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Guideline Series

Control of Volatile Organic Compound Emissions from Manufacture of Styrene-Butadiene Copolymers

Preliminary Draft

CTG

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Guideline Series

Control of Volatile Organic Compound Emissions from Manufacture of Styrene-Butadiene Copolymers

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

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GUIDELINE SERIES

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TABLE OF CONTENTS

Table of Contents (continued)

LIST OF TABLES

LIST OF FIGURES

1.0 INTRODUCTION

This needs CPPD review. Stonefield/Calcagni.

1.1 INTRODUCTION

The Clean Air Act Amendments of 1977 require each State in which there are areas in which the national ambient air quality standards (NAAQS) are exceeded to adopt and submit revised state implementation plans (SIP's) to EPA. Revised SIP's were required to be submitted to EPA by January 1, 1979. States which were unable to demonstrate attainment with the NAAQS for ozone by the statutory deadline of December 31, 1982, could request extensions for attainment with the standard. States granted such an extension are required to submit a further revised SIP by July 1, 1982.

Section 172(a)(2) and (b)(3) of the Clean Air Act require that nonattainment area SIP's include reasonably available control technology (RACT) requirements for stationary sources. As explained in the "General Preamble for Proposed Rulemaking on Approval of State Implementation Plan Revisions for Nonattainment Areas," (44 FR 20372, April 4, 1979) for ozone SIP's, EPA permitted States to defer the adoption of RACT regulations on a category of stationary sources of volatile organic compounds (VOC) until after EPA published a control techniques guideline (CTG) for that VOC source category. See also 44 FR 53761 (September 17, 1979). This delay allowed the states to make more technically sound decisions regarding the application of RACT.

Although CTG documents review existing information and data concerning the technology and cost of various control techniques to reduce emissions, they are, of necessity, general in nature and do not fully account for unique variations within a stationary source category. Consequently, the purpose of CTG documents is to provide State and local air pollution control agencies with an initial information base for proceeding with their own analysis of RACT for specific stationary sources.

An EPA CTG document is not necessary for a State to develop RACT.

2.0 PROCESSES AND POLLUTANT EMISSIONS

The purpose of this chapter is to describe current practices in the styrene-butadiene manufacturing industry, to provide brief process descriptions, and to identify the emission points and the volatile organic compounds (VOC) that are emitted from these processes. Quantitative estimates of the flow rates and VOC concentrations of pollutant streams from model plants are also contained in this chapter.

2.1 INDUSTRY DESCRIPTION

Copolymers of styrene and butadiene can be made with properties ranging from those of a rubbery material to those of a very resilient plastic. Copolymers containing less than 45 percent styrene by weight are known as SBR (styrene-butadiene rubber). As the styrene content is increased to over 45 percent by weight the product becomes increasingly more plastic. Then what's it called?

Two types of polymerization reactions are used to produce styrene-butadiene copolymers: the solution type and the emulsion type. Table 2-1 lists the U.S. producers, the plant locations, the polymerization type used, and the capacities. This CTG addresses VOC emissions from the manufacture of copolymers of styrene and butadiene ~~made~~ by the emulsion type process.

2.2 EMULSION PROCESS DESCRIPTIONS

The emulsion polymerization process was developed ~~by the U.S. government~~ during World War II to provide a substitute for the cut-off supplies of natural rubber. The ~~plants were sold to private companies after the war~~. Major innovations to the original process since that time are the use of a "cold" polymerization process and the use of oil as an extender of the rubber.

TABLE 2-1. DOMESTIC PRODUCERS OF STYRENE-BUTADIENE COPOLYMERS[2,3]

Company	Plant location	Plant type	Annual capacity (10^3 Mg)
American Synthetic Rubber Corp.	Louisville, KY	Emulsion crumb	100
Copolymer Rubber and Chemical Co.	Baton Rouge, LA	Emulsion crumb	191
Dow Chemical Co.	Bayport, TX	Emulsion latex	20
	Dalton, GA		
	Gales Ferry, CT		
	Midland, MI		
	Pittsburg, CA		
Firestone Tire and Rubber Co.	Lake Charles, LA	Emulsion crumb	331
	Lake Charles, LA	Solution crumb	
	Pottstown, PA	Emulsion latex	4
General Tire and Rubber Co.	Mogadore, OH	Emulsion latex	47
	Odessa, TX	Emulsion crumb	96
BF Goodrich Co.	Port Neches, TX	Emulsion crumb	142
	Louisville, KY	Emulsion latex	NA
Goodyear Tire and Rubber Co.	Houston, TX	Emulsion crumb	386
	Calhoun, GA	Emulsion latex	NA
W.R. Grace and Co.	Owensboro, KY	Emulsion latex	7
	South Acton, MA	Emulsion latex	7
Phillips Petroleum Co.	Borger, TX[a]	Emulsion crumb	125
	Borger, TX	Solution crumb	127
Reichold Chemicals, Inc.	Cheswold, DE	Emulsion latex	20
	Kensington, GA	Emulsion latex	25
Rubber Research Elastomers	Minneapolis, MN	Emulsion latex	NA
Polysar Resins, Inc.	Chattanooga, TN	Emulsion latex	30
	Chattanooga, TN	Emulsion latex	30
	Beaver Valley, PA	Emulsion latex	20
Shell Chemical Co.	Marietta, OH	Solution crumb	NA
Synpol Inc.	Port Neches, TX	Emulsion crumb	183
Union Oil Co. of California	Charlotte, NC	Emulsion latex	8
	La Mirada, CA	Emulsion latex	9
U.S. Steel Corp.	Scotts Bluff, LA	Emulsion latex	NA

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[a] Jointly owned with General Tire and Rubber Company.

The emulsion products can be sold in either a solid form, known as crumb, or in a liquid form, known as latex. The crumb product is used extensively in the manufacture of rubber tires, and has an average styrene content of 23.5 percent by weight.[1] The latex products have a wide variety of uses, depending on the styrene content of a particular grade. The rubbery types of approximately 23 percent styrene are used for dipped goods, carpet underlay, adhesives, and moldings. Small quantities of vinylpyridine monomer may be added if the latex is to be used for dipping tire cords. The high styrene latexes, ranging from 46 to 85 percent styrene, are used for paper coatings, paints, carpet backsizing, and adhesives. Small quantities of a carboxylic monomer, such as acrylic acid, are sometimes added to provide a "self-curing" material for use in carpet backsizing.

Emulsion polymerization provides an aqueous medium as a reaction site for the styrene and butadiene monomers. Surfactants are used to stabilize the monomer-in-water dispersion, and the initiators used are soluble in the water phase. There are only slight differences between the process used to make a crumb end product and the process used to make a latex end product.

2.2.1 Emulsion Crumb Polymerization

As shown in the general flow diagram depicted in Figure 2-1, fresh styrene and butadiene are piped separately to the manufacturing plant from the storage area and, if necessary, the butadiene stream is passed through a caustic soda (NaOH) scrubber to remove any inhibitors that have been added to prevent premature polymerization during shipment and storage. The inhibitors are removed in a 20 percent aqueous caustic soda solution, which is subjected to continual makeup and blowdown (with subsequent purging). Before entering the polymerization reactor train, fresh monomer streams are mixed with the recycle styrene and butadiene streams from the monomer recovery areas of the process.

Soap solution, activator, catalyst, and modifier are added to the monomer mixture prior to entering the polymerization reactor train. The soap solution, which is generally a mixture of a rosin acid soap and a fatty acid soap, is used to maintain the monomers in an aqueous emulsion state. The activator is usually a hydroperoxide or a peroxysulfate which initiates the

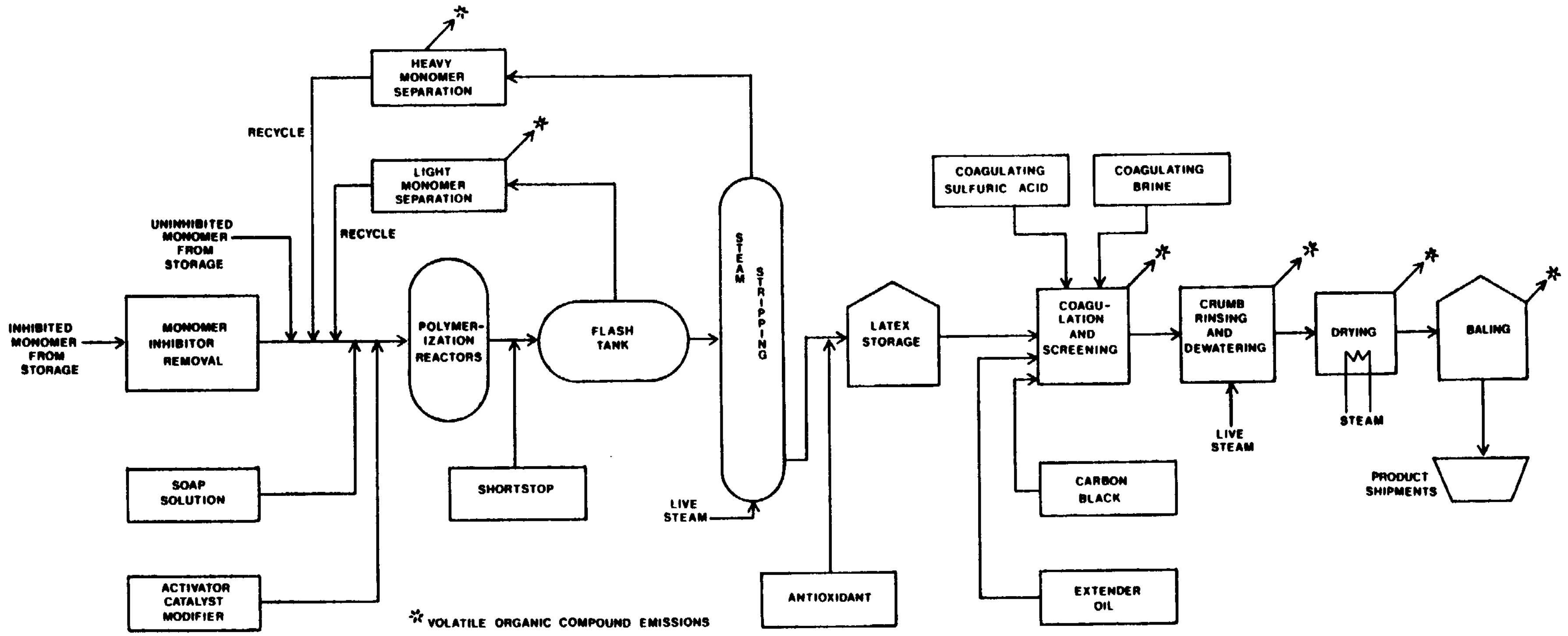

Figure 2-1. Schematic flow diagram for crumb production by emulsion polymerization.

polymerization reaction by supplying free radicals. The catalyst assists in generating the free radicals more rapidly and at lower temperatures than is possible with thermal decomposition of the activator alone. The modifier is an additive used to adjust the chain length and molecular weight distribution of the rubber product during polymerization.

Polymerization of styrene and butadiene proceeds stepwise through a train of reactors on a continuous basis, with a residence time in each reactor of approximately 1 hour. The reaction is normally carried out at a temperature of 4°C (40°F) and produces excess heat which is removed by cooling coils implanted in each reactor. Use of the reactor train system contributes significantly to the manufacturing facility's high degree of flexibility in producing different grades of crumb rubber. The overall polymerization reaction is ordinarily carried out to a 60 percent conversion of monomers to polymer, because the reaction rate falls off considerably beyond this point and production quality begins to deteriorate. The reaction product forms in the emulsion phase of the reaction mixture, yielding a milky-white emulsion called latex.

Shortstop solution, consisting mainly of sodium dimethyldithiocarbamate ($(CH_3)_2NCS_2Na$) or hydroquinone ($C_6H_6O_2$), is added to the latex leaving the reactors to stop the polymerization process at the desired conversion. The "stopped" latex is held in blowdown tanks (which function as flow-regulating holding tanks) prior to recovery operations for the unreacted monomers.

Because recovery of the unreacted monomers and their subsequent purification ~~are~~ is essential to economical operation, unreacted butadiene and styrene are recovered during the emulsion crumb polymerization process. The latex is introduced to flash tanks (using pressure flashing or vacuum flashing, either separately or in combination) where the overhead butadiene vapor stream is compressed, passed through a condenser, and finally processed in a kerosene absorber before it is vented to the atmosphere. The absorbed butadiene is recovered using either steam stripping, a pressure reduction, or a temperature increase, and is combined with the recovered butadiene from the condensers before being mixed with fresh butadiene for reactor feed.

The latex stream from the butadiene recovery area is then sent to the styrene recovery process. Styrene recovery from latex usually takes place in perforated-plate stripping columns. These operate at less than atmospheric pressure using steam injection at approximately 60°C (140°F). The steam and styrene vapor stream is condensed in a water-cooled condenser and sent to a decanter, where the water phase and styrene phase are separated. The styrene phase is sent to recycle storage tanks to be blended later with fresh styrene for reactor feed. The water from the decanter is discharged to the wastewater treatment system. Noncondensibles are directed through the butadiene compressors and are eventually scrubbed by the kerosene absorber before being vented to the atmosphere.

An antioxidant is added to the stripped latex in a blend tank to protect the polymer from attack by oxygen and ozone. The latex is then stabilized, and, as a result, different batches can be mixed.

The latex is pumped from the blend tanks to coagulation vessels where dilute sulfuric acid (H_2SO_4 with pH 4 to 4.5) and sodium chloride (NaCl) solution are added. The acid and brine mixture causes the emulsion to break, releasing the styrene-butadiene copolymer, or crumb product. Carbon black and extender oil can be added just prior to this step to produce a more specialized crumb product. The coagulation vessels are open to the atmosphere.

After leaving the coagulation process the crumb and water slurry is separated into its solid and liquid phases using screens. The crumb product is processed in rotary presses that squeeze out most of the water entrained in the crumb. The liquid streams from the screening area and the rotary presses are recycled to the polymerization area for reuse.

The partially dried crumb is finally processed in a continuous belt dryer which blows hot air of approximately 93°C (200°F) across the crumb to complete the drying of the product. Most plants still use the triple pass dryers which were installed as original equipment in the 1940's. Some plants have installed single pass dryers where space permits. (A discussion of the operation of both types of dryers is contained in Chapter 3). The dried product is baled and weighed before shipment.

2.2.2 Emulsion Latex Polymerization

Emulsion polymerization can also be used to produce latex products. These latex products have a wider range of properties and uses than the crumb products, but the plants are much smaller. Latex production, shown schematically in Figure 2-2, basically follows the same processing steps as emulsion crumb polymerization with the exception of the final product processing steps.

As in emulsion crumb polymerization, the monomers are piped to the processing plant from the storage area. At the manufacturing facility, monomer inhibitors are scrubbed out using a caustic soda solution, if necessary. Soap solution, activator, and modifiers are then added to the monomers to produce a feed emulsion that is fed to the reactors. The reaction produces excess heat which is removed by cooling coils implanted in the reactor vessel. The polymerization reaction is taken essentially to completion (about 98 to 99 percent conversion), and, as a result, the recovery of unreacted monomers is uneconomical. Process economics are directed towards maximum conversion on a once-through basis.

Because most emulsion latex polymerization is done in a batch process, the number of reactors used for latex production is usually smaller than for crumb production. The polymerization reaction is normally carried out at 49^{o}C (120^{o}F), although some manufacturers (approximately 15 percent[4]) use the "cold" polymerization process, which operates at 4^{o}C (40^{o}F). The degree of conversion of monomers to copolymer for the cold process is less than for the hot process, so recovery operations similar to those of a crumb plant are used in cold process latex plants. Also, some crumb plants produce small amounts of latex as an end product using the same equipment that produces crumb, except for the coagulation and drying equipment.

Completion of the polymerization process requires from 6 to 24 hours. The latex is sent to a blowdown tank where, under an absolute pressure of 6.5 kPa (~28 in. of mercury vacuum) and steam agitation, any unreacted butadiene and some unreacted styrene are removed from the latex. The overhead stream from the blowdown tank is sent to a water-cooled condenser where any condensibles are removed from the vapor stream and sent to a wastewater treatment facility. Noncondensibles from the condenser are discharged to the atmosphere.

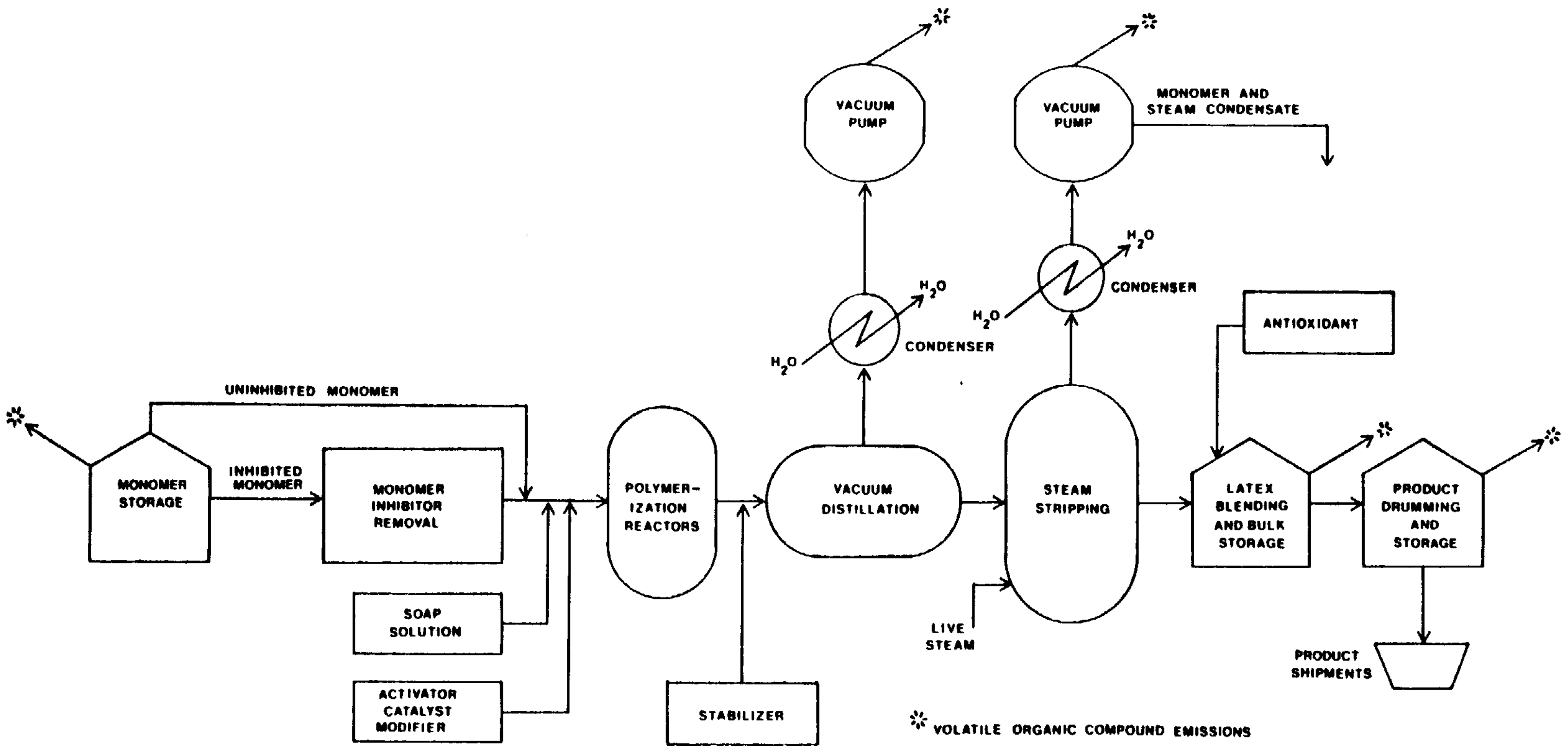

Figure 2-2. Schematic flow diagram for latex production by emulsion polymerization.

After discharge from the blowdown tank the latex is stored in process tanks where additives are introduced based upon product specifications. Starting from this point in the manufacturing process to the final product storage for shipment, latex is processed on a continuous basis.

Subsequently, the latex is screened using shaker screens to remove any large, agglomerated solids present in the latex. If the unreacted styrene content of the latex has not been reduced sufficiently to meet product specifications in the blowdown step, the latex is introduced to a series of steam-stripping steps to reduce the unreacted styrene content. Any steam and styrene vapor from these stripping steps is taken overhead and sent to a water-cooled condenser. Any noncondensibles leaving the condenser are vented to the atmosphere. The liquid stream from the condenser is discharged to the wastewater treatment system.

The stripped latex is then passed through a series of screen filters to remove unwanted large solids and is stored in blending tanks where antioxidants are mixed with the latex. Finally, latex is pumped from the blending tanks to be packaged into drums or bulk loaded into railcars or tank trucks.

2.3 MODEL PLANTS AND EMISSIONS

Model plants are parametric descriptions of the types of plants that, for the purposes of subsequent analysis, are considered representative of those currently operating within the industry. They are based on the average ("presumptive norm") plant determined from industry responses to EPA requests for data[5,6,7,8,9,10,11] and emissions test reports.[12,13] The model plants are used to facilitate analysis of the cost, energy, and environmental impacts of reasonably available control technology (RACT). Model plants were developed for the emulsion crumb and the emulsion latex processes. This section describes the operating characteristics of the model plants, the process controls used in the industry, and the magnitude and sources of the volatile organic compound emissions.

2.3.1 Operating Parameters and Emissions

The annual capacities, capacity utilizations, and annual production rates of the model plants are presented in Table 2-2. These capacities form the basis for the remainder of the model plant parameters.

Table 2-2. STYRENE-BUTADIENE COPOLYMER MODEL PLANTS

Process	Capacity, Mg/yr	Utilization, %	Production, Mg/yr
Emulsion crumb	153,000	89	136,000
Emulsion latex	30,000	90	27,000

Based on source sampling data and industry supplied information, the VOC emissions from each of the model plants were developed. These data, as shown in Table 2-3, represent the expected annual emissions from the model plants in the absence of any add-on air pollution control equipment. Because no Federal, State, or local air pollution regulations specific to styrene-butadiene copolymer manufacturing exist, these data also represent the baseline emission levels. Process control devices which are routinely applied for reasons of material recovery or plant safety are assumed present in the model plants. The use of these control devices is reflected in the emissions listed in Table 2-3.

Table 2-3 also reflects the composition of the copolymer product and the degree of conversion of monomers to polymer. For the model crumb plant, a composition of 23.5 percent styrene and 76.5 percent butadiene by weight was assumed, because this is the usual makeup of crumb used in tires.[1] The degree of conversion does not influence emissions in a crumb plant very much because of the monomer recovery and recycle system. The model latex plant is based on the use of a hot polymerization process with a 98.5 percent conversion of monomers to polymer, and a weight ratio of 46 percent styrene and 54 percent butadiene. Emissions will vary greatly in this type of plant depending upon the percent conversion, because the unreacted butadiene is vented without any reduction due to recovery operations. Emissions will also vary for different ratios of styrene and butadiene. Latex operations with less butadiene and more styrene will have lower emission rates because butadiene is more easily volatilized than styrene. (See Appendix B for a discussion of how to estimate emissions for different copolymer composition and percent conversion.) The following sections describe the sources of the emissions and the process controls assumed to be in use on the model plants.

TABLE 2-3. MODEL PLANT PARAMETERS AND EMISSIONS

Process, Production Rate	Facility	VOC emissions, Mg/yr	Vent flow, scfm	Primary compounds	Concentration Total VOC, ppm
Emulsion crumb 136,000 Mg/yr	Monomer recovery - absorber vent	35	15	Butadiene	65,000
	Coagulation/blend tanks	57	1200[a]	Styrene	700[a]
	Dryers	328	60,000	Styrene	80
Emulsion latex 27,000 Mg/yr	Monomer removal - butadiene	224	175	Butadiene	37,000
	Monomer removal - styrene	4	30	Styrene	2,000
	Blend tanks	3	275	Styrene	150

[a] assumed flow rate and concentration from capture system collecting 90 percent of VOC emitted.

2.3.2 Emulsion Crumb Model Plant Emissions

Table 2-3 lists the four major VOC emission sources for the emulsion crumb model plant. Emissions from the organic materials' storage tanks are not discussed here because they are similar to other organic materials storage emissions for substances of similar volatility.

No attempt has been made to quantify emissions from reactor vents, because these emissions only occur during a process upset and are highly variable. The reactors are normally sealed using rupture discs and discharge to an emergency flare system in case of an extreme process upset, such as a fire in the reactor. However, based on conversations with various industry representatives. the occurrence of upset conditions in the polymerization reactor is extremely rare.

Vented emissions from the monomer recovery area consist of emissions from the absorption system (normally using kerosene as the absorbent with a typical absorption efficiency of 97 to 98 percent). Fugitive emissions from the monomer recovery system result from leaks in the pumps, compressors, valves, and flanges which handle the recovered butadiene and styrene on their way back to storage.

The large majority of VOC emissions given off during the emulsion crumb polymerization process occur during the process operations that follow the monomer recovery operations. These emissions are the result of evaporation from open blend tanks and coagulation tanks and from the dryer vents. They consist mostly of unreacted styrene and, to a lesser degree, extender oils and organic emulsifiers. It is estimated that approximately 78 percent of the total emissions from the process are emitted from the product drying portion of the manufacturing line.

2.3.3 Emulsion Latex Model Plant Emissions

Table 2-3 lists the three major VOC emission sources for the emulsion latex model plant. As was the case for the emulsion crumb model plant, emissions from organic storage tanks and reactor vents are not quantified. The reactors are normally sealed using rupture discs, with venting to a flare occurring only under extreme upset conditions within the reactor.

Monomer removal produces two vent streams in the model latex plant. Vented emissions from the blowdown process step consist mainly of unreacted butadiene, though some unreacted styrene is also emitted during this step. Because this portion of the manufacturing process is a batch operation, the venting of emissions is intermittent. Emissions from this operation account for approximately 97 percent of the VOC emissions from the emulsion latex process.

After the blowdown step, the latex is stored in process tanks. Product additives may be added in these tanks. The tanks also serve as flow-regulating holding tanks, because the remaining process operations are run on a continuous basis. The latex then flows through shaker screens to remove any large agglomerated solids. These screens are open to the atmosphere, but emissions are estimated to be very small. Concentrations of 42 ppm butadiene and 44 ppm styrene have been measured above the screens, but the air flow rate could not be determined.[13]

If the residual styrene content of the latex is greater than product specifications, the latex is processed in steam-stripping columns. The overhead styrene and water vapor stream is condensed in water-cooled condensers, and the non-condensibles are vented to the atmosphere. Table 2-3 shows that this noncondensible stream carries approximately 1.7 percent of the latex plant's VOC emissions with it. The stripped latex is then held in blend tanks, where antioxidants may be added. Evaporative emissions from the vents on the blend tanks are estimated to contribute another 1.3 percent of the plant's VOC emissions.

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3.0 EMISSION CONTROL TECHNIQUES

This chapter describes the methods and equipment available to the styrene-butadiene copolymer (SBC) industry for control of volatile organic compound (VOC) emissions. The control techniques addressed include both add-on air pollution control devices and emissions reduction by process modification.

The add-on air pollution control equipment considered for SBC manufacturing plants includes carbon adsorption, condensers, thermal and catalytic incinerators, and the compression of organic vapors into fuel manifolds. The only process modification considered is improved steam stripping.

3.1 ADD-ON EMISSIONS CONTROL DEVICES

This section describes the various add-on control devices applicable to the SBC industry. For each device the emission reduction efficiency, the advantages and disadvantages, the principle behind the device, and the factors affecting performance are described.

3.1.1 Condensers

Vapor condensation involves the removal of sufficient heat to liquify one or more of the vapors contained in the exhaust stream. The most common type of condenser for this application is a surface condenser consisting of a shell and tube heat exchanger (Figure 3-1). A coolant liquid is passed through the tubes of the exchanger while the gas stream passes on the shell side. The coolant liquid employed for easily condensed vapors is normally process or chilled water. However, special coolants such as ammonia or freon may be required for highly volatile vapors.

The advantages of using condensation as an emissions control technique for VOC emissions have been proven in other industries.[1,2] The primary advantage is that the recovered condensed liquid may be used either as fuel or

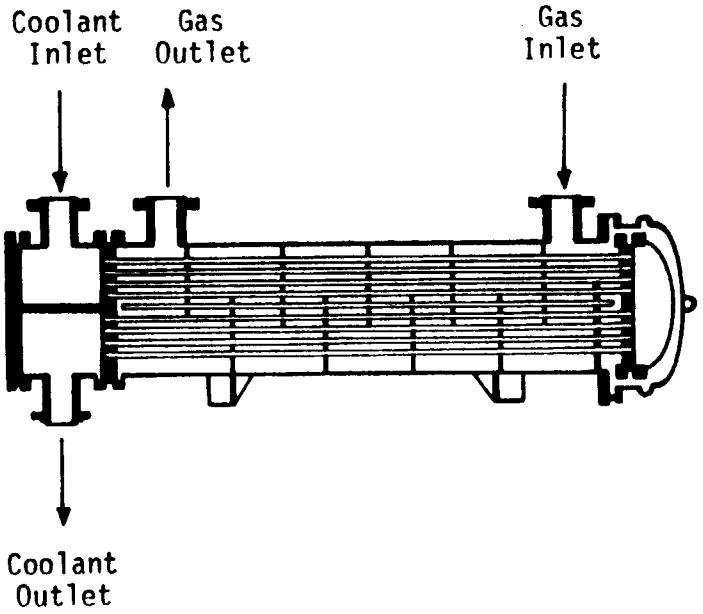

Figure 3-1. Shell and tube surface condenser.

recycled to the process. Heat exchangers can reduce heating requirements as much as 75 percent by transferring heat removed from the VOC exhaust stream to a process stream requiring heat addition.[3]

The primary disadvantage of condensation is its limited efficiency. The emissions reduction rate of a condenser is a function of the inlet VOC vapor concentration, because the outlet concentration is fixed by the temperature, pressure, and specific organic compound present. For example, an air stream at 4°C (40°F) and standard atmospheric pressure must contain over 2,800 ppm styrene before any styrene condensation will occur. This phenomenon renders condensers useless for dryer vent streams which contain less than 3,000 ppm styrene. However, condensers make excellent control devices for saturated streams (e.g., reduction of a stream containing 90 percent styrene to 6,000 ppm equals a removal efficiency of 99.3 percent). Therefore, condensers are only applicable to streams with characteristically high VOC content.

Currently, condensers are in use as process devices in all domestic SBC plants. These units are an integral part of the manufacturing process and are primarily used for organic vapor recovery to maintain desirable process economics, rather than as air pollution control devices. These condensers still provide a major reduction in VOC emissions, because the recovered organics would otherwise be released to the environment.

3.1.2 Incineration

Incineration is the oxidation (combustion) of organic vapors present in a process vent stream. The factors governing the completeness of this reaction are time, temperature, turbulence, and the type and concentration of vapor present. Combustion air must mix thoroughly with the vapor at sufficient temperatures, usually 704 to 871°C (1300 to 1600°F), and for a sufficient period of time (0.5 to 0.75 seconds) to complete the combustion reaction. Combustion of vapors resulting from SBC production results in the formation of carbon dioxide, water, carbon monoxide, and particulate.

In most applications it is necessary to provide auxiliary fuel to increase the heat content of the gas stream such that an adequate combustion temperature is reached. When sufficient VOC is present in the gas stream the combustion reaction will be self-sustaining, requiring no auxiliary fuel.

Plant insurance and safety regulations require sufficient margins of safety to ensure that the VOC vapor concentration of the gas stream remains above or below the explosive range of that vapor. This can be accomplished by keeping the stream above the explosive range by ensuring that the quantity of oxygen introduced into the gas stream (either through process requirements or inadvertent air leakage) does not create an explosive mixture or by keeping the stream below the explosive range by diluting it with ambient air. Most insurance companies require that the gas stream remain below 25 percent of the lower explosive limit in the absence of continuous VOC monitors. Concentrations as high as 50 percent of the lower explosive limit are allowed if a continuous monitor and alarm system are included in the design.

Two types of incinerators are discussed in this section, thermal (direct flame) and catalytic.

3.1.2.1 Thermal or Direct Flame Incineration. The basic thermal incinerator (Figure 3-2) consists of an insulated combustion chamber fed by the burner(s) for the vapor stream and auxiliary fuel (if necessary) and a combustion air source. The vapor-laden stream is transported by a process blower through a preheat burner to raise the stream to a temperature suitable for combustion in the combustion chamber. Based on available data, thermal incinerators can reduce VOC emissions by 98 percent, if the stream contains more than 1,000 ppm organics. For more dilute streams the reduction efficiency is governed by the fact that approximately 20 ppm is the minimum outlet concentration that can be reached.[4]

The main advantage of thermal incineration is that it can be applied to all streams. This can enhance the cost effectiveness of this option by having one large incinerator control all VOC streams. Another advantage of thermal incineration is that generated heat can be recovered using either recuperative heat exchangers or waste heat boilers. Recuperative heat exchangers preheat the incoming vapor stream to reduce the heat addition requirements for combustion. This method can reduce incinerator energy requirements by 30 to 70 percent.[5] Similarly, high temperature exhaust gas from the incinerator can be used to generate process steam using a waste heat boiler in cases where the generated steam can be used elsewhere in the SBC manufacturing process.

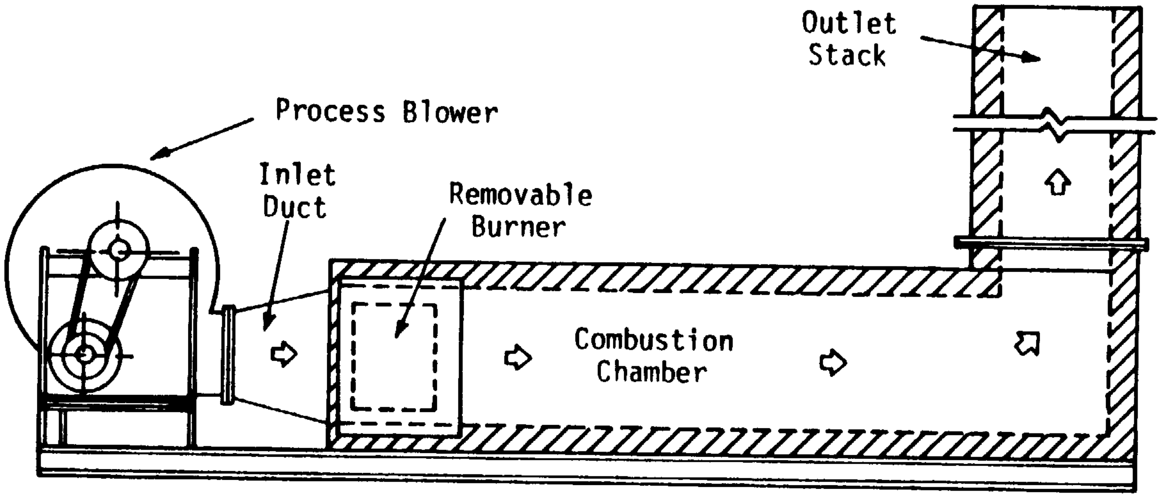

Figure 3-2. Basic thermal incinerator.

One manufacturer has developed a thermal incineration system for the oxidation of dilute VOC emission streams.[6] The oxidation occurs by passing the exhaust stream through regenerative combustion beds. Thermal recovery efficiencies of 85 to 90 percent can be achieved by this system. The system utilizes a vertical cylindrical combustion chamber surrounded by a series of packed stoneware beds. The VOC exhaust stream is preheated in a hot stoneware bed and passes through the combustion chamber, which is kept at a temperature of 760°C (1400°F). The hot combustion gases pass through other stoneware beds, transferring the heat of combustion to these beds. Inlet and exhaust valves on each bed control the gas flow as the beds are depleted or saturated with heat. This heat recovery system can substantially decrease the amount of auxiliary fuel required by the thermal incinerator.

Potential disadvantages associated with the use of thermal incineration systems at SBC plants include variations in the stream flow rate which can lead to either automatic shutdown from surges or repositioning of the flame zone in the exhaust gas entrance to the vapor control unit[7] and the possible polymerization of monomers in the vapor feed line. The flow rate problem can be minimized by proper equipment design, such as the installation of fast response temperature feedback controls for combustion air and auxiliary fuel flows to maintain constant combustion characteristics in the thermal incineration system. The monomer polymerization problem is unlikely, as long as there are no catalysts present in the lines.

One manufacturer of SBC latex has attempted to use thermal incineration as a VOC emission control technique. That particular unit was designed to reduce butadiene emissions from the monomer removal process, but problems were encountered in handling the surges of emissions from the batch process. Thermal incineration is considered a viable technique for VOC emissions reduction in the SBC industry.

3.1.2.2 Catalytic Incineration. Catalytic incineration is similar to thermal incineration in both principle and operation. The difference is that combustion is completed in a catalyst bed which reduces requirements for both temperature and residence time. The temperatures required for combustion range from 204 to 588°C (400 to 1000°F), with shorter residence times than thermal incineration.[5] The lower temperature and shorter residence time requirements can lower operating costs by as much as 30 percent compared to thermal incineration.[8]

The catalyst employed may consist of any of several compounds, deposited in thin layers on an inert substrate. The substrate is formed into specific geometric configurations designed to maximize surface area and structural integrity while simultaneously minimizing pressure drops and ensuring a uniform, evenly distributed air flow. The performance of the catalyst is dependent on the contact time and temperature, as well as the organic composition and concentration of the vapor stream.[9] Deactivation of the catalyst occurs periodically from sintering and accumulation of poisons. Although the accumulation of poisons is to some extent reversible, sintering will eventually deteriorate the catalyst, requiring its replacement. Catalyst lifetimes, in continuous use, usually range from about 1.5 to 2 years,[5] but lifetimes of up to 7 years have been reported.[9]

Catalytic incinerators have the advantages of smaller size and lowered heat requirements compared to thermal incinerators. These factors decrease both the capital costs (smaller size and lower cost materials due to lowered temperatures) and operational costs for auxiliary fuels. Similar to thermal incinerators, both recuperative heat exchangers and waste heat boilers may be employed to further reduce operating costs.

The principle disadvantage of catalytic incinerators is the high cost of the noble metals used for catalysts, which require periodic replacement. In addition to poisoning and sintering, as described earlier, the deposition of polymeric material on the catalyst could occur, decreasing the catalyst efficiency or partially plugging the catalyst bed.

Although no manufacturer of SBC has attempted to use catalytic incineration as an emissions control technique, this method is applicable to any dilute hydrocarbon stream without heavy metal poisons.[3] No proven catalytic incinerator efficiency data are available; however, catalytic incinerators should provide 90 percent removal efficiency of VOC vapors.[5]

3.1.3 Carbon Adsorption

Carbon adsorption utilizes carbon's affinity for non-polar hydrocarbons (such as styrene and butadiene) to remove these materials from a vapor stream. Hydrocarbon adsorption occurs at the surface of the carbon (including the exposed interstitial surface areas of the carbon granule) and is a physical process. Adsorption efficiency is a function of the boiling point and molecular weight of the hydrocarbon adsorbed, the residence time of the gas stream, and the capacity of the particular carbon used.[3]

A typical carbon adsorption system consists of two fixed bed carbon adsorber units. While one unit is adsorbing VOC vapors from a process exhaust stream, the VOC laden carbon in the second unit is desorbed (cleaned or stripped) through the use of low pressure steam. The steam heats the carbon bed to release the adsorbed VOC vapors, thus regenerating the activated carbon. The steam also acts as a nonflammable carrier gas for the VOC vapors. The VOC laden steam is then condensed and the two liquid phases are separated. Carbon can also be regenerated through the use of hot water or vacuum.

The major advantage of using carbon adsorption as a VOC emission control technique is that the recovered material may be recycled into the process or used as fuel. This allows for a recovery credit for the recovered material which lowers the total annualized costs for a carbon adsorption system. Carbon adsorption is very useful in that it can provide a high purity liquid organic stream for recycle from dilute VOC gas streams.

Activated carbon has a styrene capacity of 30 percent, which means that a carbon bed can adsorb up to 30 percent of its weight in styrene emissions. Activated carbon suppliers consider this a good adsorption capacity. However, styrene is difficult to desorb because of its high boiling point of 146°C (295°F) and its polymerization properties. Butadiene's low molecular weight reduces its affinity for the activated carbon, and therefore reduces the adsorption efficiency.[10,11,12] In addition to these disadvantages to using carbon adsorption as a VOC emission control technique at SBC plants, the low molecular weight extender oils present in some process emission streams may coat the surfaces of the activated carbon granules, resulting in impaired ability to adsorb VOC vapors and increased pressure drop across the carbon beds. One SBC plant has attempted to use carbon adsorption as a VOC emission control technique, with very limited success.

3.1.4 Compression of Organic Vapors to a Fuel Manifold

Organic vapor-laden streams may be injected into existing fuel manifolds or add-on burners. This may be accomplished using a positive displacement compressor to route the vapor stream into a fuel manifold or using a natural gas ejector to aspirate the vapor stream. The resulting fuel may be incinerated or used to fuel existing plant boilers.

When compressors are utilized, a means must be provided for liberating the heat generated by compression. Water-cooled compressors are available for this process.[13]

The advantages of using compression of organic vapors to a fuel manifold as an emissions control technique are that the equipment is simple and relatively inexpensive, and the heat content of the condensed organics can result in fuel cost savings for steam generation.

Disadvantages with using compression are that the vapor stream can vary in composition and concentration, resulting in varying heat content of the vapors to be compressed, and the generation of heat by the compressor can increase the possibility of explosive conditions in the compressor.

Compression of organic vapors to a fuel manifold is not currently used in the SBC industry.

3.2 PROCESS MODIFICATION

This section describes the optimization of steam stripping for the emulsion crumb polymerization process. By optimizing the steam stripping process step, more VOC can be removed from the product at this point. This results in lowering the emissions from the subsequent process steps of blending, coagulation and drying. The increased emissions from the steam stripper must be treated with one of the add-on control devices described above, but this stream would be more easily captured and controlled than those generated downstream.

In order to remove most of the unreacted styrene from the latex product from the reactors, latex and steam are countercurrently fed into a multi-plate tray tower operating at subatmospheric pressure. The steam volatilizes the styrene and carries it off overhead to a condenser.

The amount of unreacted styrene contained in the stream entering the tower is approximately 20 percent of the total styrene entering the polymerization reactors.[14,15] The concentration of unreacted styrene in the latex leaving the stripping column averages 0.3 percent of the total dry product (on a nonextended basis), based on an industry-wide mean.[16,17,18,19,20,21]

The amount of unreacted styrene removed in the steam stripping process can be increased by several methods. Based on industry comment, the residual styrene content of the dry, nonextended rubber product can be reduced approximately 27 percent to 0.24 percent by weight of the dry, nonextended product through the use of improved steam stripping.[16,17,18,19,20,21] This increased reduction can be accomplished in several ways:[17,18,22,23,24,25]

- increased steam flow rate to the stripping column;
- additional retention time in the column;
- increased steam quality;
- additional stages in the column;
- reduction in operating pressure of the column.

Reductions in the residual styrene content of the crumb beyond 0.24 weight percent can result in increased operational problems and degradation of the rubber product. Increased steaming capacity, higher operating temperatures, and lower operating pressures of the stripping column can each result in premature fouling of the tower trays through "cooking" of the rubber product onto the metal stripper surfaces.[17,18] This results in increased maintenance requirements and can also result in reduced process throughput if the plant's excess capacity cannot handle the increased downtime of the stripping columns. In addition, the possibility of degradation of the rubber product also exists due to an increase in the operating temperature of the steam strippers.

3.3 RETROFIT CONSIDERATIONS

This section discusses the problems which may be encountered in applying the emission control techniques of this chapter to existing SBC plants. Some of the problems common to all retrofit technologies are the availability and accessibility of adequate space for the control equipment, the ability of the structure to support the equipment, additional heating or cooling requirements, and the amount of process downtime needed for installation and hook-up. Additionally, there are problems specific to each industry. Those associated with the SBC industry are discussed below.

3.3.1 Dryers

The thermal dryers used for the final drying of SBC crumb employ high air flow volumes to remove 15 to 25 percent of the feed weight as water. These air streams also contain low concentrations of VOC. Because most of the control devices described previously are more effective on more concentrated streams, it is desirable to separate and control only the more concentrated portions of the total dryer air exhaust.

The separation of a slightly more concentrated stream can be accomplished for most of the existing SBC crumb dryers.[26,27] However, these streams would still be too dilute for condensation to be effective. The other control techniques outlined earlier would still be energy intensive and, therefore, very expensive.

Most of the dryers used in the industry are multi-pass apron dryers, with triple pass in particular being the most dominant. Such a dryer is shown in Figure 3-3. These dryers were originally designed so that each pass operated as a separate dryer, with the partially dried product from one pass dropping down into the next pass for further drying. In addition to this horizontal partitioning, some dryers would separate each pass into two or three zones by having insulated metal walls hang down just above the moving apron. Each zone had its own air intakes and exhausts. This provided for a greater degree of temperature control to protect the product from degradation. The exhaust streams from the wet end of the dyrer will contain more VOC per cubic foot of air than the exhaust streams from the dry end of the dryer.

Some plants have replaced the older multi-pass dryers with single pass apron dryers. A single pass dryer is shown in Figure 3-4. These dryers essentially put the separate passes of a multi-pass dryer end-to-end as one long apron. Therefore, they require more floor space than the multi-pass dryers. Each such dryer may have as many as six exhaust streams, with the streams from the wettest zones again containing higher concentrations of VOC. For energy efficiency, many of these driers are being built or modified to re-use the heated air, resulting in only one or two exhaust streams per dryer. It may be slightly more cost effective to control the first zone's exhaust from a dryer with two or more exhausts rather than all the exhaust streams.

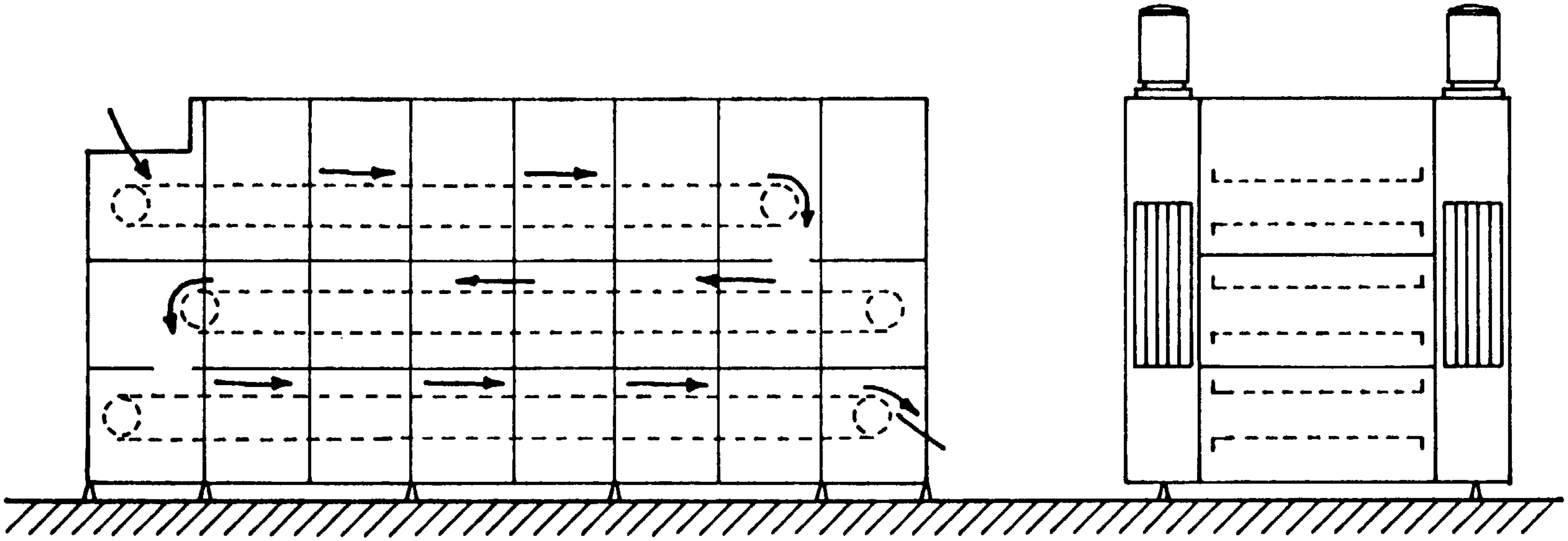

Figure 3-3. Triple-pass apron dryer.

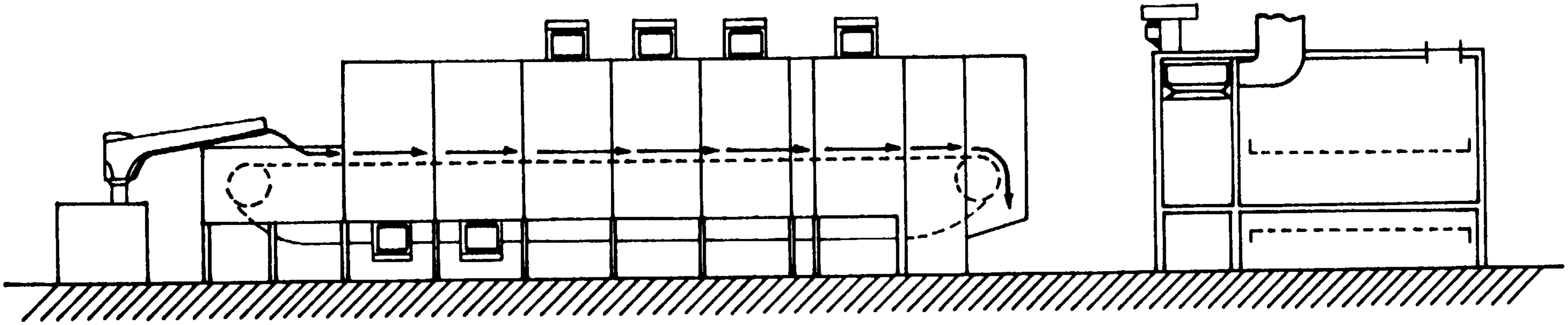

Figure 3-4. Single-pass apron dryer.

3.3.2. Steam Strippers

Improving the styrene removal efficiency of the steam strippers can be done in a number of ways, as described earlier. A major consideration here is that all of these methods will require involved testing and engineering analysis to ensure that the crumb product will not be degraded. This will result in costly engineering fees, even for a retrofit operation. Any major change to the stripping column, such as increasing the number of trays or altering the design of the trays , will require larger than normal expenditures for materials, labor, and process downtime. The minor changes, such as increasing the steam flow rate or increasing the quality of the steam, may require additional boiler capacity or even a water pretreatment operation. In addition, the maintenance required by the column will most certainly increase. All these factors significantly increase the cost for a technique which reduces emissions by only 27 percent.

3.3.3 Plant Layout

Existing SBC plants tend to separate the reactor and monomer recovery areas from the product finishing areas. This is desirable because streams within the explosive range are handled in the reactor monomer recovery areas, while the product finishing areas (coagulation, screening, drying and baling) produce much more dilute streams. Thus, if any piece of control equipment is used to control both areas, long runs of ductwork must be used. Additionally, flame arrestors will be needed in the ducts to maintain the separation of explosive and non-explosive areas.

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4.0 ENVIRONMENTAL ANALYSIS OF RACT

This chapter discusses the environmental impacts associated with the implementation of reasonably available control technology (RACT) for the styrene-butadiene copolymer (SBC) manufacturing industry. For the emulsion crumb model plant, RACT is based on an incinerator controlling the monomer recovery vent and the vapors captured from the coagulation and blend tanks through installation of a capture system designed in accordance with good engineering practice. For the emulsion latex model plant, RACT is based on an incinerator controlling the styrene and butadiene removal vents and the vapors captured from the latex blend tank through installation of a capture system designed in accordance with good engineering practice.

The beneficial and adverse impacts associated with the implementation of RACT on air pollution, water pollution, solid waste generation, and energy consumption are discussed in this chapter. These impacts are based on the model plant parameters that are presented in Tables 2-2 and 2-3 of Chapter 2.

4.1 AIR POLLUTION

The implementation of RACT would reduce volatile organic compound (VOC) emissions from emulsion crumb and emulsion latex SBC manufacturing plants in areas which are not in attainment of the ozone national ambient air quality standard. The sources of VOC emissions from emulsion crumb operations are the butadiene and styrene monomer recovery vent, evaporation from the open coagulation and blend tanks, and the vent streams from the crumb dryers. Implementation of RACT would result in controlling the butadiene and styrene monomer recovery vent to a level of 98 percent emissions reduction.[1] The open coagulation and blend tanks would be controlled to a level of 88 percent emissions reduction, and the dryers would remain uncontrolled. The resulting overall emission reduction for the emulsion crumb model plant would be 20 percent, and VOC emissions from the butadiene and styrene monomer recovery vent and the coagulation and blend tanks capture system would be reduced to 15 grams (measured as organic carbon by EPA Reference Test Method 25) per megagram of net copolymer product. Net copolymer product refers to the reacted monomer content of the copolymer product.

The sources of VOC emissions from emulsion latex operations are the butadiene blowdown tanks, the styrene stripping columns, and the latex blend tanks. Implementation of RACT would result in controlling all of these sources to a level of 98 percent emissions reduction.[1] The resulting overall emission reduction for the emulsion latex model plant would be 98 percent. However, since emissions will vary depending on the type of latex produced, implementation of RACT would reduce VOC emissions to:

$$E = (B)\,(170\,x_b + 15)$$

where:

E = Emissions in grams (measured as organic carbon by EPA Reference Test Method 25) per megagram of net copolymer product

B = Percent of charged butadiene that is unreacted

X_b = The weight fraction of butadiene in the net copolymer product

The hourly and annual VOC emissions from each model plant before and after implementation of RACT are presented in Tables 4-1 and 4-2.

The use of incineration as an emission control technique in this industry has some adverse air pollution impacts. The low BTU content of the streams to be combusted requires that natural gas be added to maintain a temperature in the incinerators sufficient for complete combustion. Combustion of natural gas produces small amounts of particulates, sulfur oxides, nitrogen oxides, hydrocarbons, and carbon monoxide.

Nitrogen oxides are the major pollutant resulting from the combustion of natural gas. The quantity of nitrogen oxides formed varies widely, depending on a number of variables, such as the amount of excess air present, the temperature of combustion, and the rate of cooling of the flue gas. Even so, the nitrogen oxides emitted by the emulsion crumb model plant as a

TABLE 4-1. HOURLY IMPACT OF RACT ON VOC EMISSIONS FROM SBC MANUFACTURING PLANTS

Model plant	Uncontrolled emissions kg/hr (lbs/hr)	Emissions with RACT kg/hr (lbs/hr)	Percent reduction due to RACT
Emulsion crumb	48 (105)	38 (84)	20
Emulsion latex	26 (58)	0.5 (1.1)	98

TABLE 4-2. ANNUAL IMPACT OF RACT ON VOC EMISSIONS FROM SBC MANUFACTURING PLANTS

Model plant	Uncontrolled emissions Mg/yr (tons/yr)	Emissions with RACT Mg/yr (tons/yr)	Percent reduction due to RACT
Emulsion crumb	420 (462)	335 (369)	20
Emulsion latex	231 (254)	5 (5)	98

result of RACT would only be from 210 to 405 kilograms (465 to 890 pounds) per year.[2] The emulsion latex model plant would emit between 270 and 520 kilograms 595 and 1,140 pounds) of nitrogen oxides per year as the result of RACT.[2]

Emissions of the other pollutants from natural gas combustion are even less significant. Only 26 kilograms (58 pounds) per year of particulates and 30 kilograms (66 pounds) per year of carbon monoxide are emitted from the emulsion crumb model plant.[2] Yearly emissions from the latex model plant are 34 kilograms (74 pounds) of particulates and 38 kilograms (84 pounds) of carbon monoxide.[2]

4.2 WATER POLLUTION

Wastewater will be generated as a result of the implementation of RACT.

4.3 SOLID WASTE DISPOSAL

No solid wastes will be generated as a result of the implementation of RACT.

4.4 ENERGY

The destruction of VOC emissions from SBC plants by incineration will require significant amounts of auxiliary fuel. In addition, electricity will be needed for the fans to move the waste gases to the incinerator. Table 4-3 lists the additional energy requirements for implementation of RACT for each of the model plants. As a point of comparison, the steam required for the drying of the crumb product in the emulsion crumb process amounts to 1100 to 1530 joules per gram (480 to 660 BTU per pound) of dry crumb.[4] No estimates are available for the amount of steam required by the stripping operations. Electrical requirements of emulsion crumb plants average an additional 1600 joules per gram (690 BTU per pound).[5,6,7,8,9] The total energy required for process operations in a latex plant is estimated at 13,600 joules per gram (5900 BTU per pound) of solids.[9,10]

TABLE 4-3. ENERGY REQUIREMENTS OF RACT

Model plant	Electricity GJ/yr (10^3 Kwhr/yr)	Natural gas 10^3 m^3/yr (10^3 ft^3/yr)	Total energy GJ/yr (10^6 Btu/yr)	Energy/unit of product J/g (Btu/lb)
Emulsion crumb	125 (35)	110 (3,875)	4,209 (3,993)	30.9 (13.3)
Emulsion latex	147 (41)	140 (4,958)	5,372 (5,097)	199.0 (85.8)

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5.0 CONTROL COST ANALYSIS OF RACT

The costs of implementing reasonably available control technology (RACT) in a styrene-butadiene copolymer manufacturing facility are presented in this chapter. The bases for both the installed capital costs and annualized costs are explained to allow the reader to develop a control cost estimate for any size facility. The emission control costs and cost effectiveness for both of the model plants presented in Chapter 2 are also developed in this chapter. All costs presented in this chapter have been updated to second quarter 1980 dollars by the use of cost indices.[1]

5.1 BASES FOR INSTALLED CAPITAL COSTS

Installed capital costs presented in this chapter represent the total investment required to install RACT in existing plants. This includes the delivered equipment cost of the control device and auxiliary equipment and the installation costs for this equipment.

For both emulsion latex and emulsion crumb plants, RACT is based on the use of a thermal incinerator capable of reducing emissions by 98 percent. A heat exchanger capable of recovering 70 percent of the flue gas heat by preheating the vapor-laden stream is included in the equipment cost. The auxiliary equipment includes the ducting, hoods, fans, and pumps needed to move the waste gas stream to the incinerator, and the instrumentation to insure proper operation. Direct installation costs include any site preparation, enclosures, or foundations necessary for the incinerator, as well as the cost of insulating, painting, and bringing utilities to the incinerator. Indirect installation costs include engineering, field expenses, construction fees, performance testing, and contingencies.

5.1.1 Delivered Equipment Costs

The delivered equipment cost is the total cost of purchasing the necessary incinerator and auxiliary equipment and having it delivered to the plant site. It consists of the purchased equipment cost, taxes, and freight. Taxes are estimated as three percent of purchased equipment cost, and freight is estimated as five percent of purchased equipment cost.[2] The bases for the purchased equipment costs are explained in the following sections.

5.1.1.1 Incinerators. The purchased equipment cost for an incinerator includes the cost of the combustion chamber, an integral fan, the burners, controls and instrumentation, and a recuperative heat exchanger. Equipment costs for these systems are available in a standard reference, "Capital and Operating Costs of Selected Air Pollution Control Systems" (EPA Cost Handbook).[3] However, this reference only presents costs for incineration systems recovering 35 percent of the flue gas heat or recovering none of the flue gas heat. Since a heat recovery of 70 percent has been shown to be economically justified for the cases considered here,[4] a supplemental reference was used to determine a "scaling factor" for estimating larger heat exchanger costs. This method was used so that the EPA Cost Handbook could be consistently used as a basis wherever possible.

The supplemental reference used was "Control Device Evaluation - Thermal Oxidation,"[5] which presents installed costs for heat exchangers capable of 30, 50, and 70 percent heat recovery for various waste gas flow rates. These costs were plotted on log-normal paper for the appropriate flow rates and yielded straight line functions. Only the ratio of the cost of a 70 percent heat exchanger to the cost of a 35 percent heat exchanger was taken from this second reference. The cost of a 35 percent heat exchanger (according to this second reference) was determined by interpolation from the above-mentioned plot. The ratio of costs was then multiplied by the cost of a 35 percent heat exchanger (according to the EPA Cost Handbook) to yield a heat exchanger cost scaled up to 70 percent heat recovery. The cost of a 35 percent heat exchanger according to the EPA Cost Handbook was assumed to be the difference in costs of the complete incineration systems for 35 percent heat recovery and for no heat recovery. The 70 percent heat exchanger cost from this scale-up was added to the cost of an incineration system with no heat recovery (from the EPA Cost Handbook) to yield the final purchased equipment cost for a system with 70 percent heat recovery.

5.1.1.2 Auxiliary Equipment. Auxiliary equipment includes the fans, motors, starters, hoods, and ductwork necessary to bring the waste gas stream to the control device and the stack necessary to vent the exhaust gases above the height of surrounding structures. The ductwork can be straight duct, elbow joints, or tees. Purchased costs for all of these components are available in the EPA Cost Handbook.[6]

5.1.2 Installation Costs

Installation costs are the costs of installing the purchased control equipment. This includes both the direct expenses such as materials and labor and the indirect expenses such as engineering costs and construction fees.

The technique used in this chapter to determine both the direct and indirect installation costs is a modified "Lang Method" of cost estimating.[2] In this method, the installation costs are computed as a percentage of the delivered equipment cost. Table 5-1 lists the cost factors for each item of installation cost for an incinerator. Adjustments must be made to some of the standard factors since they can vary widely for specific installations. Both the standard factors and the guidance for making adjustments to them are available in the EPA Cost Handbook.[7] The following sections explain the use of these adjustments for direct and indirect installation costs.

5.1.2.1 Direct Installation Costs. Direct installation costs are the costs of the materials and labor needed to install the equipment. They include costs for foundations, structural supports, erection and handling, electrical, insulation, painting, site preparation, facilities and buildings. The EPA Cost Handbook recommends adjusting the standard cost factors of Table 5-1 for erection and handling, site preparation, and facilities and buildings.[7]

The first assumption to be made about the installation of the control system is that a level and solid site is available on the plant site. No site preparation, facilities, or buildings are necessary. For erection and handling, the EPA Cost Handbook recommends an adjustment of 1.0 to 1.5 for a large system with scattered equipment and long runs of duct. As shown in Table 5-1, an adjustment of 1.2 was used, due to the fact that long runs of duct are used, but the equipment is not scattered. This yields an adjusted cost factor of 0.17 for erection and handling, and an overall factor of 0.33 for direct installation costs.

TABLE 5-1. INSTALLATION COST FACTORS FOR INCINERATORS[a]
(fraction of total delivered equipment cost)

Cost item	standard	x adjustment	= net
Direct costs			
Foundations & supports	0.08	--	0.08
Erection & handling	0.14	1.2	0.17
Electrical	0.04	--	0.04
Piping	0.02	--	0.02
Insulation	0.01	--	0.01
Painting	0.01	--	0.01
Site preparation	b	0.0	0.00
Facilities & buildings	b	0.0	0.00
Total direct costs	0.30	--	0.33
Indirect costs			
Engineering & supervision	0.10	1.0	0.10
Construction & field expenses	0.05	1.0	0.05
Construction fee	0.10	1.0	0.10
Start up	0.02	--	0.02
Performance test	0.01	--	0.01
Model study	--	--	--
Contingencies	0.03	1.0	0.03
Total indirect costs	0.31	--	0.31
Total installation costs	0.61	--	0.64

[a] Reference 2, p. 3-1 through 3-10.

[b] As required. Costs assumed to be negligible.

5.1.2.2 Indirect Installation Costs. Indirect installation costs are those costs not directly related to a specific equipment item but necessary to complete the design, construction, and startup of the control system. They include engineering and supervisory fees, construction and field expenses, the construction fee, startup expenses, performance tests, model studies, and contingencies. The EPA Cost Handbook recommends adjusting the standard cost factors shown in Table 5-1 for engineering and supervision, construction and field expenses, construction fee, and contingencies.[7]

Table 5-1 shows that no adjustments have been made to the construction fee or contingencies factors. This is because a single contractor would be able to install any one of these control devices and the process to be controlled is well understood. No adjustments have been made to the construction and field expenses factor because despite the long duct runs, these systems are only of small to medium capacity. The engineering and supervision fee was taken as the standard because of the need for automated controls to insure safe operation. The overall standard factor of 0.31 was therefore used for indirect installation costs.

5.1.3 Items Not Included in Installed Capital Costs

The installed capital costs do not include costs which could be incurred by production losses during start up of the control system or by research and development on the control system. It is assumed that these systems will be standard technology and will not require any development cost beyond the normal engineering fee. It is also assumed that the start up period will be scheduled during a routine maintenance period and, therefore, production losses will be insignificant.

No retrofit penalties have been added to account for unusual installation problems. The assumption here is that room is available outside the process buildings to locate the incinerator. No structural steel to support the incinerator on a roof is included in the costs.

5.1.4 Comparison With Actual Incurred Costs

Only one plant has used an incinerator for control of VOC emissions in the styrene-butadiene manufacturing industry. This incinerator controlled the highly variable butadiene removal stream in a latex plant, and was only half the size of the incinerator costed for the emulsion crumb model plant. The total installed cost of this incinerator was 25 percent higher than what would be estimated using the standard cost references of this chapter.[3,8,9] This difference may be due to the fact that the varying operating conditions for this incinerator required additional control instrumentation. Also another polymer manufacturing plant was tied in to this incinerator, making the system more complex and requiring over 152 meters (500 feet) of insulated ductwork.[9]

5.2 BASES FOR ANNUALIZED COSTS

Annualized costs represent the yearly cost of operating and maintaining the control system and the cost of recovering the initial capital investment. This includes annualized capital charges, direct costs, and, where appropriate, credits for product recovered. No cost for return on investment for pollution control equipment has been included.

Annualized capital charges are due to capital recovery, taxes, insurance, and administration. Direct costs are for electricity and natural gas usage and for operating and maintenance labor. No product recovery is assumed because RACT is based on a thermal incinerator. The bases for all of the annualized costs are presented in Table 5-2.

5.2.1 Annualized Capital Charges

Annualized capital charges represent the cost of owning the control system. Annualized capital charges consist of capital recovery charges and charges for taxes, insurance, and administration.

Capital recovery charges are determined by multiplying the installed capital cost by a capital recovery factor (CRF). The equation for the capital recovery factor is:

$$\text{Capital Recovery Factor} = \frac{i(1+i)^n}{(1+i)^n-1}$$

where: i = annual interest rate

n = capital recovery period = economic lifetime of capital

TABLE 5-2. BASES FOR ANNUALIZED COST ESTIMATES

Item	Cost basis
1. Annualized capital charges	
a. Capital recovery charges	0.1628[a] x installed capital cost
b. Annual charges for taxes, insurance & administration	0.04 x installed capital cost[b]
2. Direct costs[b]	
a. Operating labor	\$11.10/hr
b. Maintenance labor	\$10.90/hr
c. Electricity	\$0.0490/Kw-hr
d. Natural gas	\$2.40/1000 SCF
3. Operating hours	8760 hours/year

[a] Capital recovery factor based on 10 year life and 10 percent interest rate.

[b] Reference 12.

The capital recovery factor is a combination of depreciation and interest charges. Depreciation represents the money put aside each year to replace a system at the end of its useful life. Interest charges are the costs associated with borrowing money to pay for the system.

Taxes and insurance are each estimated as one percent of the installed capital cost, and administration costs are estimated as two percent of the installed capital cost.[10]

5.2.2 Direct Costs

Direct costs are the day-to-day expenditures required to operate the control system. This includes the utilities required to operate the system and the labor required to operate and maintain the system.

5.2.2.1 Utilities. The utilities necessary for the incinerator which RACT is based on are electricity and natural gas. Electricity is necessary to run the fans to transport the waste gases to the incinerator. Natural gas is needed to maintain a sufficient temperature for complete combustion in the incinerators.

The amount of electricity required is a function of the total pressure drop of the control stystem. The amount of natural gas needed for the incinerator is a function of the desired combustion temperature, the waste gas flowrate, the heat of combustion of the waste gas, and the amount of flue gas heat which is recovered. A heat balance around the incinerator will determine the amount of heat which must be added by the natural gas, and therefore, the amount of natural gas.

5.2.2.2 Labor. The labor necessary for the operation of an incinerator is estimated as 0.5 person-hours per eight-hour shift. The labor necessary for maintaining an incinerator is estimated as 0.5 person-hours per eight-hour shift.[11]

5.3 EMISSION CONTROL COSTS

This section will present and discuss the emission control costs of implementing RACT for both of the model plants presented in Chapter 2. Both the installed capital cost and the annualized costs for each model plant are included.

In order to determine the installed capital cost and annualized costs of control for any specific application, the purchased equipment costs must first be estimated by the procedures explained in Section 5.1.1. A number of design parameters must be specified before the cost references can be utilized. Table 5-3 presents the parameters for the model plant control systems. These parameters are based on the model plant parameters of Table 2-3, using engineering judgement and the EPA Cost Handbook.[12] The installed capital costs for each model plant are presented in Table 5-4.

The annualized costs of implementing RACT in each of the model plants are presented in Table 5-5. These costs are developed by applying the procedures explained in Section 5.2 to the design parameters of Table 5-3.

A discussion of the costs of implementing RACT for each model plant is presented below.

5.3.1 Emulsion Crumb Model Plant Costs

The installed capital cost of RACT for the emulsion crumb model plant is $316,900. This represents the total investment required to control the monomer recovery vent and the coagulation and blend tanks with one thermal incinerator. The incinerator includes a heat exchanger to preheat the incoming waste gas with the heat of the flue gas. The heat exchanger recovers 70 percent of the flue gas heat in this manner. The net annualized cost for this plant is $87,400.

5.3.2 Emulsion Latex Model Plant Costs

The installed capital cost of RACT for this plant is $331,600. This represents the total investment required to control the butadiene and styrene removal vents and the latex blend tanks with a thermal incinerator. This incinerator is also equipped with a heat exchanger to recover 70 percent of the flue gas heat by preheating the waste gas stream. The incinerator is sized for 12,000 scfm, in order to handle the surges from the butadiene vent and the dilution air that will be needed to keep the waste gas streams below 25 percent of the lower explosive limit. The annualized cost for this plant is $93,300. This is based on using the average total flow rate to calculate electricity and natural gas requirements.

TABLE 5-3. DESIGN PARAMETERS OF CONTROL SYSTEMS

Parameter	Model plant Emulsion crumb	 Emulsion latex
Production, Mg/yr (tons/yr)	136,000 (149,600)	27,000 (29,700)
Waste flowrate (scfm)	1,215	12,000 (maximum)
Average concentration (ppm)	800 butadiene 690 styrene	2,460 butadiene 40 styrene
Combustion temperature (°F)	1,600	1,600
Total pressure drop (inches H_2O)	10	10
Capture volume (of lip exhausts) (cfm per square feet of tank area)	50	--
Equivalent diameters (inches)		
Ducts, elbows, tees	8	20
Lips exhausts	48	--
Stack	8	20
Length or height (feet)		
Ducts	450	450
Lip exhausts	1	--
Stack	40	40
Number required		
Elbows	9	4
Tees	1	20
Lip exhausts	2	--
Materials of construction	1/8" stainless steel where coagulation vapors are present, including heat exchanger. 1/8" carbon steel elsewhere.	1/8" carbon steel
	(1/4" carbon steel for stacks)	

TABLE 5-4. INSTALLED CAPITAL COSTS OF IMPLEMENTING RACT[a] (June 1980 dollars)

Cost item	Model plant Emulsion crumb	 Emulsion latex
Incinerator	69,500	101,700
Heat exchanger (70% recovery)[b]	86,800	62,900
Fan	1,000	2,300
Motor and starter	500	600
Ducting	11,600	11,800
Elbows	1,800	1,500
Tees	0	2,700
Lip exhausts	4,000	--
Stack	3,700	3,700
Total FOB purchased equipment	178,900	187,200
Total delivered equipment[c]	193,200	202,200
Installation direct costs[d]	63,800	66,700
Installation indirect costs[d]	59,900	62,700
Total installed capital	316,900	331,600

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[a] Reference 15, except where noted.

[b] See Section 5.1.1.1 for discussion of methodology.

[c] Three percent for taxes, five percent for freight. Reference 2, p. 3-3.

[d] See Table 5-1.

TABLE 5-5. ANNUALIZED COSTS OF IMPLEMENTING RACT
(June 1980 dollars)

Cost item	Model plant	
	Emulsion crumb	Emulsion latex
Installed cost (4)		
Purchased equipment	178,900	202,200
Total installed	316,900	331,600
Annualized cost ($/yr)		
Annualized capital charges		
Capital recovery	51,600	54,000
Taxes, insurance, and administration	12,700	13,300
Subtotal	64,300	67,300
Direct costs		
Operating labor	6,100	6,100
Maintenance labor	6,000	6,000
Natural gas	9,300	11,900
Electricity	1,700	2,000
Subtotal	23,100	26,000
Recovery credits	--	--
Net annualized cost	87,400	93,300

5.4 COST EFFECTIVENESS

Cost effectiveness is the net annualized cost of the control system divided by the annual reduction of VOC emissions. The emissions reductions due to implementation of RACT have been computed assuming the present level of emissions control as negligible. The cost effectiveness of RACT for each of the model plants is presented in Table 5-6.

The cost effectiveness of RACT for the emulsion crumb model plant is $1,028 per megagram ($940/ton). The cost effectiveness is expected to improve for plants larger than the model, and worsen for plants smaller than the model. This is mainly due to the fact that there are economies of scale for purchasing and installing a larger incinerator which will control more VOC. Also, the operating and maintenance labor required is essentially independent of equipment size.

The cost effectiveness of RACT for the emulsion latex model plant is $413 per megagram ($375/ton). For other plant sizes, cost effectiveness is expected to vary in inverse proportion to changes in plant production. For example, the cost effectiveness of a plant only half the size of the model plant would be twice as great, or $816 per megagram ($750/ton). This is because the net annualized cost will remain essentially constant while the controlled emissions will decrease linearly with production. The net annualized cost will remain constant because the incinerator must be sized to handle the maximum surge from the butadiene removal vent. The maximum surge from this batch operation is independent of plant production. A larger plant simply produces more batches, (and therefore, more frequent surges) not larger batches.

TABLE 5-6. COST EFFECTIVENESS OF RACT

		Emulsion crumb	Emulsion latex
Net annualized cost, $		87,400	93,300
Emissions reduction,[a]	Mg/yr	85	226
	Tons/yr	93	249
Cost effectiveness,	$/Mg VOC	1,028	413
	$/Ton VOC	940	375

[a] From Table 4-2.

5.5 REFERENCES

1. Chemical Engineering. Economic Indicators. 87(21):7. October 20, 1980. 87(8):7. April 21, 1980. 85(8):7. April 22, 1978.

2. Neveril, R.B. Capital and Operating Costs of Selected Air Pollution Control Systems. U.S. Environmental Protection Agency. Research Triangle Park, North Carolina. EPA 450/5-80-002. December 1978. p. 3-2, 3-3.

3. Reference 2. p. 5-36, 5-37.

4. Memo from Mascone, D.C., EPA, to Farmer, J.R., EPA. June 11, 1980. Thermal Incinerator Performance for NSPS.

5. Blackburn, J.W. Emissions Control Options For The Synthetic Organic Chemicals Manufacturing Industry/Control Device Evaluation/Thermal Oxidation. U.S. Environmental Protection Agency. Research Triangle Park, North Carolina. Draft document under EPA contract number 68-02-2577. December 1979. 116 p.

6. Reference 2. p. 4-1 to 4-75.

7. Reference 2. p. 3-7, 3-8, 3-11.

8. Letter and attachments from Laundrie, R.W., The General Tire & Rubber Company, to Goodwin, D.R., EPA. June 6, 1978. Response to Section 114 request.

9. Telecon. Ryan, Ron, GCA/Technology Division with Novitsky, Richard, The General Tire & Rubber Company. November 5, 1980. Clarification of Section 114 response.

10. Memo from Vatavuk, W.M., EPA, to Porter, F.L., EPA. September 1980. Guidance for Developing CTGD Cost Chapters.

11. Reference 2. p. 3-14.

12. Reference 2. 277p.

APPENDIX A
EMISSION SOURCE DATA

A.1 INTRODUCTION

The purpose of this appendix is to describe the procedures used in developing emission data and to present the emission data developed for the Control Techniques Guideline (CTG) document for Control of Volatile Organic Compound (VOC) Emissions from Manufacture of Styrene-Butadiene Copolymers (SBC).

This appendix addresses VOC emissions data from two types of emulsion polymerization processes used to produce SBC. The two emulsion polymerization processes for which data are presented are the emulsion crumb and the emulsion latex processes. These processes are described in Chapter 2 of this document and a schematic flow diagram of each process is shown in Chapter 2 (Figure 2-1 shows crumb production, and Figure 2-2 shows latex production).

The primary sources of emission data presented in this appendix are industry responses to U.S. Environmental Protection Agency (EPA) requests for information under authority of Section 114 of the Clean Air Act (42 U.S.C. 7414).[1-7] Source test emission data for a latex plant are also available from an EPA sponsored source test.[8]

A.2 EMULSION CRUMB PROCESS

The sources at an emulsion crumb plant for which VOC emission data were developed are the monomer recovery operations, the coagulation and blend tanks, and the dryers. The VOC emissions from the monomer recovery operations and the dryers are considered stack emissions because they are vented through a stack or pipe. The VOC emissions from the coagulation and blend tanks are considered fugitive emissions because these tanks are open to the atmosphere and are not vented. The VOC emissions from the dryer are presented for use in determining the total emissions from an emulsion crumb plant.

A.2.1 Monomer Recovery Operations

Emission data for the monomer recovery operations at crumb plants were obtained from industry sources and were determined by source sampling tests conducted for the plant.[1-7] The emission data developed for the monomer recovery operations and used in developing this CTG are shown in Table A-1.

TABLE A-1. EMISSION DATA FOR MONOMER RECOVERY IN EMULSION CRUMB PRODUCTION[a]

Plant code	Gas flow rate, scfm	Temperature, °C	Emission factor, g VOC/kg net copolymer
A	3.6	10-16	0.08[b]
B	2.6	38	0.31[b]
C	4.0	77	0.40[b]

[a] Data obtained from information supplied by industry to D.R. Goodwin, EPA/ESED. A list of industry representatives supplying information is shown in the reference section.[1-7]

[b] Emissions determined by stack test.

A.2.2. Coagulation and Blend Tanks

Emission data for the coagulation and blend tanks at crumb plants were obtained from industry sources and were determined by calculations using material balances or engineering estimates.[1-7] The emission data developed for the coagulation and blend tanks and used in developing this CTG are shown in Table A-2.

TABLE A-2. EMISSION DATA FOR COAGULATION AND BLEND TANKS IN EMULSION CRUMB PRODUCTION[a]

Plant code	Emission factor, g VOC/kg net copolymer	
	Coagulation tanks	Blend tanks
A	0.72[b]	
B		0.16[b]
E	0.38[b,c]	

[a] Data obtained from information supplied by industry to D.R. Goodwin, EPA/ESED. A list of industry representatives supplying information is shown in the reference section.[1-7]

[b] Emissions calculated from material balance or engineering estimate.

[c] Emissions reported for coagulation and blend tanks combined.

A.2.3 Drying Process

Emission data for the drying process at crumb plants were obtained from industry sources.[1-7] These data were determined by calculations using material balances or engineering estimates and by a source test. The emission data developed for the drying process and used in developing this CTG are shown in Table A-3.

TABLE A-3. EMISSION DATA FOR DRYERS IN EMULSION CRUMB PRODUCTION[a]

Plant code	Number of vents	Gas flow rate, scfm	Temperature °C	Emission factor, g VOC/kg net copolymer
A	21	205,900 total	88	1.32[b]
B	20	150,620 total	74-93	0.51[c]
C	18	106,220 total	34-78	2.85[c]
E	unknown	8,316-11,616 per vent	60	4.94[c]

[a] Data obtained from information supplied by industry to D.R. Goodwin, EPA/ESED. A list of industry representatives supplying information is shown in the reference section.[1-7]

[b] Emissions calculated from material balance or engineering estimate.

[c] Emissions determined by stack test.

A.3 EMULSION LATEX PROCESS

The sources at an emulsion latex plant for which VOC emission data were developed are the monomer removal stream from the blowdown tank, the styrene steam strippers, and the blend tanks. The VOC emissions from the blowdown tank and the steam strippers are considered stack emissions because they are vented through a stack. The VOC emissions from the blend tanks are considered fugitive emissions because the tanks are open to the atmosphere.

For the emulsion latex model plant presented in Chapter 2, emission data were developed based on emission data obtained from industry and from an emission source test conducted by EPA.[5,8] The emission data developed for the emulsion latex model plant have been reviewed and confirmed by three industry sources as being representative of their plants.[9-11] The emission data developed for sources at emulsion latex plants and used in the development of this CTG are shown in Table A-4.

TABLE A-4. EMISSION DATA FOR EMULSION LATEX PLANTS[a]

	Emission Source		
Parameter	Blowdown tank	Steam strippers	Blend tanks
Average flow rate of vented stream (scfm)	175	30	275
Average VOC concentration of vented stream (ppm)	37,000	2,000	150
VOC emissions (Mg/yr)	224	4	3
Maximum flow rate of vented stream (scfm)	1,000	30	350
Maximum VOC concentration of vented stream (%)	60	1	<0.1
Purge stream used	steam nitrogen	steam nitrogen	air

[a] Capacity: 30,000 Mg/yr.
Production: 27,000 Mg/yr (90% utilization)
Hours of operation: 8,400 hr/yr
SBC latex composition: 54% butadiene, 46% styrene

A.4 REFERENCES

1. Letter and attachments from Rivette, D., Texas-U.S. Chemical Company, to Goodwin, D.R., EPA. May 27, 1978. Response to Section 114 request.
2. Letter and attachments from Ballard, B.F., Phillips Petroleum Company, to Goodwin, D.R., EPA. May 25, 1978. Response to Section 114 request.
3. Letter and attachments from Kaminski, K.J., B.F. Goodrich Chemical Company, to Goodwin, D.R., EPA. June 2, 1978. Response to Section 114 request.
4. Letter and attachments from Grantham, A.C., Copolymer Rubber & Chemical Corporation, to Goodwin, D.R., EPA. May 22, 1978. Response to Section 114 request.
5. Letter and attachments from Laundrie, R.W., The General Tire & Rubber Company, to Goodwin, D.R., EPA. June 6, 1978. Response to Section 114 request.
6. Letter and attachments from Burkett, E.J., The Goodyear Tire & Rubber Company, to Goodwin, D.R., EPA. July 5, 1978. Response to Section 114 request.
7. Letter and attachments from McGraw, J.L., American Synthetic Rubber Corp., to Goodwin, D.R., EPA. May 25, 1978. Response to Section 114 request.
8. Draft Emissions Test Report, General Tire Plant. U.S. Environmental Protection Agency. Research Triangle Park, N.C. EMB Report No. 79-RBM-4.
9. Letter from Kulka, A.R., Reichhold Chemicals, Inc., to Goodwin, D.R., EPA. November 14, 1980. Response to Section 114 request.
10. Letter from Arnold, S.L., Dow Chemical U.S.A., to Goodwin, D.R., EPA. December 31, 1980. Response to Section 114 request.
11. Letter from Stark, F.J., Rubber Research Elastomerics, Inc., to Goodwin, D.R., EPA. January 8, 1981. Response to Section 114 request.

APPENDIX B

EXAMPLE CALCULATIONS FOR DETERMINING REDUCTION IN EMISSIONS FROM IMPLEMENTATION OF RACT

B.1 INTRODUCTION

The purpose of this appendix is to present and explain the use of emission factors in developing emission inventories and emission reductions estimates. These factors can be used for development of emission inventories and emission reductions estimates when more specific information such as a source test or a material balance study is not available. This appendix presents an emission factor for emulsion crumb operations and an emission equation for emulsion latex operations in the absence of any RACT controls. Example calculations of the emission reductions due to RACT are shown for a crumb plant and a latex plant.

B.2 EMULSION CRUMB EMISSION FACTORS

The VOC emissions from an emulsion crumb plant in the absence of any RACT controls can be estimated from the following emission factors:

Monomer recovery vent	0.26 grams butadiene/kg net copolymer
Coagulation and blend tanks	0.42 grams styrene/kg net copolymer
Dryers	2.41 grams styrene/kg net copolymer
Total	3.09 grams VOC/kg net copolymer

These emission factors are based upon the data presented in Tables A-1 through A-3 of Appendix A. The only control device assumed to be used in developing the emission factors is a kerosene absorber on the monomer recovery vent. Although some plants use different absorbing materials and different regeneration methods, control efficiencies are usually between 95 and 98 percent.

B.3 EMULSION LATEX EMISSION EQUATION

The emission equation given in this section is for latex plants which strip unreacted monomers from the product without trying to recover these materials for reuse. These plants usually use a hot polymerization reaction (50°C) to achieve a high degree of conversion (greater than 95 percent). Emissions from latex plants which have monomer recovery operations can be estimated by using the monomer recovery vent emission factor for emulsion crumb plants.

The magnitude of emissions from latex plants without recovery operations or RACT controls varies greatly with two factors: the degree of conversion of the reaction and the butadiene content of the product. This is because all of the unreacted butadiene will be vented. Approximately 90 percent of the removed styrene will be condensed along with the steam carrying it, and this condensate will usually go to a wastewater pond. Some styrene will also remain in the final latex product. Industry data shows that styrene emissions as a percentage of the styrene purchased are only one-fifteenth of butadiene emissions as a percentage of the butadiene purchased.[1] The following formula for the emission factor accounts for different butadiene contents and percent conversions:

$$\text{Emission factor} = (100 - \text{percent conversion})\,(9.33\,X_b + 0.67)\,\frac{\text{g VOC}}{\text{kg net copolymer}}$$

where X_b = weight fraction of butadiene in net copolymer, and percent conversion is based on butadiene only.

This formula can be derived from the assumptions of the preceding paragraph as shown below.

- 100 - (percent conversion of butadiene) = percent butadiene unreacted = percent of purchased butadiene emitted to the atmosphere.
- 1% of purchased butadiene emitted = 10 grams butadiene emitted/kilogram butadiene purchased.
- Therefore, (100 - percent conversion)(10) = g butadiene emitted/kg butadiene purchased.

Using the fact that styrene emissions as a percentage of the styrene purchased are only one-fifteenth of the analogous percentage for butadiene, the following expresses the magnitude of styrene emissions:

$$(100 - \text{percent conversion})\ (10)/(15) = \text{g styrene emitted/kg styrene purchased.}$$

Neglecting the small difference in weight of styrene plus butadiene purchased and net copolymer produced, the two expressions above can be converted to a net copolymer basis by using the weight fractions of butadiene and styrene in the copolymer.

- $\frac{\text{g butadiene emitted}}{\text{kg butadiene purchased}} \times \frac{\text{kg butadiene purchased}}{\text{kg net copolymer}} = \frac{\text{g butadiene emitted}}{\text{kg net copolymer}}$
- $\frac{\text{kg butadiene purchased}}{\text{kg net copolymer}} = \text{Weight fraction butadiene in copolymer} = X_b$
- $\frac{\text{g styrene emitted}}{\text{kg styrene purchased}} \times \frac{\text{kg styrene purchased}}{\text{kg net copolymer}} = \frac{\text{g styrene emitted}}{\text{kg net copolymer}}$
- $\frac{\text{kg styrene purchased}}{\text{kg net copolymer}} = \text{Weight fraction styrene in copolymer} = 1\text{-}X_b$

$$\text{Total emissions}\left(\frac{\text{g VOC}}{\text{kg net copolymer}}\right) = \frac{\text{butadiene emissions}}{\text{kg net copolymer}} + \frac{\text{styrene emissions}}{\text{kg net copolymer}}$$
$$= (100 - \text{percent conversion})\ (10)(X_b) + (100 - \text{percent coversion})\ (10)(1\text{-}X_b)/(15)$$
$$= (100 - \text{percent conversion})\ (10X_b + 10/15 - 10X_b/15)$$
$$= (100 - \text{percent conversion})\ (9.33X_b + 0.67)$$

B.4 EXAMPLE CALCULATIONS OF EMISSION REDUCTION FROM IMPLEMENTATION OF RACT

The reduction in VOC emissions from the implementation of RACT can be determined by using the emission factors and emission equation of the preceding sections in conjunction with an assumed capture efficiency and destruction efficiency for RACT. The emission factor is multiplied by the production rate of a crumb plant to yield the mass of VOC emissions before RACT for that plant. For a latex plant, the emission equation is used to determine the proper emission factor for that plant, and this emission factor is then multiplied by the production rate to yield the mass of VOC emissions before RACT for the latex plant. These total emissions for either type of plant are multiplied by the percent of emissions which are captured and by the percent destruction or recovery of the control device to yield the emissions reduction due to implementation of RACT. Because RACT is based on a thermal incinerator, a destruction efficiency of 98 percent will be used in the calculations that follow. A collection efficiency of 90 percent for the open coagulation and blend tanks of the crumb plant will also be assumed. All other collection efficiencies are assumed to be 100 percent. Example calculations for both a crumb plant and a latex plant are shown below.

B.4.1 Emulsion Crumb Calculations

From a plant's records the following monomer usages are determined for a base year:

Butadiene	31×10^6 gallons
Styrene	14×10^6 gallons

The total weight of net copolymer is determined by converting the gallons used to kilograms and adding. (Because yields are usually greater than 98 percent, losses will be neglected for this calculation.)

Butadiene	31×10^6 gallons/yr x 2.36 kg/gallon	$= 72 \times 10^6$ kg/yr
Styrene	14×10^6 gallons/yr x 3.42 kg/gallon	$= 48 \times 10^6$ kg/yr
	Net copolymer	$=120 \times 10^6$ kg/yr

The mass of VOC emissions before RACT is determined by multiplying this production rate by the emission factor given in section B.2. The factor for total emissions is 3.09 grams VOC/kg net copolymer. Total emissions before RACT are therefore $3.09 \times 120 \times 10^6$ grams VOC = 371×10^6 grams VOC = 371 Mg VOC.

Total plant emissions after the implementation of RACT are calculated by reducing the captured emissions by the destruction efficiency of the control device. This can be done most easily be calculating a new emission factor for each source. Emissions which are not controlled by RACT are the dryers and those not captured by the hoods on the coagulation and blend tanks. The total emission factor after RACT is computed as follows, assuming 10 percent of coagulation and blend tank emissions are not captured, and 98 percent destruction of VOC is achieved by an incinerator.

Emission factor = controlled emissions + uncontrolled emissions

Emission factor = [0.26 + (0.90)(0.42)] (0.02) + [(0.10)(0.42) + (2.41)]

Emission factor = 0.01 + 2.45 = 2.46 grams VOC/kg net copolymer

Total emissions after RACT are therefore 2.46 g/kg x 120 x 10^6 kg/yr = 295 x 10^6 g/yr VOC = 295 Mg/yr VOC. The emissions reduction due to the implementation of RACT would therefore be 371 - 295 = 76 Mg/yr VOC. These calculations are summarized in Table B-1.

TABLE B-1. EMULSION CRUMB PLANT EMISSION REDUCTION FROM IMPLEMENTATION OF RACT

Annual production of net copolymer (kg)	120 x 10^6
Emission factor before RACT (g/kg)	3.09
Annual emissions before RACT (Mg)	371
Emission factor after RACT (g/kg)	2.46
Annual emissions after RACT (Mg)	295
Annual emissions reduction due to RACT (Mg)	76

B.4.2 Emulsion Latex Calculations

From a latex-producing plant's records the following monomer usages are determined for a base year:

Butadiene	3.4 x 10^6 gallons
Styrene	4.6 x 10^6 gallons

The plant's test data of unstripped latex samples shows that a 98 percent conversion of butadiene to copolymer is usually achieved.

The total weight of net copolymer is determined in the same manner as in the emulsion crumb case, above.

Butadiene 3.4×10^6 gallons/yr x 2.36 kg/gallon = 8×10^6 kg/yr
Styrene 4.6×10^6 gallons/yr x 3.42 kg/gallon = 16×10^6 kg/yr

Net copolymer = 24×10^6 kg/yr

In order to determine the emission factor for this plant, the weight fraction of butadiene in the copolymer must be known. This is available from the above weights as 8 ÷ 24 = 0.33. The emission factor can now be determined from the emission equation given in section B.3.

$$\text{Emission factor} = (100\text{-percent conversion})(9.33\ X_b + 0.67)\ \frac{\text{g VOC}}{\text{kg net copolymer}}$$

$$= (100-98)[9.33(0.33) + 0.67] = 7.56\ \frac{\text{g VOC}}{\text{kg net copolymer}}$$

The mass of VOC emissions before RACT is determined by multiplying this emission factor by the production rate found earlier. Total emissions before RACT are therefore $7.56 \times 24 \times 10^6$ grams VOC = 181×10^6 grams VOC = 181 Mg VOC.

Total emissions after the implementation of RACT are calculated by using the destruction efficiency of the control device. All of the emissions are assumed to be captured from the monomer stripping vents and the blend tanks. The emission factor after RACT would be just two percent of the uncontrolled emission factor if a 98 percent destruction were achieved by the incinerator.

Emission factor = (7.56)(0.02) = 0.15 grams VOC/kg net copolymer

Total emissions after RACT are determined by multiplying the production rate of the plant by this emission factor. Total emissions after RACT are therefore 0.15 g/kg x 24×10^6 kg/yr = 3.6×10^6 g/yr VOC = 3.6 Mg/yr VOC. The emissions reduction due to the implementation of RACT would therefore be 181 - 3.6 = 177.4 Mg/yr VOC. These calculations for the example latex plant are summarized in Table B-2.

TABLE B-2. EMULSION LATEX PLANT EMISSION REDUCTION FROM IMPLEMENTATION OF RACT

Annual production of net copolymer (kg)	24×10^6
Emission factor before RACT (g/kg)	7.56[a]
Annual emissions before RACT (Mg)	181
Emission factor after RACT (g/kg)	0.15[a]
Annual emissions after RACT (Mg)	3.6
Annual emissions reduction due to RACT (Mg)	177.4

[a] For a hot polymerization process latex plant with 98 percent conversion of butadiene to copolymer and a weight fraction of butadiene in copolymer of 0.33.

B.5 REFERENCES

1. Letter and attachments from Laundrie, R.W., The General Tire & Rubber Company, to Goodwin, D.R., EPA. June 6, 1978. Response to Section 114 request.