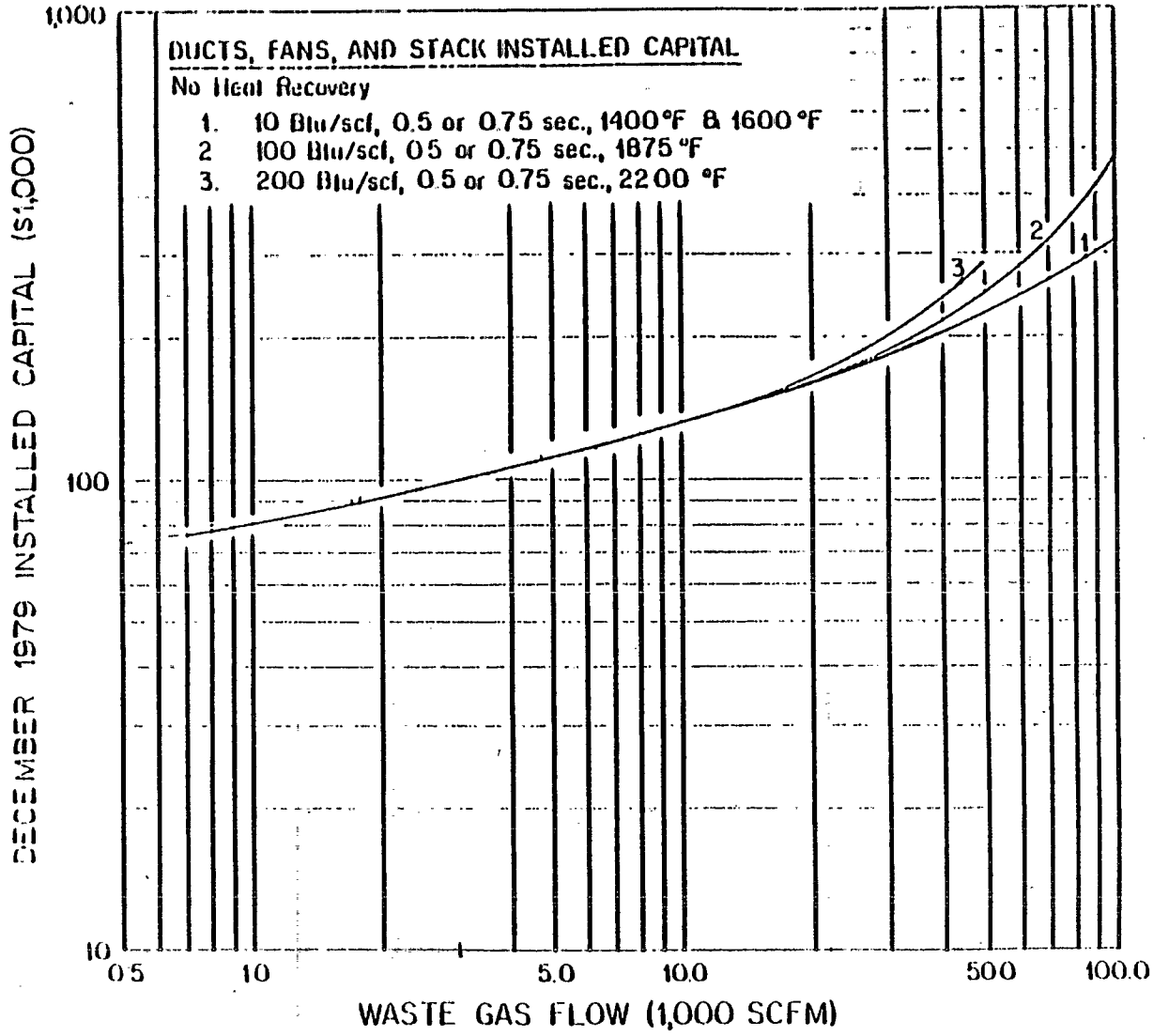

Figure E-5. Installed Capital Costs for Inlet Ducts, Waste Gas and Combustion Air Fans, and Stack for Thermal Incinerator Systems with no Heat Recovery

A minimum installed 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 noted 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 1.5kPa (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 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

TABLE E-7. CAPITAL AND ANNUAL OPERATING COST ESTIMATION PROCEDURE FOR THERMAL INCINERATORS WITHOUT HEAT RECOVERY

ITEM	VALUE
Capital Costs	
Combustion Chamber	
Purchase cost	from Figure E-4 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-5 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 $LHV_{w.g.}$ x $Q_{w.g.}$ (lb/hr)
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 for Table E-7

[a]Updated using Chemical Engineering Plant Cost Index from December 1979 (247.6) to June 1980 (259.2).

[b]Piping updated using Chemical Engineering Plant Cost pipes, valves, and fittings index from August 1978 (273.1) to June 1980 (329.3). Ducting updated using Chemical Engineering Plant Cost fabricated equipment index from December 1977 (226.2) to June 1980 (291.3).

[c]From Figure E-5 for no heat recovery from Enviroscience (Reference 28), 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 29). 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 30). 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 Chemical Engineering Plant Cost fabricated equipment index from December 1979 (273.7) to June 1980 (291.3).

[e]Cost factors presented in Table 8-3.

[f][(% aux) x $LHV_{w.g.}$/20,660 Btu/$lb_{n.g.}$](100 $lb_{n.g.}$/100 $lb_{w.g.}$) x $Q_{w.g.}$ (lb/hr) x (100 $lb_{w.g.}$)/100($lb_{w.g.}$) x (8600 hr/yr) x (lb-mole/17.4 $lb_{n.g.}$) x (379 scf at 60°F/lb-mole) x (1040 Btu(HHV)/scf at 60°F) x \$5.98/10^6 Btu (HHV) x (10^6 Btu)/10^6 (Btu).

[g]Electricity = (6 in. H_2O pressure drop) x $Q_{f.g.}$ (scfm) x (8600 hrs/yr) x (0.7457 kW/hp) x (5.204 lb/ft^2/in. H_2O) ÷ [(60 sec/min) x (550 ft-lb/sec/hp) x (0.6 kW blower/1 kW electric) x \$0.049/kWh].

[h]10 percent interest (before taxes) and 10 yr. life.

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.[31] 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 feasible heat exchanger efficiency. A maximum 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 Catalytic 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

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)] x VOC volume flow (Q_1), scfm, 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))] ÷ [Item (5) x (Item (9) - Item (6))][e]
(12)	PHR design efficiency, %	See footnote f

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] x (10^6 Btu)/10^6 Btu
(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): 11.45 for methane, 20.02 for ethane, 28.58 for propane, 54.31 for hexane, 17.15 for ethylene, and 45.73 for pentane. Values taken from p. 6-2 in Steam (Reference 18) for 100% total air and multiplied by 1.2 for 120% total air or 20% excess air.

[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 calculated maximum possible heat recovery up to an upper limit of 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 Chapters 3 and 6
Stream conditions	
Temperature, °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 1000 Mg/Gg (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) x (385 scf/lb-mole at 68°F)= 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, $W_1/100\%$)
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 ft^3 /lb mole) = 1.645 (EP/M_2W_1)
Air volume flow rate, scfm	(Total gas flow (Q_2), scfm) - (VOC volume flow (Q_1), scfm)
VOC concentration (A), % of LEL	(100) [(Volume flow of VOC, scfm) ÷ (Volume flow of air, scfm] ÷ LEL[d]
Heat content (H_2), Btu/total scf	From Chapter 6[e]

Footnotes for Table E-9

[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, butane, and pentane are 3.1, 1.32, 7.3, and 2.5, 1.9, and 1.4, respectively.

[e]Total gas heat content averages 50 Btu/scf at 100 percent LEL.

including bypass. Calculations for capital cost estimates are based on equipment purchase costs obtained from vendors [31,32,33] 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. A minimum available unit size of 500 scfm was assumed.[34,35] The installed cost of this minimum size unit (which can be used without addition of gas or air for stream flows greater than about 150 scfm[35]) was estimated to be $53,000 (June 1980). The heat exchangers for small size systems would be costly and may not be practical. Table 8-2 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 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-6.

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 E-10. CATALYTIC INCINERATOR VENDOR COST DATA[31,32]

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.

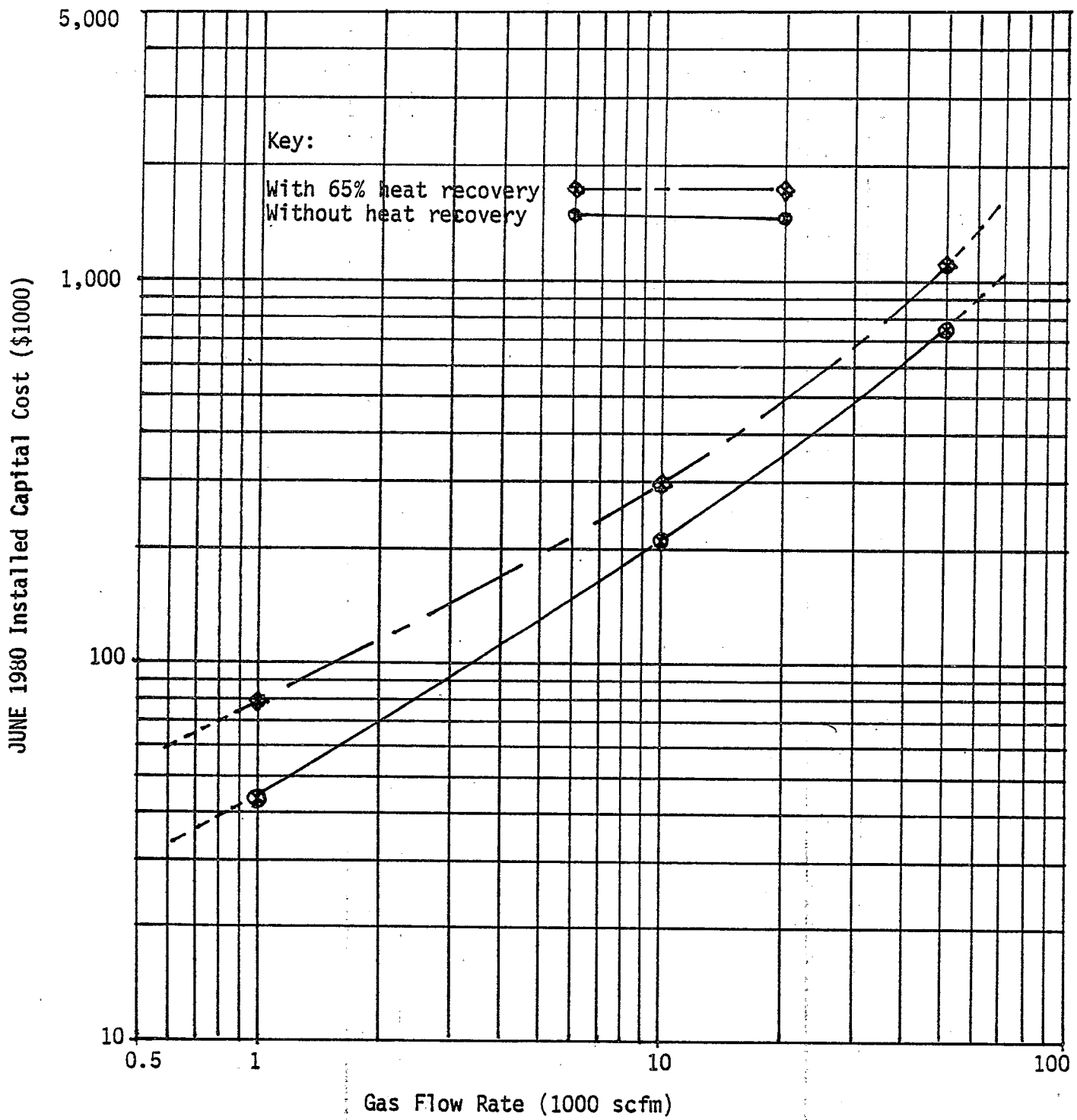

Figure E-6. Installed Capital Costs for Catalytic Incinerators With and Without Heat Recovery

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,[33] 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 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 per cubic foot per minute, 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 \$0.838/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 combined material recovery streams from the continuous process polystyrene model plant and from the DMT-process polyethylene (terephthalate) model plant. An outline of the design and costing of the condenser system for the polystyrene model plant is presented in this section as an example of the procedure used. Details of either application are given in the docket.[36]

Table E-11. CAPITAL AND OPERATIONG COST ESTIMATION FOR CATALYTIC INCINERATOR SYSTEMS

Item	Value
Capital Costs	
Incineration system Installed cost, June 1980	From Figure E-6
Piping & ducting (from sources to main incinerator duct)	
Installed cost	See Section E-7 for source flow rates, scfm.
Installed cost, June 1980[a]	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[b]	From Figure E-5 for waste gas flow (Q_2), scfm; use $70,000 minimum
Installed cost, June 1980[c]	Installed cost x 1.064
Total Installed Cost, June 1980	Sum of incineration systems, piping & ducting, and ducts, fans, & stacks
Annualized Costs	
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-6)
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, $10^3 ft^3$, Item 16 of Table E-8)[e]
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 recovery

Table E-11. CAPITAL AND OPERATIONG COST ESTIMATION FOR CATALYTIC INCINERATOR SYSTEMS (Concluded)

Item	Value
Indirect Costs	
Capital recovery	(0.1627) x (Total installed capital cost, $ from Figure E-6)
Taxes, insurance and administrative charges	(0.04) x (Total installed capital cost, administrative charges $ from Figure E-6)
Total Annualized Costs	Sum of total direct costs and total indirect costs

[a]Updated using Chemical Engineering Plant Cost Index from December 1979 (247.6) to June 1980 (259.2).

[b]Piping updated using Chemical Engineering Plant Cost pipes, valves, and fittings index from August 1978 (273.1) to June 1980 (329.3). Ducting updated using Chemical Engineering Plant Cost fabricated equipment index from December 1977 (226.2) to June 1980 (291.3).

[c]See footnote c, Table E-7 for discussion on application of these costs developed by Enviroscience (Reference 28).

[d]Updated using Chemical Engineering Plant Cost fabricated equipment index from December 1979 (273.7) to June 1980 (291.3).

[e]Total gas flow including waste gas and additional combustion air.

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 combination of the following two streams from the polystyrene model plant: the styrene condenser vent and the styrene recovery unit condenser vent. These streams may consist of styrene and steam, which are immiscible, or of styrene in air, a non-condensable. 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 condensation of styrene mixed with a non-condensable, such as air, can not be considered to be isothermal. Therefore, systems to condense in the presence of a non-condensable are usually designed by considering the heat exchanger to be divided into a number of sections, each with a portion of the total temperature drop. In general, the condensation of styrene in air is accomplished less readily, and thus more expensively, than the condensation of styrene in steam.

As new plants were assumed to use vacuum pumps, which result in styrene-in-air emissions, the costs of condensing styrene in air were estimated for regulatory alternatives of 90 percent through 98.5 percent styrene emission reduction. The condenser inlet stream was assumed to contain saturated styrene in air at 27°C (80°F) since most industry data showed the emission stream to be near that temperature, and since the stream was the outlet from previous process or emission control condensers that would emit at or near saturation. The following general procedure and assumptions were used in evaluating the condensation

systems for the combined streams containing styrene in air from the continuous polystyrene model plant. The presented procedure, however, is for the entire heat exchanger while the actual procedure[36] utilized an iterative, multiple section analysis.

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 in air. As complete vapor pressure data are not readily available, the condensation temperature is estimated by a regression equation of available data points[37] 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 and sensible heat contents of styrene and air (see Table E-12). The design requirements of the condensation system are then determined based on the heat load and stream characteristics. 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.

The styrene-in-air refrigerated condenser systems were designed according to procedures for calculating shell-side,[40] tube-side,[41] and condensation[42] heat transfer coefficients, mass transfer coefficients,[43] and, finally, an overall heat transfer coefficient for condensation in the presence of a noncondensable using a multiple section analysis.[44] Heat exchanger[45,46] and refrigeration unit[47] characteristics were developed from vendor information in conjunction with information given primarily in the Chemical Engineers' Handbook.[40,48] Refrigerant characteristics were taken primarily from the Chemical Engineers' Handbook[49,50] and publications of the American

Table E-12. PROCEDURE TO CALCULATE HEAT LOAD OF A CONDENSATION SYSTEM FOR STYRENE IN AIR

Item	Value
Heat exchanger type	Shell and tube heat exchanger with hot fluid in the shell side and the cold fluid in the tube side
Source identification	Identify the polymer industry and the vent from Chapters 3 and 6
Source production capacity (CAP), Gg/yr	From model plant in Chapter 6
Source emission factor (E), kg VOC/Mg product	From model plant in Chapter 6
Desired mass emission reduction, (% Red'n), %	From regulatory alternative in Chapter 6
Gas stream condition	Assumed saturated styrene in air at 80°F, 1 atm.
Partial pressure of styrene at inlet (P_{in})	7.959 mm Hg[a]
Composition of gas stream at inlet	0.01047 ft^3 styrene/ft^3 gas[b]; 0.002767 lb styrene/ft^3 gas[c]
Styrene mass flowrate (W_s), lb/hr[d]	0.2564 x CAP x E
Gas stream volumetric flowrate (V), acfm[e]	6.022 x W_s
Gas stream mass flowrate (W), lb/hr[f]	4.400 x V
Partial pressure of styrene at outlet (P_{out}), mm Hg	$\frac{100\%-(\%\ Red'n)}{100}$ x 7.959 mm Hg[g]
Temperature required for reduction (T'_{out}), °K[h]	4847.95 ÷ [18.2440 - ln (P_{out})]
Temperature required for reduction (T_{out}), °F	(1.8 x T'_{out}) - 459.67
Latent heat change of styrene (Q_{sty}), Btu/hr[i]	166.36 x W_s x (% Red'n)

Table E-12. PROCEDURE TO CALCULATE HEAT LOAD OF A CONDENSATION SYSTEM FOR STYRENE IN AIR (Concluded)

Item	Value
Average (bulk) gas temperature (T_b), °F	$(80 + T_{out}) \div 2$
Density of air (ρ_{air}), lb/ft^3[j,k]	$1 \div [(0.002517 \times T_b) + 1.157]$
Specific heat of air($(c_p)_{air}$), Btu/lb-°F	From API Report 44[l]
Sensible heat change of air (Q_{air}), Btu/hr	$V \times \rho_{air} \times (c_p)_{air} \times (80-T_{out}) \times$ 60 min/hr
Specific heat of styrene ($(c_p)_{sty}$), Btu/lb-°F	From API Report 44[m]
Sensible heat change of styrene (Q'_{sty}), Btu/hr	$W_s \times (c_p)_{sty} \times (80-T_{out})$
Total design heat load (Q_{tot}), Btu/hr	$(Q_{sty} + Q_{air} + Q'_{sty})$

[a]Calculated from Clausius Clapeyron curve fit

$$\left(\ln p = -4847.95 \left(\frac{1}{T,°K}\right) + 18.2440\right)$$ of styrene vapor pressure versus temperature data given on p. 3-59 of the Chemical Engineers' Handbook (Reference 37) for 80°F (see temperature required for reduction).

[b]Volume fraction of styrene = $\frac{7.959 \text{ mm Hg}}{760 \text{ mm Hg}} = 0.01047$ ft^3 styrene/ft^3 gas.

[c]Assuming ideal gas:

$$\frac{V}{n} = \frac{RT}{p} = \frac{1545 \text{ ft lb}_f/\text{lb}_m - °R \times 540 °R}{14.7 \text{ lb}_f/\text{in.}^2 \times 144 \text{ in}^2/\text{ft}^2} = 394.13 \text{ft}^3/\text{lb-mole};$$

styrene content (lb/ft^3 gas) =

$$\frac{0.01047 \text{ ft}^3 \text{ styrene}}{\text{ft}^3 \text{ gas}} \times \frac{\text{lb-mole}}{394.13 \text{ ft}^3} \times \frac{104.14 \text{ lb styrene}}{\text{lb-mole}}$$

Footnotes for Table E-12 (concluded)

[d] $\frac{\text{CAP, Gg product/yr} \times 1000\ \text{Mg/Gg} \times \text{E kg VOC/Mg product}}{8600\ \text{hr/yr} \times 0.4536\ \text{kg/lb}}$

[e] $\frac{\text{CAP, Gg product/yr} \times 1000\ \text{Mg/Gg} \times \text{E kg VOC/Mg product}}{8600\ \text{hr/yr} \times 0.4536\ \text{kg/lb} \times 0.002767\ \text{lb/ft}^3 \times 60\ \text{min/hr}} = 1.544 \times \text{CAP} \times \text{E}$

[f] $\frac{V, \text{acfm @ } 80°F \times \approx 28.9\ \text{lb/lb-mole} \times 60\ \text{min/hr}}{394.13\ \text{acf/lb-mole @ } 80°F}$

[g] Using mass emission reduction as initial approximation of volume (partial pressure) reduction. After outlet temperature is approximated, the outlet gas molar volume and partial pressure can be calculated and iterated upon until an acceptable outlet temperature is calculated.

[h] Solving Clausius Clapeyron curve fit of styrene vapor pressure data ($r^2 = 0.99995$) referred to in footnote a for temperature.

[i] Slope, m, of Clausius Clapeyron curve fit = $-\lambda/R$

latent heat of styrene, $\lambda = -m \times R =$

$$\frac{4847.95\ (°K) \times 1.9853\ \text{cal/g-mole-}°K \times 1.8\ \frac{\text{Btu-lb-mole}}{\text{cal/g-mole}}}{104.14\ \text{lb/lb-mole}}$$

[j] $T,°K = T,°C + 273.15 = \frac{5}{9}(T,°F - 32) + 273.15 = 0.5556\ T,°F + 255.37$

[k] For an ideal gas ($pV = mRT/MW$), $\frac{\rho_1}{\rho_2} = \frac{m_1/V_1}{m_2/V_2} = \frac{T_2}{T_1}$; $\rho_{air} = 0.08081\text{b/ft}^3$

at 0°C (ChE Hndbk, p. 3-72) (Reference 38)

$$\rho_{air} \text{ @ } T,°K = 0.0808 \times \frac{273.15°K}{T,°K} = \frac{0.808 \times 237.15}{0.5556 \times T,°F + 255.37}$$

[l] $(c_p)_{air} = 0.769\ (c_p)_{N_2} + 0.231\ (c_p)_{O_2}$, where $(c_p)_{N_2}$ and $(c_p)_{O_2}$ are specific heats of nitrogen and air, respectively, available by interpolation from API Report 44, p. 652 (Reference 39).

[m] $(c_p)_{sty}$ vs T,°F, values are available for interpolation on p. 682 of API Report 44 (Reference 37).

Society of Heating, Refrigerating, and Air-Conditioning Engineers (ASHRAE).[51] The total required heat transfer area and refrigeration capacity then were calculated from the total heat load, temperature difference, and overall heat transfer coefficient, and commercially available sizes were selected. A tabular procedure for calculating heat exchanger and refrigeration system size for a single section heat exchanger and a refrigeration unit is presented in Table E-13.

E.5.2 Surface Condenser Cost Estimation Procedure

Since the gas volumes of the two streams are low, the calculated required (selected) heat transfer areas are also low (about 1.7 (2.3) and 4.5 (7.2) ft^2, respectively, for 90 and 98 percent reduction of styrene emissions from a single process line). The purchase costs of the heat exchanger[43,44] and refrigeration systems [45] were estimated from data provided by vendors. An installation factor of 1.39 (see Table 8-2) was used to estimate installed condenser costs.

The corresponding required refrigeration capacities for 90 and 98 percent styrene reduction were only 0.056 and 0.080 tons (0.20 and 0.28 kW or 670 and 960 Btu/hr), respectively. For 90 percent reduction, the required capacity is much smaller than the 0.117 tons (0.412 kW or 1405 Btu/hr) available capacity for the minimum available size refrigeration unit of 1/4 compressor horsepower (0.186 kW compressor) and a coolant temperature of 0°F (-17.8°C). For 98 percent reduction, however, 1/2 compressor horsepower (0.373 kW compressor) was required to provide sufficient available refrigeration capacity for a coolant temperature of -30°F (-34.4°C).

Installed costs were put on a June 1980 basis using Chemical Engineering Plant cost indexes. No additional piping was costed since the condenser unit is so small ($\leq$ 5 in.) that it should be able to be installed adjacent to the source. Table E-14 presents the procedure for estimating capital and annual operating costs for condensation systems.

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.

Table E-13. PROCEDURES TO CALCULATE HEAT TRANSFER AREA OF A CONDENSATION SYSTEM OF STYRENE IN AIR

Heat exchanger configuration	Assume appropriate size unit.[a]
Source Identification	Identify the polymer industry and the vent from Chapters 3 and 6
Coolant temperature, T_c, °F	T_{out}-10, rounded to nearest multiple of 10°F.
Shell-side heat transfer Coefficient (h_o), $\frac{Btu}{hr\text{-}ft^2\text{-}°F}$	Calculate using procedure in Chemical Engineers' Handbook, pp. 10-25 thru 10-28 (Reference 38)[b]
Coolant	Select chilled water at $T_c \geq 60°F$; 50% ethylene glycol - 50% water solutions at T_c = 40,50°F; Freon-12 at $-40°F \leq T_c \leq 30°F$; and Freon-502, at $T_c \leq -50°F$.[c]
Tube-side Reynold's Number (N_{Re})[d]	$(12 \times r_H \times \rho) \div \mu$
Tube-side heat transfer Coefficient (h_i), $\frac{Btu}{hr\text{-}ft^2\text{-}°F}$	Calculate using appropriate equations for forced convection in pipes.[e]
Coolant flow (W_c), lb/hr	Calculate for assumed heat exchanger and coolant.[f]
Temperature change of coolant (ΔT_c), °F	$Q_{tot} \div (C_p \times W_c)$
Coolant flow (V_c), gpm	Calculate for assumed heat exchanger.[g]
Across-tube heat transfer coefficient (h_t), $\frac{Btu}{hr\text{-}ft^2\text{-}°F}$	Calculate for assumed tube size.[h]

Table E-13. PROCEDURES TO CALCULATE HEAT TRANSFER AREA OF A CONDENSATION SYSTEM OF STYRENE IN AIR (Concluded)

Condensation heat transfer Coefficient (h_c), $\frac{Btu}{hr\text{-}ft^2\text{-}°F}$	Calculate using procedure in Applied Process Design for Chemical and Petrochemical Plants, Vol. 3, p. 80 (Reference 42).[1]
Overall tube-side heat transfer coefficient (h_T)	$[(1/h_c) + (D_o/h_i D_i) + (1/h_t)]^{-1}$
Mass transfer coefficient (K_g), $\frac{lb\text{-}mole}{hr\text{-}ft^2\text{-}mm\ Hg}$	Calculate using procedure in Applied Design for Chemical and Petrochemical Plants, Vol. 3, pp. 100, 101, 104 (Reference 43).
Clean overall heat transfer coefficient (U_c), $Btu/ft^2\text{-}hr\text{-}°F$	Calculate from h_i, K_g, and h_T using procedure in Applied Process Design for Chemical and Petrochemical plants, Vol. 3, pp. 100-106 (Reference 44), which iterates on the condensate film temperature to balance the tube-side and shell-side heat transfer within ±5% and then uses the average to calculate the clean overall heat transfer coefficient.
Dirty overall heat transfer coefficient (U_d), $Btu/ft^2\text{-}hr\text{-}°F$[j]	$[(1/U_c) + 0.001 + 0.001]^{-1}$
Log mean temperature difference (LMTD), °F	$(\Delta T_1 - \Delta T_2) \div \ln (\Delta T_1/\Delta T_2)$[k]
Required heat transfer area (A), ft^2	$Q_{tot} \div (U_D \times LMTD)$[k]
Required heat transfer area with 10% safety margin (A'), ft^2	A x 1.1 if A' > A_T (available area for assumed size), try a larger size heat exchanger (see footnote a)
Required refrigeration capacity (RC'), tons[l]	$Q_{tot} \div 12{,}000$
Selected refrigeration capacity (RC), compressor horsepower	Select from vendor information (such as Reference 47) for design heat load and coolant temperature
Horsepower per ton of refrigeration (Hp/ton), Hp/ton	Based on vendor information for coolant temperature (see Reference 36).

Footnotes for Table E-13

[a] If no information is available on approximate condenser size required, assume an overall heat transfer coefficient between 3 and 15 Btu/ft^2-hr-°F, calculate required heat transfer area as noted near end of this table, and select an appropriate size condenser. Condenser and tube characteristics for large units (> 34 ft^2) can be found in pp. 11-1 thru 11-18 of the Chemical Engineers' Handbook (Reference 48). Characteristics of smaller units can be obtained from vendor information.[45,46] Characteristics of units developed for this analysis can be found in the docket.[36]

Characteristics needed for design calculations:

Tube: outer diameter, D_o (in.); inner diameter, D_i (in.); thickness, X_w (in.); cross sectional area, A_x (ft^2/tube); specific external surface area, A_e (ft^2/ft of tube); tube side hydraulic radius, r_H (ft) = $D_i/(4 \times 12$ in./ft).

Condenser: shell diameter, D_s (in.);
tube count, N_T (no. of tubes);
tube pitch, p (in.);
length, L (ft);
effective length, L_{eff} (ft) = $L - (N_{ts} \times L_{red}/12$ in/ft)
where N_{ts} = no. of tube sheets (can assume 2)
L_{red} (in.) = reduction in effective length per tube sheet (can assume = 0.25 in., for $D_s < 8$ in.; = 1.5 in., for $D_s \geq 8$ in.);
total tube area, A_T (ft^2) = $A_e \times L_{eff} \times N_T$.

[b] Assumed baffle cut, $l_c = 0.25\ D_s$, $D_s < 10$ in.; $= 0.33\ D_s$, $10 \leq D_s < 18$ in.; and $= 0.45\ D_s$, $D_s \geq 18$ in.; radial clearance between shell and outer tube limit = 1/4 in., $D_s < 8$ in.; = 7/16 in., for $8 \leq D_s < 25$ in.; = 1/2 in., for $D_s \geq 25$ in.; baffle spacing, $l_s = 0.6\ D_s$; diametrical shell-to-baffle clearance = 0.05 in., for $D_s < 4$ in.; = 0.075 in., for $4 \leq D_s < 7.5$ in.; = 0.10 in., for $7.5 \leq D_s < 14$ in.; = 0.125 in., for $14 \leq D_s < 18$ in.; = 0.150 in., for $18 \leq D_s < 25$ in.; = 0.30 in., for $25 \leq D_s < 42$ in.; = 0.35 in. for $D_s \geq 42$ in.; number of sealing strips $\geq 0.5 \times N_c$

[c] Coolant characteristics can be interpolated or extrapolated for the coolant temperature, T_c, from the Handbook of Chemistry and Physics, p. F-36 (Reference 50) and from The Chemical Engineers' Handbook: pp. 3-71, 126, & 214 (Reference 49) for water; and pp. 12-46 thru 12-48 (Reference 49) for ethylene glycol-water solutions; and from Thermophysical Properties of Refrigerants pp. 9 thru 11 for Freon-12 and pp. 105 and 106 for Freon-502 (Reference 51). Characteristics required are dynamic viscosity (μ_c), lb/ft-hr; density (ρ_c), lb/ft^3; specific heat [$(c_p)_c$] Btu/lb-°F; thermal conductivity (k_c), Btu/hr-°F-ft; and specific gravity (γ_c), dimensionless = $\rho_c/62.42$ lb/ft^3.

Footnotes for Table E-13 (Continued)

[d]For coolant velocity, V = 3 fps (3-10 fps recommended by Kern in Process Heat Transfer (Reference 52)).

[e]From The Chemical Engineers' Handbook, pp. 10-12 thru 10-15 (Reference 41).

Assumed $\left(\frac{\mu b}{\mu w}\right)^{0.14} \approx 1$ since little change in coolant temperature.

(1) For turbulent flow ($N_{Re} \geq 10{,}000$) (from Eq. 10-51):

$$h_i = \frac{0.023 \times V, \text{ ft/hr} \times \rho \text{ lb/ft}^3 \times c_p, \text{ Btu-lb-°F}}{(N_{Pr})^{2/3} (N_{Re})^{0.2}} \times \left(\frac{\mu b}{\mu w}\right)^{0.14}$$

if $0.7 < N_{Pr} < 700$ & $L/D > 60$;

(2) For transition flow ($2000 < N_{Re} < 10{,}000$) (from Eq. 10-49):

$$h_i = \frac{0.029\ k}{r_H} (N_{Re}^{2/3} - 125) N_{Pr}^{1/3} \left[1 + \left(\frac{D_i}{L}\right)^{2/3}\right] \left(\frac{\mu b}{\mu w}\right)^{0.14}$$

(3) For laminar flow ($N_{Re} < 2100$) (from Eq. 10-40):

$$h_i = \frac{0.465\ k}{r_H} N_{Gz}^{1/3} \left(\frac{\mu b}{\mu w}\right)^{0.14} \times 0.87\ (1 + 0.015\ N_{Gz}^{1/3})$$

where $N_{Gz} = (N_{Re} \times N_{Pr} \times 4 \times r_H) \div L$.

[f]For coolant velocity of 3 fps,
W_c, lb/hr = $A_x \times N_T \times 10{,}800$ ft/hr $\times \rho_c$

[g]For coolant velocity of 3 fps.
V_c, gpm = $A_x \times N_T \times 180$ ft/min $\times 7.48$ gal/ft^3.

$$h_{h_t} = \frac{K_t\ D_L}{D_o\ X_w}$$

where: K_t = thermal conductivity of tube = 64.2 Btu/ft-hr-°F for Admiralty brass given on pp. 23-49 of The Chemical Engineers' Handbook (see Reference 53).

$$D_L = (D_o - D_i) \div \ln (D_o/D_i)$$

[i]Must assume temperature of condensate film on tubes, which is the variable that is iterated upon to balance overall shell-side and overall tube-side heat transfer.

Footnotes for Table E-13 (Concluded)

[j]Assuming both shell-side and tube-side fouling coefficients equal to 0.001 as given for cooling tower water or refrigerants and for industrially clean gases, clean hydrocarbon vapors, or atmospheric air in Applied Process Design for Chemical and Petrochemical Plants, Vol. 3, Tables 10-10 and 10-11 (Reference 54).

[k]More complicated relationships, considering the average and end point values for individual sections and the entire exchanger, were used for the multiple section analysis (see References 36 and 44).

Table E-14. CAPITAL AND ANNUAL OPERATING COST ESTIMATION PROCEDURE FOR CONDENSERS WITH REFRIGERATION

Item	Value
Capital Costs	
Purchase cost, July 1984	[(16.816 x A_T) + 125.151] if Freon coolant, x 1.3[a]
Purchase cost, June 1980[b]	Purchase cost, Sept. 1982 x 0.871
Total Installed Cost, June 1980[c]	Total Purchase Cost x 1.39
Annualized Costs[d]	
Operating labor	See footnote e
Maintenance materials & labor	0.05 x total installed cost
Utilities Electricity, pumping[f]	6.104 x Y_C x V_C
Electricity, refrigeration	See footnote g
Coolant, make-up	See footnote h
Capital recovery[i]	0.1627 x total installed cost
Taxes, administration & insurance	0.04 x total installed cost
Total annualized cost without recovery credit	Operating labor & maintenance + utilities + capital recovery + taxes, administration & insurance
Styrene recovery credit[j]	2767 x W_S x (% Red'n. ÷ 100)
Net Annualized Cost after recovery credit	Total annualized cost - styrene recovery credit

Footnotes for Table E-14

[a]Based on vendor information (References 45 and 46).

[b]Deflated using Chemical Engineering Fabricated Equipment Cost Index from July 1984 (estimated) (334.6) to June 1980 (291.3).

[c]Breakdown of installed cost factor given in Table 8-2.

[d]Cost factors presented in Table 8-3.

[e]For 20 to 34 ft^2 units: Operating labor cost = 1 hr/wk x 52 wk/yr x 1.15 (with supervision/without supervision) x $18/hr (including overtime). For < 20 ft^2 units: Operating labor cost = 2 hr/mon x 12 mon/yr x 1.15 (with supervision/without supervision) x $18/hr (including overtime).

[f]Using Equation 6-2, p. 6-3 in The Chemical Engineers' Handbook (Reference 55) for V = 3 fps, assuming a pumping height of of 50 ft. and a pump efficiency of 65%:

$$\frac{50 \text{ ft} \times \gamma_c \times V_c}{3960 \text{ gpm ft/hp}} \times \frac{0.7457 \text{ kW}}{\text{hp}} \times \frac{8600 \text{ hr/yr}}{0.65 \text{ pump efficiency}} \times \$0.049/\text{kwh}$$

where γ_c = specific gravity of coolant = $\rho_c \div 62.42$ lb/ft^3 (the density of water)

[g]
$$\frac{RC' \times (\text{hp/ton of refrigeration for } T_c)}{0.85 \text{ compressor efficiency} \times 0.85 \text{ motor efficiency}} \times \frac{0.7457 \text{ KW}}{\text{hp}} \times 8600 \text{ hr/yr} \times \$0.049/\text{kwh}$$

[h]For coolant velocity of 3 fps:

For ethylene glycol-water solutions, Freon-12, and Freon-502, assume one replacement per year of coolant in condenser and refrigeration system and coolant volume in condenser and refrigeration unit twice that of condenser alone.

$$\text{coolant volume, gal} = A_X \times N_T \times L \times (2 \text{ vol. in cond \& refrig/vol. in cond}) = 2 \times A_X \times N_T \times L.$$

For ethylene glycol-water solutions:

$$\text{cost of coolant} = \{[X_W (\$0.30/1000 \text{ gal} \times 7.48 \text{ gal/ft}^3)] + [X_{EG} (\$0.27/\text{lb} \times \rho_{EG} \text{ lb/ft}^3)]\} \times \text{coolant volume, ft}^3 \times 1 \text{ replacement per year} = (\$0.00224 X_W + \$0.27 X_{EG} \rho_{EG}) \times \text{annual coolant replacement, ft}^3/\text{yr}.$$

Footnotes for Table E-14 (Concluded)

where,

X_W = volume fraction of water in solution
X_{EG} = volume fraction of ethylene glycol in solution

For Freon-12:

Cost of coolant = (coolant volume, ft^3 x 1 replacement/yr x \$1.31/lb) ÷ ρ_c, lb/ft^3

For Freon-502:

Cost of coolant = (coolant volume, ft^3 x 1 replacement/year x \$3.11/lb) ÷ ρ_c, lb/ft^3

For chilled water, assume 0.1% make-up (99.9% recycle):

Cost of coolant = \$0.30/1000 gal x V_c, gpm x 60 min/hr x 8600 hr/yr x 0.001 make-up/total
= 0.1548 x V_c

[i] 10 percent interest (before taxes) and 10 yr. life.

[j] W_s, lb styrene emitted/hr x 8600 hr/yr x (% Red'n in condenser ÷ 100) x 0.90, fraction of reduction recovered x \$0.3575/lb styrene.

E.6.1 Ethylene Glycol Recovery System Design

The baseline system for a plant producing a low viscosity product or high viscosity product with a single end finisher was represented by a system that recovers ethylene glycol (EG) emitted from the polymerization reactors through use of EG spray condensers and from the esterifiers through use of reflux condensers. The baseline system for a plant producing a high viscosity product was represented by a system that recovers ethylene glycol emitted from the polymerizers through the use of EG spray condensers on the initial end finishers and a distillation column on the cooling water tower and from the estrifiers through the use of reflux condensers.

The regulatory alternative control systems utilized distillation columns. For those PET plants producing a low viscosity product or a high viscosity product with a single end finisher, further control of EG emissions was obtained by the installation of a distillation column that reduces the EG concentration in the cooling water tower. For those PET plants producing a high viscosity product with multiple end finishers, further control of EG emissions was obtained by increasing the flow rate of cooling water to the existing distillation column, which results in a lower EG concentration in the cooling tower.

The equipment selected to comprise the ethylene glycol recovery systems, as well as the design and operating parameters, was based on information provided by industry sources. The industry information (much of which was considered confidential) was used in conjunction with standard engineering references such as the Chemical Engineers' Handbook[17], and McCabe and Smith's Unit Operations of Chemical Engineering[56], and engineering judgment.[57] These design procedures are summarized in the footnotes to Tables E-15 and E-16 for the baseline systems.

E.6.2 Ethylene Glycol Recovery System Cost Estimation Procedure

The cost estimates and their bases are presented in Table E-15 for the baseline ethylene glycol recovery system for PET plants producing a low viscosity product or a high viscosity product with a single end finisher and in Table E-16 for PET plants producing a high viscosity product with multiple end finishers. The costs of the baseline systems

Table E-15. EG RECOVERY COSTS FOR BASELINE SYSTEM FOR PLANTS PRODUCING A LOW VISCOSITY PRODUCT OR A HIGH VISCOSITY PRODUCT WITH A SINGLE END FINISHER (June 1980 dollars)

Item (Number of Item)	Model Plant	Process Line
Capital Costs		
Spray Condensers (28 @ $17,384 each)	$486,750[a]	-
Reflux Condensers (14)	$243,400[b]	-
Pumps (28)	$23,300[c]	-
Heat Exchangers (42)	$64,300[d]	-
EG Recovery System (1)	$386,300[e]	-
Refrigeration System (14)	$128,700[f]	-
Installed Capital Cost Factor	4.24[g]	-
Total Installed Capital Cost	$5,234,000	$1,628,400[h]
Annualized Costs		
Operating Labor	$173,500[i]	$22,100[j]
Operating Materials	0	0
Maintenance Materials and Labor	$21,700[k]	$3,100[j]
Electricity	$86,650[l]	$12,400[j]
Steam	$917,930[m]	$131,100[j]
Water	$4,650[n]	$664[j]
Taxes, Insurance and Administration	$209,400[o]	$65,100
Capital Recovery	$853,100[p]	$265,400
Recovery Credit	($1,218,300)[q]	($174,000)[j]
Total (Annual Costs - Recovery Credit)	$1,048,700	$325,900

Footnotes for Table E-15

[a]Based on size estimate from Tennessee Eastman and cost estimate from Missouri Boiler. Two spray condensers plus two spares 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 plus one spare per line. The size and cost were assumed to be the same as for EG spray condensers on the reactors.

[c]See Reference 57, p. 14, for pump sizing. A 1 hp pump is required per line at a cost of $831 per pump.

[d]See Reference 57, pp. 7-9, for 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. Seven scraped surface exchangers (plus 7 spares) for end finishers at $1,130 each; 7 scraped surface exchangers for prefinishers (plus 7 spares) at $1,930 each; 7 exchangers (plus 7 spares) for esterifiers at $1,530 each (average of $1,130 plus $1,930).

[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]See Reference 57. Seven systems for prefinisher at $9,283 each, and 7 for finishers at $9,100 each.

[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. Does not apply to refrigeration systems.

[h]Model plant is comprised of seven process lines. Estimate of total installed capital cost of equipment for just one process line was obtained with the following equation:

$$\left(\frac{1}{7}\right)^{0.6} \times \$5{,}234{,}000 = \$1{,}628{,}400$$

[i]Based on 9,640 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.

[j]Estimate of cost for a single process line was obtained by the following equation:

$$\left(\frac{1}{7}\right) \times AC_i$$

where: AC_i = operating labor, operating materials, maintenance materials and labor, electricity, steam, and water.

Footnotes for Table E-15 (Concluded)

[k]Based on maintenance labor requirements and maintenance materials cost given in confidential industry information for a larger system.

[l]This cost comes from 4 sources: pumps, recovery system, heat exchanger to chill water, and for the reflux condensers. The 14 pumps at 1 hp use 91,500 kWh per year. Recovery system was estimated to use 412,800 kWh per year based on confidential industry information for a larger system. 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 is 660,100 kWh per year for a plant producing a high viscosity product. Electricity usage by the exchangers for the reflux condensers was assumed to be the same as that required by the spray condensers (i.e., 603,900 kWh per year). Assuming all the lines in a model plant produce high tenacity (high viscosity) industrial resins, total electricity usage in the model plant is 91,500 plus 412,800 plus 660,100 plus 603,900, which is equal to 1,768,300 kWh per year. Cost of electricity is $0.049/kWh.

[m]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. An estimated additional 14.532×10^6 lbs of steam per year would be required over the baseline system by the vacuum system servicing the polymerization reactors. A cost of $6.18/1,000 lb of steam was used.

[n]Based on water consumption of 15.5×10^6 gallons per year (from confidential industry information scaled down by proportioning relative capacities).

[o]Based on 0.04 x Installed Capital Cost.

[p]Based on a capital recovery factor of 0.163.

[q]Based on a total EG recovery of 19.5 kg of EG/Mg of product and a recovery credit of $0.595/kg ($0.27/lbs of EG from Chemical Marketing Reporter). The 19.5 kg of EG/Mg of product is an increase in the recovery of ethylene glycol from the polymerization reactors plus ethylene glycol recycled from the estifiers that would otherwise have to be replaced with fresh feed if the baseline system was used.

TABLE E-16. EG RECOVERY COSTS FOR BASELINE SYSTEM FOR PET PLANTS PRODUCING A HIGH VISCOSITY PRODUCT WITH MULTIPLE END FINISHERS (June 1980 dollars)

Item (Number of Item)	Model Plant	Process Line
Capital Costs		
Spray Condensers (4)	$82,640[a]	$41,320
Pumps (4)	3,945[b]	1,970
Heat Exchangers (4)	8,220[c]	4,110
Reflux Condensers (4)	82,640[d]	41,320
Distillation Column, cooling tower	661,800[e]	568,000
Refrigeration system (2)	22,065[f]	11,030
Installed Capital Cost Factor	3.24[g]	3.24
Total Installed Capital Cost	1,258,800	866,480
Annualized Costs		
Operating Labor	29,500[h]	14,750
Operating Materials	-	-
Maintenance Materials and Labor	3,940[i]	1,970
Electricity	21,640[j]	10,820
Steam	5,170[k]	2,635
Water	-	-
Taxes, Insurance and Administration	50,350[l]	34,660
Capital Recovery	204,800[m]	140,980
Recovery Credit	(233,490)[n]	(116,745)
Total (Annual Costs - Recovery Credit)	81,910	89,070

Footnotes for Table E-16

[a]The number of spray condensers is based on one per line with one spare per line and two lines per plant. The cost per spray condenser was estimated by multiplying the spray condenser cost in Table E-15 by the relative process line sizes (20 Gg vs. 15 Gg) raised to the 0.6 power as follows:

$$\$17,384 \times \left(\frac{20}{15}\right)^{0.6} = \$20,660$$

[b]Two pumps plus two spares per process line. Cost per pump adjusted from Table E-15 as follows:

$$\$830 \times \left(\frac{20}{15}\right)^{0.6} = \$986$$

[c]Two heat exchangers plus two spares per process line. Cost per heat exchanger adjusted from Table E-15 as follows:

$$[\text{2 scraped surface exchanges for prefinishers at \$1,930 each plus 2 exchangers at \$1,530 each for esterifiers}] \times \left(\frac{20}{15}\right)^{0.6} = \$8,220$$

[d]One reflux condenser plus one spare per process line. Cost per reflux condenser assumed the same as the spray condenser in footnote a.

[e]Based on adjusting size and cost information claimed confidential from an industry source. In general, the size was adjusted based upon proportioning the flow to the distillation column directly on plant capacity and then sizing the distillation column on the basis of the new flow divided by the flow to the original distillation column and the result raised to the 0.6 power. The cost was adjusted to June 1980 dollars using a cost index factor of 0.8208 based upon the Chemical Engineering Plant Cost Index of 320.3 for January 1984 and 262.9 for June 1980.

[f]Cost per refrigeration system adjusted from Table E-15 as follows:

$$\$9,283 \times \left(\frac{20}{15}\right)^{0.6} = \$11,030$$

[g]See Table 8-2. Applies to spray condensers, pumps, heat exchangers, and reflux condensers. Does not apply to distillation column and refrigeration system.

[h]Confidential industry information on operating labor scaled down to assume 1,640 hours per year at $18 per hour.

[i]Based on maintenance labor requirements and maintenance materials cost given in confidential industry information for a larger system. Assumes $1,330 for materials and 145 hours for maintenance labor.

[j]Electricity usage per source was assumed to be proportional to plant capacity. The 4 pumps at 1.33 hp each would use 17,385 kWh/year. The electricity requirement to maintain the chilled water (refrigeration

Footnotes for Table E-16 (Concluded)

system) is 212,070 kWh/year. Electricity used by the exchangers for the reflux condensers was assumed to be the same as that required by the spray condensers (i.e., 212,070 kWh/year). Total electrical usage in the model plant for these controls is 17,385 plus 212,070 plus 212,070, which is equal to 441,525 kWh/year. The cost of electricity is $0.049/kWh.

[k]Includes only steam used in distillation column. Distillation steam total equals 836,625 lb/yr (97,575 lb/yr to heat the water; 5,853 lb/yr to heat the ethylene glycol; and 733,197 lb/yr to vaporize the water). Steam cost of $6.18 per 1000 lb of steam.

[l]Total installed capital cost x 0.04.

[m]Total installed capital cost x 0.1627.

[n]Based on total EG recovery of 17.65 lbs of EG/1,000 lbs of product and a recovery credit of $0.15 per lb. Of the total recovered EG, about 16.68 lbs EG/1,000 lbs comes from the initial end finishers and 0.97 lb EG/1,000 lbs product from the distillation column.

were estimated based on the design estimate developed and standard engineering procedures.

The costs of the regulatory alternative distillation columns were derived from confidential cost data provided by an industry source for a similar system on a larger capacity plant. In general, the distillation columns were sized and costed on the basis of the flow rate from the cooling tower to the column using the ratio of the target flow rate divided by the base flow rate and the result raised to the 0.6 power. Utility costs, which are primarily steam costs, were calculated based on the flow rate from the cooling tower to the distillation column.

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-17. For flows less than 700 scfm, an economic pipe diameter was calculated based on an equation in the Chemical Engineer's Handbook[59] and simplified as suggested by Chontos.[60,61,62] 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[30] as combined for 70 ft. source legs and 500 ft. and 2,000 ft. pipelines for the cost analysis of the Distillation NSPS.[63] (See Tables E-18 and E-19). Ducting costs were calculated based on the installed cost equations given in the GARD Manual.[64] (See Table E-20).

Table E-17. PIPING AND DUCTING DESIGN PROCEDURE

Item	Value
(1) Pipe diameter, D	
(a) Piping[a]	For Source Legs: 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 For Pipeline Legs: D (in.) = 0.024 x Q (scfm) + 0.549, for Q < 65 scfm D (in.) = 4.8 $\log_{10}$ [Q (scfm)] - 7.33, for 65 ≤ Q < 700 scfm Select next larger inner diameter of schedule 40 pipe (given in Table E-19) unless calculated size with 10 percent of interval to next larger size.
(b) Ducting[b]	D (in) = (0.3028) $\sqrt{Q(acfm)}$, for D > 12 in. or Q ≥ 700scfm and Q ≤ 60,000 acfm D (in.) = (0.1915) $\sqrt{Q(acfm)}$, for Q > 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 (≥ 35,000 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 (Reference 28) which assumes a 150-ft. duct cost based on the GARD Manual (Reference 64).

[a]Economic pipe diameter equations from Reference 62 (which is based upon References 59 and 60).

[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-18. PIPING COMPONENTS[a]

Equipment Type	Number of Equipment Type in Pipe Leg Type			
	Source	Compressor	Pipeline (500 ft)	Pipeline (2,000 ft)
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	2	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 Length, (Schedule 40) (ft)	70	20	500	2,000

[a] From Reference 62.

Table E-19. INSTALLED PIPING COSTS[a]

Pipe Diameter (ft)	Cost in August 1978 Dollars for Pipe Leg Type			
	(70 ft) Source Leg	(20 ft) Compressor Leg	500 ft Pipeline	2,000 ft Pipeline
0.0411	2,200	1,050	2,275	4,360
0.0518	2,262	1,085	2,555	4,955
0.647	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 63) based on component costs from Reference 30.

Table E-20. INSTALLED DUCTING COST EQUATIONS, DECEMBER 1977 DOLLARS[a]

Pipe Wall Pipe Wall Thickness (in)[b]	Source Leg	Compressor Leg	500 ft. Pipeline	2,000 ft. Pipeline	Permissible Diameters (in.)
1/2	$1866 + 546.6\ D + 4.80\ D^2$	$963.4 + 269.4\ D + 3.20\ D^2$	$-894.9 + 1722\ D + 3.36\ D^2$	$-6685 + 6237\ D + 3.36\ D^2$	$50 \leq D \leq 180$
3/8	$1966 + 466.4\ D + 4.50\ D^2$	$1005 + 232.7\ D + 3.00\ D^2$	$-405.2 + 1382\ D + 3.15\ D^2$	$-4800 + 4952\ D + 3.15\ D^2$	$40 \leq D \leq 170$
1/4	$2074 + 396.8\ D + 2.40\ D^2$	$1054 + 202.5\ D + 1.600\ D^2$	$67.53 + 1058\ D + 1.68\ D^2$	$-3007 + 3713\ D + 1.68\ D^2$	$28 \leq D \leq 122$
3/16	$2083 + 352.2\ D + 2.050\ D^2$	$1052 + 183.5\ D + 1.367\ D^2$	$204.3 + 846.7\ D + 1.435\ D^2$	$-2436 + 2901\ D + 1.435\ D^2$	$30 \leq D \leq 90$
1/8	$2198 + 285.8\ D + 1.76\ D^2$	$1121 + 146.7\ D + 1.174\ D^2$	$420 + 673.9\ D + 1.232\ D^2$	$-1770 + 2309\ D + 1.232\ D^2$	$12 \leq D \leq 81$

[a]From equations for carbon steel pipe components (straight duct, elbows, tees and transitions) given in the GARD Manual (Reference 64) combined for piping components in each pipe leg type as shown in Table E-18.

[b]After examining dependence of wall thickness, t, upon pressure (psi), and diameter, D (in.):

$$t(\text{in.}) = \frac{p \times D}{2\ [\ (18{,}200 \times 0.50) + (0.4 \times p)]} + 0.016$$

for cases where pressure was known.
assumed = 1/4" for $p \geq 74.7$ psia or $D \geq 8.5$ ft.
= 3/16" for $Q \geq 30{,}000$ acfm, (0-10 psig)
= 1/8" for $Q \leq 30{,}000$ acfm

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.

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15. SUPPLEMENTARY NOTES

16. ABSTRACT

Standards of performance for the control of volatile organic compound emissions from the polymer manufacturing industry are being proposed under the authority of Section 111 of the Clean Air Act. These standards would apply to new, modified, and reconstructed facilities that manufacture polypropylene, polyethylene, polystyrene, or poly(ethylene terephthalate). This document contains background information and environmental and economic impact assessments of the regulatory alternatives considered in developing the proposed standards.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution Volatile Organic Compounds Polymers Resin Polyethylene Polypropylene Polystyrene Poly(ethylene terephthalate)	Air Pollution Control	13B

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