The Production of Methanol and Hydrogen Fuels from Municipal Solid Waste

Jeffrey S. Chen

PU/CEES Report No. 289

March 1995

Submitted in partial fulfillment of the requirements for the degree of
Master of Science in Engineering from Princeton University

Center for Energy and Environmental Studies
School of Engineering and Applied Science
Princeton University
Princeton, New Jersey 08544-5263

Financial support for this research was provided by Argonne National Laboratory, the Energy Foundation, the Air and Energy Engineering Research Laboratory of the US Environmental Protection Agency, the W. Alton Jones Foundation, and the Merck Fund

Abstract

This thesis assesses the feasibility of producing methanol and hydrogen from municipal solid waste (MSW) to be used as fuel for fuel cell vehicles (FCVs). Replacing the gasoline fueled internal combustion engine vehicle (ICEV) with the FCV could dramatically reduce urban air pollution, greenhouse gas emissions, and petroleum imports.

Fuels production from MSW is similar to fuels production from biomass, which has been examined by others. After removing the inorganic material, MSW is very similar in composition to biomass. For both feeds, the gasifier is the only non-commercial component in the fuel production system. Three distinct gasifier designs, all of which have been tested with MSW, are analyzed here. The thermodynamics of producing fuel from MSW are analyzed using chemical process simulation software and pinch analysis. The calculated overall fuels production energy efficiencies are similar for the three gasifiers. The thermal efficiency is roughly 50% for methanol production and 60% for hydrogen production. These efficiencies are only slightly lower than the efficiencies that have been calculated elsewhere for biomass feeds.

Emissions measurements of one of the gasifiers analyzed show that it emits fewer gaseous pollutants than modern incinerators and produces liquid and solid wastes that are nonhazardous. Additional empirical measurements of gasifiers' environmental performances are needed to lend confidence to the generalizability of this result.

Using a baseline set of assumptions, the costs of producing fuels from MSW are competitive with biomass feeds (assuming $2.5/GJ biomass cost) with tipping fees of $20-$60/tonne, depending on the gasifier technology used. In some cases with tipping fees lower than the average landfill fee for the Northeast ($70/tonne), fuels

from MSW will compete with fuels from natural gas (at the future gas price of $3.8/GJ).

Using vehicle performance and cost characteristics developed by others, the total levelized lifecycle cost of driving the FCV using MeOH or H_2 produced from MSW is calculated to be lower than the levelized cost of an ICEV fueled by gasoline because the FCV would be much more efficient than the ICEV.

Acknowledgements

First of all, I wish to thank the Center for Energy and Environmental Studies for letting me tackle such important research and giving me support throughout my project. Many people at the CEES and the Department of Mechanical and Aerospace Engineering have contributed to this work.

I especially thank my three advisors for their assistance and confidence in my work. Dr. Eric Larson patiently reviewed many revisions of this report and helped strengthen the thesis with his tireless effort and constant encouragement. Dr. Robert Williams contributed to this thesis and my education through his guidance, insight, enthusiasm, and thoughtful feedback. Professor Robert Socolow gave the thesis direction with his invaluable questions and constructive comments.

I am also grateful to Professor Judy Wornat, Dr. Ernst Worrell, Dr. Joan Ogden, Dr. Sivan Kartha, and Dr. Valerie Thomas for reviewing parts of the thesis and providing useful suggestions. I also want to acknowledge Ryan Katofsky for helping me get started on this project.

Many people outside of Princeton also contributed to this research, especially Dr. Durai-Swamy from MTCI, Mark Paisley from Battelle Columbus Laboratories, David Runyon from Thermoselect, Inc., Jim Solomon and Bob Moore from Air Products, Inc., Gary Fenner from Praxair, Paul Mazzaccaro from the Mid-Connecticut Resource Recovery Facility, and the many helpful people at the California Integrated Waste Management Board.

Finally, I thank my family and all my friends for helping me keep my sanity and making my experience at Princeton fun and memorable.

This thesis carries the identification 2024-T in the records of the Department of Mechanical and Aerospace Engineering.

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Glossary and List of Abbreviations

bbl: barrel of oil = 42 gallons

BIG/GT: Biomass integrated gasification/gas turbine power plant

BPEV: Battery powered electric vehicle

CIWMB: California Integrated Waste Management Board

Cellulose: The main chemical component of much biomass that is a high molecular weight hydrocarbon polymer

Char: Solid carbon produced during gasification

Cold Gas Efficiency: A performance measure of a gasifier defined as the higher heating value (HHV) of the product gas divided by the HHV of the input feed

EPA: Environmental Protection Agency

Energy Ratio (ER): A performance parameter defined for the fuel production process as the higher heating value (HHV) of the fuel output divided by the HHV of the input feed

Exajoule (EJ): 10^{18} joules

Exergy (or Availability): The maximum useful work that can be extracted from a substance

FCV: Fuel cell vehicle

Gigajoule (GJ): 10^9 joules

Higher Heating Value (HHV): A thermodynamic measure of the energy content of a fuel when the H_2O in the products of combustion is in the liquid state

ICEV: Internal combustion engine vehicle

IGCC: Integrated gasification combined cycle power plant

Lower Heating Value (LHV): A thermodynamic measure of the energy content of a fuel when the H_2O in the products of combustion is in its gaseous state

MAF: Moisture- and ash-free

MSW: Municipal solid waste

Makeup gas: The fresh feed gas to a methanol synthesis reactor

Neat alcohol fuels: 100% alcohol fuels

Normal Cubic Meter (Nm^3): The volume of a gas at 273.15 K and 101,325 Pa

PCDD: Polychlorinated dibenzo-p-dioxin (dioxin)

PCDF: Polychlorinated dibenzo-furan (furan)

PM: Particulate matter

PEMFC: Proton exchange membrane fuel cell. A fuel cell that is promising for vehicular applications. The PEMFC has an operating temperature of roughly 100 °C.

Petajoule (PJ): 10^{15} joules

Refuse-Derived Fuel (RDF): Pre-processed MSW that has been shredded and has had much of the inorganic material removed.

STIG: Steam injected gas turbine

STLC: Soluble threshold limit concentrations. These are the maximum concentrations allowed for given hazardous materials (e.g., heavy metals) in the leachate of a substance. If the concentrations of hazardous materials in the leachate exceed the STLC, the substance is treated as hazardous by the EPA

Standard Cubic Foot (scf): The volume of a gas at 60 °F and 14.7 psia.

Synthesis Gas: A gas that is primarily CO and H_2 that is produced either through reforming natural gas or gasifying solid carbonaceous matter, like coal, biomass, or MSW.

TCLP: Toxicity characteristics leaching procedure. Test currently used by the EPA to extract the leachate from a potentially hazardous substance (e.g. ash). If the leachate from the TCLP exceeds the STLC for any materials, then the substance is treated as a hazardous substance

Thermal Efficiency (TE): A performance parameter defined for the fuel production process as the higher heating value (HHV) of the fuel output divided by the HHV of all the system energy inputs.

Volatile Matter: The components of solid carbonaceous matter that are converted into gases or tars/oils during pyrolysis.

Zero Emission Vehicle (ZEV): An automobile that emits no tailpipe emissions (CO, NOx, unburned hydrocarbons).

Chapter 1: Introduction

1.1 Introductory Remarks

Municipal solid waste (MSW) management is a contentious issue. Siting landfills is becoming increasingly difficult due to public opposition and more strict regulation of liquid effluents. Incinerators are perceived to emit toxic pollutants into the air. Though recycling targets have been set in a few states, millions of tonnes of garbage still must be safely disposed of every year in the US.

An alternative to conventional disposal measures is thermochemical gasification of MSW with conversion of the resulting gas into hydrogen or methanol for use as transportation fuels. These fuels are suitable for use in a fuel cell vehicle (FCV), which is a low polluting and highly efficient future alternative to the internal combustion engine vehicle (ICEV). An assessment of fuels production from MSW is the focus of this thesis.

1.2 Alternatives to the ICEV

1.2.1 Motivations to find alternative vehicles and fuels

Today, petroleum is the dominant transportation energy resource, and the ICEV is the dominant vehicle transportation technology. However, concerns of energy security, pollution, and greenhouse gas (GHG) emissions are driving the development of alternative fuels and vehicle technologies.

Urban air pollution is today the primary driver of efforts to develop alternatives to the gasoline fueled ICEV. Carbon monoxide (CO), unburned hydrocarbons (UHCs), and nitrogen oxides (NO_x) are emitted from the tailpipes of gasoline fueled ICEVs. CO is a noxious gas that is dangerous to humans. NOx and UHCs react in sunlight to form photochemical smog. To prevent the further deterioration of urban air quality, dramatic changes in automobile emissions standards are being imposed. In

California for example, radical reductions in tailpipe emissions have been mandated. In 1998, 2% of new cars must be "zero-emission vehicles" (ZEVs). 10% of new autos must be ZEVs by 2003. Twelve states in the Northeast are considering similar standards. Many cities around the globe would benefit from similar measures.

Security of petroleum supplies is also a concern for the future of transportation. Much of the world's petroleum is currently supplied by the politically volatile Middle East, and it is expected that this share will rise because there are limited supplies of low cost oil outside the Middle East. Though it has been over 20 years since the 1973 oil crisis, US petroleum supplies are not secure. The 1991 war in the Persian Gulf and the recent tension between Iraq and the US are merely reminders of this fact. Nevertheless, over half of the petroleum consumed in the US today is imported, compared to 39% in 1973 (EIA, 1993).

Finally, global warming is a seemingly insurmountable challenge for the future of the automobile. At present the transportation sector in the US accounts for over 30% of the nation's carbon emissions to the atmosphere (EIA, 1993). Should the world community decide to stabilize greenhouse gas levels, drastic reductions in carbon dioxide emissions will be required.[1]

1.2.2 The Fuel Cell Vehicle

A promising alternative technology to the ICEV with the potential to address environmental and security problems is the fuel cell vehicle. The fuel cell is an electrochemical device that converts the chemical energy of a fuel directly into electricity. It works in a manner similar to a battery, but is constantly supplied with fuel. These devices were developed in the 1960s to provide reliable on-board electricity to rockets in the Gemini and Apollo programs. They were too expensive for

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[1] The Intergovernmental Panel on Climate Change (IPCC) suggests a 60% reduction in carbon emissions would be required (IPCC, 1990).

broad commercial use, but further development has brought down the costs of production dramatically, especially during the last ten years. Now, the fuel cell is seriously being considered for road transportation as a replacement for the IC engine.

Of the many different types of fuel cells, the proton exchange membrane fuel cell (PEMFC), which operates at about 100 °C, is currently the leading contender for vehicular applications. The fuel cell is made up of an anode and a cathode separated by an electrolyte (Figure 1.1). Fuel (hydrogen for the PEMFC) is delivered to the anode, where it splits into two ions and two electrons over a catalyst. At the cathode, or oxidant side, oxygen catalytically reacts with two hydrogen ions and two electrons to form water. The electrolyte is permeable to the hydrogen ion (proton) but impermeable to the electrons and to hydrogen and oxygen. Thus, the hydrogen ions travel through the electrolyte while the electrons are delivered as current to an external load. The process can be simplified into the reactions:

$$H_2 \rightarrow 2H^+ + 2e^- \qquad \text{(anode)}$$

$$2H^+ + 2e^- + \frac{1}{2}O_2 \rightarrow H_2O \qquad \text{(cathode)}$$

Fuel cells are much more efficient than IC engines because conversion to electricity does not require the intermediate generation of heat, and efficiency is not limited by the Carnot constraint applicable to heat engines. Hydrogen fuel cells can theoretically achieve thermodynamic efficiencies over 80% (Kartha and Grimes, 1994) and have demonstrated over 40% efficiencies in a prototype bus developed by Ballard (Ballard, 1993; Prater, 1992). For comparison, an IC engine used in today's automobiles has an average thermal efficiency of about 15% over an urban driving cycle. Since peak efficiencies of internal combustion engines are greater than 30%, this large difference is due largely to the fact that for fuel cells, the efficiency increases as the load is reduced, in contrast to the decline in efficiency of the internal combustion engine with decreasing load. This makes it possible for a fuel cell car to be 2.5 - 3 times as fuel efficient as an internal combustion engine car, when performance is averaged over the entire driving cycle, since most of the time a car is operated at a small fraction of its peak output.

Besides providing a tremendous improvement in efficiency, a FCV fueled with pure hydrogen will produce no local pollutants. Water is the only product. The California tailpipe emissions mandates are powerful drivers of FCV development. Though the battery powered electric vehicle (BPEV) is a candidate as a ZEV, slow recharging times and limited travel ranges make it less attractive than FCVs.[2] Other attractive features of fuel cells are that they have no moving parts and designs that could be realized by the turn of the century will also be compact.

The hydrogen fuel takes up more space than a gasoline tank because of its low volumetric energy density. Many methods of storing hydrogen have been proposed: as a compressed gas, as a cryogenic liquid, or in metal hydrides. An alternative to direct hydrogen storage is storing a hydrogen carrier, which could be a liquid hydrocarbon. A liquid carrier would be easier to store and handle. The drawback is that the hydrogen carrier would have to be transformed into hydrogen gas and CO_2 on board the vehicle. Methanol is an excellent hydrogen carrier because it is relatively easy to steam reform onboard the vehicle.[3] Partial oxidation (POX) reformers could be used to reform a wide range of liquid fuels, but they are less energy efficient than steam reformers.[4]

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[2] The GM Impact, which is a super lightweight electric car that uses lead acid batteries, has a highway travel range of 90 miles (150 km). Recharging requires 2 to 3 hours (GM, 1993)

[3] Methanol can be steam reformed into CO_2 and H_2 at 250°C. The needed heat is provided by burning the unconsumed hydrogen in the "spent fuel" exhausted from the anode. Ethanol, another candidate FCV fuel, requires temperatures of over 500°C to be reformed. Since the PEMFC operates around 100°C, raising the fuel temperature to 500°C would incur an energy penalty.

[4] In a POX, fuel is burned in fuel-rich conditions generating a mixture of CO, H_2, CO_2, and N_2. The CO and H_2O can be combined in shift reactors to form more CO_2 and H_2. The shift reactor products can then be fed to the fuel cell anode. Since it is difficult to use the unconsumed hydrogen in the anode exhaust, POX converters are less energy efficient than steam refomers.

More efficient FCVs that operate on steam reformed methanol or pure hydrogen will eventually enter the market. However, when FCVs are initially introduced, POX reformers will likely be used since they allow FCVs to operate on conventional hydrocarbon fuels.

Methanol and hydrogen are far less attractive when used as neat fuels in ICEVs. The vehicle range of an ICEV running on H_2 is limited by the low efficiency of the ICEV and the low volumetric energy density of hydrogen. Methanol used in ICEVs has not shown significant emission benefits in comparison to gasoline fuels (Calvert et al., 1993). Also MeOH has a lower energy density than gasoline.

1.3 Biomass and MSW as feedstock for fuel production

Both hydrogen and methanol can be produced from fossil fuels. Today, hydrogen and methanol are primarily produced from natural gas. If fuel markets develop for methanol or hydrogen, natural gas will likely be the initial feedstock of choice, since it will be the least expensive option. Natural gas prices are projected to rise, however, and at sufficiently high prices, other energy sources, including coal and biomass, could become competitive feedstocks for fuel production.

Methanol and hydrogen from biomass would typically be only half as costly as hydrogen derived electrolytically from wind or advanced photovoltaic power sources (Ogden and Nitsch, 1993), and from nuclear power as well (Williams, 1993). Moreover, recent studies indicate that methanol and hydrogen produced from biomass would cost about the same as these fuels produced from coal (Wyman et al., 1993; Katofsky, 1993; Williams et al., 1994). But biomass feedstocks offer substantial global warming benefits. While coal-derived methanol and hydrogen would generate substantial lifecycle carbon dioxide emissions, lifecycle emissions would be small for biomass grown sustainably on dedicated energy plantations: an equivalent amount of CO_2 released into the atmosphere by gasification or combustion would be captured through photosynthesis in growing the biomass.

The favorable results obtained in analyses of fuels production from biomass have led to the current study of MSW feedstocks. The feedstock costs of biomass, natural gas, and coal represent a large fraction of the total costs of fuel production.

Municipal solid waste (MSW) has a negative cost at present because communities, businesses, and institutions pay a waste management facility to dispose of their refuse. This characteristic makes MSW a potentially interesting feedstock for fuels production.

Table 1.1 - Comparison of the Chemical Compositions (dry basis) of Wood and MSW

	% mass	
	Wood[a]	MSW[b]
Carbon	50.65	36.02
Hydrogen	6.18	4.88
Oxygen	43.01	26.51
Nitrogen	0.06	0.62
Ash	0.10	31.97

Notes:
a) Douglas fir. Taken from Jenkins (1989)
b) from Herrmann (1989)

How similar are MSW and biomass? An analysis of MSW in comparison to biomass is displayed in Table 1.1. Not shown in the table are inorganic components and trace constituents, of which there are many in MSW. Also, the composition of MSW can vary greatly over time at the same collection point and from one collection point to another.

Table 1.2 - Comparison of Wood and MSW on a Moisture & Ash-Free (MAF) Basis

	% mass	
	Wood	MSW
Carbon	51	53
Hydrogen	6	7
Oxygen	43	39
Nitrogen	0	1

Removal of the non-organic fraction and mixing of the MSW to form a more uniform feedstock is a well established process, known as RDF (refuse-derived fuel) production. Once the non-organics are removed, MSW shows a remarkable similarity to biomass (Table 1.2).

1.4 Solid Waste Management

Though the prospect of using MSW as a feedstock for fuel production may be interesting from an energy perspective, it is imperative to examine issues of solid waste management and how MSW gasification and fuels production would differ from conventional MSW disposal means.

1.4.1 Overview of Solid Waste Management

Though it has been relatively easy to dispose of organic wastes in rural areas, refuse management has been historically more difficult for large cities. In many densely populated areas, it was common practice to dump refuse into streets or waterways in earlier eras. This was the case for ancient Rome, and the practice persisted in many cities even through the nineteenth century (Wilson, 1977). Some cities collected refuse in town dumps outside of the municipal center to concentrate the unpleasant sights and odors away from residences. With growing urbanization, greater amounts of solid waste needed to be disposed. It was not until the mid-19th century that the public took aggressive measures to manage the garbage, after direct connections were made between trash disposal and disease and the contamination of water supplies (Neal and Schubel, 1987).

Since the first ordinance dealing with solid waste in the U.S. was enacted in 1795 by the Corporation of Georgetown (Washington DC), prohibiting the dumping of refuse in streets, solid waste management has mainly been a local issue in this country. The first refuse collection scheme financed by the public began in 1856 in Washington DC. Throughout the 19th century, most cities, including New York, Philadelphia, Chicago, and Boston, carted their trash to open dumps. Some coastal cities would dump non-recyclable waste and municipal sludge into the ocean.

Burning refuse in open dumps has been practiced since the middle ages to increase the lifetimes of landfills. But this practice often led to uncontrolled fires lasting for days and producing smoke and unpleasant stenches that would reach the communities. The year-1874 marks the construction of the first closed incinerator in England. In 1885, the first US incinerators were built, operating in Allegheny, PA and Governor's Island, NY. This was perceived at the time to be the most sanitary way of reducing the volumes of solid waste, but air emissions from these simple incinerators were a public health problem that was eventually recognized as such (Neal and Schubel, 1987).

In the past century, great improvements have been made in the technologies for solid waste disposal. The internal combustion engine has made it possible to collect garbage by truck and haul it over longer distances more easily. Tractors have made it possible to move large amounts of dirt to dig and cover landfills. With modern incinerators operated at high temperatures and fitted with emission controls, air pollution has been reduced. Even with these advances in technology, however, harmful effects of solid waste disposal remain.

The 1976 Resources Conservation and Reclamation Act (RCRA) set up rules and guidelines for handling hazardous and non-hazardous wastes. This statute was enacted after studies revealed that the nation annually discarded 135 million tons of MSW and spent between $4 - $6 billion on MSW collection and disposal. Though the burden of nonhazardous waste disposal is placed on each state, RCRA institutionalized federal support for managing waste in an environmentally sound manner.

Some 80% of the nation's present landfills are projected to reach capacity within the next twenty years (OTA, 1989), and new landfills are difficult to site. To address this problem, education and awareness programs along with integrated recycling and alternative disposal measures are being implemented across the country.

In general, recycling is a top priority for state waste management boards. Source reduction is another option that is getting some attention. Though source reduction reduces the amount of MSW that ends up in a landfill, few market incentives exist in the US to promote this measure. Suggestions have been made to place the burden of waste disposal on manufacturers, which could help reduce the amount of packaging that must be discarded. But these types of programs have not yet been implemented in the US. Much rhetoric has been dedicated to the subject; however, per capita MSW generation has risen in the last decade. Recycling has gained support as a primary waste reduction measure. Community source separation and curbside pickups, which were practiced in only a handful of municipalities fifteen years ago, are now in place throughout the country. Many states have radical reduction schedules legislated to divert MSW from landfills. California leads the way, mandating a 50% reduction goal by the year-2000 (CIWMB, 1992). Other states, like Connecticut, have specified recycling goals. These legislative measures help decrease the amount of MSW dumped into landfills but must be coupled with proper market development of recycled materials and price structures to be truly effective. Even with extensive recycling programs, however, there will still be massive amounts of waste that need to be disposed.

1.4.2 Waste Disposal Technologies

The primary methods currently used for disposing of MSW are landfilling and incineration. Other methods that are not discussed here include composting, anaerobic digestion, aerobic digestion, and ocean dumping.

Landfills

Landfilling is the oldest form of MSW disposal. Today, over 80% of US MSW is still disposed of in landfills. This is the only method for final disposal of wastes; combustors, digestors, and composting facilities all produce some sort of solid residue that is discarded to a landfill.

Landfills are basically large pits into which solid wastes are dumped. Within the last century, problems of foul odors, groundwater contamination, and gathering of vermin led to the development of sanitary landfill techniques. This engineering method consists of spreading the MSW in a thin layer at the landfill and covering it with soil every day. When the landfill reaches a certain height, it is covered with several meters of dirt and closed down. These sanitary landfills greatly reduce odor and pest problems, but they do not completely eliminate them.

Well managed landfills use sophisticated methods to protect surface and groundwater supplies. Water collection systems prevent surface runoff from the landfill, which could pollute surface waters. The greatest concern, however, is that the leachate from the landfill will seep into the groundwater. The leachate is the liquid effluent from the MSW that can absorb contaminants as it percolates through the landfill. Advanced synthetic linings can prevent the leachate from leaving the landfill. The contaminated water must then be collected and treated separately. Clay liners impede the leachate flow and provide a large surface area to collect and treat some of the toxic constituents.

Gas retrieval systems recover gases produced by the decomposing organic wastes in the anaerobic environment of the landfill. This landfill gas consists of methane, carbon dioxide, water vapor, and other trace components. At high levels, the methane-rich landfill gas poses a fire and explosion hazard at the landfill. Usually these gases are collected and flared, but many landfills are now producing heat and power with landfill gas.

Selecting a good site for a landfill involves studying the geology and topography of the region as well as considering its distance from the serviced communities. The site's hydrological characteristics, soil conditions, environmental conditions, and potential use when closed must all be taken into account. Naturally, sufficient land must be available, and the public must approve its use for the landfill to be sited (Tchobanoglous et al., 1977). If the site meets all the specified requirements of being available and far from ground or surface water supplies, and if the waste is primarily biodegradable, a well managed landfill can be an efficient and economical method of waste disposal.

However, there is significant public opposition to landfills because of the offensive odors and possible contamination of water supplies. As more restrictions are placed on landfill leachate and gas emissions and finding available sites gets more difficult, the costs of setting up landfills will increase.

Incinerators

Incinerators are large furnaces that burn MSW to reduce the volume and mass of the waste. Incinerators have been used since the 19th century, but coupling incinerators with heat recovery systems was not practiced until the 1920s (Neal and Schubel, 1987). Incinerators that recover the energy from the MSW are known in the industry as Waste-to-Energy systems (WTEs). Small scale apartment-type combustors were predominantly used in urban areas through the 1960s and were not regulated for emissions. At that time 33% of all disposed MSW was burned (US EPA, 1992). These were shut down in the 1970s and 1980s due to pollution regulations. Interest in large scale WTEs was sparked by the 1970s energy crisis, and WTE construction rapidly rose in the 1980s. In 1991, 137 WTEs burned 18% of the nations disposed MSW. The construction of WTEs, however, has recently come to a halt due to renewed health concerns (Schneider, 1994).

Modern WTEs have various designs. They can be generally categorized as mass burn, modular, and RDF burn incinerators.

The mass burn incinerator is a large scale incinerator that usually burns over 200 tonnes/day of MSW. At this scale, mass burn incinerators must be field-erected. They typically handle several counties or waste management districts. Another characteristic is that they combust waste that has been minimally processed. MSW is delivered to the mass burn facility into a pit or floor. A loader or crane will break up and mix the larger waste components. The waste is dropped into the furnace onto a grate. Various proprietary technologies move, mix, and shuffle the MSW on the bottom grate. Underfire air is blown in from under the grate. Over-fire air is also blown in from above the grate to increase turbulence within the chamber. The total amount of air blown into the reactor is much greater than stoichiometric to provide good mixing and ensure as close an approach to complete combustion as possible. The temperature within the incinerator is usually kept around 1000°C with the intent of destroying pathogens and large organic molecules.

Modular incinerators burn raw MSW, like mass burn incinerators, but they are much smaller so they are not field erected. Most parts of modular incinerators are pre-fabricated and manufactured off-site, like the furnace or boiler. The distinction between modular and mass burn facilities, however, is becoming more blurred because many components of mass burn plants, like the steam generator or superheater, are manufactured off-site.

Like mass burn incinerators, RDF facilities are usually very large (>700 tonnes/day). Most of these facilities have an on-site RDF production facility integrated with the incinerator/boiler. RDF can be burned in conventional spreader stoker, suspension, or fluidized bed combustors. These combustors do not need the proprietary grate designs that are used for mass burn incinerators. In fact, RDF facilities often make use of already existing old boilers and co-fire RDF with wood or coal. Although RDF is easier handle than raw MSW, the high costs of producing RDF can make the RDF facility economically unattractive. Which type of incinerator is most economical in any given case depends on a variety of local factors.

WTEs have reduced their air pollution emissions as federal and state environmental legislation has become more stringent. Now that more is understood about the various pollutants in the incinerator, guidelines for good combustion practice have been developed to destroy toxic substances. The Clean Air Act Amendments of 1970 specified New Source Performance Standards (NSPS) for particulate matter emissions from combustors. In 1991, the NSPS were amended for large incinerators, adding CO, NOx, acid gases, and dioxins as regulated pollutants. Amendments to the NSPS regulating heavy metal emissions are expected. Also, dioxin/furan emissions from MSW incinerators are a topic of debate and may be regulated more stringently.

Currently at issue is the solid residue (ash) from the incinerators. Incinerator ash has been treated in the past as non-hazardous waste and was managed as such.

Table 1.3 - Toxicity Test Data of MSW Incinerator Combined Ash Leachate for Lead and Cadmium[a]

	Lead	Cadmium
Number of samples analyzed	366	272
Number of samples over the toxicity limit[b]	171	54
Percent of samples over the toxicity limit	47%	20%

Notes:

a) Data from 26 facilities were taken from a database compiled by the Environmental Defense Fund (Denison and Silbergeld, 1988).

b) Ash leachate was taken using the Extraction Procedure (EP) Toxicity test. The US EPA defines a substance to be a hazardous waste when the EP test yields concentrations greater than 5.0 mg/liter for lead and 1.0 mg/liter for cadmium.

In 1994, the Supreme Court ruled in EDF vs. City of Chicago that incinerator ash would no longer be exempt from hazardous waste regulations. Standard toxicity tests must be run on the ash to see if the ash or ash leachate is considered hazardous. Table 1.3 shows sampling tests of ash toxicity characteristics. If the ash must be disposed as a hazardous waste, the costs of incineration will dramatically rise. Hazardous waste disposal can be three to ten times more expensive than non-hazardous waste disposal. Though incinerators can greatly reduce volumes of MSW, safety, health, and economic issues have raised public concern and opposition against them.

1.5 Summary of This Study

MSW disposal and petroleum fuel consumption are issues of concern today. Methanol and hydrogen production from municipal solid waste (MSW) could be a means of safely disposing MSW and producing domestic fuel supplies for transportation systems that produce little or no air pollution.

This thesis examines the technical and economic feasibility of methanol and hydrogen production from MSW. Consistent thermodynamic and economic frameworks are used to compare fuel production using three different MSW gasification technologies. These are consistent with the study by Katofsky (1993) on fuels production from biomass, coal, and natural gas, enabling some comparisons to be made. The environmental characteristics of each of the three gasifiers are discussed and compared to conventional incinerator technologies.

Chapter 2 motivates this report by examining the characteristics and generation trends of MSW. The chapter ends by providing a resource assessment of MSW in the US on a national, state, and county basis. Chapter 3 discusses principles of gasification and feed preparation. It also gives a history of MSW gasifiers and supplies basic information about the three gasifier technologies analyzed in this thesis. Chapter 4 explains in detail the entire fuel production process and the various model assumptions used to simulate the thermochemical process. Chapter 5 details the results of the thermodynamic analysis of the three processes. Chapter 6 examines some of the environmental characteristics of gasifiers, with comparisons to incineration. Chapter 7 investigates the economics of fuels production from MSW, with comparisons to others' results with other feedstocks. Chapter 8 summarizes the thesis and provides recommendations for further work.

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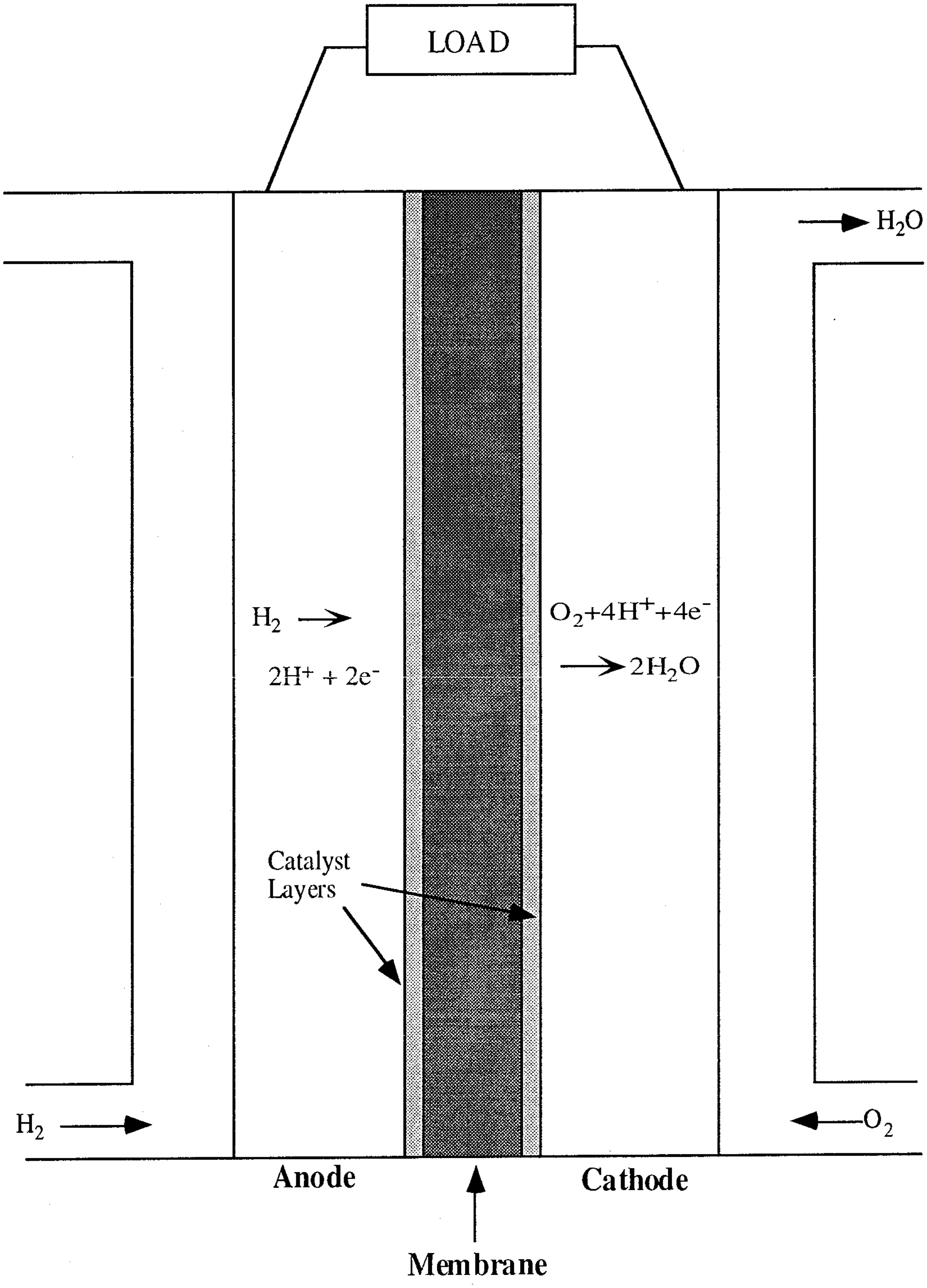

Figure 1.1 - Fuel Cell Schematic

Chapter 2: MSW Feedstock Analysis and Resource Assessment

How much fuel can municipal solid wastes (MSW) provide? To answer this question, the MSW energy resources for the United States are assessed on a national, state, and county level. California and Los Angeles County have been chosen for detailed examination because these regions are leading candidates for fuel cell vehicles owing to tight air quality regulations there.

2.1 Overview

2.1.1 What is MSW?

MSW is defined by the U.S. Environmental Protection Agency (EPA) as solid wastes that are "durable goods, nondurable goods, containers and packaging, food waste, yard wastes, and miscellaneous inorganic wastes from residential, commercial, institutional, and industrial sources." This includes household trash like appliances, clothing, newspaper, etc. According to the EPA, municipal solid waste is only one of several subcategories of wastes defined by Subtitle D of the Resource Conservation and Recovery Act (RCRA), which deals with the management of non-hazardous wastes (figure 2.1).[1] Agricultural, construction, and industrial wastes are *not* defined as MSW, but they are often discarded in the same disposal facilities along with MSW.

Therefore, given the ambiguity of the EPA definition, resource assessments of MSW must be clear in defining what is included in the estimates. Since MSW resource estimates and composition studies are mostly taken from actual landfills, incinerators, and recycling centers, they include solid wastes that are not considered MSW by the EPA definition. To be consistent with most of the available data, MSW is defined in this thesis as the solid wastes that are handled in MSW processing facilities.

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[1] Subtitle C of RCRA is the framework established to govern hazardous waste management.

2.1.2 Definition of Terms

Some frequently used terms associated with MSW are generation, recovery and disposal. *Generation* is defined as the MSW that is produced before any of it is recovered or combusted. Generated MSW is also referred to as "gross discards" in past literature. *Recovery* is defined as the MSW that is recovered for recycling or composting. Recovered material does not equal the amount of recycled and composted material because recovery processes usually produce some residues. *Disposed* MSW, also referred to as the "net discards," is the MSW that is not recovered, which is essentially all the MSW that goes to the landfill or incinerator. Incinerating MSW to generate process steam or electricity is not considered recovery (U.S. EPA, 1992).

2.2 MSW Disposal and Composition Estimates

2.2.1 Methods of Estimating MSW

Two primary methodologies are reported in the literature for estimating the quantities and composition of the MSW generated, recovered, and disposed. The first methodology is based on taking samples, sorting and weighing the various components of MSW, which provides an estimate of the MSW composition. Records of the weigh stations at MSW disposal facilities give the total weight of waste disposed. This methodology is good for site-specific cases and is necessary for a local waste management project. Many samples, however, are necessary to prevent skewed results. For estimates on a national or state-wide scale, it would be very costly to take consistent samples from various places around the country over an entire year and gather weigh station data from every MSW disposal facility.

The second method is known as the material flows methodology, which has been used and developed by the EPA and Franklin Associates, Ltd. for over 20 years to estimate national MSW disposal trends (US EPA, 1992). This method uses a model based on production and end use data from the Department of Commerce, trade and manufacturing associations, and elsewhere. After adjusting for product lifetimes, imports and exports, and other factors, the model determines the total quantity of MSW generated and the average composition for a given year. MSW recovery and disposal are estimated on a material-by-material and product-by-product basis and aggregated for total findings. The model results have been corroborated with various sampling studies and have shown close agreement (US EPA, 1992).

2.2.2 Estimates for Three Cases

Table 2.1 and figure 2.2 show the quantities and composition of MSW in 1990 for the U.S., California (CA), and Los Angeles (L.A.). The estimates for these three cases will later be used to determine MSW's potential as an energy resource. The estimate for the U.S. case is based on the EPA's material flows model, thereby including only wastes defined by the EPA as MSW. The California and Los Angeles MSW disposal estimates are consistent with the definition given earlier in section 2.1.2, including all the wastes discarded in landfills or incinerators. The data are from the California Integrated Waste Management Board (CIWMB), based on an aggregation of independent county-level estimates, which in turn are based on a mix of landfill data, site samples, and models. The CA and L.A. figures are used here because the CIWMB maintains one of the most comprehensive MSW databases.

The per capita MSW disposal rate for California is double the national rate in part because the CIWMB includes more types of waste in their estimates than the EPA does. By using the EPA definition, the figures for the US case underestimate the amount of MSW disposed. Unfortunately, no detailed analyses have been made on a national level of EPA's other categories of subtitle D wastes. A brief study in 1986 of all subtitle D wastes estimated roughly the amount of waste disposed for each of the EPA categories (Table 2.2). That analysis shows that much more industrial, oil and gas, and mining wastes are generated than MSW. But most of these wastes are

Table 2.1 - 1990 MSW Generation, Recovery, and Disposal (Three cases)

	USA case[a]				California case[b]				Los Angeles case[b]			
	million tonnes Generated	% Recovered	million tonnes Disposed	% Comp. Disp.	million tonnes Gen.	% Recov.	million tonnes Disp.	% Comp. Disp.	million tonnes Gen.	% Recov.	million tonnes Disp.	% Comp. Disp.
TOTAL	177.5	17.1	147.2		40.89	11.6	36.16		11.22	16.3	9.40	
Paper and Paperboard	66.5	28.5	47.5	32.3	12.75	19.0	10.32	28.5	4.02	27.4	2.92	31.0
Glass	12.0	19.9	9.6	6.5	1.74	27.9	1.26	3.5	0.50	28.8	0.36	3.8
Total Metals	14.7	23.0	11.3	7.7	2.01	7.4	1.86	5.2	0.59	6.4	0.55	5.9
Plastics	14.7	2.2	14.4	9.8	2.55	2.7	2.49	6.9	0.78	5.5	0.74	7.8
Rubber, leather, textile	9.3	4.4	8.8	6.0	1.28	7.4	1.20	3.3	0.55	7.1	0.51	5.4
Wood Wastes	11.2	3.2	10.8	7.3	3.50	12.4	3.06	8.5	0.93	11.0	0.83	8.8
Other	2.9	23.8	2.2	1.5	1.80	2.8	1.75	4.8	0.38	1.1	0.37	3.9
Food Waste	12.0	0.0	12.0	8.1	2.88	10.8	2.57	7.1	1.14	17.8	0.94	10.0
Yard Waste	31.8	12.0	27.9	19.0	5.80	8.9	5.29	14.6	1.33	10.2	1.19	12.7
Misc. Inorganic	2.6	0.0	2.6	1.8	6.56	21.0	6.37	17.6	1.01	2.6	1.00	10.6
kg/capita/day disposal	1.6				3.3				2.9			

Notes: estimates on as received (wet) basis
a) USA estimates from US EPA (1992)
b) CA and L.A. estimates from CIWMB (1992)

separately handled and are not disposed of in MSW landfills or incinerators. Since it is unclear how much of these subtitle D wastes are disposed with MSW, the EPA figures of table 2.1 remain as the best estimates of MSW generation and disposal on a national level.

Table 2.2 - Estimated Annual Generation Rates for Various Subtitle D Wastes

Waste Category	**Generation (million tonnes)**
Industrial non-hazardous waste[a]	6900
Oil and Gas Waste[b]	
drilling	117 - 790
produced waters	1780 - 2480
Mining waste[c]	1300
MSW[d]	149
Utility Wastes[e]	77
Construction and Demolition Waste[f]	29
Municipal Sewage Sludge[g]	9
Incinerator ash[h]	3 - 7
Small quantity generator hazardous wastes[i]	0.2
Agricultural Wastes[j]	unknown

(from U.S. EPA, 1988)

Notes:

a) Based on surveys of waste from major industries, including pulp and paper, iron and steel, chemical, food products, textiles, etc. (1986 data)

b) Residues from the exploration and production of petroleum and gas. (1985 data)

c) Waste generated from extraction, beneficiation, smelting, and refining of metals, minerals, shale, etc. (1985 data)

d) Municipal solid waste *generated* in 1985 as estimated by the EPA.

e) Wastes generated from the combustion of fuels for power generation. (1984 data)

f) C&D wastes include concrete, asphalt, stone, wallboard, glass, piping and other building materials. (1970 data)

g) Data from American Water Works Association. Sludge is disposed in lagoons and landfills, incinerators, the ocean, and land applications (1986 data)

h) 1988 data of incinerator ash generation. This might be considered hazardous waste in the future.

i) Hazardous waste from industries that generate less than 100 kg per month. These industries include vehicle maintenance shops, laundries, photo processors, laboratories, etc. (1986 data)

j) Includes crop residues, irrigation wastes, and manure.

2.2.3 Reliability of Composition Estimates

MSW composition varies considerably. On a local level, each residence, business, or institution generates waste of a different composition, i.e. - restaurants dispose mostly food wastes whereas an office complex discards mainly paper products. Thorough mixing of the refuse can alleviate this local irregularity, and is practiced in modern incinerators to make the MSW feedstock more homogeneous.

Table 2.3 - MSW Disposal Compositions in Different Regions

	% composition by mass	
Material	Santa Clara	Stanislaus
Paper	40%	24%
Plastics	11%	7%
Glass	3%	4%
Metals	6%	7%
Yard	13%	17%
Food	9%	12%
Other[a]	18%	29%
per capita disposal (kg/capita-day)	3.5	2.6

Notes: (from CIWMB, 1992)
a) Other includes wood wastes, textiles, rubber, inert solids, bulky items, and other miscellaneous wastes.

On a regional level, differences in quantity and composition of waste are evident in some data. For example, Table 2.3 shows MSW estimates from urbanized Santa Clara, CA and rural Stanislaus, a neighboring county which includes a national park.

Table 2.4 - Seasonal Variation in MSW Composition

		% composition by mass			
Material	Summer	Fall	Winter	Spring	Avg.
Paper	31.0	39.9	42.2	36.5	37.4
Yard	27.1	6.2	0.4	14.4	13.9
Food	17.7	22.7	24.1	20.8	20.0
Glass	7.5	9.6	10.2	8.8	9.8
Metal	7.0	9.1	9.7	8.2	8.4
Wood	2.6	3.4	3.6	3.1	3.1
Textiles	1.8	2.5	2.7	2.2	2.2
Rubber & Leather	1.1	1.4	1.5	1.2	1.2
Plastics	1.1	1.2	1.4	1.1	1.4
Misc.	3.1	4.0	4.2	3.7	3.4

Notes: from Niessen (1972) Compositional estimates are averages of MSW studies conducted in municipalities throughout the United States in 1970.

The time of the year plays another important role in determining MSW composition. Table 2.4 displays the seasonal changes in MSW composition. People stay inside and buy more packaged goods in the winter, but they produce more food and yard wastes in the spring and summer.

Table 2.5 compares US, Japanese, and German MSW. The MSW composition is different for each country, with the US disposing of more plastics and paper than the others. Also, the US generates and disposes of substantially more waste per capita.

Table 2.5 - Comparison of International MSW Generation

	% composition of total weight		
Material	U.S.A.[a]	Japan[b]	Germany[c]
Paper	37%	46%	18%
Glass	8%	1%	9%
Metals	10%	1%	3%
Plastics	7%	8%	5%
Other	38%	44%	64%
Total generation (10^6 tonnes)	149	48	19
per capita generation (kg/year)	597	394	318

Notes:
a) U.S. composition and total generation are based on 1985 data (U.S. EPA, 1992)
b) Japanese composition based on 1989 estimates. Total generation estimate is for 1988 (World Resources Institute, 1993)
c) (West) German composition is from 1985 estimates. Total generation estimate is for 1987 (World Resources Institute, 1993)

Japan and many European nations are much more waste conscious because they have less space into which they can dump their trash.[2]

Though MSW compositions vary between regions, seasons, and sectors, some underlying patterns can be identified. The compositions of the L.A., CA, and US disposed waste have many important similarities. The paper fraction in each case is about 30%, and the plastics range between 7-10% of the total waste stream (Figure 2.2). Though the metal and glass percentages are higher for the US, the total inorganic fractions are higher for CA and L.A. The U.S. has an inorganic fraction of 16%, whereas the total inorganic fraction is 20% for L.A. and 26% for CA. The low inorganic fraction for the US case reflects the use of the EPA's definition of MSW,

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[2] Less developed countries (LDCs) generate much less MSW and have fewer non-organic materials in their MSW than OECD nations because they do not use as much expensive metal or glass containers and are economically driven to re-use materials (Niessen, 1978).

which does not include many materials like inert solids, sludges, and ashes that are dumped into MSW landfills.

Table 2.6 compares the MSW composition of the US, CA, and L.A. cases (Table 2.1) with other studies and estimates. The L.A., CA, and U.S. compositions all fall within the bounds of the nine Office of Technology Assessment (OTA) samples, the two Franklin Associates studies, and the Cornell assessment. But the 40 studies by OTA and the analyses by Hickman and National Ecology Inc. (NEI) show much higher paper fractions and lower wood waste fractions.[3]

It cannot be overemphasized that MSW streams show significant variation, and care must be taken in any MSW management strategy. The nonhomogeneous characteristic of MSW is especially important for a waste treatment facility like a Waste-to-Energy (WTE) facility. Since each component of MSW has different physical and chemical characteristics, wide variations in the MSW composition will affect the heat rate and temperature of a WTE. Of greater concern, a large change in MSW composition can damage waste processing equipment, e.g. - a sudden increase in sulfur concentrations (from tires).

Though the U.S., CA, or L.A. cases give good rough estimates of MSW composition and have been corroborated with several sampling studies, they can be very different from a local MSW stream. Local sampling studies are necessary to properly analyze the MSW feedstock in preparation for the construction of a specific treatment facility.

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[3] The NEI analysis was probably based on winter samples or urban MSW because the yard and food waste fractions are especially low.

Table 2.6 - Comparison of MSW Compositions by Material Components

Material Component	**% composition by weight**									
	Three cases									
	US[a]	CA[b]	L.A.[b]	*OTA*[c]		*Franklin Assoc*[d]		*NEI*	*Cornell*	*Hickman*
	1990	1990	1990	9 local studies	40 studies	9 studies	8 studies	analysis[e]	study[f]	study[g]
Paper and Paperboard	32.3	28.5	31.0	29.9 - 45.0	36.5 - 54.7	29.9 - 45.9	29.1 - 43.8	45.5	36.0	43.0
Glass	6.5	3.5	3.8	3.6 - 12.9	6 - 13.7	3.6 - 12.9	3.3 - 5.9	11.5	8.0	9.0
Metals	7.7	5.2	5.9	1.5 - 9.4	4 - 14.7	1.5 - 9.4	4.4 - 8.8	7.6	9.0	7.0
Plastics	9.8	6.9	7.8	5.3 - 12.6	2 - 9.0	5.3 - 12.6	6.3 - 10.2	7.2	7.0	3.0
Rubber	2.7	1.2	1.4	0 - 1		1.1 - 7.2	3.2 - 5.6	1.9		2.0
Textiles	3.3	2.2	4.0	1.1 - 6.2				3.1	2.0	3.0
Wood	7.3	8.5	8.8	0.7 - 8.2	0.7 - 5	0.7 - 8.2	4.5 - 15.1	0.8	4.0	3.0
Food	8.1	7.1	10.0	1.3 - 28.8	0.9 - 18.2	1.3 - 28.8	6.5 - 9.8	3.7	9.0	10.0
Yard	19.0	14.6	12.7	0.0 - 39.7	0.4 - 25	0.0 - 39.7	5.1 - 19.8	1.1	20.0	10.0
Diapers					0.5 - 2.9					
Total Other wastes				3.8 - 16.2	0.5 - 10	3.8 - 16.6			5.0	
Other combustibles	1.5	4.8	3.9					17.5		
Other Inorganics	1.8	17.6	10.6							10.0

Notes:

a) U.S. estimates from the U.S. EPA (1992)
b) Estimates from CIWMB (1992)
c) From the Office of Technology Assessment (1989)
d) Franklin Associates' 8 studies are from 1984-87. 9 studies are from 1987-89 (U.S.EPA, 1992)
e) National Ecology Inc. (NEI) analysis (Herrmann, 1989)
f) From Cornell Waste Management Institute (Ditz, 1990)
g) From the *Waste-to-Energy Compendium* (1988).

2.3 Trends and Projections of MSW Disposal Composition and Quantities

MSW generation, recovery, and disposal trends over the past thirty years have shown that the U.S. MSW composition has changed (Table 2.7).[4] Importantly, the organic fraction has increased (Figure 2.3), as a result of increased recycling and decreased use of glass and metals for packaging. The largest component change has

Table 2.7 - Disposal Composition Trends

	million tonnes						
Material	1960	1965	1970	1975	1980	1985	1990
Paper	22.2	29.3	33.4	31.6	38.8	43.9	47.5
Glass	6.0	7.8	11.3	11.9	12.9	11.1	9.6
Metals	9.4	9.7	12.4	12.3	12.1	11.5	11.3
Plastics	0.4	1.3	2.8	4.1	7.1	10.4	14.4
Rubber & Leather	1.5	2.1	2.6	3.4	3.8	3.3	4.0
Textiles	1.5	1.7	1.8	2.0	2.4	2.5	4.8
Wood	2.7	3.2	3.6	4.0	6.1	7.4	10.8
Other	0.1	0.0	0.5	1.2	2.2	2.6	2.2
Food	11.1	11.5	11.6	12.2	12.0	12.0	12.0
Yard	18.1	19.6	21.0	22.9	24.9	27.2	27.9
Misc. inorganics	1.2	1.5	1.6	1.8	2.0	2.3	2.6
Total	74.3	87.6	102.8	107.2	124.2	134.3	147.2
Material	% of disposal						
Paper	29.9	33.4	32.5	29.4	31.3	32.7	32.3
Glass	8.1	8.9	11.0	11.1	10.4	8.2	6.5
Metals	12.7	11.1	12.1	11.5	9.7	8.6	7.7
Plastics	0.5	1.4	2.7	3.8	5.7	7.8	9.8
Rubber & Leather	2.1	2.4	2.6	3.1	3.1	2.4	2.7
Textiles	2.1	2.0	1.8	1.9	1.9	1.9	3.3
Wood	3.7	3.6	3.5	3.7	4.9	5.5	7.3
Other	0.1	0.0	0.4	1.1	1.8	2.0	1.5
Food	14.9	13.1	11.3	11.3	9.6	8.9	8.1
Yard	24.4	22.4	20.5	21.3	20.1	20.3	19.0
Misc. inorganics	1.6	1.7	1.6	1.7	1.6	1.7	1.8

from US EPA (1992)

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[4] Since the national trends are taken from EPA estimates, the term MSW used in this section follows the EPA definition. MSW, as otherwise defined in this report, may show similar trends.

been for plastic, the fraction of which has increased by an order of magnitude. This is significant from an environmental and energy standpoint because plastic occupies a large volume in landfills, often contains chlorine, and has a high energy content.

By examining these trends, the EPA has projected MSW generation and disposal. Table 2.8 shows the EPA's forecasts of MSW disposal assuming a 25, 35, and 50% overall recovery rate (versus 17% recovery in 1990). Consistent with current practice, most of the recovered products in these projections are metals, glass, and paper, which would continue the trend of the disposed MSW having a higher organic fraction.

Table 2.8 - Projections of MSW Disposal for the U.S. in the Year 2000

	Average Recovery Rates					
	25%		35%		50%	
Material	10^6 tonnes disposed	% comp.	10^6 tonnes disposed	% comp.	10^6 tonnes disposed	% comp.
Paper and paperboard	51.4	34%	44.0	34%	26.9	26%
Glass	8.8	6%	7.4	6%	4.3	4%
Metals	10.3	7%	9.1	7%	3.1	3%
Plastics	20.8	14%	19.9	15%	18.0	18%
Textiles	5.7	4%	5.3	4%	4.9	5%
Wood	13.5	9%	12.6	10%	12.3	12%
Food	12.0	8%	12.0	9%	10.8	11%
Yard	17.9	12%	10.7	8%	10.5	10%
Rubber & leather	5.6	4%	5.7	4%	5.0	5%
Other	5.3	3%	4.5	3%	5.9	6%
TOTAL	151.2		131.2		101.6	

Notes:
25%, 35% recovery scenarios from U.S. EPA (1992)
50% scenario uses the same generation rates predicted by the EPA and is based on similar trends in recovery rates.

No historical trends or projections have been studied for CA or L.A. MSW disposal, but these trends might follow the national trends under a 50% recovery scenario, since the CIWMB has recently mandated that every county must recover 50% of its generated waste by the year 2000.

Though projections are not necessarily predictive, some MSW projections have been reasonably good. Table 2.9 shows some of the forecasts from 1972 made of 1990 MSW generation. The 1990 composition projected in 1972 is more or less consistent with the actual 1990 composition. The plastic fraction did indeed increase the most; the paper fraction increased slightly; the metals and glass fraction stayed fairly constant; and the food and yard waste fraction decreased. The per-capita generation rate for 1990 was overestimated in 1972, as was the total generated waste. Considering the uncertainty associated with MSW generation data, this forecast must be considered relatively good.

Table 2.9 - Predicted MSW Generation Trends versus Actual

	1970 Composition[a]	Predicted 1990 Composition[a]	Actual 1990 Composition[b]
Paper and Paperboard	37.4	43.4	37.4
Glass	9.0	9.5	6.7
Total Metals	8.4	8.6	8.3
Plastics	1.4	3.9	8.3
Rubber, leather, textiles	3.4	3.9	5.3
Wood Wastes	3.1	2.0	6.3
Food Waste	20.0	14.0	6.7
Yard Waste	13.9	12.3	17.9
Other	3.4	2.4	3.1
Ratio to 1970 values			
Per capita refuse generation	1.0	1.44	1.30
Total generated MSW quantity	1.0	1.89	1.61

Notes:
a) from Niessen (1972)
b) from U.S. EPA (1992)

2.4 MSW Chemical Characteristics

The chemical composition of a MSW stream can be calculated by examining each component in the waste stream. While MSW as a whole can be extremely variable, each individual component will have similar characteristics regardless of the global composition of the MSW stream it is part of. Niessen has compiled a table of

Table 2.10 - Chemical Characteristics of MSW Components

	Dry mass %						HHV (MJ/kg)		
	C	H	O	N	S	Ash	As rec'd	Dry	Moisture %
Paper, mixed	43.41	5.82	44.32	0.25	0.20	6.00	15.81	17.61	10.24
Newsprint	49.14	6.10	43.03	0.05	0.16	1.52	18.54	19.72	5.97
Magazines	32.91	4.95	38.55	0.07	0.09	23.43	12.22	12.74	4.11
Junk mail	37.87	5.41	42.74	0.17	0.09	13.72	14.16	14.83	4.56
Corrugated boxes	43.73	5.70	44.93	0.09	0.21	5.34	16.38	17.28	5.2
Brown paper	44.9	6.08	47.34	0.00	0.11	1.07	16.88	17.92	5.83
Food cartons	44.74	6.10	41.92	0.15	0.16	6.93	16.88	17.97	6.11
Mixed Plastics	60	7.20	22.6	0.00	0.00	10.20	32.79	33.41	2.0
Polyethylene	84.54	14.18	0.0	0.06	0.03	1.19	45.76	45.88	0.2
Polystyrene	87.1	8.45	3.96	0.21	0.02	0.45	38.18	38.58	0.2
PVC*	45.14	5.61	1.56	0.08	0.14	2.06	22.68	22.73	0.2
Plastic Film	67.21	9.72	15.82	0.46	0.07	6.72		32.20	3-20
Textiles	46.2	6.40	41.85	2.20	0.20	3.20		18.69	15-31
Tires	79.1	6.80	5.9	0.10	1.50	6.60	32.09	32.34	1.02
Leather	60	8.00	11.5	10.00	0.40	10.10	18.51	20.58	10
Rubber	77.65	10.35			2.00	10.00	26.05	26.35	1.2
Waste wood	49.4	6.10	43.7	0.10	0.10	0.60	14.95	16.98	15
Food Wastes	44.99	6.43	28.76	3.30	0.52	16.00	5.51	19.73	72
Yard Wastes	40.31	5.64	39	2.00	0.05	13.00	6.25	16.46	62
Manure	45.4	5.40	31	1.00	0.30	15.90	3.08	15.42	73-88
Field & seed	42.5	5.20	41.7	1.40	0.10	9.10	9.70	17.30	8-80
Fruit & nut	49.3	5.6	42.1	0.6	0.02	2.00	10.80	19.60	35-55

NOTES:
Taken from Niessen (1978) and California Energy Commission (1993)
*PVC is 45.4% Cl

chemical properties of various components of MSW (Table 2.10). Certain components of MSW, however, are much more variable (on a wet basis), like yard wastes or food wastes. For these materials, "typical" values have to be taken based on sampling or averaging. Table 2.10 reveals how each material affects the overall MSW chemical makeup. Clearly, the food and yard wastes have the greatest moisture content. Plastics have the highest heating value, and the polyvinyl chloride (PVC) plastics have the highest chlorine content. Much of the sulfur in the MSW comes from rubber and tires.

Using Niessen's estimates (Table 2.10), the chemical composition, average moisture, and HHV can be calculated for the three cases of Table 2.1. The results are presented in Table 2.11.

The first three columns of table 2.11 show the results of summing up the weighted chemical characteristics of each MSW material for the three cases. For comparison, Table 2.11 also shows others' independent analyses of MSW, revealing that the MSW chemical composition is relatively consistent. Even the chemical analyses by Niessen and Kaiser, which were done over 20 years ago, are similar to recent studies. Apparently, the increased plastic content does not have much effect on the chemical characteristics. Only the ash and moisture contents vary considerably, ranging from 19-30% and 20-40% respectively. The calculated compositions for L.A., CA, and the U.S. cases have moisture and ash contents between 20-25%. Table 2.11 also shows the moisture and ash-free (MAF) compositions. On a MAF basis the chemical composition varies very little among these MSW analyses, so that the average MAF molar formula is well estimated as $CH_{1.6}O_{0.6}$. This C/H/O ratio is very similar to that of biomass. The nitrogen content is smaller than biomass', while the chlorine and sulfur contents are much higher.

The higher heating value (HHV) of a solid fuel is usually given on a dry weight basis because the dry HHV is independent of the fuel's moisture content. But in the field of MSW disposal, the HHV is conventionally given on an as received (wet) basis since most estimates of the MSW resource are also on a wet basis (as they are in Table 2.1). To prevent confusion, it will be explicitly stated throughout this thesis whether the HHV is given on a dry, wet, or moisture and ash free (MAF) basis.

Since the moisture and ash contents of MSW are variable, the higher heating values (HHVs) of MSW on a wet basis (MJ per wet kg) span a wide range. For example, the calculated as-received HHVs of the U.S. and L.A. cases are 19% and 13% respectively greater than the mean of the others' estimates (10.4 MJ/kg) because

Table 2.11 - Ultimate Analyses of Various MSW Samples and Studies

Ultimate analysis	*three cases*											
% mass basis	**U.S.**[a]	**CA**[b]	**L.A.**[b]	*NEI*[c]	*EPA*[d]	*Runcie*[e]	*Niessen*[f]	*Hickman*[g]	*Kaiser*[h]	*Lakeland*[i]	*EPRI*[j]	*NSP*[k]
Carbon	28.76	25.11	27.99	26.65	26.73	20.51	25.62	25.60	29.83			
Hydrogen	3.76	3.34	3.74	3.61	3.60	2.72	3.45	3.40	3.99			
Nitrogen	0.47	0.35	0.44	0.46	0.17	0.32	0.64	0.50	0.37			
Oxygen	22.45	19.79	21.41	19.61	19.74	16.86	21.21	20.30	25.69			
Chlorine	0.87	0.35	0.33	0.55	0.12			0.50			0.45	
Sulfur	0.12	0.09	0.11	0.17	0.12	0.52	0.10	0.20	0.12		0.15	
nonorganics	16.00	26.20	20.30									
Moisture	23.25	21.00	21.90	25.30	27.14	40.00	28.16	25.20	20.00	29	25.00	30.00
Ash	20.37	29.94	24.06	23.65	22.38	19.06	20.82	24.40	20.00	23	25.00	20.00
HHV (MJ/kg)												
As received	12.36	10.60	11.80	11.00	10.47	9.54	10.35			10.47	10.35	10.47
Dry	16.10	13.42	15.11	14.72	14.36	15.89	14.41			14.74	13.80	14.95
MAF[l]	21.92	21.61	21.84	21.54	20.73	23.29	20.29			21.80	20.70	20.93
MAF composition												
Carbon	50.97	51.21	51.81	52.20	52.95	50.11	50.22	50.69	49.72			
Hydrogen	6.66	6.81	6.92	7.07	7.13	6.65	6.76	6.73	6.65			
Oxygen	39.78	40.36	39.63	38.41	39.10	41.19	41.57	40.20	42.82			
Nitrogen	0.83	0.71	0.81	0.90	0.34	0.78	1.25	0.99	0.62			
Chlorine	1.54	0.71	0.61	1.08	0.24	0.00	0.00	0.99	0.00			
Sulfur	0.21	0.18	0.20	0.33	0.24	1.27	0.20	0.40	0.20			

Notes:

a) my calculations based on data from U.S. EPA (1992)
b) my calculations based on data from CIWMB (1992)
c) National Ecology Inc. analysis (Herrmann, 1991)
d) analysis used by U.S. EPA (1988)
e) from Runcie (1989) used in Epri's Fluidized Bed Combustion of Alternate Fuels (1993)
f) from Hickman (1983) used in the *Waste-to-Energy Compendium* (1988)
g) analysis by Niessen (1970)
h) analysis of Long Island MSW by Kaiser (1965)
i) analysis of MSW for McIntosh 3 Incinerator in FL (Bryk et al., 1991)
j) given by the WTE Permitting Handbook (1992)
k) Northern States Power analysis (Brobjorg, 1993)
l) Moisture and Ash Free basis

the U.S. and L.A. wastes have low ash and moisture contents. As was the case for the chemical composition, the HHV for MSW is fairly consistent on a moisture and ash free (MAF) basis. Except for the analysis by Runcie, the moisture and ash free (MAF) higher heating value (HHV) of MSW seems to be between 20-22 MJ/kg.

2.5 The MSW Resource

To estimate the energy resource base of MSW for the US, CA, and L.A., the dry weight basis HHVs (MJ per dry kg) calculated for the three case studies (Table 2.11) are used. The MSW disposal estimates (Table 2.1), however, are given on a wet basis. To estimate MSW disposal for the three cases on a dry weight basis, a range of moisture contents is assumed instead of using the calculated moisture contents of the three cases, which are low in comparison to other studies. By varying the moisture content for each case between 15-40%, which encompasses all the estimated moisture contents shown in table 2.11, a broad range of values for the energy resource is calculated. The results of the resource analysis for the three cases are shown in Table 2.12.

Table 2.12 - MSW Energy Resource

	L.A. county	**California**	**U.S.**
MSW disposed (10^6 wet tonnes)[a]	9	36	147
HHV (MJ/dry kg)[b]	16.1	13.4	15.1
Range of energy resource (PJ)[c]	85 - 121	291 - 412	1423 - 2015

Notes:
a) Table 2.1 cases, using the data from CIWMB (1992) and US EPA (1992)
b) from Table 2.11, using a weighted dry HHV for each component
c) Range calculated by using the dry HHVs from table 2.11 and assuming a moisture content variation of 15-40%. PJ = petajoules

Another method used to estimate the MSW energy resource base is to simply multiply the MSW disposal estimates with an average value for the wet HHV. Empirical estimates of the HHV on an as-received (wet) basis have been given in past studies: Arnasson used 10.7 MJ/kg (1983), Urban et al. assumed 11.6 MJ/kg

(1979), Niessen used 10.4 MJ/kg (1970) and Wilson used 10.1 MJ/kg (1981). These older studies combined with the more recent analyses shown in table 2.11 give a range of wet HHV's roughly between 10-11 MJ/kg. Using an average heating value of around 10.5 MJ/kg and using the data from Figure 2.2, the nation's disposed MSW for 1990 is estimated to contain 1.5 EJ (10^{18} joules), which is within the range calculated in Table 2.12. For future reference, using 10-11 MJ/kg for the wet HHV can provide quick good estimates of U.S. energy resources without any knowledge of the MSW composition.

Recall that the per capita disposal rates for CA and L.A. are much higher than for the US because the national estimates use the EPA definition of MSW. Thus, the values given in Table 2.12 most likely underestimate the actual MSW energy resource for the nation.

2.6 Fuel from MSW

MSW can be converted to methanol or hydrogen, as described in the rest of this thesis. To provide some motivation for the analysis in the chapters that follow, consider how many vehicles might be operated on fuels from MSW.

If the fuels derived from MSW were burned in internal combustion engine vehicles (ICEVs), they could account for a nontrivial share of the total fuel resource. Since hydrogen and methanol have much higher octane ratings than gasoline, ICEs could operate at higher compression ratios, thereby increasing the fuel economy of the ICEV (on a km/GJ basis).

Even considering the improvements of the ICEV running on MeOH or H_2, it is not as efficient as a fuel cell vehicle (FCV). Models of FCVs have shown that a hydrogen fueled FCV has almost three times the HHV efficiency of a gasoline fueled ICEV of the same size. A methanol fueled FCV is nearly two and a half times more efficient than the ICEV (DeLuchi, 1992; Ogden et al., 1994). These high efficiencies effectively enlarge the scope for MSW to be a significant fuel resource.

Table 2.13 - MSW as a Potential Transportation Fuel Resource

	L.A. county	California	US
MSW energy resource			
energy content (exajoules)	0.09 - 0.12	0.29 - 0.41	1.4 - 2.0
Light-Duty Vehicle Travel (1990)			
billion vehicle-miles (v-m)[a]	63	229	1764
Potential Displacement with ICEVs (% of 1990 v-m)[b]			
H_2 from waste[c]	17 - 25%	17 -24%	11 - 15%
MeOH from waste	15 - 21%	14 - 20%	9 - 12%
Potential Displacement with Fuel Cell Vehicles (% of 1990 v-m)[b]			
H_2 from waste[d]	43 - 62%	41 - 58%	26 - 36%
MeOH from waste	31 - 45%	29 - 42%	19 - 26%

Notes:

a) The CA and L.A. data are the vehicle miles traveled by light duty trucks and passenger vehicles. Data taken from the State of CA Business, Transportation, and Housing Agency (1991) and SCAQMD (1993). The national v-mt data accounts for passenger vehicles. Data from US Dept. of Energy (1991).

b) The conversion efficiency of MeOH and H_2 production from MSW is taken from the thermodynamic analysis in chapter 5 of this report.

c) The HHV efficiencies of the ICEVs running on H_2 and MeOH are assumed to be slightly higher than a gasoline fueled ICEV due to the higher compression ratio that these fuels allow: 29.9 and 29.1 gasoline-equivalent mpg for H_2 and MeOH, respectively, compared to 25.8 mpg for a gasoline fueled ICEV (year-2000 Ford Taurus) (Ogden et al., 1994).

d) The HHV efficiencies of the FCVs running on H_2 and MeOH are respectively assumed to be 2.8 and 2.4 times greater than a gasoline fueled ICEV -- a year-2000 Ford Taurus (DeLuchi, 1992).

Table 2.13 shows the number of vehicle-miles that could be displaced by MSW derived fuels for the three case studies for both ICEVs and FCVs. The inclusion of the L.A. county and California case studies is particularly appropriate because California's South Coast Air Quality Management District's (SCAQMD) stringent vehicle emissions standards are likely to make southern California the first market for fuel cell vehicles and methanol and hydrogen fuels.

If the fuels are used in ICEVs, around 10% of the total vehicle-miles traveled by light duty vehicles in the US could be fueled by methanol from MSW. In Los Angeles, up to a quarter of the light duty vehicle-miles traveled could be fueled by hydrogen. MSW-derived fuels would indeed be a nontrivial fuel resource.

If the fuels from MSW are used in fuel cell vehicles (FCVs), the number of vehicle miles that are fueled by MSW-derived fuels would dramatically increase. For the nation, up to 26% of the total vehicle-miles traveled by light duty fuel cell vehicles could be fueled by methanol from MSW; 36% by hydrogen from MSW. For California, over 40% and 58% of the total passenger vehicle-miles traveled could be fueled by methanol and hydrogen from MSW, respectively. In Los Angeles, the MSW could provide enough hydrogen for over 60% of the vehicle-miles traveled. Undoubtedly, MSW is a significant resource when it is considered as a transportation fuel for fuel cell vehicles.

Table 2.14 - Projections for the Year-2000 of MSW as a Transportation Fuel Resource in the US: Low and High Recovery Cases

	No recycling	50% recovery
MSW energy resource		
energy content (exajoules)[a]	1.6 - 2.2	1.1 - 1.6
Light-Duty Vehicle Travel		
billion vehicle-miles (v-m)[b]	2148	2148
Potential Displacement with Fuel Cell Vehicles[c]		
H_2 from waste	24 - 33%	17 - 25%
MeOH from waste	17 - 24%	13 - 18%

Notes:

a) Energy content uses data from the US EPA (1992), which are displayed in Table 2.8, and is calculated with the same algorithm used in Table 2.12.

b) from the U.S. Dept. of Energy (1991).

c) using same assumptions from Table 2.13.

With increased MSW recovery in the country, the size of the MSW energy resource would decrease. But the projections of MSW generation for the year-2000 (Table 2.8) reveal that much of the organic material will still be disposed of even in the high recovery case (50% recovery). Even with this high recovery rate, hydrogen from MSW could fuel up to 25% of the vehicle-miles traveled by light duty vehicles in the country (Table 2.14). Assuming similar waste disposal and driving trends for Los Angeles, 40% of the vehicle-miles in L.A. county could still be fueled by hydrogen from MSW. Therefore, MSW would still be a significant energy resource even with high recovery rates.

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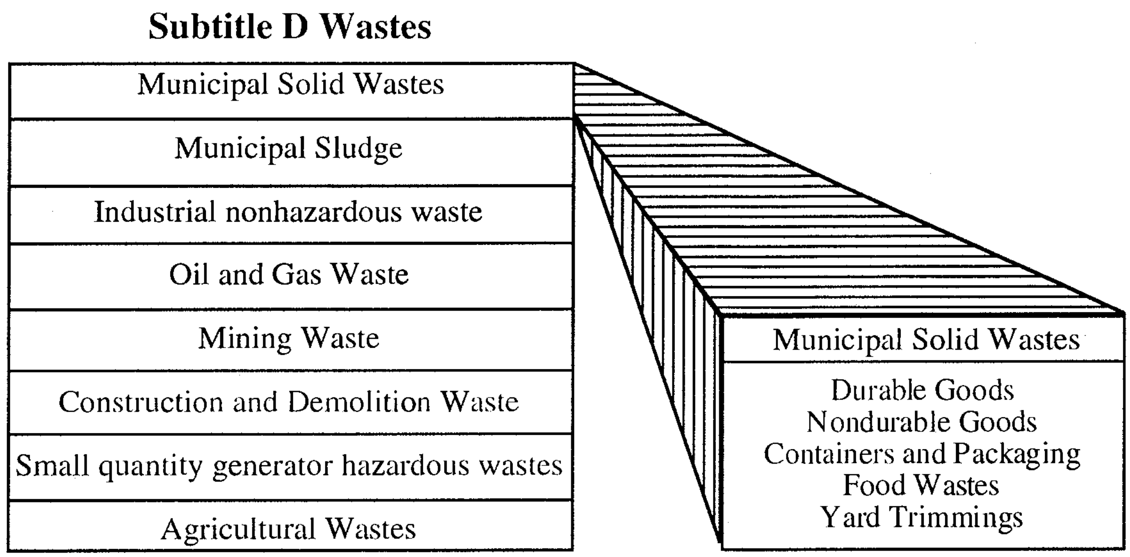
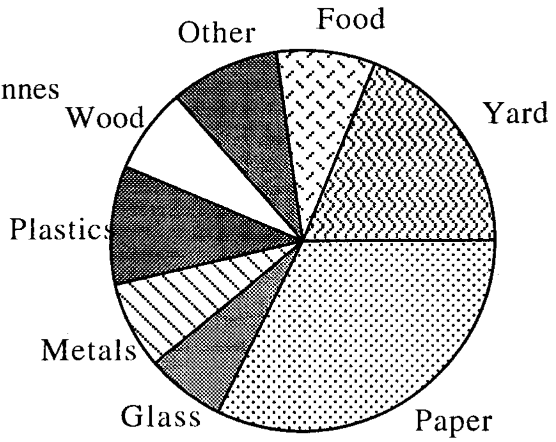
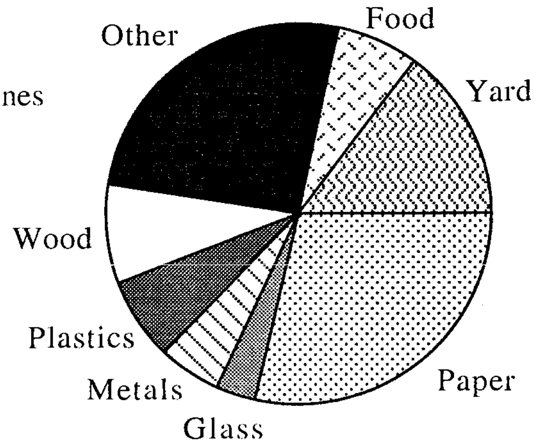


Figure 2.1 - MSW in the world of Subtitle D wastes
(from the US EPA, 1992)

USA
Total MSW disposal = 147 million tonnes
(1.6 kg disposed/capita-day)



California
Total MSW Disposal = 36 million tonnes
(3.3 kg/capita-day)



Los Angeles
Total MSW Disposal = 9 million tonnes
(2.9 kg/capita-day)

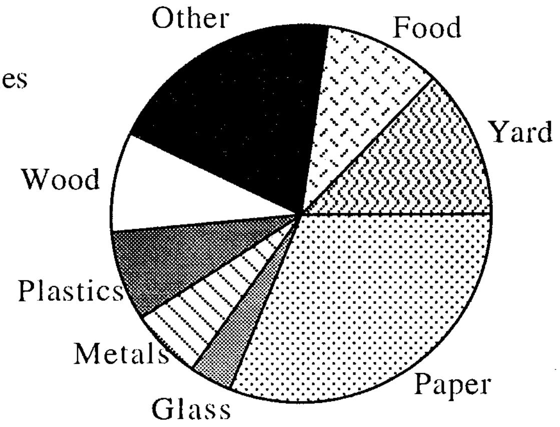


Figure 2.2 - 1990 MSW disposal composition
(from US EPA, 1992; CIWMB, 1992)

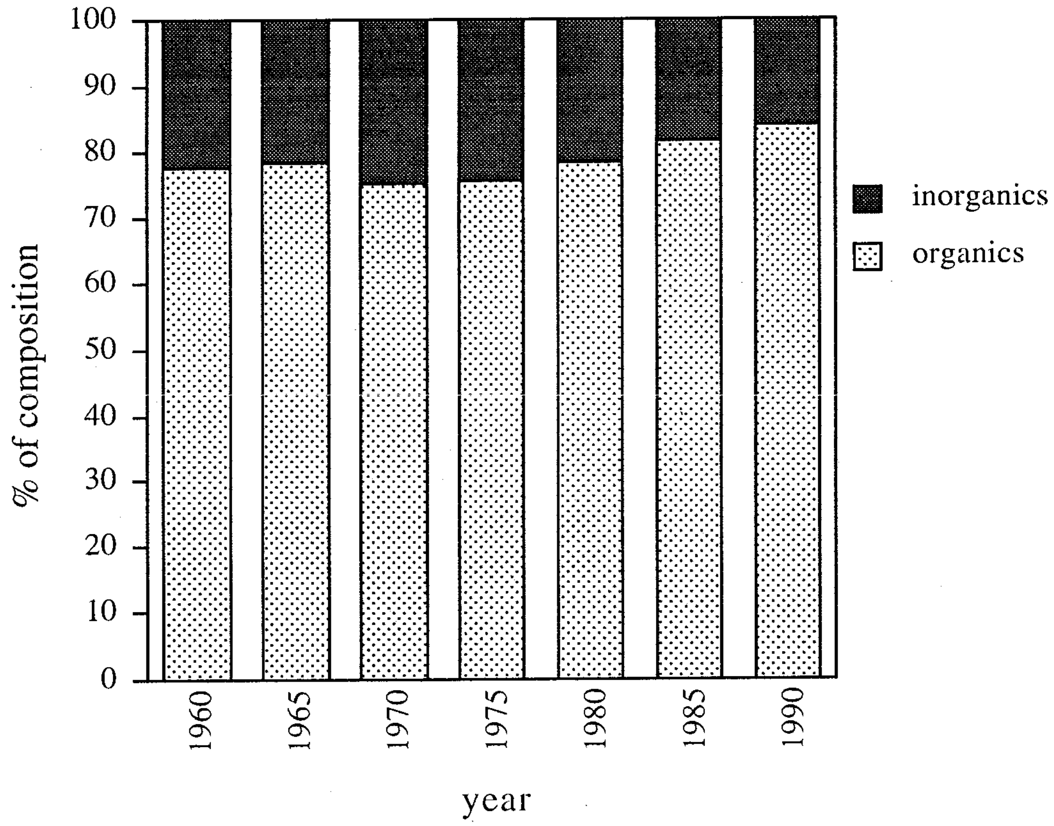


Figure 2.3 - Trends in Organic and Inorganic Fractions of MSW
(based on data from US EPA, 1992)

Chapter 3: Gasification of MSW

3.1 Introductory Remarks

In producing methanol or hydrogen from MSW, the MSW must first be gasified. A gasifier converts the organic solids of the MSW feed into a product gas that can be thermochemically processed into H_2 or MeOH using commercially established technology. Figure 3.1 schematically displays the entire fuel production process. Before entering the gasifier, the MSW can first be processed to RDF, which has less bulk, has fewer nonorganic materials, and is more uniform than raw MSW. Both RDF and MSW gasifiers have shown promise but are not yet commercially established. The gasifier is given focused attention in this thesis because it is the only fuel production system component that is not in widespread commercial use. In this chapter, the three prospective gasifiers that are examined for the fuel production analysis - MTCI, BCL, and Thermoselect - are described in detail. MTCI and BCL require RDF feed; Thermoselect uses raw MSW. Before entering the details of gasification, however, the key step of feed preparation is discussed.

3.2 Feed Preparation - RDF production

3.2.1 Overview

The first step in any gasification process is feed preparation. Typically, coal feeds are first dried, pulverized, and stored. Biomass is usually chipped, dried, and possibly ground. MSW can require extensive processing before it is gasified, because of the large fraction of metals, glass, and other inert solids. Any materials that might foul a reactor must be removed. In general, low temperature gasifiers require more extensive feed preparation than high temperature slagging gasifiers. Gasifiers operating at extremely high temperatures can be fed unprocessed MSW because the noncombustible materials melt into a molten slag. For commercial operation in the

U.S., at least large bulky items and explosive canisters would normally be removed manually from the MSW feed before entering a gasifier, as they are removed for mass burn incinerators.

Processed MSW is called refuse-derived fuel (RDF). RDF production is a well established process with over 20 years experience. RDF is a more homogeneous fuel than MSW, and it also has a higher heat content per unit mass because many of the non-combustible materials have been removed. The degree to which MSW is processed is dependent upon the reactor characteristics. RDF facilities basically remove ferrous metals and reduce the bulk of the received waste. Depending on the required quality of the fuel, the RDF facility may have other processes to remove aluminum, glass, and grit (dirt, gravel, etc.). In the waste processing industry, there are three major classifications of RDF: coarse, fluff, and powder. The major difference among these is in the particle size and ash content. The Electric Power Research Institute (EPRI) loosely categorizes the three types of RDF as follows:

Table 3.1 - Categorization of RDF (loose definitions)

	Coarse	**Fluff**	**Powder**
ash %	21	8	11
% weight in RDF	94	51	42
% energy in RDF	99	67	72
higher heating value (MJ/kg)	10.9	13.6	17.8
Size, mm	<150	<50	<0.45

from WTE Permitting Sourcebook (1993)

The RDF can also be densified into pellets or briquettes, making it easier to handle. Producing coarser RDF requires less complex systems and thus entails less cost, but gasifiers will not perform as well with coarser RDF. The ASTM definitions for RDF are listed in table 3.2. RDF-2 through RDF-5 are the categories commonly thought of as RDF. The other categories are not often used in the industry.

The RDF combustor was designed as an alternative to a mass burn system. Since RDF is more homogeneous and has less ash than raw MSW, it is easier to handle. RDF can be co-fired with other fuels, such as coal and wood, in existing boilers. RDF production plants work well with recycling programs, since lower maintenance is required, and a higher quality fuel can be produced if glass and aluminum are not present in the waste stream.

Table 3.2 - ASTM Categorizations of RDF

RDF type	Definition
RDF-1	wastes used as fuel in an as-discarded form with only bulky wastes removed
RDF-2	wastes processed to coarse particle size with or without ferrous metal separation
RDF-3	combustible waste fraction processed to particle size with 95% passing through a 2 inch square screening
RDF-4	combustible waste fraction processed into powder form with 95% passing through a 10 mesh screen
RDF-5	combustible waste fraction densified (compressed) into the form of pellets, slugs, cubettes, or briquettes
RDF-6	combustible waste fraction processed into liquid fuel
RDF-7	combustible waste fraction processed into gaseous fuel

(ASTM STP 832 taken from Hollander, 1983)

3.2.2 The RDF Production Process

The RDF production process can be divided into five stages: tipping, size reduction, ferrous metal removal, size separation, and RDF storage. Figure 3.2 shows a typical process flow. The process begins at the *tipping floor* where the hauling trucks dump the MSW. Floor inspectors immediately remove any bulky or otherwise undesirable materials. The rest of the MSW can then be front loaded into a storage pit or fed directly onto a belt conveyor leading to the primary shredder.

The primary *size reduction* stage involves only minimal shredding. Nominal sizes leaving this shredder are 15-30 cm. The major purpose of this stage is to break open bags and boxes. Flail mills and hammer mills, which originally were designed to shred automobiles, are used in this process. Flail mills are often used as primary shredders, since they require less energy and maintenance than other shredders. This mill can be enclosed behind metal and concrete barriers to protect the rest of the facility from flying objects or an explosion. Well-run modern facilities are equipped with fire extinguishers. Most facilities also have secondary shredders to reduce the oversized waste coming out of size-separators. Hammer mills, which require more power than flail mills, are often used as secondary shredders, reducing the MSW to 5-15 cm in size.

The *ferrous metal removal* stage follows primary shredding. The MSW coming from the primary shredder passes by a set of magnets that removes the ferrous material. A good RDF production plant might also have another set of magnetic separators further downstream in the process to recover additional ferrous material.

Although recovered aluminum is valuable, aluminum recovery is relatively difficult. Fortunately, since most states have bottle bills or other source separation programs for aluminum, there is not an appreciable amount of aluminum in most waste streams. Aluminum can be removed with eddy-current separators. This device subjects the stream to a magnetic field that induces an electric field in the aluminum, creating a magnetic field which opposes the original magnetic field. These have had numerous maintenance problems, however, so that manual separation has been a more successful option.

The MSW stream then goes through *size separation* stages. Size separation is usually carried out by a trommel or sets of disc screens. The trommel is a rotating cylindrical screen, which is divided into three or four distinct sections. Each section has a different screen size, so different sized materials fall through different sections (Figure 3.3). Disc screens consist of horizontal shafts, onto which discs are attached. The shafts are lined up, with the discs of adjacent shafts overlapping to form a screen. Large materials are rolled over the rotating discs, and small particles fall through the screen (Figure 3.3). In this step, the waste is usually divided into three categories: small (< 2 cm), intermediate, and oversize (>15 cm). The oversize fraction can be recycled back to the beginning of the process or passed to another shredder. The small-sized fraction has a high percentage of dust and glass and is usually taken to a landfill. Some processes also use air density classifiers, in which an air stream entrains the less dense particles (paper, plastic) to separate them from the denser materials (metals, dirt) which fall through the air stream. Air classifiers recover more combustible material and reduce the ash content of the RDF, but consume more power (Municipal WTE Tech. Assess., 1992). If the glass can be manually separated from the rest of the grit, it can be resold in recycling markets.

The RDF is then *stored* in a bin or chamber. In *storage* RDF poses more health and fire risks than coal and does not stack as well due to its low bulk density. Traditional coal storage designs are therefore generally not appropriate. Generally, RDF can be stored for only about a week before bacteria growth is excessive and becomes a health risk. Some storage facilities solve this problem by receiving RDF from the top and discharging from the bottom of the chamber and using sweep buckets or screws to turn over the RDF. Many facilities, however, merely empty out their storage chambers on a weekly basis.

3.2.3 Costs and Efficiencies of RDF production

The first boiler that co-fired RDF was demonstrated in St. Louis, Missouri in 1972, and the first commercial dedicated RDF system started operating in 1975 at

Ames, Iowa. Since then, many RDF-fired combustion plants have operated throughout the country, and improvements have been made in the RDF production process.

Various operating processes have different costs and produce varying qualities of RDF. To quantify some of the important parameters of RDF production, some of the modern processes that produce high quality RDF have been investigated. Two of the three gasifiers analyzed for fuel production (MTCI and BCL -- discussed in detail later) used RDF feeds in test runs. MTCI bought fluff RDF for its test runs from Reuters, Inc., but data from Reuters' RDF production process are not available. The BCL gasifier tests were run with fluff RDF from Cockeysville, Maryland. National Ecology Inc. (NEI) designed and built the RDF plant in Cockeysville and has had considerable experience in RDF processing. NEI has helped design, construct or operate RDF plants for, among others, Baltimore Gas & Electric (BG&E), the Solid Waste Authority of Palm Beach County, Florida, and Northern States Power. The plant in Cockeysville was an older plant that began operating in 1976. Though the ash content of the RDF from Cockeysville was very low (9%), the mass yield[1] was too low to be economical (Herrmann, 1991). Since operating the Maryland plant, NEI has greatly evolved in its RDF processing systems (Herrmann, 1991). Table 3.3 compares the RDF from old and new processes. Present processes produce a low-

Table 3.3 - NEI's Old versus New RDF

	Cockeysville, MD		Palm Beach, FL
	phase II	phase III	
% mass yield	40	60	83
% energy yield	n.a.	n.a.	96
HHV (MJ/kg)	13.7	13.5	12.8
% ash	9.5	8.3	9

(Herrmann, 1991) n.a. = not available

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[1] the mass yield is the percent of the original MSW mass that is retained in the RDF. The energy yield is similarly defined as the percent of the total energy content of the MSW that the RDF still has. For a MSW feed that has 20% ash content, it would be ideal to have RDF with 0% ash, 80% mass yield, and 100% energy yield.

ash RDF that retains a high percentage of the combustible fraction of MSW. New configurations require fewer electricity inputs and have lower maintenance and labor costs.

Table 3.4 compiles information from some operating RDF production facilities. The plants in Florida, Connecticut, and Virginia all have similar costs and qualities of RDF and have demonstrated reliable operation. From this information, the costs and energy losses associated with processing MSW into RDF have been estimated. It is assumed that modern RDF facilities, like those mentioned above, produce a product that can be used in a gasifier.

3.3 Gasification Principles

3.3.1 What is Gasification?

The next stage of the fuel production process is gasification. Gasification is broadly defined as the thermochemical conversion of a solid carbonaceous feedstock into a combustible gaseous product. Sometimes (usually in the case of coal) gasification is accomplished by partial oxidation, i.e. - oxygen-starved combustion. As will be described later, some newly designed gasifiers for biomass feeds do not use any oxygen in a process termed as pyrolytic gasification.

Pyrolysis is loosely defined as the thermal degradation of the feed into non-condensable gases, condensable liquids and solid char residues in the absence of air. In other words, the volatile matter in a feed is vaporized or liquefied, leaving only fixed carbon and non-combustible material; pyrolysis is thus often termed *devolatilization*. Pyrolysis is the first step in most thermal conversion processes (Milne, 1981), including pyrolytic gasification and partial oxidation gasification. Sometimes pyrolysis is referred to as *carbonization* because it leaves a carbon and mineral solid product; this is usually the case when a solid product is desired, like charcoal. Pyrolysis of biomass feeds usually begins around 200 °C.

Table 3.4 - Summary of RDF Production Plant Data

	Ames, IA[a]	Madison, WI[b]	SPSA[c] Portmouth, VA	UPA[d] Elk River, MN	NSP[e] French Island, WI	NSP[f] Newport, MN	CRRA[g] Hartford, CT	BCRRF[h] Cockeysville, MD	Palm Beach, FL[i]
plant size (tonnes raw MSW/day)	181	1090	1814	1360	363	1270	1814	1090	1814
wet HHV (MJ/kg)	14.8	13.3	13	13.1	13		12.4	13.5	13
% ash	9	12	9	12	12		12	9	13
% moisture	22.5	25	24	20	20			27.5	27
% mass yield	77	60	82	75	75	70	83	60	83
Capital (million $)	11.5	3.9	56	29.3	9.4	32.3	52	31	54
$/(t/day)	63400	3580	30900	21530	25900	25400	28700	28500	29800
O&M (million $/yr)	1.7	0.48	4.8	4.2	1	n.a.	5.5	n.a.	6.2
($/yr)/(t/day)	9400	440	2650	3090	2760		3030		3400
electrical inputs (kwh/wet tonne)	55		30	27.5		26.5	27.5	22	27.5

Notes:

a) Data taken from Joensen (1986), Wiegand (1993), and City of Ames (1989)
b) Data from Oashgar (1993), Jacobus (1991), Casey (1993), and Madison (1993)
c) Southeasten Pacific Public Servic Authority of Virginia. Data from SPSA (1993) and Stevens (1993).
d) United Power Association. Cost data is from Shurtz (1993). RDF specs. are from Kaas (1993).
e) Northern States Power plant data is from Sullivan (1993)
f) NSP facility cost data is from Smith (1993). Other information is from the WTE Tech. Assess. (1992)
g) The Connecticut Resources Recovery Authority. RDF specs. are from Boley (1991). Other data comes from Mazzaccaro (1993).
h) Baltimore County Resource Recovery Facility designed by NEI. Data from Herrmann (1991, 1993) and BG&E (1984).
i) NEI's newly designed facility. Data from Herrmann (1990, 1991, 1993) and Arvan (1993).

The conversion of the solid carbon residue of pyrolysis to a gas, *char gasification*, is accomplished in various ways, depending on the design of the gasifier. In partial oxidation gasification, the heat needed to drive the endothermic reactions involved is provided by combustion of some of the feed within the gasifier:

$$C + 1/2O_2 \rightarrow CO \qquad \Delta H^\circ = -110\ \mathrm{MJ/kmol} \qquad (3\text{-}1)$$
$$C + O_2 \rightarrow CO_2 \qquad \Delta H^\circ = -394\ \mathrm{MJ/kmol} \qquad (3\text{-}2).$$

In pyrolytic gasification, heat is provided from an external source.

The *char gasification* reactions proceed rapidly only at temperatures greater than 700 °C (Antal, 1981). These reactions are known as steam gasification (3-3), the Boudouard reaction (3-4), and hydrogasification (3-5).

$$C + H_2O \rightarrow CO + H_2 \text{ (steam gasification)} \qquad \Delta H^\circ = +131\ \mathrm{MJ/kmol} \qquad (3\text{-}3)$$
$$C + CO_2 \rightarrow 2CO \text{ (Boudouard reaction)} \qquad \Delta H^\circ = +172\ \mathrm{MJ/kmol} \qquad (3\text{-}4)$$
$$C + 2H_2 \rightarrow CH_4 \text{ (hydrogasification)} \qquad \Delta H^\circ = -75\ \mathrm{MJ/kmol} \qquad (3\text{-}5)$$

The steam gasification and Boudouard reaction are highly endothermic and are thermodynamically favored at high temperatures. Hydrogasification is exothermic, but the reaction is kinetically favored by high temperatures.

Along with the solid-phase reactions of oxidation and char gasification, many secondary gas-phase reactions occur at temperatures over 600 °C if residence times are long enough (Antal, 1981). Key gas-phase reactions involve the decomposition of the higher hydrocarbon products of pyrolysis. These reactions are similar to the processes that thermally crack hydrocarbons in petroleum into lighter products. Other important gas-phase reactions are the water-gas shift reaction (3-6) and methanation (3-7):

$$CO + H_2O \rightarrow CO_2 + H_2 \text{ (water-gas shift)} \qquad \Delta H^\circ = -41\ \mathrm{MJ/kmol} \quad (3\text{-}6)$$
$$CO + 3H_2 \rightarrow CH_4 + H_2O \text{ (methanation)} \qquad \Delta H^\circ = -206\ \mathrm{MJ/kmol} \ (3\text{-}7)$$

3.3.2 Thermodynamic considerations

Only the primary reactions are listed above, but innumerable other reactions occur during gasification. In contrast to the situation for modern high-temperature coal gasifiers, residence times and temperatures in most biomass gasifiers are not sufficient to reach equilibrium. The operating parameters of biomass gasifiers are such that it is nearly impossible to predict output gas characteristics using thermodynamic considerations alone (Probstein, 1982). The design and operating parameters of a gasifier can be described qualitatively, however. Some of the important parameters are residence time, temperature, heating rate, pressure, oxygen rate, steam rate, feed rate, and gasifier configuration.

A longer residence time lets the product gas approach chemical equilibrium more closely. For the operating temperatures of the gasifiers that are examined later (700-1200 °C), a sufficiently long residence time would lead to near elimination of the condensable gases (tars and oils) and higher hydrocarbons from the product gas. A longer residence time also increases the carbon conversion of the process. Long residence times and high reactor temperatures have similar impacts on the approach to equilibrium. High temperature processes have faster reaction rates so that less time is required for the products to approach equilibrium. Thus, reactors that operate at very high temperatures need very short residence times (on the order of seconds to convert all the carbon to gas). Reactors that operate at much lower temperatures are designed to have longer residence times (a couple of minutes).

Higher temperatures favor pyrolysis and char gasification reactions, so that more hydrogen is volatilized and more carbon is converted to gas. At very high temperatures the reverse shift (3-6) and forward Boudouard (3-4) reactions are favored, giving rise to more CO at the expense of CO_2. Low temperatures favor methanation, which is not surprising since reaction (3-7) is highly exothermic. The heating rate is closely related to the reactor temperature and greatly affects the pyrolysis products. Slow heating rates at low temperatures produce large amounts of solid products, whereas rapid heating produces liquid and gaseous products.

Pressure is also a very important parameter for gasification reactions. Le Chatelier's principle explains why high pressures adversely affect the char gasification reactions and favor methanation. It may seem unwise to operate at high pressures since less char would be gasified. But increased throughput rates and reduced plant sizes can sometimes decrease the plant costs so as to make the efficiency tradeoff worthwhile. Also for fuels production processes that begin with gasification, pressurized gasification can reduce downstream compression loads. Despite potential advantages of pressurization, all three gasifiers examined in this fuel production analysis, however, are being developed for atmospheric pressure operation because of the greater mechanical challenges involved in pressurization.

High steam rates should increase the carbon conversion through hydrolysis. More steam favors hydrogen formation and CO_2 over CO through the water-gas shift reaction. Raising steam, however, requires a substantial amount of energy that must be provided by either waste process heat or burning some of the gasifier output.

Limiting oxygen input (in a partial oxidation gasifier) to the vessel will prevent complete combustion of the char to CO_2, so that more CO will be produced through the Boudouard reaction and partial oxidation. Since complete combustion releases more energy than partial oxidation, higher temperatures can be achieved with more oxygen. If air is used as the oxidant, the nitrogen acts as a diluent and will absorb much of the energy released by the exothermic oxidation reactions, so that more feedstock must burn to provide the same temperature in the gasifier, compared to using pure oxygen. None of the gasifiers that are analyzed use air as the oxidant. For fuel production processes, air-blown gasifiers reduce the overall system efficiency because of the extra fuel burned in the gasifier. Also, the presence of nitrogen greatly increases downstream compression loads and required reactor volumes, as well as preventing optimal operation of the methanol synthesis reactor or hydrogen separation beds.

3.4 Gasifier Classification

Biomass and waste gasifiers can be divided into two types, directly and indirectly heated. Directly heated (partial oxidation) gasifiers use air or oxygen to combust some of the feed to supply heat necessary to gasify the rest of the feed. Indirectly heated (pyrolytic) gasifiers are externally heated by using heat exchanger tubes or an inert medium like sand that can transfer the heat.

3.4.1 Directly heated gasifiers

Directly heated gasifiers are generally adapted from coal gasification technologies. Over one hundred years of experience has been accumulated with coal-fed directly heated gasifiers. Oxygen is preferred to air as the oxidant for reasons discussed before. The drawback is that a plant that produces high purity oxygen is expensive and consumes a lot of electricity. There are three major configurations for directly heated gasifiers: fixed-bed, entrained-bed and fluidized-bed (Figure 3.4).

The *fixed bed* gasifier is the simplest and was the first to operate successfully on coal and MSW. Until WWII, almost all coal gasifiers were of this configuration (Hos and Groeneveld, 1987). One variant of the fixed bed gasifier is an updraft process; the feed is fed from the top and drops to the bottom, while the gases flow upward, countercurrent to the feed. The feed forms a packed bed, where distinct zones are formed (Figure 3.5). As the feed falls, it is first dried, then pyrolyzed, and then gasified. At the bottom of the bed, air or oxygen is injected and the remaining char is combusted. The hot combustion gases flow upward and provide the heat to gasify, pyrolyze and dry the feed.

The solid residue from the gasifier may be either slag or dry ash. Dry ash gasifiers must keep the peak temperatures below the ash fusion temperatures (1000-1300 °C). Slagging gasifiers melt the residue into a molten slag which drains into a quench tank. High peak temperatures are required for slagging and usually require oxygen as the oxidant. Though the peak temperatures at the bottom grate of a fixed bed gasifier may reach 2000 °C, the product gas usually leaves the top of the gasifier at low temperatures below 600 °C (Probstein, 1982). Steam can be injected to assist the char gasification reactions and cool the reactor. Since the feed continually moves downward, the fixed bed gasifier is also known as a "moving bed" gasifier. Some of the early work in MSW gasification was based on this configuration, but no fixed-bed gasifiers are considered for fuel production in this report. They produce large amounts of oils and condensable gases, making them unattractive for fuel production because the oils contain a significant fraction of the products' energy and will condense out in the low temperature clean up steps in the fuel processing system. Tars can also clog and foul catalyst beds and valves in the system. Only one oxygen-blown fixed bed reactor that gasifies MSW or RDF feeds is believed to be operating currently, but no recent data are available (Aoki, 1994).

Entrained-bed gasifiers are usually oxygen blown and are characterized by very high temperatures, so that a molten slag residue is formed. The bottom combustion zone can reach 2000 °C and the product gas exits the gasifier between 1000-1500 °C. This is a co-flow device. Oxygen and steam[2] are blown into the gasifier and entrain the feed. The main aim of these gasifiers is to gasify the feed rapidly (i.e. - in seconds). This is quite different from some fixed bed gasifiers, which have residence times of over one hour (for coal feeds). Due to the fast reaction times, entrained-bed gasifiers can have high throughput rates at modest reactor sizes,

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[2] Some work has been done with using hot product gases or hydrogen, rather than oxygen and steam, to entrain and gasify the feed (Probstein, 1986 and Dong and Steinberg, 1994).

especially when operating at high pressures. Entrained flow gasifiers can achieve complete carbon conversion and elimination of tars or oils from the product gas. Their drawbacks are that the reactors must withstand severe conditions, and oxygen is required. To entrain the feed in the gas stream, the feed must be fed as very fine particles (100-600 µm). Various feeding methods have been used for coal gasifiers. For the Texaco gasifier, coal is fed as a liquid slurry. The Shell and Koppers-Totzek gasifiers use dry pulverized coal feed. In comparison to coal, grinding biomass into a fine powder requires large capital and energy inputs (Katofsky, 1993). This is one reason entrained-bed gasification has not been seriously considered for biomass feeds. The other reason for the lack of interest in entrained-beds is that biomass is much more reactive than coal, so that peak temperatures need not be as high to achieve the same level of gasification.

Table 3.5 shows a comparison of proximate analyses of coal to typical biomass feeds. Refuse-derived fuel (RDF) is also included to display its similarity to biomass. Both have volatile matter fractions that are about twice the value for coal. For dry coal, only 40% of the feed will be gasified through pyrolysis reactions. Most of the coal must undergo the char gasification reactions. About 80% of dried biomass and RDF, on the other hand, will be gasified at the lower pyrolysis temperatures. The

Table 3.5 - Compositional Analysis of Coal and Biomass

Proximate Analysis (% mass dry basis)	coal[a]	biomass[b]	RDF[c]
Volatile Matter	37.5	81.3	77.8
Fixed Carbon	43.4	17.2	11.0
Ash	18.2	1.5	11.2
HHV (MJ/kg)	26.7	19.4	18.8

Notes:
a) coal is Illinois no. 6 bituminous coal. Taken from Williams and Larson (1992)
b) biomass is white oak. Taken from Jenkins (1989)
c) RDF data taken from Paisley et al. (1989)

extreme temperatures of entrained-bed gasifiers are not necessary to gasify biomass or RDF. Thus, the oxygen production and slag removal systems add unnecessary costs to the process.

Though low temperature processes could sufficiently gasify the organic fraction of MSW, higher temperatures are preferred to destroy organic compounds. Also, the solid residue from slagging gasifiers will form a vitreous slag instead of ash. Due to the recent concerns about the hazardous nature of incinerator ash, high temperature slagging gasifiers are being considered as an alternative MSW disposal technology to incinerators (see Chapter 6). In this analysis of fuel production, the Thermoselect gasifier is examined as a representative of high temperature processes.

The *fluidized-bed* reactor is the third basic type of directly heated reactors. Of the directly heated gasifiers, this configuration will probably be the favored candidate for fuel production from biomass because of its ability to accept feeds of different sizes and densities, its low tar/oil output, and other reasons described below. Crushed feedstock is fed through the side of the vessel and is mixed and suspended by oxygen or air blown in from the bottom. Sand or other inert materials can be used in the bed to carry heat and prevent agglomeration. Steam can also be injected to act as fluidizing material and to gasify the feed. The fluidized-bed gasifier is characterized by good solids mixing which ensures a nearly isothermal bed and efficient energy transfer within the bed. Fluidized-beds help prevent the formation of hot spots where ash can fuse and cold channels where feed can travel through unreacted. Ash and char are carried by the product gas through the top of the gasifier and separated by cyclones. Bubbling fluidized-bed gasifiers (BFBG) separate the solid residue. Circulating fluidized-beds (CFB) recycle the captured solids through a standpipe back to the gasifier, which promotes higher carbon conversions. Typical operating temperatures for fluidized-bed gasifiers are 900-1000° C, which is also the product gas exit temperature. Though this gasifier type looks promising for biomass feeds, it is not examined in this analysis for MSW. Almost all of the experimental work and commercial applications for directly heated fluidized-bed refuse gasification use air as the oxidant to produce a low energy value gas. These air-blown gasifiers are interesting for electricity or steam generation but not for fuel production.

3.4.2 Indirectly heated gasifiers

Indirectly heated gasifiers are heated externally in various ways (Figure 3.6). They usually operate at lower reactor temperatures (600-900 °C) than directly heated gasifiers. The higher reactivity of biomass has driven the designs of these indirectly heated gasifiers because even at their lower temperatures biomass can be gasified. Reactor temperatures are lower than ash fusion points, so that clinkering is prevented. Since the heat is provided without the direct introduction of combustion air to the gasifier, the product gas that forms is usually of medium energy value (10-20 MJ/Nm^3): nitrogen from air does not dilute the product gas. Also, the use of expensive oxygen is avoided. This is especially important because oxygen costs are high at the modest scales that characterize biomass systems or MSW disposal units. These indirectly heated processes have only recently been developed since they are not based on coal gasifier designs; no commercial plants are in operation. Most designs are still in the pilot and demonstration stages. Recent analyses of fuel production from woody biomass have shown that the lowest costs of production may well be achievable using indirectly heated gasifiers. (Katofsky, 1993 ; Williams et al., 1994).

There is some experience with RDF feeds for two indirectly heated gasifiers analyzed by Katofsky, the BCL and MTCI gasifiers. Since for these gasifiers fuel production costs were found to be lowest among various biomass gasifiers considered (Katofsky, 1993), the BCL and MTCI gasifiers are selected for the present analysis as representatives of indirectly heated processes.

3.4.3 Definitions of Gasifier Parameters

Some typical measures of merit for gasifiers are described here. The *carbon conversion* of a gasifier is defined as the carbon in the product gas divided by the carbon in the original feedstock. This can get complicated for indirectly heated gasifiers which recycle char or gas to the gasifier. It is defined in this report as the carbon in the net product gas (subtracting out any carbon in the recycled gas or char) divided by the feed carbon. Typical values for high temperature gasifiers are 95%., while for low temperature gasifiers, the carbon conversion can be as low as 60%.

The *cold gas efficiency* is defined as $\frac{\text{HHV of product gas}}{\text{HHV of the feed}}$. Again, for indirectly heated gasifiers, the HHV of the net product gas is taken as the numerator. These two measures are the simplest of performance parameters, but are sufficient to characterize the gasifiers in this analysis. Thermodynamic efficiencies that account for steam inputs and oxygen production loads are not necessary, since all heat and electricity loads are accounted for in the fuel production analysis of Chapter 5.

3.5 Overview of Past Experience

3.5.1 Fuels Production from MSW

How much experience is there with fuel production from MSW? There are no commercially operating methanol or hydrogen plants that use biomass feeds, let alone MSW. Katofsky's thorough study of methanol and hydrogen production from biomass provided a starting point for the present work. A key finding in Katofsky's work was the apparent attractiveness of indirectly heated gasifiers for fuels production from biomass.

Besides Katofsky's (1993) thorough study, there have been other comprehensive studies of fuel production from biomass. Analyses by Wyman (1993) and the OPPA (1990) examined methanol production from biomass and were starting points for Katofsky. Williams et al. (1994) updated Katofsky's original work, including a lifecycle cost analysis of fuel cell vehicles fueled by hydrogen or methanol produced from biomass. The above mentioned studies showed that biomass grown on dedicated energy plantations will not become competitive with natural gas as a feedstock for fuels production until the natural gas price rises to roughly double the present level. MSW is an economically attractive feedstock because its cost is negative.

There have been few studies of fuel production from MSW. The city of Seattle commissioned a cursory feasibility study in 1974 to see if methanol or ammonia could be produced from MSW using the Purox gasifier, a fixed-bed gasifier that will be discussed later (Paul, 1978). Arnasson (1983) examined MSW and biomass feeds for methanol production using the Wright Malta gasifier. His thermodynamic calculations and economic case studies showed that under many scenarios, methanol from MSW would be competitive as a peaking fuel for electricity production. Arnasson calculated that methanol as a neat fuel could even be competitive with gasoline.[3]

There is a dearth of fuel production studies because few technologies have successfully gasified MSW or RDF. All of the fuel production system downstream of the gasifier is based on commercial technology, so there should be no fundamental problems associated with converting clean syngas to methanol or hydrogen. The MSW gasifier is the least established component in the system. Experience is now being gained with dedicated biomass gasifiers. Lurgi, Ahlstrom, TPS, and Uhde all

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[3] Arnasson's analysis (1983) uses data for the Wright Malta gasifier, an indirectly heated rotary kiln, that was fed biomass to reach his conclusions. Katofsky (1993) showed that the Wright Malta gasifier was very inefficient in comparison to other gasifier technologies because it had enormous external heat requirements. The kiln operated at high pressures (1.5 MPa), which caused problems in sealing the rotating parts. Also, the long feed residence times (1 hour) would greatly limit the throughput of a large scale facility (Katofsky, 1993). The Wright Malta gasifier is not examined in detail in this thesis.

have commercial plants available, and there are many others in the pilot or demonstration stages that use biomass (Consonni and Larson, 1994).

3.5.2 MSW Gasification Experience

The experience with MSW or RDF gasification is sparse in comparison to biomass feeds. In the 1970's, there were some attempts in this country to pyrolyze and gasify MSW. Unfortunately, these demonstration projects had multiple problems and were all closed. The low oil prices of the 1980's along with plant failures certainly deterred private sector and government support for commercializing MSW gasification. Today there are no commercial MSW or RDF gasifiers operating in the United States. A common belief of the past was expressed in the Solid Waste Handbook. The authors acknowledged the flexibility of pyrolysis and gasification products. "However, the major limitation is that the technical and economic feasibility of a large-scale facility has yet to be proven." (Schlotthauer et al., 1986). More recent experiments with MSW and RDF feeds, however, have shown promise in new gasifier designs. There are currently a handful of commercially operating MSW and RDF gasification plants in the world.[4] Experience learned from past mistakes and the advent of new technologies are starting to make MSW gasification a viable waste management alternative. Table 3.6 outlines some of the noteworthy MSW/RDF gasification projects of the past 20 years. More detailed descriptions of these projects are provided below.

3.5.2.1 Early MSW Gasification Experience

The most notable of the 1970's projects was the *Monsanto Landgard* project. Monsanto Enviro-Chem began operating its 0.3-0.5 tonne/day pilot plant in Dayton,

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[4] Two commercial units are in Italy (Riegel et al., 1993; Bridgwater and Evans, 1993). There is a Torrax plant still operating in Creteil, France (Bridgwater, 1993). A plant in Chichibu city, Japan, erected in 1981 is still in operation (Aoki, 1994; Nakazato, 1994). The Nippon Steel unit in Ibaraki, Japan was closed in the 1980's but was re-opened after a complete retrofit (Nakazato, 1994).

Ohio in 1969. Later that year, a prototype kiln was built in St. Louis to process 32 tonnes/day. With the support of the Environmental Protection Agency (EPA) and Maryland Environmental Services, Monsanto opened the Baltimore plant in 1975. The plant was rated to process 907 tonnes raw MSW/day, but it never operated at capacity. The plant failed to meet two of the three guarantees: particulate loadings and reliability. In 1977 Monsanto recommended the plant closure and later withdrew

Table 3.6 - Summary of Selected Waste Gasification Projects

Gasifier Technology	Scale (tpd)	Type of gasifier and feed	Comments
Monsanto Landgard	900	air-blown rotary kiln MSW feed	Baltimore demonstration (1975) Did not meet PM loadings or reliability guarantees
Torrax	400	air-blown updraft fixed-bed MSW feed	6 plants set up, including Disney World, FL (1982) All except one (Creteil, France) were closed.
Union Carbide's Purox process	180	oxygen-blown updraft fixed-bed RDF feed	S. Charleston, WV (1974) demonstration. Expensive. 73 tpd plant in Chichibu City, Japan still operating
Occidental	180	flash pyrolysis vertical shaft (powder) RDF feed	San Diego demonstration (1976) Oil product. Extremely low availability and capacity
Stardust '80 (Ebara)	100	indirectly heated dual bed RDF feed	Yokohama city, Japan demonstration (1979-1982) showed adaptability
Tsukishima Kikai	450	indirectly heated dual bed RDF feed	Commercial plant in Funabashi City, Japan (1982-1988). Closed due to high operating costs
Siemens KWU	70	indirectly heated rotary kiln with in-bed heat exchangers. MSW feed	Ulm-Wiblingen, Germany demonstration (1988 - present).
Lurgi GmbH	100	air-blown CFBG RDF feed	Pols, Austria demonstration with bark feeds. Ran tests with RDF
TPS	100	air-blown CFBG RDF feed	Commercial plant in Greve-in-Chianti, Italy (1991-present)
Thermoselect	100	oxygen blown entrained-bed. MSW feed	Commercial plant in Fondotoce, Italy (1992 - present)

its involvement with the facility. The plant reopened in 1979 under the city of Baltimore's supervision, but closed soon afterwards. A Wheelabrator mass burn incinerator was erected in its place the 1980's (Turner, 1988).

The Landgard process was an air-blown horizontal rotary kiln pyrolysis plant. Oil was added at the discharge end to help generate enough heat to gasify the rest of the feed. The product gas contained 3-4 MJ/Nm^3, exiting the kiln at 650°C. An after burner, known as the gas purifier, burned some of the product gas to turn the flyash into slag. The product gas was burned to raise steam for Baltimore Gas & Electric. At the pilot scale, the net thermal efficiency of the process was only 40% (Helmstetter and Sussman, 1978).

The problems that the Baltimore project encountered reflected the dearth of experience in the waste disposal field. The process had only one residue conveyor, even though this device required a high level of maintenance. The feed rams were exposed to high temperatures which distorted the throats and jammed the feed. Flyash loadings were higher than expected, which in turn clogged the slag tap in the afterburner. The operators heated up the tap to keep the slag liquid, but this damaged the burner's refractory lining. A major conclusion from this demonstration was that more expertise and experience were needed in the gasification field. Many of the problems could have been avoided if people with experience in waste management had helped design and operate the systems (Helmstetter and Sussman, 1978).

Andco, Inc. and Carborundum Company formed the joint venture Torrax Systems, Inc. dedicated to the *Torrax* waste gasification process. A demonstration plant processing 68 tonnes/day of MSW was set up in Orchard Park, NY in 1969. Six plants were constructed in the late 1970's and early 1980's. They were in Luxembourg, Frankfurt, Grasse and Créteil (France), Disney World (Florida), and Hammamatsu City (Japan) (Wilson, 1981 and Bridgwater, 1993). The Disney World plant was a project supported by the EPA (SRI, 1992). The Créteil 400 tonne/day plant consists of two units and is the only Torrax system still operating (as of 1992). It was built by Sofresid/Caliqua and is owned and operated by Créteil Incinération Énergie. Recent plans call for only using one unit for medical waste disposal, and setting up an incinerator for MSW disposal (Bridgwater and Evans, 1993).

The Torrax system is an air-blown, updraft, fixed-bed, slagging gasifier, which is fed unprocessed MSW. The product gas leaves the gasifier at 430 °C, and a fraction is burned with auxiliary fuel to pre-heat the air going to the gasifier.

The commercial plants all had many problems and were shut down for various reasons. Slag rates were higher than expected, which clogged the slag removal system. The bed was not properly mixed, so hot gases would rise through open channels without heating the unreacted feed. Corrosion problems were pre-eminent. Besides the early operating problems, the Torrax system had enormous fuel costs which kept it non-competitive with conventional waste disposal measures.

Purox is a fixed-bed process developed by Union Carbide, which began operating a 4.5 tonne/day pilot plant in Tarrytown, NY in 1971. A 180 tonne/day plant was installed in S. Charleston, WV in 1974, and Showa Denko K.K. constructed a 73 tonne/day Purox plant in Chichibu City, Japan in 1981. Purox is an oxygen blown, updraft, fixed-bed slagging gasifier. It is very similar to the Torrax process, except that it uses oxygen and is fed coarse RDF. Purox is characterized by very high temperatures (1650°C) so the solid residues melt into a slag at the bottom. The product gas has medium energy value (12 MJ/Nm^3) (Fisher et al., 1976).

Though the Purox process, unlike other projects, had no major problems, it was the most capital intensive of the early MSW/RDF gasification processes because it required pure oxygen (Wilson, 1981 and Paul, 1978). It was never commercialized in the U.S. because of its high capital cost (Turner, 1988). The Japanese plant has been operating satisfactorily since its start up, but no recent data are available (Aoki, 1994).[5] Like the Créteil Torrax process, the Japanese unit needs extra fuel for the slagging process.

The Occidental Petroleum Company designed the *Occidental pyrolysis process*, which is sometimes referred to as the Garrett Process[6]. Though its principal product was a liquid oil, it is briefly mentioned here because it was a highly publicized MSW pyrolysis project. A pilot plant processing 4 tons/day was tested in Vancouver, WA in 1971 (Domino, 1979). With EPA support, a 180 ton/day plant was constructed in 1976 in El Cajon, CA which is in San Diego county. This plant was set to go into operation for 1978, but the entire project was disbanded due to similar problems that the other projects experienced.

The Occidental process was a vertical shaft, entrained-bed "flash" pyrolysis reactor. RDF was dried to 3% moisture and finely ground to powder. 80% of the feed needed to be smaller then 1.2 mm. Hot char acted as the heating agent for pyrolysis. No. 2 fuel oil was sprayed into the product gas stream to cool it and prevent any further cracking of the gas. The oil fraction, called pyrofuel, condensed out in a decanter. Only 47% of the feed energy was recovered in the oil (Wilson, 1981). The process required high energy inputs, especially for the RDF process (over 100 kwh/ton) (Wilson, 1981).

Like the Landgard process, the San Diego project never performed as well as the small scale predecessors. Due to various problems during shakedown testing, the longest run lasted less than four hours and the plant only reached 20% of its capacity (Chatterjee, 1978).

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[5] Showa Denko asserts that this plant has been operating (Aoki, 1994), but the most recent paper or article about this plant was dated from 1981 (Yasui and Masuda, 1981).

[6] Garrett Research and Development Co. is the research division of Occidental. This should not be confused with the GERE (Garrett Energy Research and Engineering Co.) process, which uses a multiple hearth.

3.5.2.2 International Experience

The early concurrent failures of the Occidental and Landgard projects were setbacks for the U.S. R&D effort in MSW gasification. Many ongoing R&D projects at the time never evolved beyond the pilot scale stages. Some projects worth mentioning are the Urban Research and Development Corporation (URDC) fixed-bed slagging gasifier, Hamilton Standard Refu-cycler, Battelle's fixed-bed gasifier, and Devco Management's rotary kiln. Though there were no operating demonstration plants in the U.S. after the initial failed attempts, successful large scale projects were demonstrated overseas in Japan and Europe.

Stardust '80 was a project supported by the Japanese government that encouraged innovative waste disposal methods. The Agency of Industrial Science and Technology (AIST) of the Ministry of International Trade and Industry (MITI) sponsored research in MSW disposal technologies from 1973-1976. This included classification, shredding, and gasification technologies. Beginning in April 1976, they began the national project known as the "R&D of Municipal Refuse Resources Recovery System," or Stardust '80. This was to coordinate all the new sophisticated developments in solid waste disposal.

A pilot plant was constructed in Yokohama City rated at 100 tonne/day capacity in 1978. This resource recovery plant included composting, pulping, and gasification. The gasification system was designed by *Ebara Corp.* This was a dual bed indirectly heated reactor, much like the BCL system, which will be described in detail later. It operated from 1979-1982 and logged 8000 hours of run time (Ishii et al., 1983).

This test plant showed great adaptability, and a large scale (450 tonnes/day) dual bed pyrolysis plant began operation in Funabashi City in 1982. The Funabashi City plant was constructed by *Tsukishima Kikai Co., Ltd.* After initial troubles with the gas scrubber and residue removal system, the Funabashi City plant operated regularly for several years. Though it accomplished its goals of disposing waste cleanly, it was closed after 1988.[7] High operating cost is the most likely reason for its shut down. Also, the costs of processing MSW to RDF may have been large at this scale. Unlike the Yokohama City plant that used a diesel engine, the Funabashi city plant burned its gas in a conventional boiler.

In Europe, various research groups that experimented with MSW gasification showed positive results. *Siemens KWU* built a 3 tonne/hour test facility in Goldshöfe, Germany in 1985 that was successfully tested for over 10,000 hours. Their process is an indirectly heated rotary kiln that uses axial heat exchanger tubes lying within the kiln. A fraction of the product gas is burned in these tubes to provide the heat for the kiln. This process has a long residence time (one hour), so that a large fraction of the gasifier product is solid fuel.[8] A 0.2 tonne/hour demonstration plant was built in 1988 in Ulm-Wiblingen and is operating today (KWU, 1985; Siemens, 1993).

Lurgi GmbH has operated since 1986 a 4 tonne/hour commercial plant in Pöls, Austria, that uses bark from a pulp mill as feed. This is an air-blown circulating fluidized-bed gasifier (CFBG) and produces a low Btu product gas that is burned in a lime kiln. They claim that their gasifier is suitable for a wide variety of feeds, and tests with RDF have shown encouraging results (Herbert et al., 1988a). Lurgi has extensive experience in developing fixed-bed coal gasifiers.

TPS (formerly Studsvik) is the largest R&D lab in Sweden doing research on new energy technologies. They have operated a pilot scale air-blown CFBG since 1986 and have done extensive research with their circulating fluidized-bed cracker

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[7] Some sources indicate that it was operational in 1988 (SRI, 1992; Turner, 1988), but Japanese officials state that it was closed many years ago (Nakazato, 1994; and Yokoyama, 1994).

[8] This process is not examined for fuel production because data on the product gas composition and gasifier heating requirements were not available. Though it is a viable candidate, this process would not perform as well as the BCL and MTCI gasifiers because it has a low carbon conversion and cold gas efficiency.

(CFBC) as a hot gas clean up system. Various biomass feeds have been tested, including RDF in the pilot scale gasifier and cracker (Rensfelt, 1989). The first commercial RDF gasification plant was built in Grève-in-Chianti, Italy, and is based on the TPS design. Aerimpianti, a subsidiary of Ansaldo, built these two 100 tonnes/day CFBGs (Bridgwater, 1993). The RDF gasifier began operations in 1992. The product gas is of low energy value and is burned in a cement kiln and boiler. Since the TPS gasifier is an air-blown process, it is not considered for fuel production in this thesis.

Thermoselect, a Swiss company, recently converted their demonstration plant in Fondotoce, Italy, into a commercially operating plant. This is an entrained-bed gasifier and is analyzed in this study. The Thermoselect and TPS processes are two plants that currently operate on a commercial scale and have demonstrated reliable operation. These systems are very different from the Landgard and Occidental processes that operated twenty years ago. Experience in MSW management coupled with increased R&D efforts in gasification have greatly improved MSW and RDF gasifier technologies.

3.6 Characteristics of Gasifiers Analyzed for Fuel Production

The fuel production analysis depends on reliable output data from MSW (RDF) gasifiers. Since the experience with RDF and MSW feeds is limited, good data are scarce. As seen from the discussion above, only a few of the plants that have been built have not been shut down. Only processes that are currently operating at least at the pilot scale and manifest commercial potential have been chosen to be examined. Processes that are in operation today but are air-blown (TPS, IGT[9]) were not considered for this fuel production analysis for reasons stated before. Three gasifiers

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[9] Institute for Gas Technology (IGT) has extensive experience with their pressurized BFBG that has been analyzed by OPPA (1990), Wyman (1993), and Katofsky (1993) among others. They have run only one test with RDF feed with air blowing (Onischak et al., 1988).

Table 3.7 - Gasifier Characteristics

Gasifier	BCL[b]		MTCI[c]		Thermoselect[d]	
Type of bed	Fast Fluidized		CFB		Entrained	
Feedstock	RDF		RDF		MSW	
Feed moisture (%)	22.5		17.0		20.0	
Feed dry HHV (MJ/kg)[a]	18.1		17.5		15.0	
Additional inputs						
steam (kg/kg dry feed)[a]	0.31		1.00		0.00	
oxygen (kg/kg dry feed)[a]	0.00		0.00		0.41	
Dry feed throughput (kg/m^2-s)	2.7		0.08		n.a.	
Exit Temperature (°C)	804		788		1200	
Residence Time	seconds		minutes		seconds	
Product gas composition (volume %)	Wet	Dry	Wet	Dry	Wet	Dry
H_2O	61.2		52.4		20.0	
H_2	7.2	18.8	22.3	46.9	32.6	41.6
CO	16.2	42.2	6.8	14.3	33.9	43.2
CO_2	4.2	10.9	11.8	24.9	11.9	15.2
CH_4	6.5	16.9	3.9	8.2	0.0	0.0
C_2+	4.3	11.2	2.7	5.7	0.0	0.0
N_2	0.4		0.0		1.6	
Yield (kmol/dry tonne)[a]	67.4		108.7		69.4	
HHV (MJ/kg wet gas)	10.7		10.0		10.2	
HHV (MJ/Nm^3 dry gas)	21.3		15.2		10.6	
Cold gas efficiency (%)	79.1		84.8		87.9	
Carbon conversion (%)	63.2		70.8		96.0	

Notes: n.a. = not available

a) For MTCI and BCL this is in terms of dry RDF feed. For Thermoselect it is in terms of dry MSW feed

b) BCL data taken from Paisley (1989) and Paisley (1994a)

c) MTCI data taken from MTCI (1992) and Durai-Swamy (1994a)

d) Thermoselect data from Runyon (1993) and Riegel and Runyon (1993)

that have used MSW (or RDF) feeds are currently in operation, produce a medium energy value gas, and had available data are used for this analysis: BCL, MTCI,[10] and Thermoselect. The BCL and MTCI systems are low temperature indirectly

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[10] The BCL and MTCI gasifiers have only limited experience with RDF, so they are not discussed in the previous section (3.5).

heated gasifiers, while the Thermoselect gasifier is a high temperature oxygen-blown entrained-bed gasifier. The primary characteristics of these three gasifiers are summarized in Table 3.7.

Before describing each of the three gasifiers that were examined, the feeds assumed for the characterizations in Table 3.7 are compared. Since MSW is such a variable feed, available data reflect somewhat different feedstock compositions for the different gasifiers. Table 3.8 compares the average composition of each feed on a moisture- and ash-free (MAF) basis.

These differences in feedstock composition affect the gasifier product outputs. The MTCI feed has a high hydrogen/carbon ratio in comparison to the MSW chemical composition estimates of Chapter 2 (table 2.11). The MTCI feed also has a very high sulfur content. The Thermoselect feed has a fairly low H/C ratio. The BCL feed has a fairly "typical" C/H/O ratio. Both of the RDF feeds have unusually low chlorine contents.

Table 3.8 - Gasifier Feed Characteristics[a]

Process	MTCI	BCL	Thermoselect
Feed	RDF	RDF	MSW[b]
	% mass (MAF basis)		
C	47.0	50.0	55.6
H	7.6	6.4	6.5
O	43.8	43.0	36.9
N	0.8	0.4	1.0
S	0.87	0.15	N/A
Cl	0.01	0.00	N/A
moisture %	17.0	22.5	20.0
ash %	12.7	9.1	25.5
dry HHV (MJ/kg)	17.5	18.1	15.0

Notes: N/A = not available

a) Data taken from MTCI (1992), Paisley et al. (1989), and Runyon (1993).

b) Ultimate analyses were not carried out for Thermoselect feeds. The composition here is estimated from mass balances, given the outputs and inputs to the gasifier, and has been corroborated with Thermoselect (Runyon, 1993).

Battelle Columbus Laboratories (BCL)

Battelle Columbus Laboratories has successfully produced medium-Btu syngas with their indirectly heated gasifier using a wide variety of wood feedstocks including hardwood and softwood chips, bark, sawdust, and others. In 1989, BCL ran a series of tests using RDF feedstock and found that performance was similar to that with biomass feed.

The BCL atmospheric pressure, indirectly heated gasifier consists of two separate reactors. One reactor is a circulating fluidized-bed gasifier that converts the feed into a medium-Btu product gas. The other is an air-blown fluidized-bed combustor that burns the residual char to provide the necessary heat to gasify the feed. Sand is circulated between the two reactors to transfer the heat. Steam is injected into the gasification chamber to act as a fluidizing agent (see Figure 3.7). Since the oxidizing and reducing environments are separated into two vessels, the process generates a medium energy value gas without the use of pure oxygen. The process is called a fast fluidized process because it has much higher throughputs than typical fluidized-bed gasifiers. Residence times are on the order of one second (Katofsky, 1993). Char gasification reactions do not occur to any significant extent (Paisley, 1989).

The process operates most efficiently when it is in *balanced* mode (Paisley, 1994), i.e., when no auxiliary fuel is used in the combustor to provide additional heat (which would raise the gasifier temperature). In *balanced* mode, the unreacted char from the gasification vessel provides enough heat when burned to gasify the rest of the feed. The operating temperatures are low (800 °C) in this case, so that the carbon conversion rates are low. In truth, nearly all of the carbon in the feed is converted into gas, but considerable CO_2 is in the flue gas rather than in the product gas. Higher temperatures can be reached if auxiliary fuel is added, but this lowers the overall process efficiency.[11]

The BCL pilot plant based in Columbus, OH operates with biomass at a dry feed rate of 12 tonnes/day. For their tests with RDF fluff, the feed rates had to be reduced because the fluff density is much less than that of wood.[12] The limiting factor in achieving high throughputs is the capacity of the feed system. Commercial plants would overcome this by adding extra feed lines.

The product gas characteristics are adapted from a scale-up model developed by BCL. The most conspicuous feature of the product gas is that it has an extremely high moisture content (61%). The RDF is not dried, as it would be for biomass feeds, and a large amount of steam is injected into the gasifier as a fluidizing medium.[13] The product gas has the highest percentage of methane and higher hydrocarbons of the three gasifiers. This is not surprising considering the short residence time of the feed. The concentrations of methane and higher hydrocarbons are greater (on a dry basis) for RDF feeds than biomass, which is most likely due to the high plastic content of RDF. The tar/oil content of the gas is similar to that for biomass feeds. Less than 1% of the feed carbon ends up as tars/oils. Another important feature is that there is much more CO than H_2 in the gas.

BCL has operated the pilot plant for over 20,000 hours. Recently a 200 kW gas turbine has been added to the pilot plant, and over 50 hours of tests have been run. A 200 tonne/day biomass gasifier/gas turbine plant to be set up in Burlington, Vermont is currently in the design and permitting stages. This project is receiving support from

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[11] The Funabashi City plant had a similar design but did not operate as efficiently. It operated at 650°C with auxiliary fuel to provide enough heat (unbalanced mode). The cold gas efficiency for the Funabashi City plant was only around 40% (Igarishi et al., 1984).

[12] Though the RDF fluff has a packed bulk density of 8 lbs/ft^3 (130 kg/m^3), which is similar to dry biomass, it only has a density of 1 lb/ft^3 (16 kg/m^3) after handling (Paisley et al., 1989).

[13] BCL has improved its fluidization system since running these RDF tests. The gasifier now requires much less fluidizing steam - only 0.02 kg H_2O/kg dry biomass feed compared to the old steam/feed ratio of 0.3 (Paisley, 1994a). No data for RDF feeds have been provided since making these changes however. Only the high steam/feed ratio case will be considered in this analysis.

the US Department of Energy, and is expected to begin operating in early 1996 (Paisley, 1994b).

Manufacturing and Technology Conversion International (MTCI)

MTCI is a small US company based in Columbia, MD. MTCI has designed an indirectly heated gasifier that has demonstrated the capability to produce a medium energy value gas from biomass feeds. The company has tested a wide variety of unusual feedstocks, ranging from pistachio shells to RDF.

The MTCI design uses a fluidized-bed gasifier with in-bed heat exchanger tubes (Figure 3.8). A fraction of the product gas is burned with air in a pulse combustor, and the hot combustion products travel through the resonance heater tubes to provide heat for gasifying the feed. The pulse combustor creates high frequency fluctuations, which induces a significant turbulent flow in the heater tubes. This pulse combustion greatly reduces the convective boundary layer of the flow, leading to high heat transfer rates in the heat exchanger tubes. The heat transfer rates can be up to five times greater than conventional heater tubes' rates (Durai-Swamy et al., 1991). Hence, fewer firetubes are required and less fuel has to be burned than with conventional designs (e.g. - Siemens KWU). A large amount of steam (triple the steam/feed ratio of BCL) is injected into the gasifier as the fluidizing medium.[14]

The gasifier temperature is quite low (650-800°C), but the residence times here are much longer than for the BCL gasifier. The gas residence time is of the order of minutes and the solids are recirculated to the bed. The long residence time coupled with large steam inputs leads to high carbon conversion rates.[15]

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[14] In the bench scale test, the gasifier needed over 3 kg water/kg dry feed. The commercial scale gasifier would require much less fluidizing steam per dry feed than small scale tests. As for MTCI's scale up experience with other feedstocks, it is assumed in this analysis that the steam/feed ratio is 1.0 (Durai-Swamy, 1994a).

[15] MTCI found that RDF gave lower carbon conversions when gasified without solids recirculation compared to biomass. A carbon conversion of 83% was reported for the bench scale test. It is assumed that the char caught in the cyclone would be continuously recirculated and converted to gas to

Since the RDF tests were done on MTCI's bench scale reactor, certain characteristics can be expected to change for a large scale operation. With the guidance of MTCI staff and their test data, the gasifier characteristics and performance parameters for a large commercial plant have been estimated.

Though the biomass fed cases resulted in low methane yields, RDF feeds yielded more higher hydrocarbons and methane. The dry product gas has the highest concentration of hydrogen and the lowest CO concentration for the three gasifiers analyzed here. It seems that the moderately long residence time allows for the water-gas shift reaction to occur in the gasifier. Analyses of the product gas revealed that it contained low amounts of condensable gases (<0.01% of the feed carbon). As was the case for biomass feeds, the low tar/oil yield can be attributed to the moderately long gas residence times and high steam rates.

MTCI has operated a 24 tonne/day unit fed with pulp mill sludge in Ontario, CA. They also have a 48 tonne/day pilot plant in Erode, India that gasifies black liquor, a residue from paper processing. A similar plant gasifying 90 tonnes/day black liquor is undergoing start up tests in New Bern, North Carolina. In Pennadam, India, a 110 tonne/day plant that is fed sugar mill distillery residues will begin full commercial operation in March of 1995 (Durai-Swamy, 1994b).

Thermoselect S.A.

This Swiss based company has designed a high temperature process specifically for MSW and other waste feedstocks. Thermoselect has a 4 tonne/hr plant running in Fondotoce, Italy that processes raw MSW. Thermoselect is the only one of the processes modeled in this thesis that is already commercialized. The commercial Thermoselect design is a modular unit with a processing capacity of 480 tonnes/day. Multiple modules would be used to increase capacity.

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give a 96.3% gross carbon conversion rate. This has been corroborated by MTCI (Durai-Swamy, 1994).

The Thermoselect process consists of a pre-processing oven and an oxygen-blown entrained-bed slagging gasifier (Figure 3.9). The MSW is dumped into a baling chamber that compresses the feed into packages that are 10% the volume of the original feed. The compressed feed is fed to a channel that is heated to 600 °C. This channel acts as a coking oven, where the feed is dried and pyrolyzed. The feed remains in this coking oven for over one hour. Water vapor and pyrolysis gases flow ahead to the gasifier. The carbonized briquettes break apart when fed to the gasifier and are entrained by oxygen injected into the reactor. The gas residence time is around four seconds. Temperatures in the oxidation zone reach 2000 °C, which is high enough to melt all the non-organic matter into liquid slag. The product gas leaves the top of the gasifier at over 1200 °C. The slag is drained to water baths below (Riegel and Runyon, 1993; Runyon, 1993).

The process requires oxygen (0.4 kg O_2/ kg dry MSW). Some of the cleaned product gas can be burned to provide heat for the coke-oven. To ensure high temperatures for the slagging process, auxiliary fuel can be injected into the gasifier. In this analysis, it is assumed that a small fraction of the medium energy value product gas is recycled to the gasifier to provide the high temperatures required.[16]

The product gas from this high temperature process closely matches the chemical equilibrium composition, having high CO and H_2 contents, negligible amounts of methane, and no tars and oils. Essentially all the carbon is converted to gas at these high temperatures.[17] The water vapor levels are also low since no steam is injected in this process.

The Fondotoce plant is now in commercial operation, processing nearly 100 tonnes raw MSW/day. Possible sites for Germany and Switzerland have been discussed, and the US is another targeted region (Runyon, 1994). The marketed

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[16] The adiabatic flame temperature of the clean product gas in oxygen is about 2700 K.

[17] Some of the gas must be recycled to the gasifier, which reduces the carbon conversion to 96%.

plants are fully operational, having extensive gas clean up and water treatment systems. The product gas fuels a 10 MW_e diesel engine (for a 480 tonne/day raw MSW plant) (Riegel and Runyon, 1993).

3.7 Summary

A crucial step in producing hydrogen and methanol from solid wastes is gasifying the feedstock. The feed preparation and downstream syngas processing steps can use commercial off-the-shelf technologies. Though coal gasification has been practiced for over one hundred years, there has been limited success with gasifying MSW. Demonstrations in the US twenty years ago were plagued with operational problems, but more recent tests have shown significant improvements. The higher reactivity of MSW and biomass compared to coal allows for new gasifier configurations that operate at lower temperatures. These indirectly heated gasifiers are potentially cost effective and efficient, but are in the early stages of development. High temperature, directly heated processes that are based on coal gasifiers are better established, and at least two directly heated plants are in commercial operation abroad. One of these directly heated commercial gasifiers and two indirectly heated processes still in the demonstration stages have been examined and characterized. Each of these three gasifiers requires a different system configuration to produce fuel. This affects the performance and costs of fuel production and will be discussed in the following chapters.

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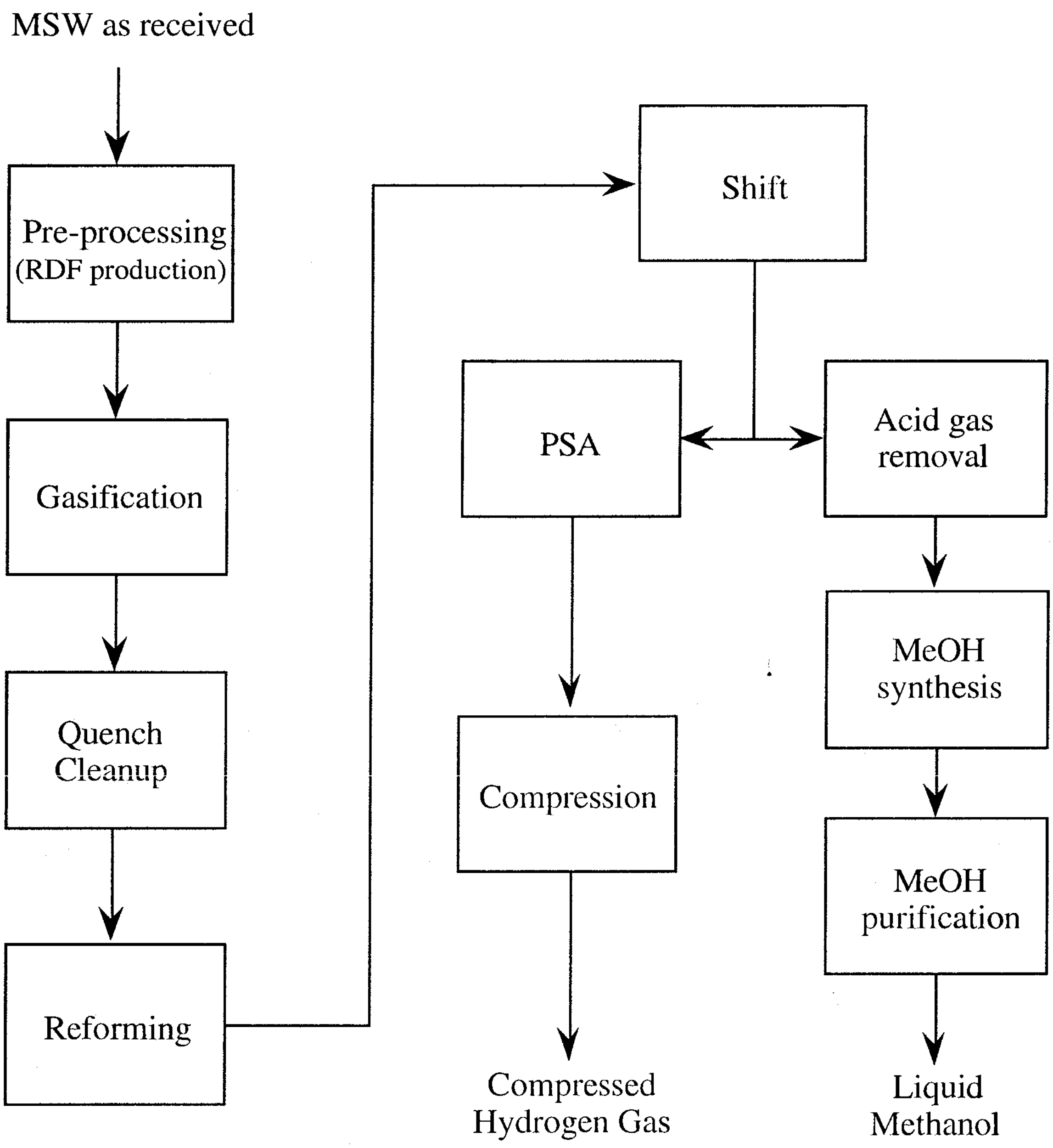


Figure 3.1 - Process Diagram of Methanol and Hydrogen Production from Municipal Solid Waste

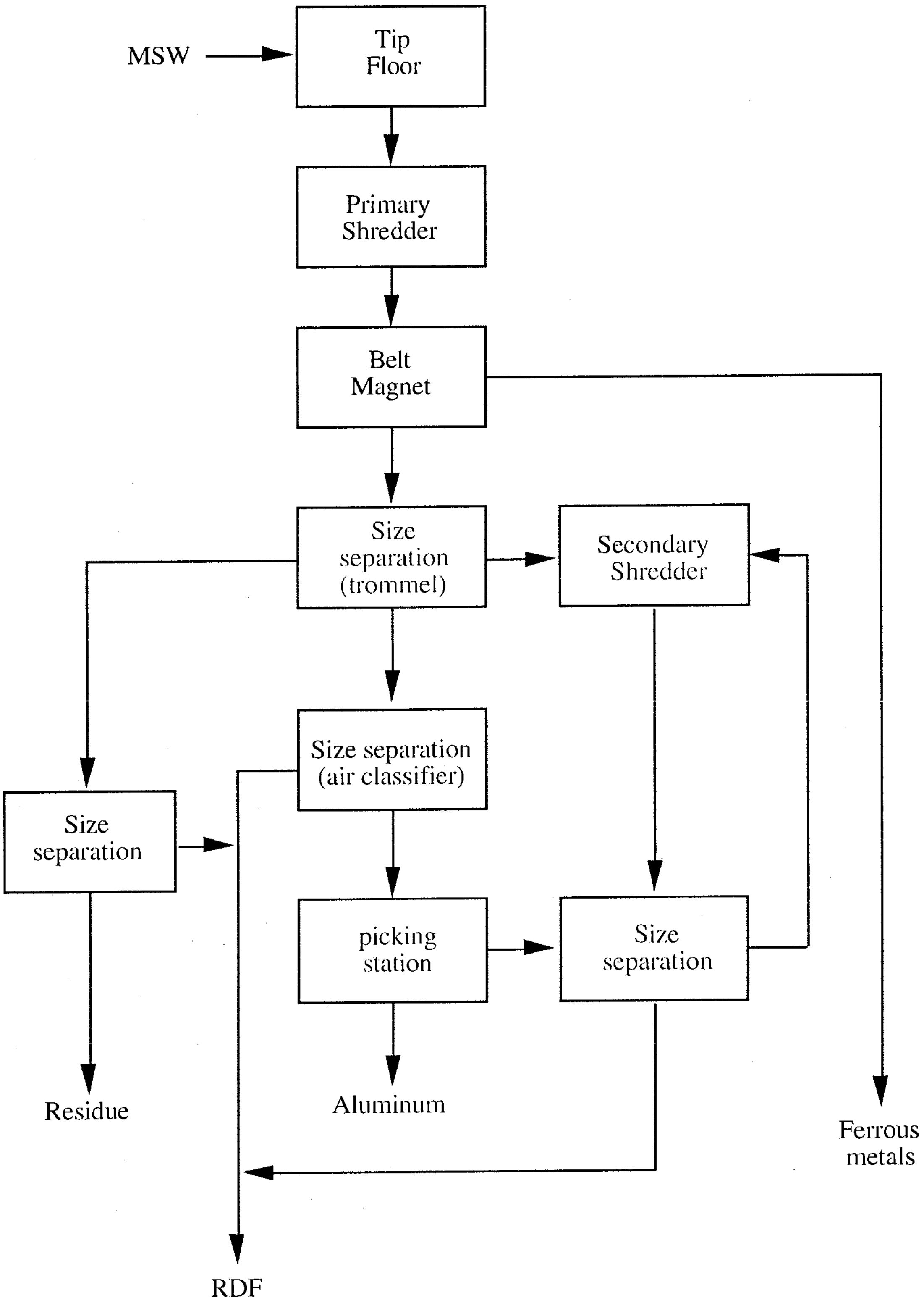
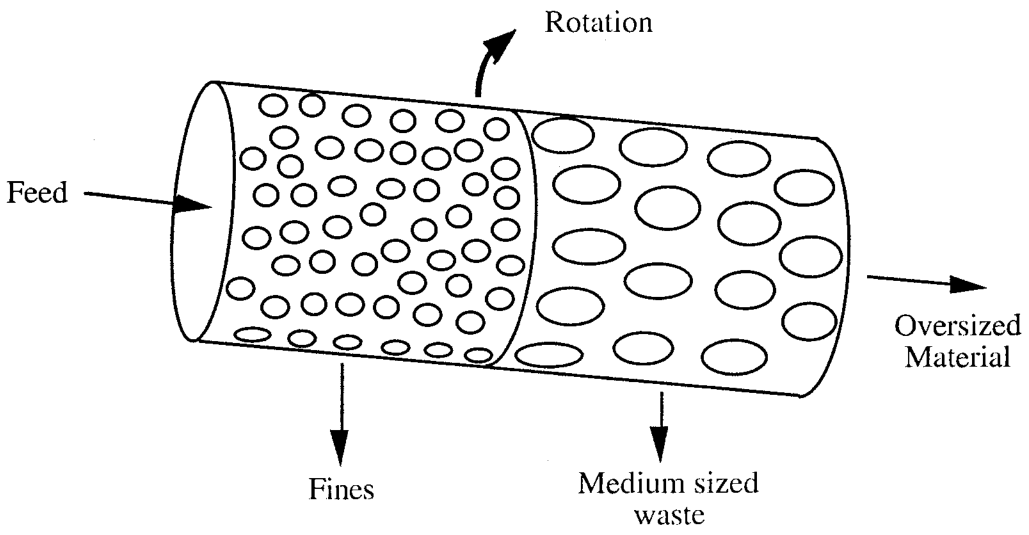
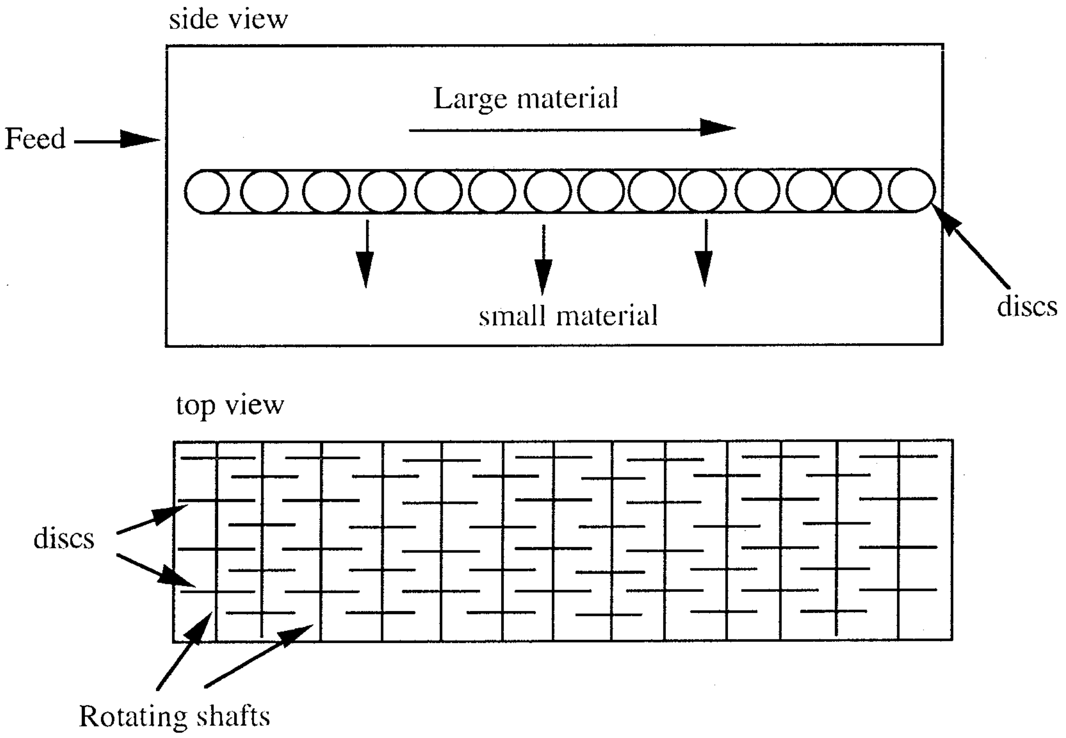


Figure 3.2 - Flowsheet of a typical RDF production facility (based on Palm Beach County plant)

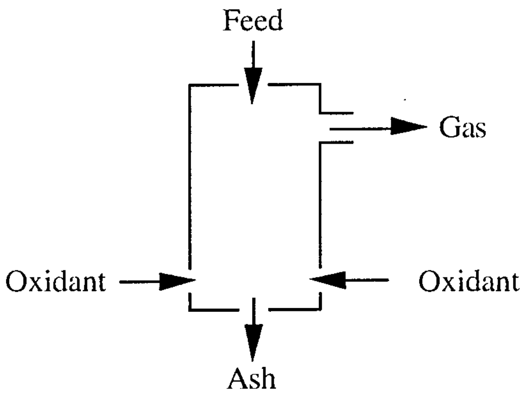


Trommel Screen Use

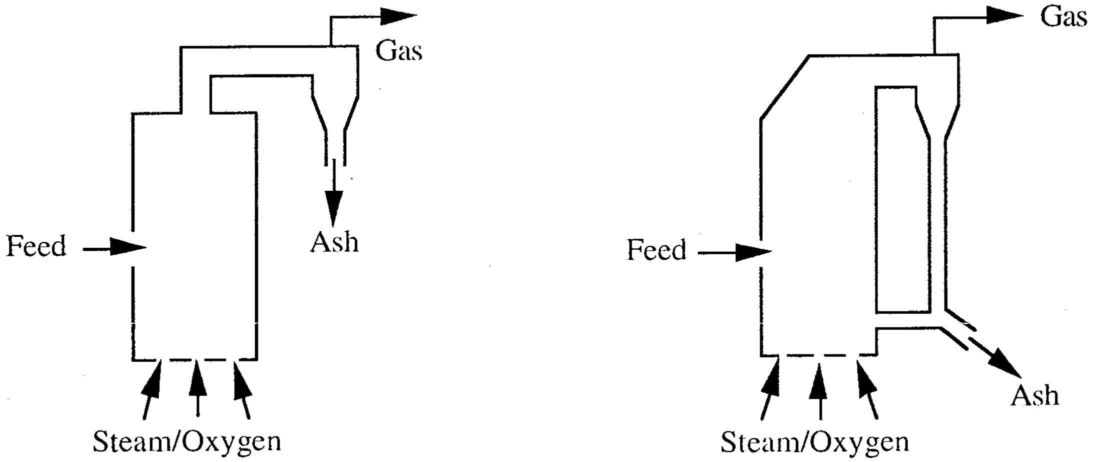


Disc Screen Use

Figure 3.3 - Diagrams of size separation equipment

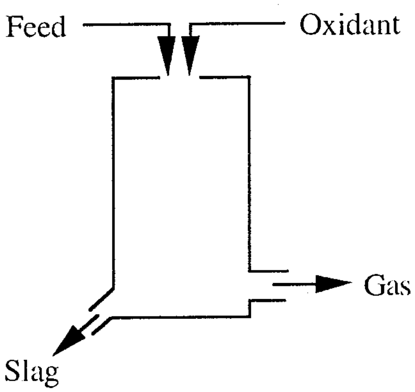


Updraft fixed-bed



Bubbling fluidized-bed

Circulating fluidized-bed



Downdraft entrained-bed

Figure 3.4 - Directly heated Gasifiers

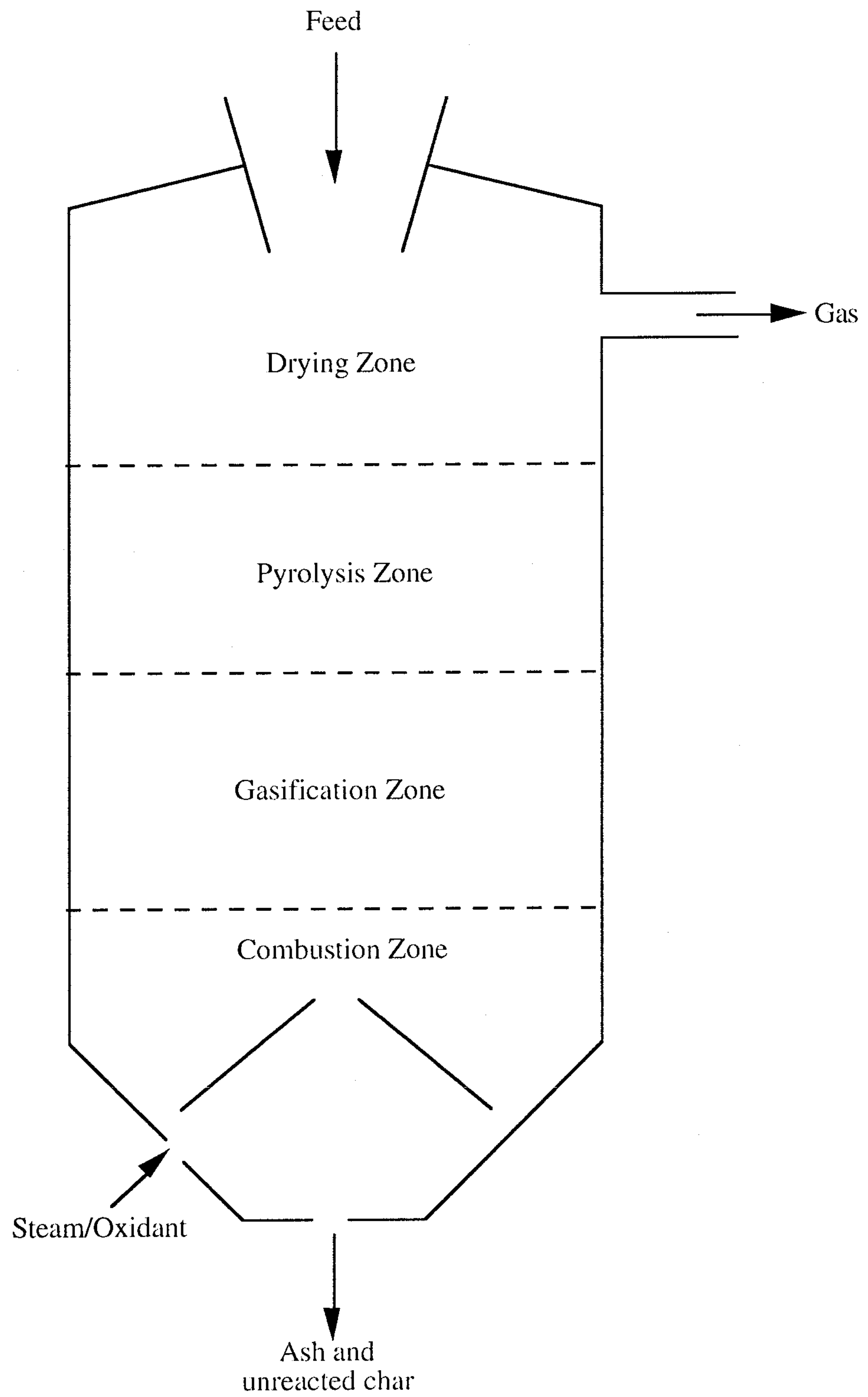
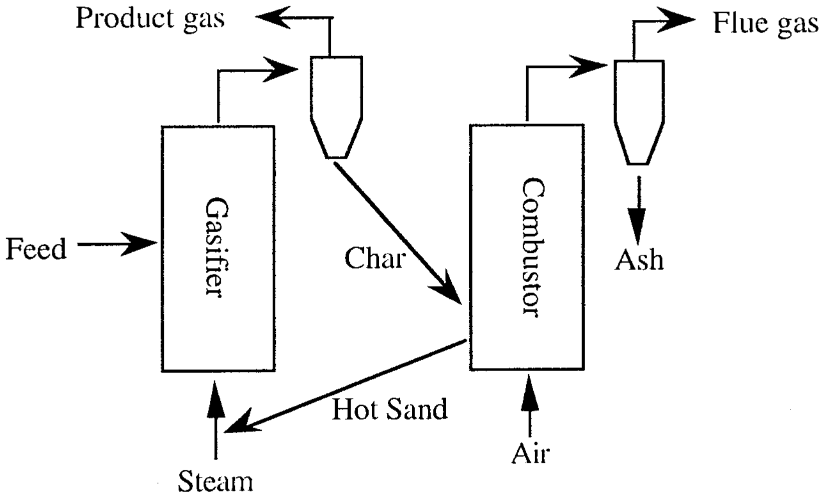
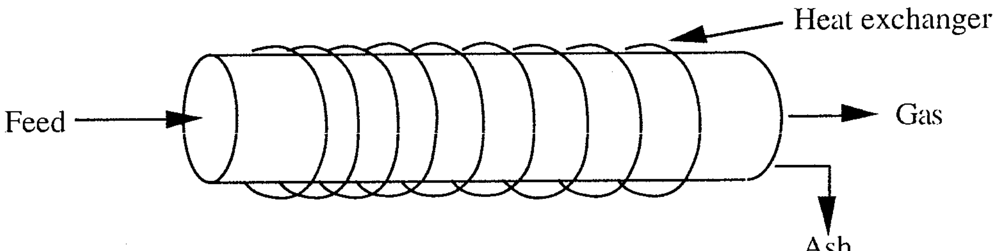


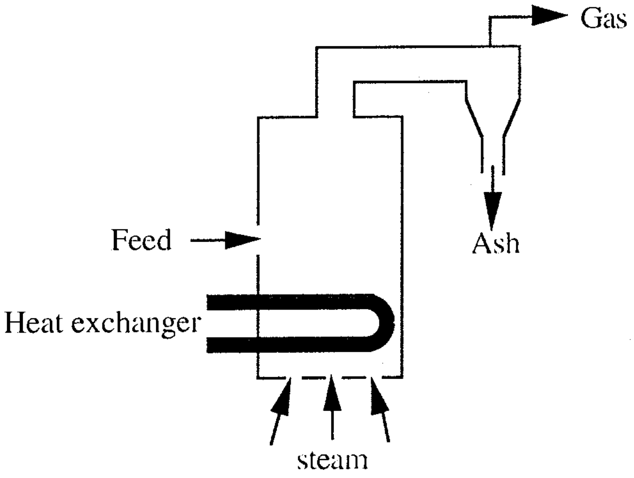
Figure 3.5 - Schematic of the different zones in a fixed-bed gasifier



Fluidized bed gasifier with circulating heating medium (sand)



Rotary kiln with external heat exchangers



Fluidized bed with in-bed heat exchangers

Figure 3.6 - Schematics of heating methods for indirectly heated gasifiers

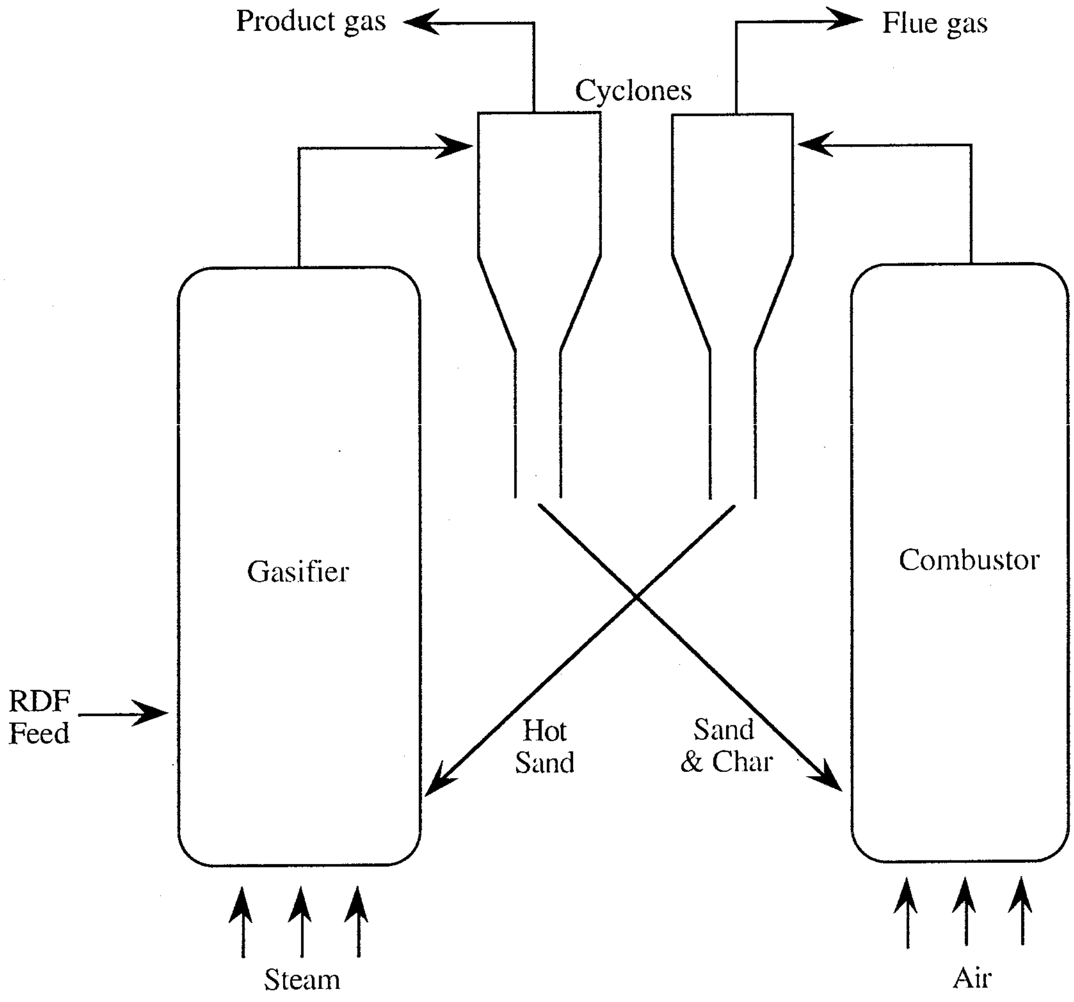


Figure 3.7 - BCL indirectly heated gasifier

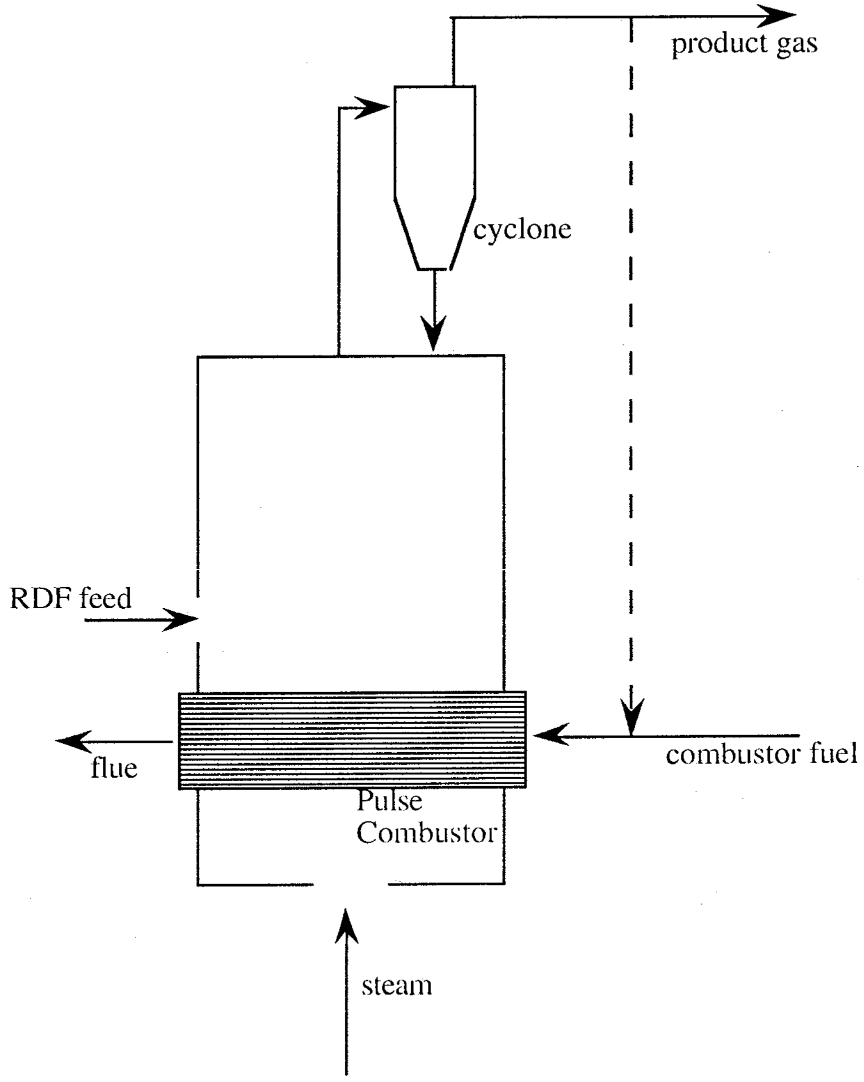


Figure 3.8 - MTCI indirectly heated gasifier

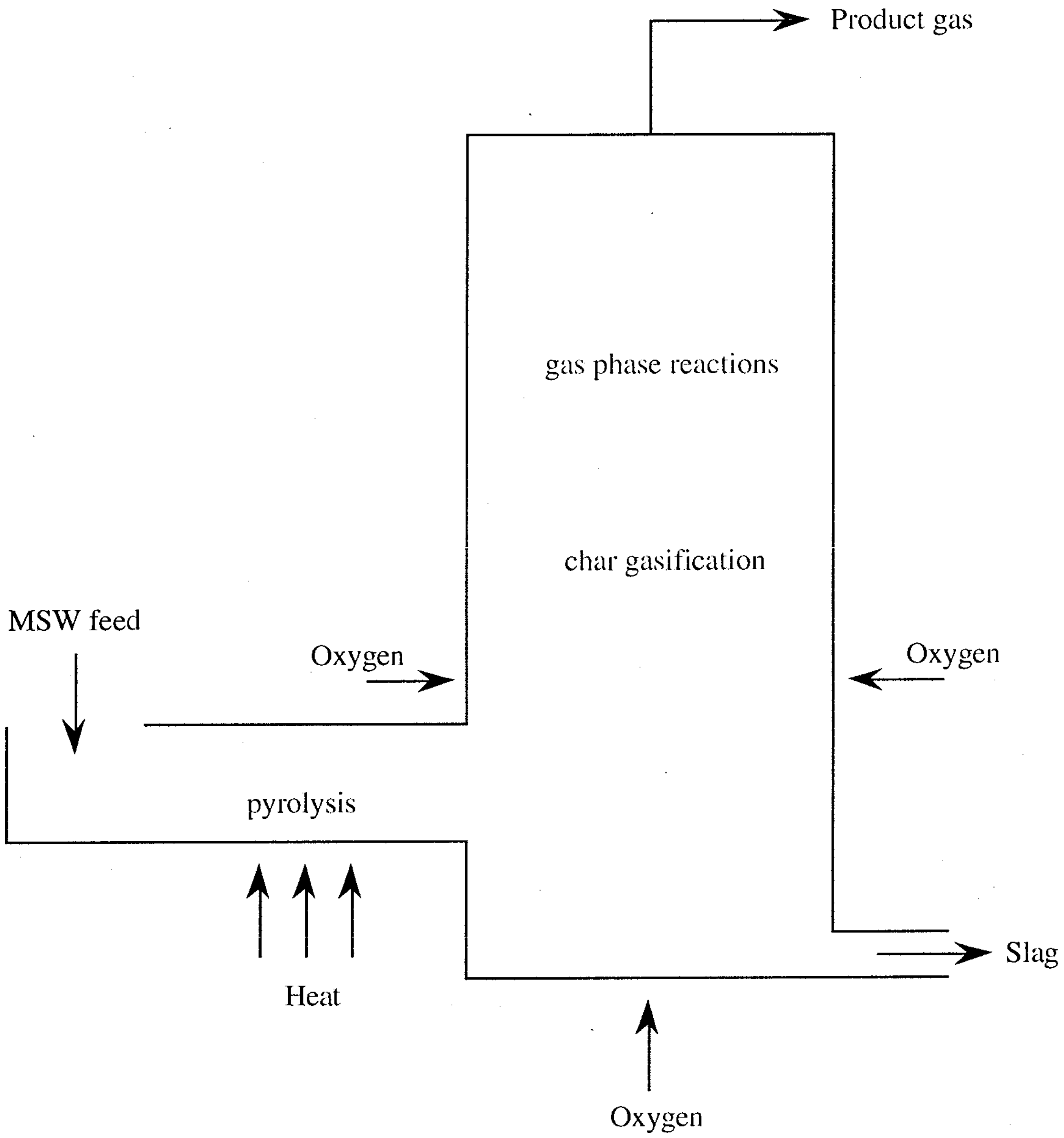


Figure 3.9 - Thermoselect high temperature slagging gasifier

Chapter 4: The Fuel Production Model

4.1 Overview

After gasification, the product gas derived from MSW undergoes a series of commercially established gas treatment steps (See Figure 3.1). In Katofsky (1993), the gas processing reactors that convert the product gas into MeOH or H_2 are discussed in detail. Based on his analysis of the various gas processing reactors, Katofsky developed models of the fuel production process for biomass, coal and natural gas feeds using ASPENPLUS™ process simulation software.[1] Since it is too difficult to model gasifiers accurately, product gas data are used as inputs to the simulations. Given that the fuel production process after gasification is basically the same for biomass and MSW feeds, Katofsky's simulations are used to model fuel production from MSW. In these MSW simulations, the inputs are data from BCL, MTCI, and Thermoselect. Modifications made to the process configurations that account for the different product gas compositions and gasifier operating characteristics are discussed in Chapter 5.

In this chapter, the gas processing equipment that upgrade product gas into fuels is briefly discussed. Also, the reactor design parameters and other assumptions used in the base case models are described. More details on each of the gas processing units and the fuel production model can be found in Katofsky (1993).

4.2 General Remarks on the Fuel Production Model

The output from the gas process simulations is analyzed in Chapter 5 to calculate the energy efficiencies of MeOH and H_2 production from MSW using the three gasifier technologies. A summary of all the thermodynamic states assumed for each step in the fuel production model can be found in the process flow diagrams and mass balance tables of Appendix 5A. Details concerning technological aspects that

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[1] ASPENPLUS™ is a widely used simulator for industrial and chemical process design.

are important for practical applications but do not have significant impacts on the thermodynamic performance have not been modeled.

Consistent with Katofsky's analysis, only widely used commercialized processes are modeled here. New gas processing technologies (e.g. - catalytic autothermal reformer, partial oxidation reformer, one pass methanol synthesis reactor, etc.) have not been considered for this model because they are not yet in widespread commercial use and would only marginally improve the system efficiencies (Katofsky, 1993), although they might have more significant impacts on reducing capital costs.[2]

Table 4.1 - Exergy Balances (in MW) for Fuels Production from Biomass Feeds Using the BCL Gasifier[a]

	Methanol			Hydrogen		
exergy inputs						
feed	402.22			402.22		
fuel[b]	11.7			57.7		
oxidant[c]	0.2			0.2		
exergy output	245.4			259.1		
exergy losses		% losses	% input		% losses	% input
gasification	62.3	37.0	15.0	62.3	30.9	13.5
cleanup	3.4	2.0	0.8	3.4	1.7	0.7
heat transfer	38.3	22.8	9.3	42.6	21.1	9.3
reforming	11.9	7.1	2.9	12.2	6.1	2.7
MeOH synthesis	9.4	5.6	2.3	----	----	----
Hydrogen separation	----	----	----	12.3	6.1	2.7
effluents[d]	17.1	10.2	4.1	13.7	6.8	3.0
other[e]	25.8	15.3	6.2	54.9	27.3	11.9
Total	168.2	100.0	40.6	201.4	100.0	43.8
exergy efficiency (%)	59.3			56.3		

Notes:
a) from Katofsky (1993)
b) Fuel is burned to provide electricity
c) Air burned in reformer furnace
d) Effluents includes methanol leaving system
e) This includes the exergy losses for external electricity requirements.

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[2] Information concerning new designs in gas processing reactors and a more detailed discussion of the technological constraints in the process can be found in Katofsky (1993).

In his analysis of biomass feeds, Katofsky shows that the exergy losses due to the gasifier account for 30-50% of the total exergy losses (14-22% of input exergy), by far the single largest source of irreversibilities in the complete fuel processing system (Table 4.1). Since only 1-4% of the input exergy is lost through each of the gas processing reactors, improvements in the technologies will have minor effects on the process efficiencies.

4.3 Quench and Cleanup

(i) Particulate Matter (PM) and Halide Gas Removal

The gasifier product gas must be cleaned before it enters any gas upgrading equipment. Cyclones are simple vessels used for the bulk removal of dust particles. Other control equipment is necessary to clean out the finer particulates and gaseous pollutants. Katofsky assumed that a simple quench was the clean up step for biomass and coal feeds. In the base case analyses for MSW feeds it is assumed that a water quench which saturates the product gas, cooling it to 70-100 °C, can sufficiently force all the tars and oils, heavy metals, and particulate matter to condense out in a waste water stream. More extensive cleanup processes that may be required to remove submicron particulate matter and acidic halide gases (HCl and HF) are discussed in Chapter 6.[3]

Scrubbing the hot synthesis gas from the gasifier with water would waste the sensible enthalpy of the gas. Therefore, it is assumed that the gas is cooled to 400 °C before it is sprayed with water, allowing for over 60% of the sensible enthalpy to be extracted. At temperatures below 400 °C heavy tars and oils would begin to condense and foul the process equipment (Feldmann, 1989).

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[3] More extensive cleanup processes have little effect on the process performances, so they are not modeled here (see Chapter 6).

(ii) Sulfur Removal

Sulfur removal in the MSW cases is the only gas upgrading step that significantly differs from the biomass cases. Sulfur compounds (mainly in the form of H_2S) must be removed from the product gas because they poison many of the reactor catalysts.

H_2S cannot be removed from the product gas with a simple scrubber. For biomass feeds, Katofsky assumed that zinc oxide guard beds would remove the trace amounts of H_2S in the product gas. Unlike biomass, MSW has a fairly high sulfur concentration (0.2-1% by dry mass).[4] A sulfur removal system that can handle larger quantities of sulfur than the ZnO beds would therefore be necessary. For the base case analyses, it is assumed that a liquid redox sulfur removal system, operating at the quench exit temperature and atmospheric pressure, follows the quench. The liquid redox and other sulfur removal systems are discussed in Chapter 6.

4.4 Reformer

For the low temperature gasifiers, a significant fraction of the gas heating value is entrapped in methane and other hydrocarbons. These are upgraded for methanol and hydrogen production through steam reforming.[5] In fact, the most widely used method to produce hydrogen is by steam reforming natural gas (Probstein, 1982). The steam reformer catalytically converts methane and other paraffins to hydrogen and carbon monoxide through highly endothermic reactions such as:

$$CH_4 + H_2O \rightarrow CO + 3H_2 \qquad \Delta H^{\circ}_{298} = 210 \text{ MJ/kmol} \qquad (4\text{-}1)$$

$$C_2H_6 + 2H_2O \rightarrow 2CO + 5H_2 \qquad \Delta H^{\circ}_{298} = 347 \text{ MJ/kmol} \qquad (4\text{-}2)$$

---

[4] Biomass is 0.01-0.1% S by mass (Katofsky, 1993).

[5] An alternative scenario is not to have a reformer and to let the methane go through the system as an inert gas. Though the fuel output would be significantly reduced for the cases studied here, the extra methane could be burned or sold. These potentially profitable configurations are not considered here because this study focuses on the cases that maximize fuel output.

The reformer is modeled as a near-equilibrium reactor. Based on Bain's and Kuo's models of the steam reformer, the approach temperature is assumed to be -10°C (Katofsky, 1993). This means that the equilibrium reaction is evaluated 10 °C below the actual temperature of the reactor, but the thermodynamic state properties are evaluated at the reactor temperature. The heat duty of the reformer combustor is calculated by assuming a temperature difference of 20 °C between the reformer tubes and surrounding furnace and 2% heat losses.

Katofsky showed for the BCL gasifier output, the optimal reformer temperatures would be around 870 °C. At this temperature, 90% of the methane in the feed gas would be reformed. Although reformers can operate at 1000 °C, lower operating temperatures would extend the catalyst lifetimes (Katofsky, 1993). The reformer is assumed to operate in a temperature window of 760-870 °C in the simulations. At 760 °C, the methane slip (% methane that does not get reformed) is high, but all the higher paraffins are reformed. High methane slip is desirable in some cases considered here, because methane is a vital component of the purge gas that is combusted to supply the external heat input for the reformer or other endothermic processes.

The number of moles of products exceeds the number of moles of reactants for the steam reforming reactions. Le Chatelier's principle says, therefore, that the reformer will perform best at low pressures. Reformer costs, however, scale with the volume flows, making higher pressures desirable from an economic point of view. For this reason, industrial reformers usually operate at pressures between 1-3.5 MPa (Katofsky, 1993). In the models the cleaned product gas is compressed to 1.5 MPa prior to reforming.

Carbon deposition (sometimes called coking) is a problem in the reformer that is also encountered throughout the fuel processing system. Reactions 4-1 and 4-2 suggest that a steam:carbon molar ratio of 1:1 would be sufficient to reform the syngas, but commercial steam reformers maintain molar steam:carbon ratios of 3-5:1 to prevent carbon deposition in the reactor (Katofsky, 1993). Katofsky showed through thermodynamic analysis that indeed high steam:carbon ratios would prevent solid carbon formation in the reformer. In the simulations, a 3:1 molar ratio is maintained by raising steam and adding it to the gas stream prior to the reformer.

Steam Reformer Guard Beds

Unsaturated hydrocarbons, like olefins or alkynes, will cause carbon deposition and foul the reformer catalyst bed, so it is necessary to hydrogenate these hydrocarbons in the reformer feed gas into paraffins. A cobalt-molybdenum guard bed will catalytically convert all the olefins into saturated hydrocarbons through reactions similar to

$$C_2H_4 + H_2 \rightarrow C_2H_6 \qquad \Delta H^{\circ}_{298} = -137 \text{ MJ/kmol} \qquad (4\text{-}3).$$

The cobalt-moly bed requires a supply of hydrogen and will be effective only if small amounts of unsaturated hydrocarbons are in the gas stream. A maximum of 10% olefins by volume is tolerable (Bochow, 1994). The product gases from the three gasifiers examined have abundant hydrogen and minimal olefin concentrations. The BCL product gas has the highest fraction of olefins at 4%.

The hydrogenating reactions are exothermic, increasing the gas stream temperature as gases pass through the bed. Since the gas stream exits the feed compressor around 300 °C, the cobalt-moly bed is self-cooled by incoming gases and operates adiabatically near 400 °C, its optimal operating temperature (Bochow, 1994). The cobalt-moly bed is modeled as a perfect selective catalyst for the hydrogenating reaction, converting all the olefins into paraffins.

The steam reformer uses a nickel based catalyst that is poisoned by trace amounts of sulfur (0.5 ppm). To ensure that no sulfur reaches the reformer catalyst, a zinc oxide guard bed (ZnO) is used to scavenge the trace amounts of hydrogen sulfide that a bulk sulfur removal system (e.g. - liquid redox) does not remove.[6] As is the case with the cobalt-molybdenum bed, the gas stream is fed to the ZnO guard bed at 400 °C (Bochow, 1994).

4.5 Water Gas Shift Reactors

To upgrade the syngas further, shift reactors catalytically convert CO to H_2 through the water gas shift reaction:

$$CO + H_2O \rightarrow H_2 + CO_2 \qquad \Delta H^{o}_{298} = -41.2 \text{ MJ/kmol} \qquad (4\text{-}4)$$

For hydrogen production, it is desirable to shift as much CO to H_2 as possible. For methanol production, it is desirable to shift enough CO to make the H_2:CO ratio a little greater than 2:1.

The shift reaction is thermodynamically favored at lower temperatures, but the kinetics are favored by higher temperatures. Also, the catalytic activity of the shift reactor decreases at low temperatures. Typical shift reactors use a chrome or iron oxide to selectively catalyze the water gas shift reaction at temperatures of about 500°C. Copper-zinc or cobalt-molybdenum catalysts remain active at lower temperatures and can operate as low as 200 °C (Taylor, 1984). For hydrogen production, the model involves two shift reactors in series. The first reactor is a high temperature shift reactor, which operates between 325-600 °C, catalyzing the shift reaction for the bulk of the CO. The second reactor is an isothermal low temperature

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[6] The combination of the two guard beds can remove trace amounts of organic sulfur, like mercaptans and carbonyl sulfide (COS). The cobalt-moly bed will hydrogenate the organic sulfur into H_2S, which the ZnO bed will then remove (Bochow, 1994).

shift reactor that operates at 227 °C. The first reactor operates adiabatically because precise temperature control is not necessary. The second operates isothermally because it is important to maintain the low temperature to shift as much CO as possible. The combination of the two shift reactors converts over 95% of the CO to H_2. For methanol production, only one adiabatic high temperature shift reactor is used to shift a fraction of the reformer output.

The reaction is independent of pressure, so the entering gas stream is kept at 1.3-1.5 MPa (the reformer pressure). The shift reactors are modeled as near-equilibrium reactors, with approach temperatures of 10 °C and 20 °C for the high and low temperature reactors, respectively.[7] As with the reformer, the steam:carbon molar ratio is kept at 3:1 by feeding in extra pressurized steam to prevent carbon deposition in the reactor.

4.6 Hydrogen Separation through Pressure Swing Adsorption

Though there are many different types of hydrogen separation techniques, the Pressure Swing Adsorption (PSA) reactors seem to be the best suited to provide a pure hydrogen fuel. With fuel cell vehicles, it is extremely important that the hydrogen fuel has no CO because the PEM fuel cell is poisoned by CO concentrations of 10 ppm. The PSA is a commercially available technology that can meet such strict requirements yielding >99.9+% pure H_2.

The PSA uses highly porous zeolite beds to adsorb molecules at high pressures and desorb them at low pressures. Polar molecules, like water and carbon dioxide are strongly adsorbed, whereas nonpolar molecules like hydrogen and nitrogen are easily desorbed. Thus each different molecule will be desorbed from each PSA bed at different pressures. Each bed must go through adsorption, depressurization,

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[7] It is assumed that the low temperature shift reactor has a greater approach temperature because the reaction rates are slower at the lower temperature (Katofsky, 1993).

desorption, and repressurization steps, but many beds can be sequenced together to continuously provide hydrogen. PSA reactors can have long lifetimes of up to ten years (Corr, 1979).

Katofsky's models assume the use of the Gemini-9 PSA system from Air Products Industries, which has two sets of beds in series and a recycling loop. The first set of beds provides CO_2 at 40 °C and 0.13 MPa, and the second splits off the pure H_2 stream. The remaining stream from the second bed is recompressed and recycled to the first set of PSA beds (Figure 4.1). This method recovers about 96% of the H_2 in the gas stream at 99.999% purity and at the pressure of the incoming feed to the PSA. It also provides a separate carbon dioxide stream. There are two major drawbacks to the Gemini-9 system: (1) it is energy intensive, requiring over 25% of the total electricity inputs for hydrogen production from biomass (Williams et al., 1994); (2) it is expensive in comparison to simple PSA systems.

A one pass PSA system that provides a hydrogen stream and an aggregate purge stream has been modeled in this analysis (Figure 4.1 compares the two PSA systems). Without the recycling loop, the PSA system recovers only 89.5% of the H_2, but it requires negligible electricity and is about half the capital cost of the Gemini-9. The hydrogen stream is 99.9% pure, and methane is the only trace impurity (Solomon, 1993).[8] In the model, the gas stream is fed to the PSA at 40 °C and over 1.1 MPa. The purge stream leaves the PSA at atmospheric pressure and is used as fuel to heat various other processes (reformer, Thermoselect coke-oven, MTCI pulse combustor). The H_2 stream leaves the PSA at the inlet conditions and is further compressed to 75 MPa for storage or pipeline transmission. The single pass PSA system is assumed to use two sets of beds operating in parallel.

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[8] Higher purities of hydrogen can be attained with the penalty of lower hydrogen recovery (Solomon, 1993).

4.7 Acid Gas Removal

The quench could scrub out all of the HCl, HF, HNO_3 and other acid gases. But more sophisticated removal systems are required to remove H_2S and CO_2. H_2S removal prior to the reformer was discussed earlier. CO_2 does not poison the reformer or shift catalysts but does hamper the MeOH synthesis reactor. Therefore a gas purification system is required to separate CO_2 from the fresh feed entering the methanol synthesis reactor.

There is substantial experience with acid gas removal and many commercial processes exist. Consistent with Katofsky's analysis, the Selexol process from Union Carbide is modeled as the CO_2 removal system. The Selexol process is a physical absorption system that uses dimethyl ether of polyethylene glycol as its solvent to remove H_2S and CO_2 from gas streams. The Selexol solvent is effective at absorbing CO_2 at moderate temperatures (Katofsky, 1993). In the models, the gas stream is fed to the Selexol at 1.3-1.4 MPa and 127 °C. The CO_2 can be desorbed in a stripper or flash regenerator, which reduces the pressure or raises the temperature of the solvent so that the CO_2 vapor rises out of the liquid. The purified gas stream has a CO_2 concentration around 2% for optimal methanol synthesis conditions.

4.8 MeOH Synthesis

The main methanol formation reactions are

$$CO + 2H_2 \rightarrow CH_3OH \qquad \Delta H^{o}_{298} = -90.7 \text{ MJ/kmol} \qquad (4\text{-}5)$$

$$CO_2 + 3H_2 \rightarrow CH_3OH + H_2O \qquad \Delta H^{o}_{298} = -49.5 \text{ MJ/kmol} \qquad (4\text{-}6)$$

The shift reaction also participates:

$$CO + H_2O \rightarrow H_2 + CO_2 \qquad \Delta H^{o}_{298} = -41.2 \text{ MJ/kmol} \qquad (4\text{-}4)$$

Reaction 4-5 is by far the dominant MeOH synthesis reaction. Too much CO_2 will block the catalyst sites for the CO reaction; thus, most of the CO_2 must be removed prior to the MeOH synthesis reactor. In practice the CO_2 concentration is maintained at roughly 2% in the fresh feed stream for high MeOH yields (Klier, 1982). The two reactions 4-5 and 4-6 suggest an optimal amount of CO, CO_2, and H_2 in the MeOH synthesis feed. The "stoichiometric number" is a ratio of the molar quantities involved:

$$R = \frac{H_2 - CO_2}{CO + CO_2}$$

R = 2.01 - 2.1 for optimal results. For methanol production, only a fraction of the gas stream passes through a shift reactor. This fraction is adjusted until the feed has the proper ratio of H_2:CO and H_2:CO_2.

The widely used ICI methanol synthesis reactor is modeled for this analysis, as in Katofsky's analysis. This is a low pressure (5-10 MPa) reactor that operates most efficiently at 260 °C (Rogerson, 1984). Chrome, copper, and zinc oxides are the catalysts for this reactor, where the copper is the active component at low temperatures. The feed gas is compressed and mixed with gas recycled from the reactor output before entering the ICI reactor (Figure 4.2). To control the temperature, some of the gas is cooled and injected into the reactor at different points. Crude liquid methanol, which includes trace amounts of other hydrocarbons, is condensed out of the product stream. The rest of the gas is recycled back and mixed with fresh feed gas to the MeOH synthesis reactor. A fraction of the gas is purged to prevent the buildup of inert gases (e.g. - methane) and used as heating fuel for system processes. With this recycle loop, typically 98% of the carbon in the feed gas is converted to methanol.

In the models, four moles of recycled gas are combined with every mole of fresh feed. The synthesis reactor operates adiabatically and maintains a temperature of 260 °C. The feed gas is compressed to 10.8 MPa before it enters the synthesis loop. The crude liquid methanol is condensed out at 27 °C. Only the reactions above are modeled, and a 12.5 °C approach temperature is used, which is consistent with Kuo's (1993) model of methanol production from natural gas.

4.9 Compressors and Pumps

A maximum compression ratio of four is assumed for individual compressors or stages of compression. The major compression steps, like the compression for MeOH synthesis or the feed compression for the reformer, require three-stage, intercooled compressors. The intercoolers reduce compression duties. The compressors and pumps are assumed to have 85% polytropic efficiencies.

4.10 General Model Considerations

4.10.1 Pressure Losses

It is assumed that almost every reactor and heat exchanger contributes a 0.5 bar pressure loss to capture the effects of system pressure losses (Katofsky, 1993). This is similar to the 0.34 bar (5 psi) pressure drop approximations recommended for preliminary plant designs (Baasel, 1990). The 0.5 bar loss represents a $\Delta p/p$ range of 0.5 - 4%. For streams with pressures of one atmosphere that are not part of the main fuel process stream (e.g. - combustor flue gas streams), negligible pressure drops through heat exchangers are assumed.

4.10.2 Equations of State

ASPENPLUS™ has several built-in equations of state that can be used to calculate the thermodynamic properties of process streams. The ideal gas law is assumed for gas streams at atmospheric pressure and high temperatures (e.g. - heated air streams and combustion product streams). The ASME steam tables are used to calculate state properties for all pure water streams. For all other streams, including vapor-liquid mixtures and gas streams at higher pressures, the Redlich-Kwong-Soave equation of state is employed.

4.11 Heat Recovery & Power Generation

A pinch analysis is performed to maximize the heat recovery for each process configuration. Pinch analysis is a tool that is used to analyze and optimize heat recovery for thermochemical processes. Linhoff (1982), Karp (1990) and Katofsky (1993) are excellent references for a more detailed description of pinch analysis. Basically, the pinch analysis optimizes the heat exchange for each different process, providing a consistent basis to thermodynamically compare the various case studies.[9]

With the pinch analysis, the electricity, heat, and cooling duties for each fuel production system can be determined. If extra heat of sufficient quality is available, it is used to raise steam for power generation in a simple rankine cycle. This on-site power is assumed to provide electricity for other processes (RDF production, compression, etc.).

The thermal efficiency of the steam power plant is 27.7% using the following assumptions: The steam is raised to 6.2 MPa and 400 °C and expanded in a turbine with an isentropic efficiency of 75% to 0.005 MPa. The feedwater pump is assumed to be isentropic and the generator efficiency is 95%.

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[9] The pinch analysis does not take into account the practical considerations, such as cost and plant geometry, that may prevent the use of some heat exchangers.

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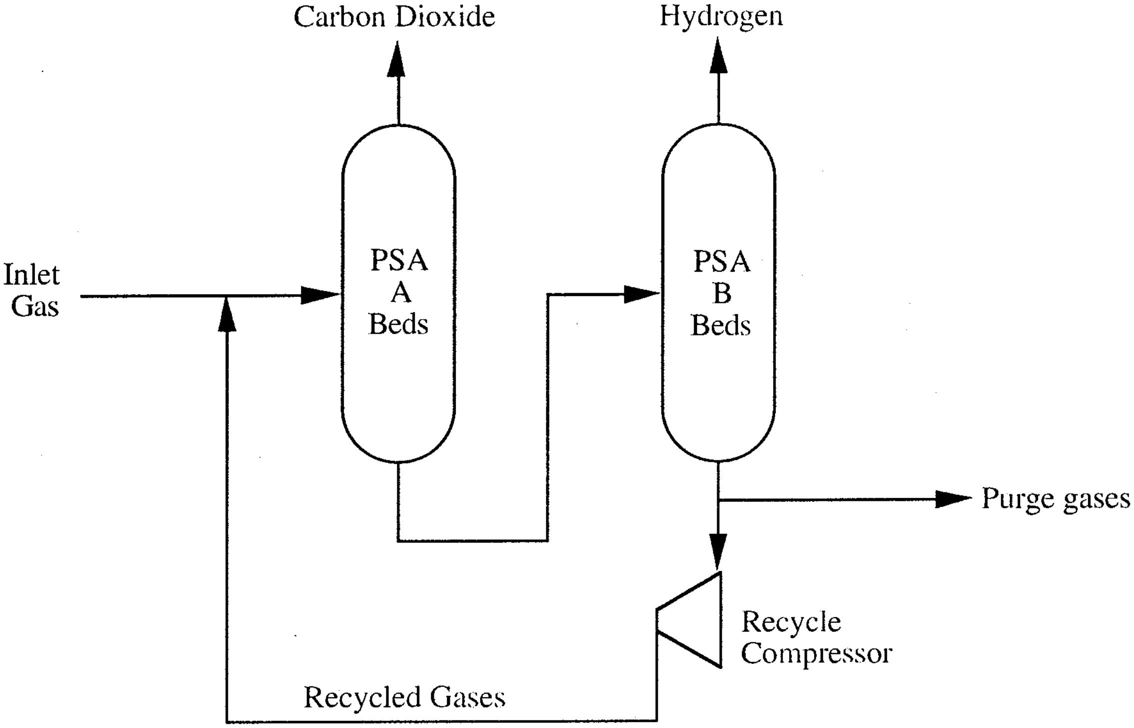
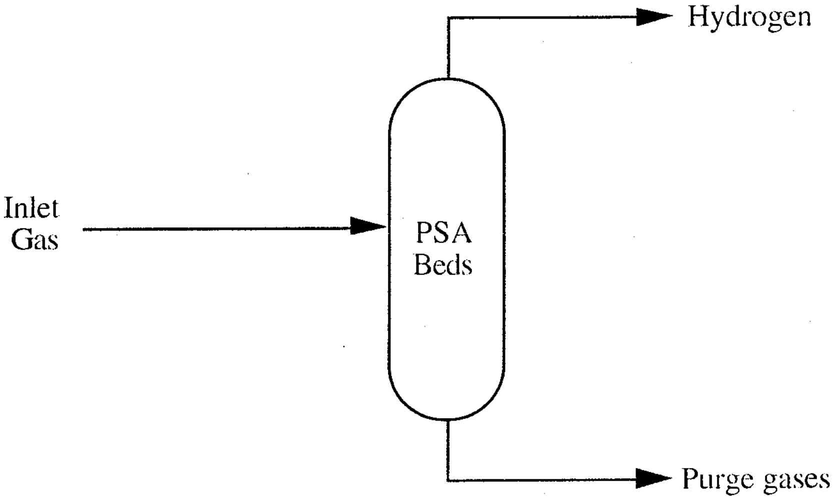


Gemini-9 system with recycle loop and carbon dioxide stream



One pass PSA system

Figure 4.1 - Pressure Swing Adsorption (PSA) configurations

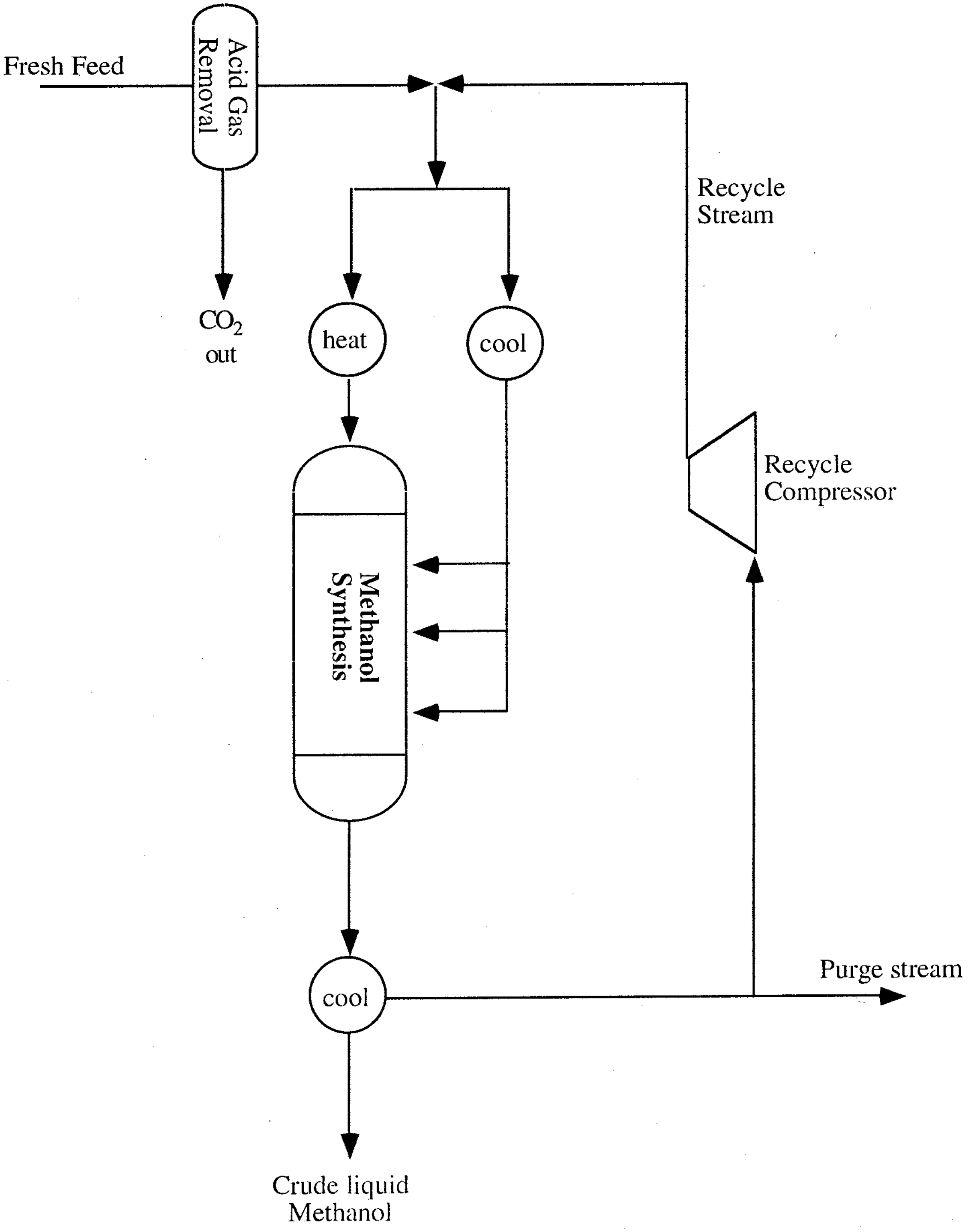


Figure 4.2 - Schematic of MeOH synthesis loop

Chapter 5: Thermodynamic Analysis of Fuel Production from MSW

5.1 Introduction

Using the model assumptions and reactor specifications described in Chapter 4, six base case process flow systems were built (methanol and hydrogen production using the BCL, MTCI, and Thermoselect gasifiers). These models are based on Katofsky's designs (Katofsky, 1993) but have been modified to reflect differences in the simulation inputs, i.e. - the product gas compositions and gasifier characteristics.

In this chapter the thermodynamic performance of each fuel production process is analyzed. Since each system configuration is slightly different, a consistent set of parameters is used in this analysis to make meaningful comparisons. Detailed flowsheets and the pinch analyses for the base case configurations are given in appendix 5A.

The analyses are carried out in the same steps used by Katofsky (1993). Each fuel production simulation is designed using ASPENPLUS™, which calculates the heat and mass balances for every reactor and the thermodynamic properties for every process stream. The system reaction-steam requirements, work (electricity) duties, and hydrogen or methanol produced by each process are displayed in the simulation results. A pinch analysis is performed for each system using the heat and mass balance information to determine the process heating requirements. As mentioned in Chapter 4, steam generated electricity from excess heat provides power for processes in the fuel production system (i.e.- feed preparation, O_2 production, compression and pumping).

5.2 Framework for Thermodynamic Analysis

5.2.1 Size of Facility

Conventional waste processing facilities range in size from 100 tonnes/day (tpd) modular combustors to 2000 tpd mass burn facilities. Capacities are rated by the solid waste disposal industry in terms of as received (wet) tonnes MSW/day because disposal costs are based on the weight of the raw MSW. But for an energy analysis, the dry feed rate is more pertinent, especially for a feed with such a variable moisture content as solid waste.

The reference plant assumed for the base case studies is a fuel production plant that processes 1050 dry tonnes MSW/day (1500 tpd raw MSW assuming 30% moisture content), the size of a typical mass burn incinerator in the US. These plants would be large enough to handle the MSW disposed by over 700,000 people. The scale of the plant will primarily affect the cost of fuel production. There effects are briefly discussed in Chapter 7.

5.2.2 Thermodynamic Performance Parameters

Two characteristic measures are used to compare the performances of the different case studies. The first is the energy ratio (ER), which is defined as the higher heating value (HHV) of the fuel output divided by the higher heating value of the input feed:

$$ER = \frac{HHV\ of\ fuel\ output}{HHV\ of\ feed\ input}. \qquad (5\text{-}1)$$

The energy ratio shows how much methanol or hydrogen each process can produce for a given feed input but does not account for any external energy inputs.

A process thermal efficiency (TE) is defined to take into account all the external energy requirements. As one way to ensure consistency for this analysis, it is assumed that the fuel production plant is isolated from external heat and electrical sources and must provide these energy inputs using MSW as the primary energy source. The thermal efficiency defined in Katofsky (1993) is also used here:

$$TE = \frac{HHV\ of\ fuel\ output}{HHV\ of\ feed + \dfrac{net\ electricity\ inputs}{\eta_{el}} + \dfrac{net\ heat\ inputs}{\eta_{ht}}}\ . \tag{5-2}$$

η_{el} is the HHV efficiency of producing electricity from MSW. The power is assumed to be produced on site with an integrated gasification combined cycle power plant (IGCC). Katofsky used linear regressions from data of expected performance (the technology is not yet commercial) to determine η_{el} (in %) for a power plant fueled by biomass:

$$\eta_{el} = 37.0 + 0.041 * Output_{MW} \tag{5-3}$$

The output in this case is the net electricity inputs to the system (7-20 MW in this thesis). It is assumed here that an RDF- or MSW-fed IGCC plant would have similar efficiencies, and the above formula is used in the calculations. η_{ht} is the higher heating value efficiency of providing high quality heat from MSW feedstocks. For the gasifiers considered here there are no cases that require external heating. “Free” external cooling is assumed to be available (e.g., from river water), so cooling needs are not included in the thermal efficiency.

5.3 Thermodynamic Results for Each Gasifier

Tables 5.1 and 5.2 summarize the energy balances for methanol and hydrogen production from MSW using the BCL, MTCI, and Thermoselect gasifiers. Figure 5.1 graphically shows the key performance results for the six fuel production processes. For either methanol or hydrogen, all three processes have similar Thermodynamic Efficiencies (TEs), but have widely varying Energy Ratios (ERs). Details on the thermodynamic performances for each process configuration follow. The BCL and MTCI systems are compared to biomass-fed systems, since fuel production from biomass has also been analyzed using these gasifiers (Williams et al., 1994; Katofsky, 1993). Process flowsheets, mass balances, and pinch analysis results for each system are detailed in Appendix 5A.

5.3.1 Methanol and Hydrogen Production from the BCL Gasifier

For both methanol and hydrogen production, the BCL cases have by far the lowest energy ratios (ERs): 0.49 for methanol production and 0.58 for hydrogen production. These low ERs can be directly attributed to the low carbon conversion and cold gas efficiency of the BCL gasifier. The BCL cases, however, have comparable TEs compared to the other base case scenarios because they export electricity, and

Table 5.1 - Energy balances for Methanol Production from MSW Simulations

	BCL[c]	MTCI	Thermoselect
Gasifier Temperature (K)	1077.0	1060.8	1473.0
Gasifier Pressure (bars)	1.0	1.5	1.0
Feed throughput (dry tonnes RDF/day)	872.8	872.8	n.a.
Feed throughput (dry tonnes MSW/day)	1051.5	1051.5	1050.0
Gasifier RDF Input (MW)	182.8	176.6	n.a.
Process MSW Input (MW)	190.5	183.9	182.3
MeOH output (MW)	93.0	102.7	117.4
MeOH output (tonnes/day)	354.3	391.2	447.4
Total Electricity Inputs (MW)	16.33	19.78	27.30
Waste Heat Suitable for Raising Steam (MW)	63.13	39.59	35.42
Electricity from Waste Heat (MW)[a]	17.48	10.96	9.81
Net Electricity Needs (MW)[b]	-1.15	8.82	17.49
Cooling needs (MW)	7.10	4.70	17.23
Heating needs (MW)	0	0	0
Energy Ratio	0.488	0.558	0.644
Thermal Efficiency	0.497	0.492	0.514

Notes: n.a. = not applicable

a) The electricity is generated in a rankine cycle with 27.7% efficiency

b) The electricity is generated by a MSW/RDF IGCC

c) The BCL system is a net exporter of electricity, so has negative electricity needs thus have negative electricity requirements. The TEs are higher than the ERs in these cases.

The BCL configurations have large amounts of high quality process heat that come from cooling the two flue gas streams (from the reformer and gasifier combustors), the reformer output stream, and the gasifier product gas. The pinch analysis shows that for both methanol and hydrogen production the BCL cases can raise the most steam. The electricity generated from this steam is more than enough to power the RDF production plant, compressors, and pumps.

The BCL cases require the least electricity inputs of the base cases. Without the need of an oxygen plant, the BCL cases have far less electricity requirements than the Thermoselect cases (Table 5.3). In comparison to MTCI, the BCL cases have

Table 5.2 - Energy balances for Hydrogen Production from MSW Simulations

	BCL[c]	MTCI	Thermoselect
Temperature (K)	1077.0	1060.8	1473.0
Pressure (bars)	1.0	1.5	1.0
Feed throughput (dry tonnes RDF/day)	872.8	872.8	n.a.
Feed throughput (dry tonnes MSW/day)	1051.5	1051.5	1050.0
Gasifier RDF Input (MW)	182.8	176.6	n.a.
Process MSW Input (MW)	190.5	183.9	182.3
Hydrogen output (MW)	110.9	118.6	136.9
Total Electricity Inputs (MW)	12.95	16.15	26.20
Waste Heat Suitable for Raising Steam (MW)	53.59	30.29	25.25
Electricity from Waste Heat (MW)[a]	14.84	8.39	6.99
Net Electricity Needs (MW)[b]	-1.88	7.76	19.21
Cooling needs (MW)	0.98	7.72	21.39
Heating needs (MW)	0.00	0.00	0.00
Energy Ratio	0.582	0.645	0.751
Thermal Efficiency	0.599	0.577	0.588

Notes:

n.a. = not applicable

a) The electricity is generated in a rankine cycle with 27.7% efficiency

b) The electricity is generated by a MSW IGCC

c) The BCL system is a net exporter of electricity, so has negative electricity needs lower reactor-feed compressor duties because the BCL gasifier requires fewer steam inputs, which reduces the product gas volume.

The abundance of excess heat and low electricity requirements allow the BCL cases to export electricity. For both methanol and hydrogen production, slightly more

Table 5.3 - Energy Inputs for Producing Methanol and Hydrogen from MSW

	Methanol			Hydrogen		
	BCL	**MTCI**	**ThS**	**BCL**	**MTCI**	**ThS**
Product output (GJ/hr)	334.68	369.6	422.65	399.32	427.1	492.78
Electricity requirements (kWh/GJ product)						
Pumps	0.00	0.01	0.03	0.02	0.01	0.05
Compressors	43.97	49.15	37.59	28.39	34.03	29.98
RDF processing[a]	4.81	4.36	6.47	4.03	3.77	5.55
Oxygen Production[b]	0.00	0.00	20.50	0.00	0.00	17.58
Total (kWh/GJ product)	48.78	53.52	64.59	32.44	37.81	53.16
Fraction of electricity requirements						
Pumps	0.00	0.00	0.00	0.00	0.00	0.00
Compressors	0.90	0.92	0.58	0.88	0.90	0.56
RDF processing	0.10	0.08	0.10	0.12	0.10	0.10
Oxygen Production	0.00	0.00	0.32	0.00	0.00	0.33
Fraction provided by						
Waste heat[c]	1.07	0.55	0.36	1.15	0.52	0.27
External sources	-0.07	0.45	0.64	-0.15	0.48	0.73
Reaction Steam (kg/kg dry feed)	1.00	0.31	0.35	1.00	0.42	0.76
Energy Ratio	0.49	0.56	0.64	0.58	0.65	0.75
Thermal Efficiency	0.50	0.49	0.51	0.60	0.58	0.59

Notes:

a) RDF production is assumed to need 36.7 kWh/dry tonne MSW processed (see Table 3.4). The Thermoselect pre-processing coke-oven requires 62.5 kWh/dry tonne (Riegel and Runyon, 1993).

b) Production of 99.5% pure oxygen is assumed to require 480 kWh/tonne O_2 (Williams et al., 1994).

c) Electricity is assumed to be provided by a steam rankine cycle with 27.7% thermal efficiency.

than one megawatt of power is exported for a 1050 dry tonnes MSW/day facility, which is less than 1% of the input energy.

BCL Configuration Notes

Due to the BCL gasifier's high steam:feed ratios, its product gas has the highest water content of the three gasifiers at 61.2%. With such a high steam fraction, the process requires no extra steam for methanol production and very few additional steam inputs for hydrogen production to prevent carbon deposition in the reactors.

The BCL product gas is rich in hydrocarbons (17% methane and 11% ethylene), signaling the need for a reformer to exploit the extra chemical energy. Without a reformer, the system would be much less efficient. In the case of methanol production, a fraction of the cleaned product gas is used to help fuel the reformer because the purge gas is inadequate to provide all the heat needed. For hydrogen production, the reformer temperature is lowered to 820 °C, so that the purge gases are adequate to provide the entire energy duty.[1] At this temperature, only 57% of the methane is reformed. Still, the hydrogen mole flow increases sixfold because all of the C_2+ compounds are reformed. The reformer temperature in the MeOH case is maintained at 867 °C, the practical maximum (see Chapter 4).

For MeOH production, no shift reactors are used because the reformer product gas is hydrogen-rich. Thus, the MeOH synthesis feed is CO limited, and the optimal R value of 2.01-2.1 mentioned in Chapter 4 is not achieved. 23% of the hydrogen that leaves the reformer is not used by the MeOH synthesis reactor and ends up in the purge gas.

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[1] The reformer temperature is varied in each system to maximize the fuel output.

Comparison to Biomass-fed Systems Using the BCL Gasifier

Table 5.4 summarizes the thermodynamic performances of the biomass and MSW cases. The ERs for the biomass cases are around 25% greater than the MSW cases. The TEs for the biomass and MSW cases are closer (the biomass cases are 6% and 15% greater for MeOH and H_2 production, respectively) because the MSW cases export electricity while the biomass cases need external sources of power.

A key difference between the two processes is the pre-processing step. RDF production discards a small fraction of the MSW organic material, losing 4% of the input energy. The biomass-fed cases do not lose any of the original feed, so the ERs for the MSW cases are 4% lower than the biomass cases before the feed enters the gasifier.

Table 5.4 - Fuel Production Using the BCL Gasifier: Biomass vs. MSW

	Biomass[a]		MSW	
Gasifier Characteristics				
Temperature (K)	1136		1077	
Composition				
H_2O	19.9		61.2	
H_2	16.7		7.2	
CO	37.1		16.2	
CO_2	8.9		4.2	
CH_4	12.6		6.5	
C_{2+}	4.8		4.3	
Cold gas Efficiency (%)	80.1		79.1	
Fuel Production Energy Balances	**MeOH**	**H_2**	**MeOH**	**H_2**
Total Electricity Inputs (kWh/GJ product)	29.74	31.79	48.78	32.44
Fraction of Electricity Input From:				
Waste heat	0.696	0.317	1.07	1.15
External sources	0.304	0.683	-0.07	-0.15
Energy Ratio	0.606	0.732	0.488	0.582
Thermal Efficiency	0.576	0.636	0.497	0.599

Notes: a) biomass values taken from Williams et al. (1994)

The differences in the product gases for the two feeds explain some of the differences in system performance. The gasifier product gas in the MSW case has three times as much steam as the biomass case. This is due to high steam:feed ratios that were used in the MSW case but not in the biomass case.[2] Also, the RDF feed has a 22.5% moisture content before it enters the gasifier, while the biomass is dried to 10% moisture. The RDF is not dried because it becomes too difficult to handle when it is dried.

More char must be burned in the MSW case (the carbon conversion for biomass is 19% greater than for MSW) to heat up the enormous amounts of steam within the gasifier. Though the cold gas efficiencies are similar for biomass and MSW feeds, the latent heat of water constitutes 13% of the product gas higher heating value in the MSW case, compared to 3% for biomass. On a dry gas basis, the biomass-fed gasifier cold gas efficiency is 14% greater than the RDF-fed case. A benefit of the large steam fraction in the product gas of the MSW cases is that the system heating load is low, since few additional steam inputs are required.

In both the MSW and biomass cases, the reforming step is crucial because product gases from the two feeds have large fractions of methane and higher hydrocarbons. Due to the high plastic content of MSW, the MSW derived product gas has far more higher hydrocarbons (on a dry basis), making the reformer duty greater. The heating requirements for the reformer in the biomass cases is 10% of the energy in the feed, whereas in the MSW cases it is 12-14% of the feed energy. The greater reformer duties and the lower gasifier efficiencies of the MSW cases basically account for the lower ERs in comparison to the biomass cases.

The great amount of excess heat in the MSW cases makes the TEs closer together for the two feeds. The MSW cases have more waste heat available for raising steam because there is no separate drying step, which requires 12% of the feed energy in the biomass-fed systems.

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[2] As mentioned in Chapter 3, BCL recently improved its fluidization process to reduce the amount of steam required to fluidize the gasifier bed. No recent data for RDF feeds using reduced steam inputs are available. Though the gasifier performance would improve with reduced steam inputs, the fuel production process thermal efficiency would not necessarily increase because steam is required for downstream processes (e.g., the reformer and shift reactors)

For MSW, the feed is essentially dried in the gasifier, which lowers the ER of the process. But by not having a separate energy intensive drying step, more excess steam can be raised to generate more power. Thus, the TE is insensitive to drying the feed in a separate step or within the gasifier.

For hydrogen production, the small difference in the TEs of the biomass and MSW cases can be attributed to the feed losses[3] and electricity requirements in the feed preparation step for MSW. The 15% difference in the TEs of the MeOH systems is due to the MeOH synthesis feed gas in the MSW case being hydrogen rich. As mentioned earlier, the MeOH synthesis reactor does not use much of the hydrogen in the reformer product gas. The MeOH synthesis feed has a high H_2:CO ratio because the MSW-derived product gas has a large fraction of hydrocarbons. The steam reforming reactions (4-1, 4-2) show that more hydrocarbons in the reformer feed gas increases the H_2:CO ratio in the reformer product. For biomass, there are fewer hydrocarbons in the product gas; hence, the reformer product gas is CO rich. With a shift reactor, CO is converted into H_2 to reach the optimal feed ratios for MeOH synthesis.

5.3.2 Methanol and Hydrogen Production from the MTCI Gasifier

The energy ratios for the MTCI case are fairly high, at 0.56 and 0.65 for methanol and hydrogen, respectively. Unlike the BCL cases, the MTCI cases require net electricity inputs, reducing the thermal efficiencies to 0.49 and 0.58 (the lowest of all the base case scenarios).

Like the BCL cases, the MTCI case has a large amount of high quality heat. But recall from Chapter 3 that the MTCI gasifier requires the most steam of the three gasifiers, around three times more than the BCL case. The massive heating load

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[3] Feed losses refers to the loss of a fraction of organic material in MSW when producing RDF (4% of the energy in the MSW feed).

needed to raise the gasifier steam (32 MW) reduces the amount of waste heat available to generate power. As can be seen in Table 5.3, the MTCI cases have similar electricity requirements to the BCL cases but generate very little electricity from waste heat.

The base case scenarios shown in tables 5.1 and 5.2 assume that the unconverted char can be washed and used as fuel for the pulse combustor (Durai-Swamy, 1994a). Since the char provides only 9% of the necessary energy for the gasifier, purge gases from the PSA or the methanol synthesis loop are the primary pulse combustor fuels.

Table 5.5 - Comparison of the MTCI Base Case Scenario to a Case with No Recycling of the Char to the Pulse Combustor

	Methanol		Hydrogen	
	Base Case	No Char	Base Case	No Char
Product output				
(kmol/hr)	508.8	489.1	1494	1419
(MW)	102.7	98.7	118.6	112.7
Gasifier Input (MW)	176.6	176.6	176.6	176.6
Energy Ratio	0.558	0.537	0.645	0.613
Total electricity requirements (MW)	19.78	19.67	16.15	16.01
Electricity from waste heat (MW)[a]	10.96	10.67	8.39	8.49
Net Electricity Needs (MW)	8.82	9.00	7.76	7.51
Cooling needs (MW)	4.70	4.47	7.72	7.34
Heating needs (MW)	0.00	0.00	0.00	0.00
Thermal Efficiency	0.492	0.472	0.577	0.550

Notes: a) The electricity is generated in a rankine cycle with 27.7% efficiency

An interesting modification to the base case is to treat the char as residue and recycle only the clean purge gases to the pulse combustor. By not burning the char, the process is more attractive from an air emissions standpoint. As Table 5.5 shows, the ERs for the cases with no char recycling are only 4-5% lower than the MTCI base cases, which are still higher than the BCL system ERs. The TEs for the no char cases are also 4-5% lower than the base cases, since the amount of waste heat available does not vary between the cases. Though these TEs are lower than the TEs of any base case scenario, the environmental benefits of not burning the char may be more important than the efficiency losses. These cases with no char recycling will be discussed further in Chapter 6.

MTCI Configuration Notes

For RDF feeds, the product gas from the MTCI gasifier has significant amounts of methane and hydrocarbons (8.2% methane and 5.7% higher hydrocarbons on a dry basis). These account for half of the dry gas heating value. Reformers are therefore used in the process configurations to use the energy of the hydrocarbons. If reformers were not used in the configurations, much less fuel would be produced.

For methanol production it is assumed that the reformer operates at 786°C because the purge gases are not adequate to maintain the reformer temperature at 870°C. Though only 26% of the methane is converted, the carbon monoxide in the reformer output gas is nearly double that in the input gas because all the higher hydrocarbons are reformed. No shift reactors are necessary, since the gas leaving the reformer is hydrogen rich, as is the case for the BCL system. 36% of the hydrogen entering the methanol synthesis loop is not converted and ends up in the purge gas.

For hydrogen production the reformer operates at the lowest feasible limit of 765°C. At this temperature none of the methane is converted, but all the higher hydrocarbons are reformed, which doubles the amount of hydrogen in the reformer output relative to the input. Even at this low reformer temperature, there is not enough purge gas to fuel the two combustors. In order to satisfy both conditions, the PSA is modified in this case to recover only 77% of the hydrogen, instead of 89.5% as specified in Chapter 4.

Without reformers, the systems are less efficient. As shown in table 5.6, using a reformer plus an inefficient PSA still recovers almost 50% more hydrogen than a configuration that does not include a reformer. The base case methanol production scheme produces nearly 60% more methanol than a variation with no reformer. For the configurations that do not include reformers, it is assumed that the methane and higher hydrocarbons act as inert gases through the gas processing reactors and become components of the purge gases. The purge gases have more than enough energy to supply the pulse combustor, and the extra gases are burned to produce electricity.[4] Since a greater fraction of the purge gases are used to generate

Table 5.6 - Comparison of the MTCI System with and without a Reformer

	Methanol		**Hydrogen**	
	Reformer	No Reformer	Reformer	No Reformer
Product output				
(kmol/hr)	508.8	319.8	1494	1025
(MW)	102.7	64.5	118.6	81.4
Gasifier Input (MW)	176.6	176.6	176.6	176.6
Energy Ratio	0.558	0.351	0.645	0.443
Total electricity requirements (MW)	19.78	17.21	16.15	8.56
Electricity from waste heat (MW)[a]	10.96	7.39	8.39	2.36
Electricity from purge gases (MW)[b]	0.00	20.73	0.00	20.30
Net Electricity Needs (MW)[c]	8.82	-10.91	7.76	-14.11
Cooling needs (MW)	4.70	2.60	7.72	23.25
Heating needs (MW)	0.00	0.00	0.00	0.00
Thermal Efficiency	0.492	0.422	0.577	0.567

Notes:
a) The electricity is generated in a rankine cycle with 27.7% efficiency
b) The electricity is generated by a MSW IGCC
c) The cases without reformers are net exporters of electricity so have negative electricity needs

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[4] It is assumed that the excess gas is burned in a gas turbine/steam turbine combined cycle with the HHV efficiency based on Katofsky's (1993) linear regression:
Efficiency (%) = 39.58 + 0.134 * Output(MW).

electricity than heat the gasifier, these systems without reformers are net exporters of electricity. Unlike the BCL base cases which exported only one MW_e, these systems export between 10-15 MW_e (around 17% of the energy in the product output); hence, the TEs are significantly higher than the ERs. Nevertheless, these configurations have lower TEs than the respective base cases. Since the cases without reformers produce far less fuel and are less efficient than the base cases, they are not considered here for fuel production.

Comparison to Biomass-fed Systems Using the MTCI Gasifier

Like the MSW cases, the biomass cases have moderately high ERs and low TEs (table 5.7). The differences in the process performances for the two feeds can basically be attributed to the same factors discussed for the BCL systems (e.g. - RDF production losses, lower gasifier efficiencies, reformer duty, etc.).

Table 5.7 - Fuel Production Using the MTCI Gasifier: Biomass vs. MSW

	Biomass[a]		MSW	
Gasifier Characteristics				
Temperature (K)	970		1061	
Composition				
H_2O	49.5		52.4	
H_2	25.3		22.3	
CO	11.2		6.8	
CO_2	9.9		11.8	
CH_4	4.0		3.9	
C_{2+}	0.2		2.7	
Cold gas Efficiency (%)	90.0		84.8	
Fuel Production Energy Balances	**MeOH**	**H_2**	**MeOH**	**H_2**
Total Electricity Inputs (kWh/GJ product)	35.39	35.45	53.52	37.81
Fraction of Electricity Input From:				
Waste heat	0.61	0.03	0.55	0.52
External sources	0.39	0.97	0.45	0.48
Energy Ratio	0.62	0.76	0.56	0.65
Thermal Efficiency	0.57	0.61	0.49	0.58

Notes: a) biomass values taken from Williams et al. (1994)

The most prominent difference in the process configurations is that the MSW cases have reformers whereas the biomass cases do not. Like the BCL cases, the significant difference in the gasifier product gases is the hydrocarbon concentration. In Katofsky's analysis of biomass feeds, it is assumed that the 4% methane and 0.2% higher hydrocarbons in the product gas end up in the purge gas that would be burned in the pulse combustor.[5] In the MSW case, the product gas has much more methane and higher hydrocarbons; thus, a reformer is required to increase the fuel output and improve the system efficiency, as discussed above.

5.3.3 Methanol and Hydrogen Production from the Thermoselect Gasifier[6]

The Thermoselect cases have the highest energy ratios by far: for methanol, 0.64 and for hydrogen, 0.75. Due to large electricity requirements the thermal efficiencies are much lower at 0.51 and 0.59, comparable to the BCL and MTCI cases.

The high ERs in the Thermoselect cases can be attributed to the absence of reformers, which have tremendous heat duties in the BCL and MTCI cases (21-27 MW). No reformers are needed because the product gas from the high temperature Thermoselect gasifier has only trace concentrations of methane and other hydrocarbons. The gasifier's higher carbon conversion percentage in comparison to the indirectly heated gasifiers also contributes to the high ERs.

The Thermoselect process has the least waste heat available to generate electricity. Though the gasifier product gas generates some heat as it is cooled, there are no combustion products from a reformer or gasifier to provide as much high quality heat as the BCL and MTCI processes.

Besides generating the least amount of electricity, the Thermoselect systems also consume the most electricity. Oxygen production constitutes 30% of the total gross electricity needs (Table 5.3). The energy demand required for pre-processing the feed is also a major contributing factor. The combination of consuming the most

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[5] Unlike the MSW to MeOH configuration, the purge gases in the biomass system fuel only the pulse combustor.

[6] Since Thermoselect has not been tested with biomass, no comparison to biomass-fed systems using the Thermoselect gasifier is made.

power and producing the least makes the Thermoselect process have the largest net electricity requirements.

Thermoselect Configuration Notes

The most notable detail in the Thermoselect process configuration is that no reformers are used. As mentioned before, the product gas does not have to be reformed because it has almost no methane.

Another key difference in the Thermoselect system (discussed in Chapter 3) is that the gasifier uses MSW as the feed, unlike the two indirectly heated gasifiers. Though the system pre-processes the feed in a coking oven, none of the feed heating value is lost before it enters the gasifier (4% of the feed energy is lost in RDF production). The drawbacks to the coking oven are that it requires twice as much electricity as an RDF plant and over one GJ/dry tonne MSW of process heat (7% of the feed energy). The purge gases from the PSA or methanol synthesis loop are recycled to the coke oven and burned to provide the heat to dry and pyrolyze the MSW. For methanol production, the purge gas cannot adequately heat the coking oven. In this case, some of the cleaned product gas is combined with the purge gas and burned to supply enough heat to the coking oven.

In the lower section of the gasifier, the temperature is kept at 2000 °C to melt down the inorganic material in the MSW into slag. To prevent temperature fluctuations in this region, a fraction of the clean product gas is recycled to the gasifier and combusted with oxygen.

5.4 General Remarks Concerning Fuel Production Thermodynamics

5.4.1 Characteristics of Methanol and Hydrogen Production

The thermal efficiencies and energy ratios for hydrogen production are all higher than for methanol production. This can be understood through the comparison of the methanol synthesis reaction to the equivalent reaction for hydrogen production, the shift reaction (Katofsky, 1993). Starting with the same composition of two H_2 molecules reacting with H_2O and CO, the reactions are

$$CO + 2H_2 + H_2O \rightarrow 3H_2 + CO_2 \qquad \Delta H^{\circ}_{298} = -41.2 \text{ MJ/kmol} \qquad (5\text{-}4)$$

$$CO + 2H_2 + H_2O \rightarrow CH_3OH + H_2O \qquad \Delta H^{\circ}_{298} = -210 \text{ MJ/kmol} \qquad (5\text{-}5)\ .$$

Methanol synthesis is much more exothermic, so less of the input enthalpy shows up in methanol as compared to hydrogen. The ratio of higher heating values for three hydrogen molecules to one methanol molecule is

$$\frac{3 * 285.8}{726.4} = 1.18$$

For a given gasifier, the energy ratio for hydrogen production is indeed nearly a factor of 1.18 greater than the ER for methanol production.[7] Since methanol synthesis releases more energy than the equivalent shift and PSA steps, methanol production should have more waste heat available. Inspecting the pinch analyses reveals that with methanol production more steam can be raised for electricity production. Thus, the differences between the thermal efficiencies and the energy ratios are smaller for methanol production than hydrogen production.

5.4.2 Discussion on Process Performances

The Thermoselect systems exhibit the highest ERs of the systems, while the BCL systems have the lowest. The ERs basically follow the same trends as the gasifier cold gas efficiencies and carbon conversions. As discussed in Chapter 3, the Thermoselect gasifier has much higher carbon conversions than indirectly heated gasifiers because it operates at much higher temperatures. The use of reformers in

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[7] Katofsky (1993) found that the same is true for natural gas, coal, and biomass feeds.

the BCL and MTCI systems is the other major reason for the relatively low ERs in comparison to Thermoselect's.

The Thermoselect gasifier exhibits the highest thermal efficiencies (TEs) for methanol production, and the BCL gasifier has the highest for hydrogen. But surprisingly, all three of the systems for a given fuel output share similar TEs. For methanol production, the ratio of TE's for the most to least efficient process is only 1.04. The ratio is also around 1.04 for hydrogen production.[8] The TEs are remarkably similar because the definition of the TE takes into account all the external heat and work requirements which balance out the high ERs of some processes. For example, the high temperature Thermoselect gasifier can produce more fuel than the other two gasifiers for a given amount of feed, but it needs more energy inputs to do so, which lowers the system TE.

The most significant determinants of the process performances are the input data to the simulations, i.e - the product gas composition and gasifier operating characteristics supplied by each gasifier manufacturer. As mentioned in Chapter 4, losses from the gasifier dwarf all other process losses; thus, changes in gasifier efficiency have major impacts on the process performance, whereas improvements in other reactors have only marginal effects. This underscores the importance of obtaining reliable input data for the models. In this fuel production analysis the data from the three gasifiers yield similar TEs.[9] This does not mean that the type of gasifier does not have any effect on the fuel production TE. Recall from chapter 3 that many types of gasifiers were not even considered for this fuel production analysis. Air-blown gasifiers would have much higher compression duties, which would dramatically lower the TEs. Fixed-bed gasifiers produce a gas that has many tars and oils, which would be scrubbed out in the quench step of a fuel production system.

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[8] For comparison, the ER's for the Thermoselect process are 30% greater than those of the BCL cases.

[9] In comparing high temperature and indirectly heated gasifiers for fuels production from biomass, the TEs are also very similar, differing by only 5-7% (Williams et al., 1994).

Such a system would therefore have a low ER and TE in comparison to the systems analyzed here.

Using what appear to be today's best gasification technologies for fuels production, the TE of producing hydrogen from MSW is about 0.6, which is expected because it compares closely with the highest TE for biomass feeds analyzed by Williams et al. (0.636). The slightly lower TE for MSW feeds in comparison to biomass feeds can be mainly attributed to pre-processing the MSW feed (RDF production; coking oven).

For methanol production from MSW using the gasifiers analyzed, the TE is around 0.5. As discussed in section 5.3, this could actually be improved if the hydrocarbon content were reduced in the product gases from the indirectly heated gasifiers. Still, the TE is close to the highest biomass system TE of 0.576.[10]

Though the TEs for fuel production from MSW given above represent the process performances using the best modern technologies, improvements are certainly possible since the MSW (RDF) gasifier is a nascent technology. If the oxygen plant duties decreased significantly, for example, the TE for the Thermoselect systems would rise accordingly. Similarly, if the BCL or MTCI systems could increase the carbon conversions of their processes by improving their respective heat transfer mechanisms, the fuel production TE would rise.

5.4.3 Comparison of Performances Using Other Solid Feedstocks

Fuel production from MSW is less efficient than from biomass or coal feedstocks, but the differences in the TEs are small. The lower TE in the MSW case is mainly a consequence of the feed preparation energy requirements and feed losses.

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[10] The TE for MSW feeds would be around 0.54, if the only difference between biomass and MSW fed systems was the feed preparation step.

The thermal efficiency calculated for methanol production from MSW is about 50%. Estimates for coal to methanol plant efficiencies range from 51-61% (Probstein, 1982; Wyman, 1993; Williams et al., 1994), and biomass to methanol spans 54-58% (Williams et al, 1994 and Katofsky, 1993).

For hydrogen production, the MSW-fed systems also exhibit similar efficiencies to those for coal and biomass systems. The hydrogen from MSW systems yield a TE of nearly 60%. Using coal and biomass feeds, the TEs range from 56-64% (Williams et al, 1994). With MSW as feed for producing fuel, nearly the same thermal efficiencies can be achieved as with other solid feedstocks.

5.5 Conclusions

The thermodynamic performances of methanol and hydrogen production from MSW using the BCL, MTCI, and Thermoselect gasifiers have been calculated and presented in this chapter. The Thermoselect systems have the highest ERs, while systems using the BCL gasifier have the lowest. But all three of the systems studied share similar TEs: around 50% for methanol production and 60% for hydrogen production.

With MSW as feed nearly the same thermal efficiency can be achieved in producing methanol and hydrogen as with biomass or coal feedstocks. The primary efficiency differences between using MSW and biomass for fuel production come from the feed preparation step. For methanol production, the higher concentration of hydrocarbons in the BCL and MTCI product gases also contributes to lower process TEs. Potential improvements in the process TEs mainly depend on future advances in gasification technologies.

The calculated efficiencies were used in Chapter 2 as the basis for assessing the fuel resource base of MSW. The results of this thermodynamic analysis will also be used in Chapter 7 to parameterize the economic analysis of methanol and hydrogen production from MSW.

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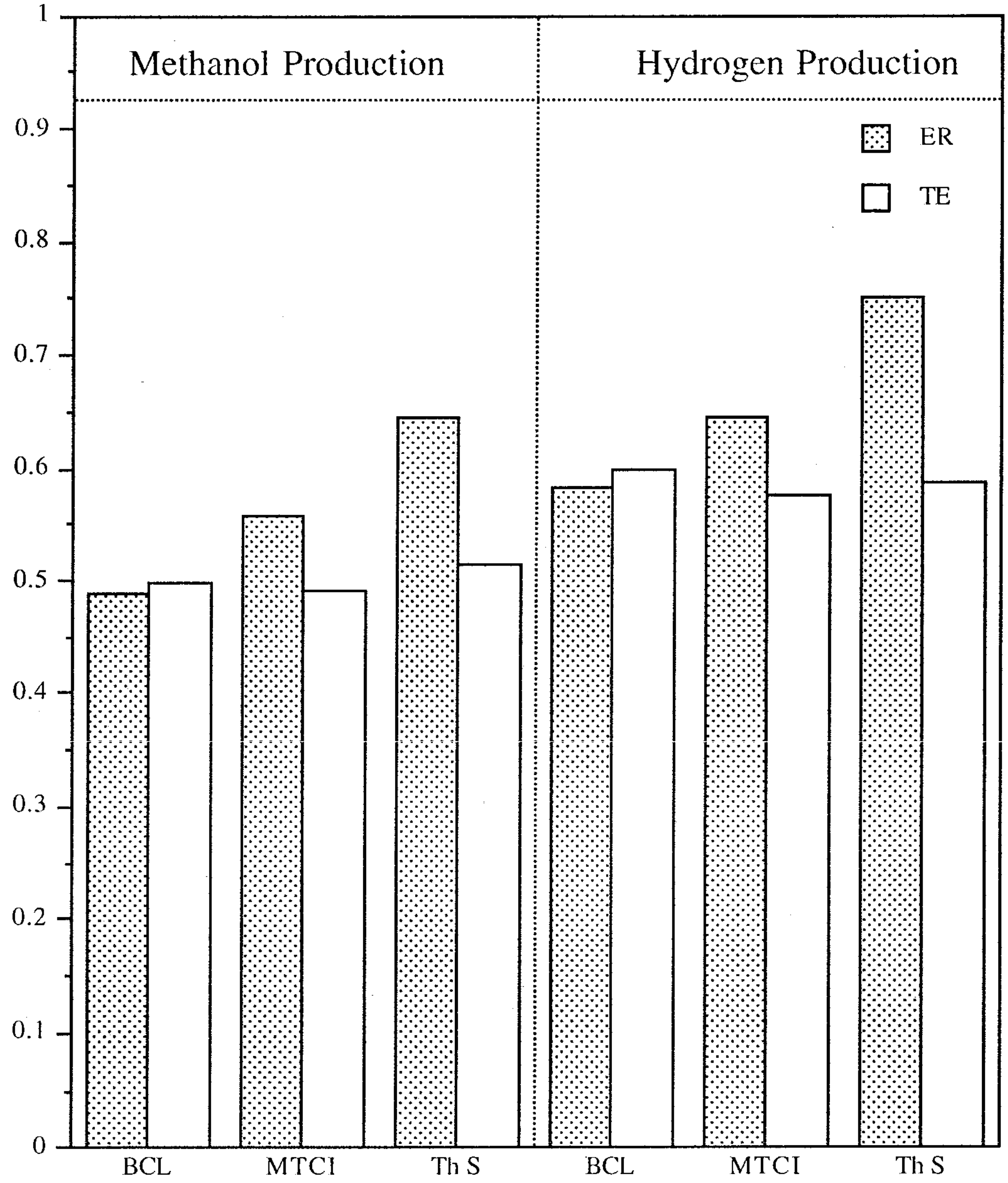


Figure 5.1 - Performance Characteristics of Different Systems for Fuel Production from MSW

Appendix 5A

Detailed Process Diagrams, Material Flows, and Heat Integration Analyses for Fuel Production Simulations

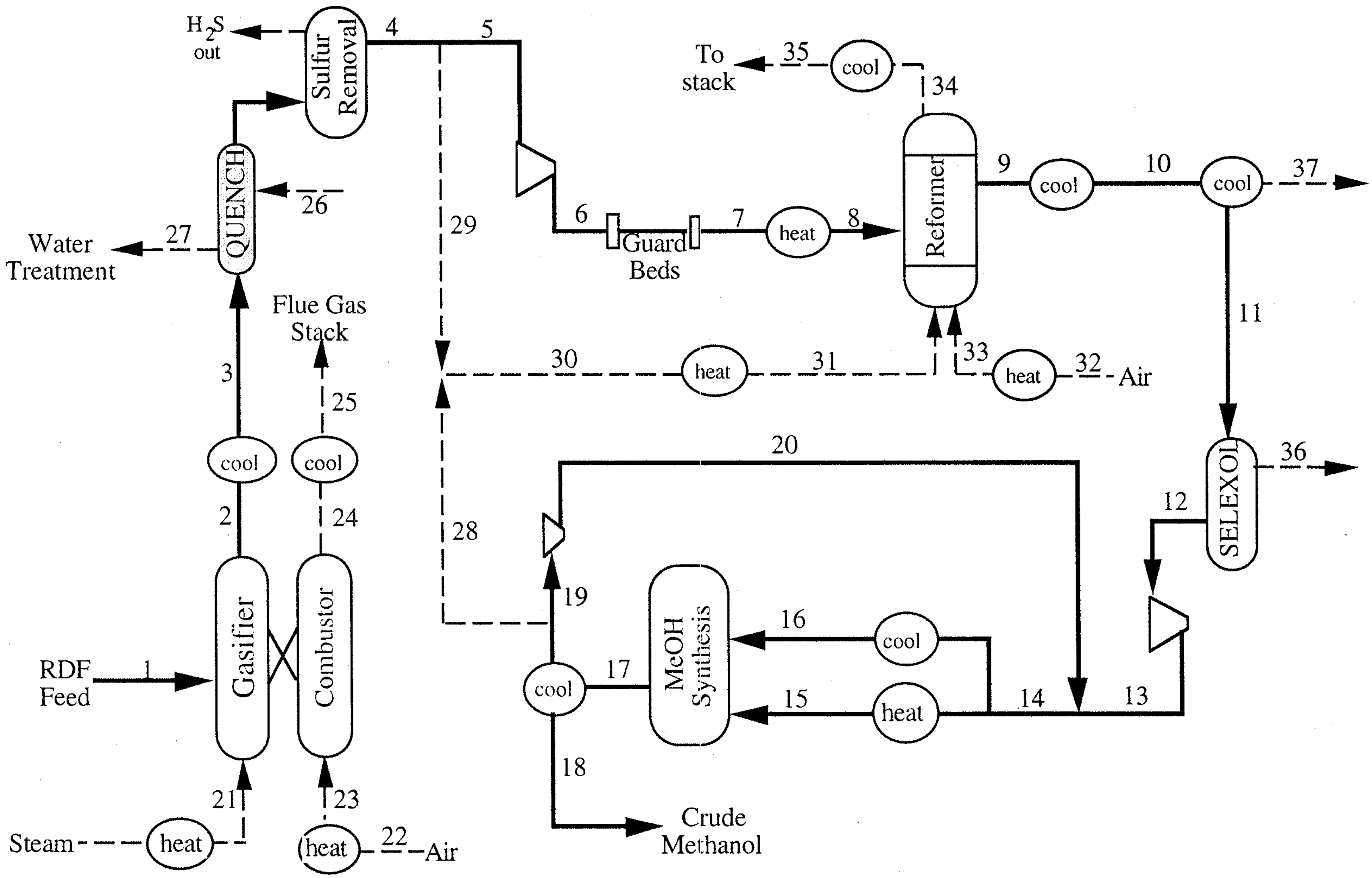


Figure 5A.1 - Flow Diagram for Methanol Production from the BCL Gasifier

Table 5A.1 - Mass flows for Methanol Production from RDF using the BCL gasifier

				Molar Flowrates	(kmol/hr)									
Stream	T (K)	P (bars)	kg/h	Total	H_2	CO	CO_2	CH_4	C_2H_4	C_2H_6	MeOH	H_2O	N_2	O_2
1 RDF feed			872.75 dry	tonnes/day										
2 gasifier product gas	1077	1.0	48856	2452	177	397	103	159	101	5	0	1501	10	0
3 cooled gasifier product gas	673	1.0	48856	2452	177	397	103	159	101	5	0	1501	10	0
4 cleaned gas	366	1.0	58383	2981	177	397	103	159	101	5	0	2030	10	0
5 feed to reformer compressor	366	1.0	55958	2857	169	381	99	153	96	5	0	1945	9	0
6 compressed gas	576	15.5	55958	2857	169	381	99	153	96	5	0	1945	9	0
7 cool reformer feed gas	700	15.4	55958	2761	73	381	99	153	0	101	0	1945	9	0
8 hot reformer feed gas	850	14.9	55958	2761	73	381	99	153	0	101	0	1945	9	0
9 reformer product gas	1140	14.4	55958	3431	1261	432	383	20	0	0	0	1326	9	0
10 cool reformer product gas	623	13.9	55958	3431	1261	432	383	20	0	0	0	1326	9	0
11 selexol feed	400	13.4	40276	2561	1261	432	383	20	0	0	0	456	9	0
12 selexol output	400	13.4	16766	1757	1261	432	35	20	0	0	0	0	9	0
13 makeup	501	105.9	16766	1757	1261	432	35	20	0	0	0	0	9	0
14 synfeed	348	105.9	44616	8786	7600	508	56	408	0	0	15	0	199	0
15 hotfeed	523	105.4	9658	1902	1645	110	12	88	0	0	3	0	43	0
16 coldfeed	320	105.4	34958	6884	5955	398	44	320	0	0	12	0	156	0
17 MeOH synthesis product	533	97.3	44616	7863	6644	79	23	408	0	0	476	34	199	0
18 crude MeOH product	300	96.8	15433	498	1	0	1	1	0	0	461	33	0	0
19 recycled synprod	300	96.8	27850	7029	6339	76	21	388	0	0	15	0	190	0
20 compressed recycled synprod	309	105.9	27850	7029	6339	76	21	388	0	0	15	0	190	0
21 gasifier steam	430	1.3	11335	629	0	0	0	0	0	0	629	0	0	0
22 cool gasifier air	298	1.0	79883	2769	0	0	0	0	0	0	0	0	2187	581
23 hot gasifier air	672	1.0	79883	2769	0	0	0	0	0	0	0	0	2187	581
24 gasifier hot flue	1244	1.0	85863	2795	0	0	495	0	0	0	0	56	2191	53
25 gasifier cool flue	393	1.0	85863	2795	0	0	495	0	0	0	0	56	2191	53
26 quench water	293	1.5	27023	1500	0	0	0	0	0	0	0	1500	0	0
27 condensate	366	1.0	17496	971	0	0	0	0	0	0	0	971	0	0
28 purge stream	300	96.8	1333	337	303	4	1	19	0	0	1	0	9	0
29 fuel split from product gas	366	1.0	2425	124	7	17	4	7	4	0	0	84	0	0
30 cool reformer comb. feed	335	1.0	3758	460	311	20	5	25	4	0	1	84	10	0
31 hot refomer comb. feed	850	1.0	3758	460	311	20	5	25	4	0	1	84	10	0
32 cool reformer comb.air	298	1.0	36184	1254	0	0	0	0	0	0	0	0	991	263
33 hot reformer comb. air	1000	1.0	36184	1254	0	0	0	0	0	0	0	0	991	263
34 reformer hot flue	1160	1.0	39942	1549	0	0	59	0	0	0	1	455	1000	34
35 reformer cool flue	393	1.0	39942	1549	0	0	59	0	0	0	1	455	1000	34
36 CO2 out	400	13.4	23510	804	0	0	348	0	0	0	0	456	0	0
37 condensed water	400	13.4	15682	870	0	0	0	0	0	0	0	870	0	0

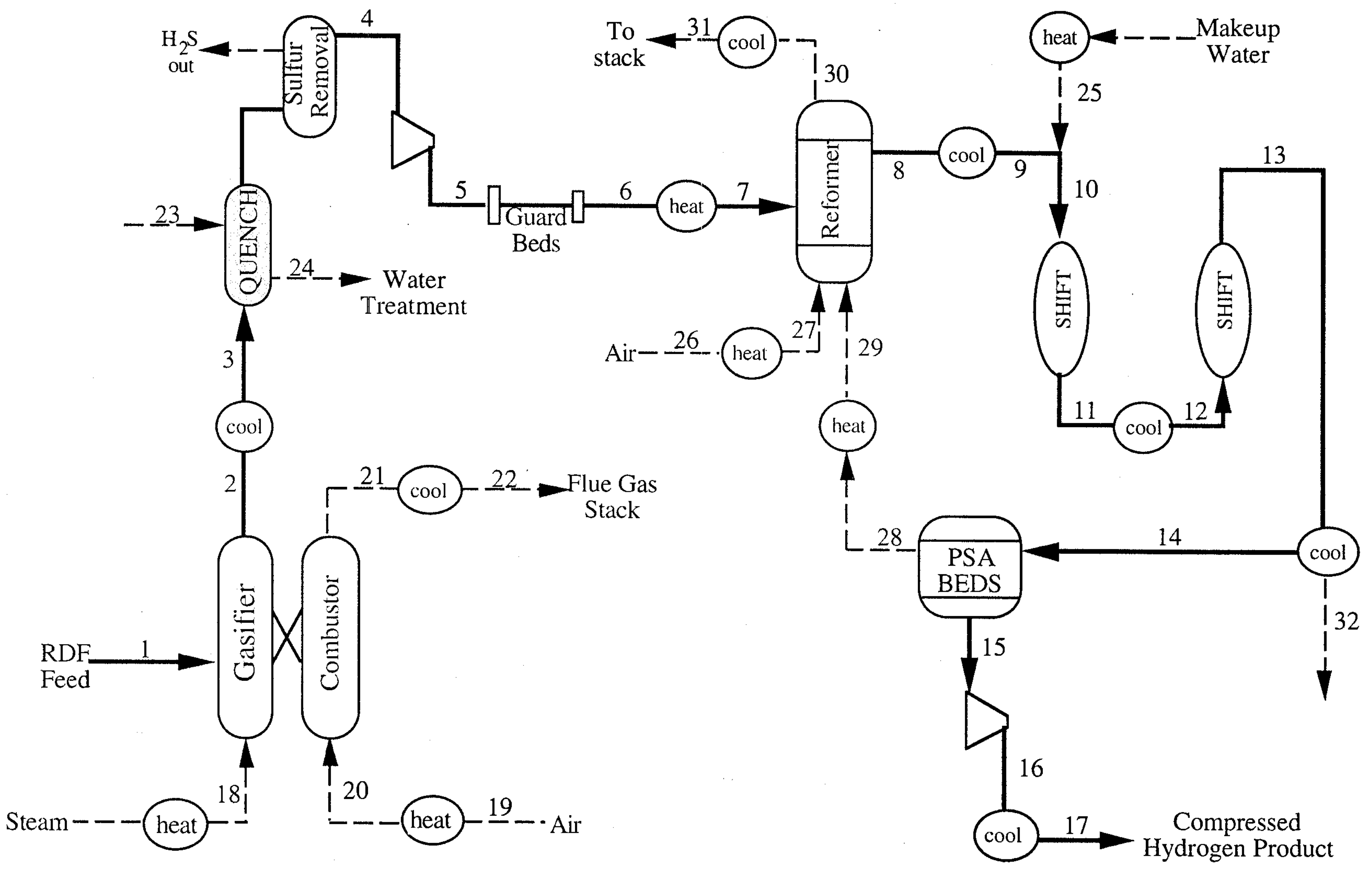


Figure 5A.2 - Flow Diagram for Hydrogen Production from the BCL Gasifier

Table 5A.2 - Mass flows for Hydrogen Production from RDF using the BCL gasifier

				Molar Flowrates (kmol/hr)									
Stream	T (K)	P (bars)	kg/h	Total	H_2	CO	CO_2	CH_4	C_2H_4	C_2H_6	H_2O	N_2	O_2
1 RDF feed			872.75 dry tonnes/day										
2 gasifier product gas	1077	1.0	48856	2452	177	397	103	159	101	5	1501	10	0
3 cooled gasifier product gas	673	1.0	48856	2452	177	397	103	159	101	5	1501	10	0
4 cleaned gas	364	1.0	51235	2584	177	397	103	159	101	5	1633	10	0
5 compressed gas	576	15.5	51235	2584	177	397	103	159	101	5	1633	10	0
6 cool reformer feed gas	717	15.4	51235	2484	76	397	103	159	0	105	1633	10	0
7 hot reformer feed gas	850	14.9	51235	2484	76	397	103	159	0	105	1633	10	0
8 reformer product gas	1092	14.4	51235	3089	1155	422	380	68	0	0	1053	10	0
9 cool reformer product gas	623	13.9	51235	3089	1155	422	380	68	0	0	1053	10	0
10 high temp. shift feed	623	13.9	55081	3302	1155	422	380	68	0	0	1267	10	0
11 hi-temp shift product	711	13.4	55081	3302	1442	135	668	68	0	0	979	10	0
12 low-T shift feed	500	12.9	55081	3302	1442	135	668	68	0	0	979	10	0
13 low-T shift product	500	12.4	55081	3302	1561	16	786	68	0	0	861	10	0
14 cooled PSA feed	313	11.9	39795	2454	1561	16	786	68	0	0	12	10	0
15 hydrogen	313	11.6	2835	1398	1397	0	0	1	0	0	0	0	0
16 compressed hydrogen	408	75.0	2835	1398	1397	0	0	1	0	0	0	0	0
17 hydrogen product	313	75.0	2835	1398	1397	0	0	1	0	0	0	0	0
18 gasifier steam	430	1.3	11335	629	0	0	0	0	0	0	629	0	0
19 gasifier air	298	1.0	79883	2769	0	0	0	0	0	0	0	2187	581
20 hot gasifier air	672	1.0	79883	2769	0	0	0	0	0	0	0	2187	581
21 gasifier hot flue	1244	1.0	85863	2795	0	0	495	0	0	0	56	2191	53
22 gasifier cool flue	393	1.0	85863	2795	0	0	495	0	0	0	56	2191	53
23 quench water	293	1.0	136340	7568	0	0	0	0	0	0	7568	0	0
24 condensate	364	1.0	133970	7436	0	0	0	0	0	0	7436	0	0
25 shift steam	623	13.9	3846	213	0	0	0	0	0	0	213	0	0
26 reformer comb. air	298	1.0	35317	1224	0	0	0	0	0	0	0	967	257
27 hot reformer comb. air	1000	1.0	35317	1224	0	0	0	0	0	0	0	967	257
28 purge stream	313	1.3	36959	1055	164	16	786	67	0	0	12	10	0
29 hot purge	850	1.3	36959	1055	164	16	786	67	0	0	12	10	0
30 reformer hot flue	1112	1.0	72276	2189	0	0	869	0	0	0	310	977	33
31 reformer cool flue	393	1.0	72276	2189	0	0	869	0	0	0	310	977	33
32 condensed water	313	11.9	15287	848	0	0	0	0	0	0	848	0	0

Table 5A.3 - Pinch analysis for methanol production with the BCL gasifier

stream	tsource K	ttarget K	hin GJ/hr	Hout GJ/hr	delta T K	delta H kw	CP kw/K	Tsource corrected	Ttarget corrected
flue (hot)	1244.4	393.0	-116.5	-200.1	-851.4	-23210.4	27.3	1239.4	388.0
comb1 h	1160.0	393.0	-88.3	-129.0	-767.0	-11287.1	14.7	1155.0	388.0
refprod h	1140.0	623.0	-418.7	-484.3	-517.0	-18221.8	35.2	1135.0	618.0
gfrprod h	1077.4	673.0	-379.8	-421.7	-404.4	-11622.2	28.7	1072.4	668.0
shift6a	623.0	426.4	0.0	-23.1	-196.6	-6405.8	32.6	618.0	421.4
synproda h	533.0	388.5	0.0	-21.2	-144.5	-5898.8	40.8	528.0	383.5
shift6b	426.4	400.0	-23.1	-61.9	-26.4	-10794.4	408.6	421.4	395.0
synprodb h	388.5	300.0	-21.2	-77.9	-88.5	-15730.4	177.7	383.5	295.0
cold1 h	347.9	320.0	-76.9	-82.7	-27.9	-1595.0	57.2	342.9	315.0
air1 (cold)	298.0	1000.0	0.0	27.3	702.0	7577.8	10.8	303.0	1005.0
reffeed4 c	699.9	850.0	-531.5	-514.1	150.1	4837.2	32.2	704.9	855.0
fuel2 c	334.9	850.0	-26.2	-18.3	515.1	2200.1	4.3	339.9	855.0
Gair c	298.0	672.0	0.0	31.0	374.0	8612.9	23.0	303.0	677.0
hotfeed1	347.9	523.0	-21.2	-11.2	175.1	2794.8	16.0	352.9	528.0
gfrsteamc	380.4	430.3	29.5	30.7	49.9	319.0	6.4	385.4	435.3
gfrsteamb	380.3	380.4	4.1	29.5	0.1	7044.8	55558.6	385.3	385.4
gfrsteama c	293.0	380.3	0.0	4.1	87.3	1151.3	13.2	298.0	385.3
						-70228.0			

Interval	T(i) K	T(i+1) K	del T K	CPhot kw/K	CPcold kw/K	diff kw/K	delta H kw	Cascade kw	Corrected cascade
0	1239.4	1155.0	84.4	27.3	0.0	27.3	2300.9	2300.9	2300.9
1	1155.0	1135.0	20.0	42.0	0.0	42.0	839.5	3140.4	3140.4
2	1135.0	1072.4	62.6	77.2	0.0	77.2	4831.1	7971.5	7971.5
3	1072.4	1005.0	67.4	106.0	0.0	106.0	7145.9	15117.4	15117.4
4	1005.0	855.0	150.0	106.0	10.8	95.2	14274.7	29392.1	29392.1
5	855.0	704.9	150.1	106.0	47.3	58.7	8805.9	38197.9	38197.9
6	704.9	677.0	27.9	106.0	15.1	90.9	2535.9	40733.8	40733.8
7	677.0	668.0	9.0	106.0	38.1	67.9	610.8	41344.6	41344.6
8	668.0	618.0	50.0	77.2	38.1	39.1	1956.4	43301.0	43301.0
9	618.0	528.0	90.0	74.6	38.1	36.5	3282.1	46583.1	46583.1
10	528.0	435.3	92.7	115.4	54.1	61.3	5687.5	52270.6	52270.6
11	435.3	421.4	13.9	115.4	60.5	54.9	761.1	53031.7	53031.7
12	421.4	395.0	26.4	491.4	60.5	430.9	11385.0	64416.7	64416.7
13	395.0	388.0	7.0	82.8	60.5	22.4	156.5	64573.2	64573.2
14	388.0	385.4	2.6	40.8	60.5	-19.6	-51.0	64522.1	64522.1
15	385.4	385.3	0.1	40.8	55612.7	-55571.9	-7046.5	57475.6	57475.6
16	385.3	383.5	1.8	40.8	67.2	-26.4	-46.4	57429.2	57429.2
17	383.5	352.9	30.6	177.7	67.2	110.5	3384.8	60814.0	60814.0
18	352.9	342.9	10.0	177.7	51.3	126.4	1264.3	62078.2	62078.2
19	342.9	339.9	2.9	234.9	51.3	183.6	537.3	62615.6	62615.6
20	339.9	315.0	24.9	234.9	47.0	187.9	4688.4	67303.9	67303.9
21	315.0	303.0	12.0	177.7	47.0	130.7	1568.4	68872.3	68872.3
22	303.0	298.0	5.0	177.7	13.2	164.5	822.6	69694.9	69694.9
23	298.0	295.0	3.0	177.7	0.0	177.7	533.1	70228.0	70228.0
24	295.0	293.0	2.0	0.0	0.0	0.0	0.0	70228.0	70228.0

Figure 5A.3 - Grand Composite Curve for Methanol Production from the BCL Gasifier

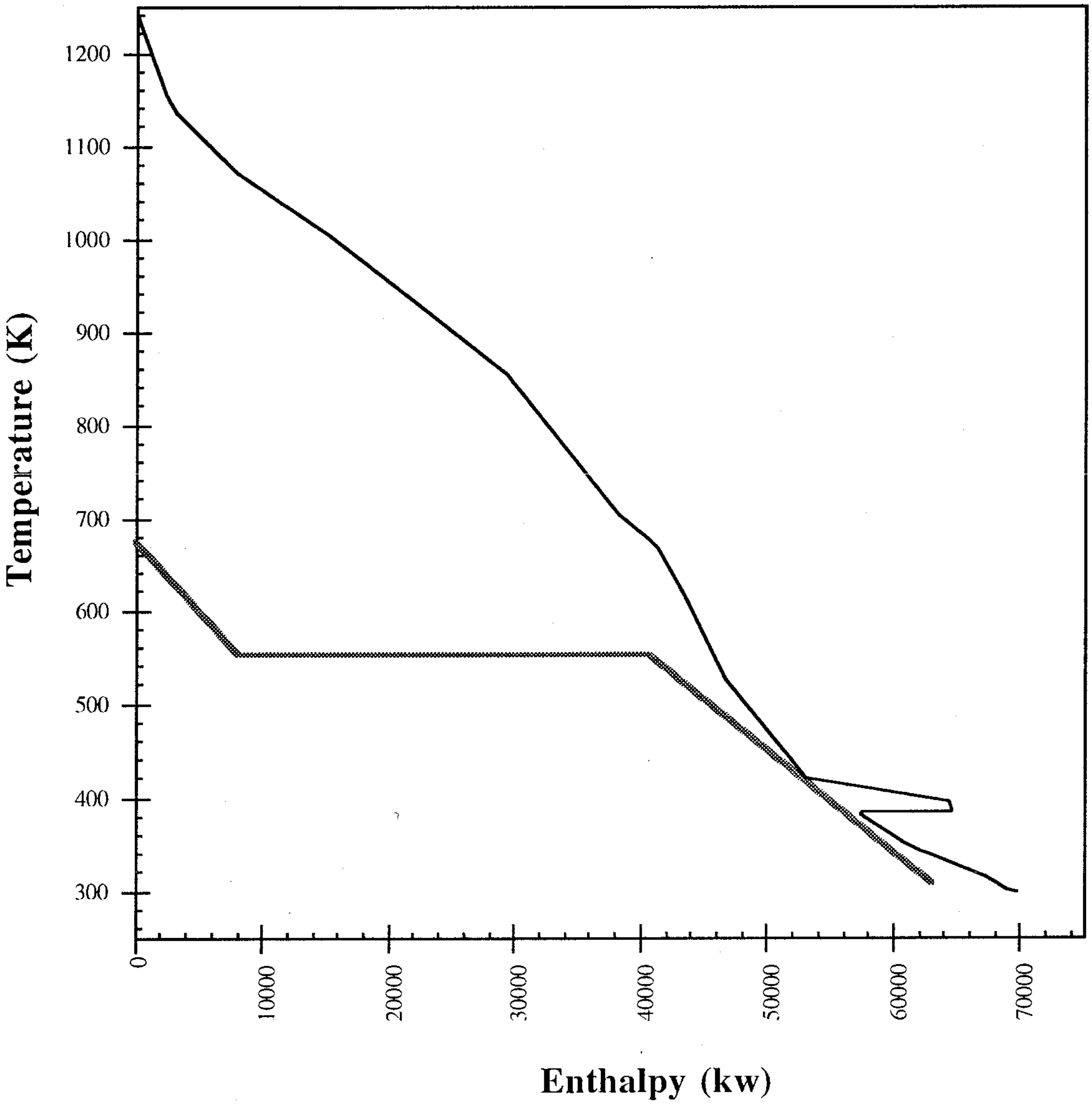
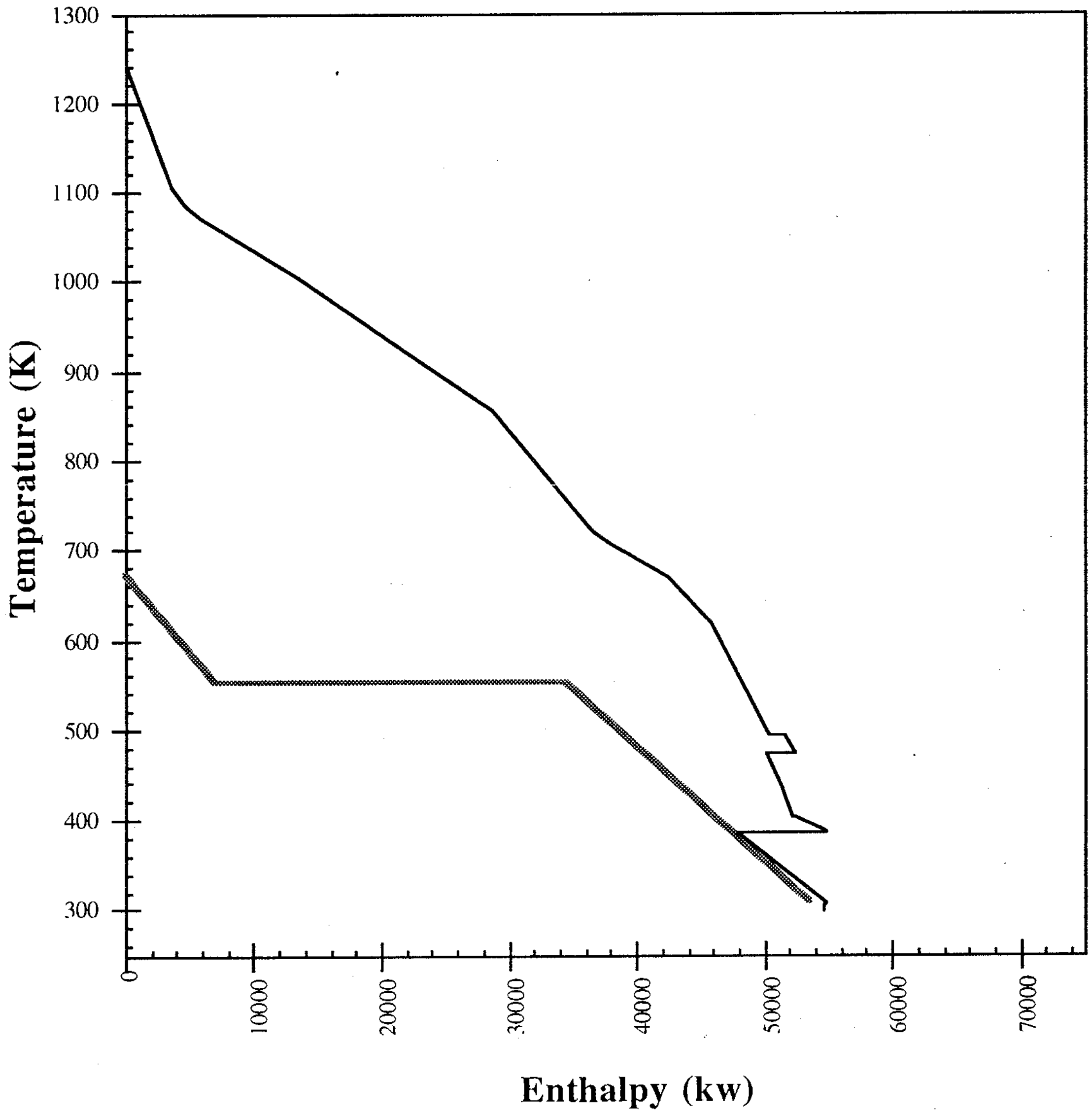


Table 5A.4 - Pinch analysis for hydrogen production with the BCL gasifier

stream	tsource K	ttarget K	hin GJ/hr	Hout GJ/hr	delta T K	delta H kw	CP kw/K	Tsource corrected	Ttarget corrected
flue (hot)	1244.4	393.0	-116.5	-200.1	-851.4	-23210.4	27.3	1239.4	388.0
comb1 h	1111.8	393.0	-347.9	-410.3	-718.8	-17323.6	24.1	1106.8	388.0
refprod h	1091.8	623.0	-369.5	-423.3	-468.8	-14961.4	31.9	1086.8	618.0
gfrprod h	1077.4	673.0	-379.8	-421.7	-404.4	-11622.2	28.7	1072.4	668.0
shift3 h	711.0	500.1	-473.0	-498.1	-210.9	-6958.5	33.0	706.0	495.1
shift4 h	500.1	500.0	-498.1	-502.8	-0.1	-1302.5	13025.0	495.1	495.0
shift5a h	500.0	408.5	0.0	-10.5	-91.5	-2913.4	31.8	495.0	403.5
shift5b h	408.5	313.0	-10.5	-60.0	-95.5	-13754.8	144.0	403.5	308.0
h22 h	407.6	313.0	4.5	0.6	-94.6	-1063.2	11.2	402.6	308.0
air1 (cold)	298.0	1000.0	0.0	26.6	702.0	7395.9	10.5	303.0	1005.0
reffeed4 c	717.4	850.0	-462.2	-448.2	132.6	3896.6	29.4	722.4	855.0
fuel2 c	312.9	850.0	-318.9	-294.2	537.1	6852.7	12.8	317.9	855.0
Gair c	298.0	672.0	0.0	31.0	374.0	8612.9	23.0	303.0	677.0
shsteamc	468.0	623.0	10.4	11.8	155.0	387.8	2.5	473.0	628.0
shsteamb	467.9	468.0	2.8	10.4	0.1	2092.5	15010.8	472.9	473.0
shsteama c	293.8	467.9	0.0	2.8	174.1	791.2	4.5	298.8	472.9
gfrsteamc	380.4	430.3	29.5	30.7	49.9	319.0	6.4	385.4	435.3
gfrsteamb	380.3	380.4	4.1	29.5	0.1	7044.8	55558.6	385.3	385.4
gfrsteama c	293.0	380.3	0.0	4.1	87.3	1151.3	13.2	298.0	385.3
						-54565.4			

Interval	T(i) K	T(i+1) K	del T K	CPhot kw/K	CPcold kw/K	diff kw/K	delta H kw	Cascade kw	Corrected cascade
0	1239.4	1106.8	132.6	27.3	0.0	27.3	3614.2	3614.2	3614.2
1	1106.8	1086.8	20.0	51.4	0.0	51.4	1027.2	4641.4	4641.4
2	1086.8	1072.4	14.4	83.3	0.0	83.3	1197.9	5839.3	5839.3
3	1072.4	1005.0	67.4	112.0	0.0	112.0	7554.0	13393.3	13393.3
4	1005.0	855.0	150.0	112.0	10.5	101.5	15221.3	28614.6	28614.6
5	855.0	722.4	132.6	112.0	52.7	59.3	7864.6	36479.2	36479.2
6	722.4	706.0	16.5	112.0	23.3	88.7	1461.8	37941.0	37941.0
7	706.0	677.0	29.0	145.0	23.3	121.7	3523.8	41464.7	41464.7
8	677.0	668.0	9.0	145.0	46.3	98.7	888.2	42352.9	42352.9
9	668.0	628.0	40.0	116.3	46.3	70.0	2798.1	45151.0	45151.0
10	628.0	618.0	10.0	116.3	48.8	67.5	674.5	45825.5	45825.5
11	618.0	495.1	122.9	84.4	48.8	35.5	4367.6	50193.1	50193.1
12	495.1	495.0	0.1	13076.4	48.8	13027.5	1302.8	51495.9	51495.9
13	495.0	473.0	22.0	83.2	48.8	34.4	756.3	52252.2	52252.2
14	473.0	472.9	0.1	83.2	15057.1	-14973.9	-2087.4	50164.8	50164.8
15	472.9	435.3	37.6	83.2	50.9	32.3	1215.3	51380.1	51380.1
16	435.3	403.5	31.8	83.2	57.3	25.9	824.0	52204.2	52204.2
17	403.5	402.6	0.9	195.4	57.3	138.1	130.2	52334.4	52334.4
18	402.6	388.0	14.6	206.6	57.3	149.4	2174.7	54509.1	54509.1
19	388.0	385.4	2.6	155.3	57.3	98.0	254.8	54763.9	54763.9
20	385.4	385.3	0.1	155.3	55609.5	-55454.2	-7031.6	47732.3	47732.3
21	385.3	317.9	67.3	155.3	64.1	91.2	6141.6	53873.9	53873.9
22	317.9	308.0	9.9	155.3	51.3	104.0	1033.2	54907.1	54907.1
23	308.0	303.0	5.0	0.0	51.3	-51.3	-256.5	54650.6	54650.6
24	303.0	298.8	4.2	0.0	17.7	-17.7	-74.8	54575.8	54575.8
25	298.8	298.0	0.8	0.0	13.2	-13.2	-10.4	54565.4	54565.4
26	298.0	293.0	5.0	0.0	0.0	0.0	0.0	54565.4	54565.4

Figure 5A.4 - Grand Composite Curve for Hydrogen Production from the BCL Gasifier



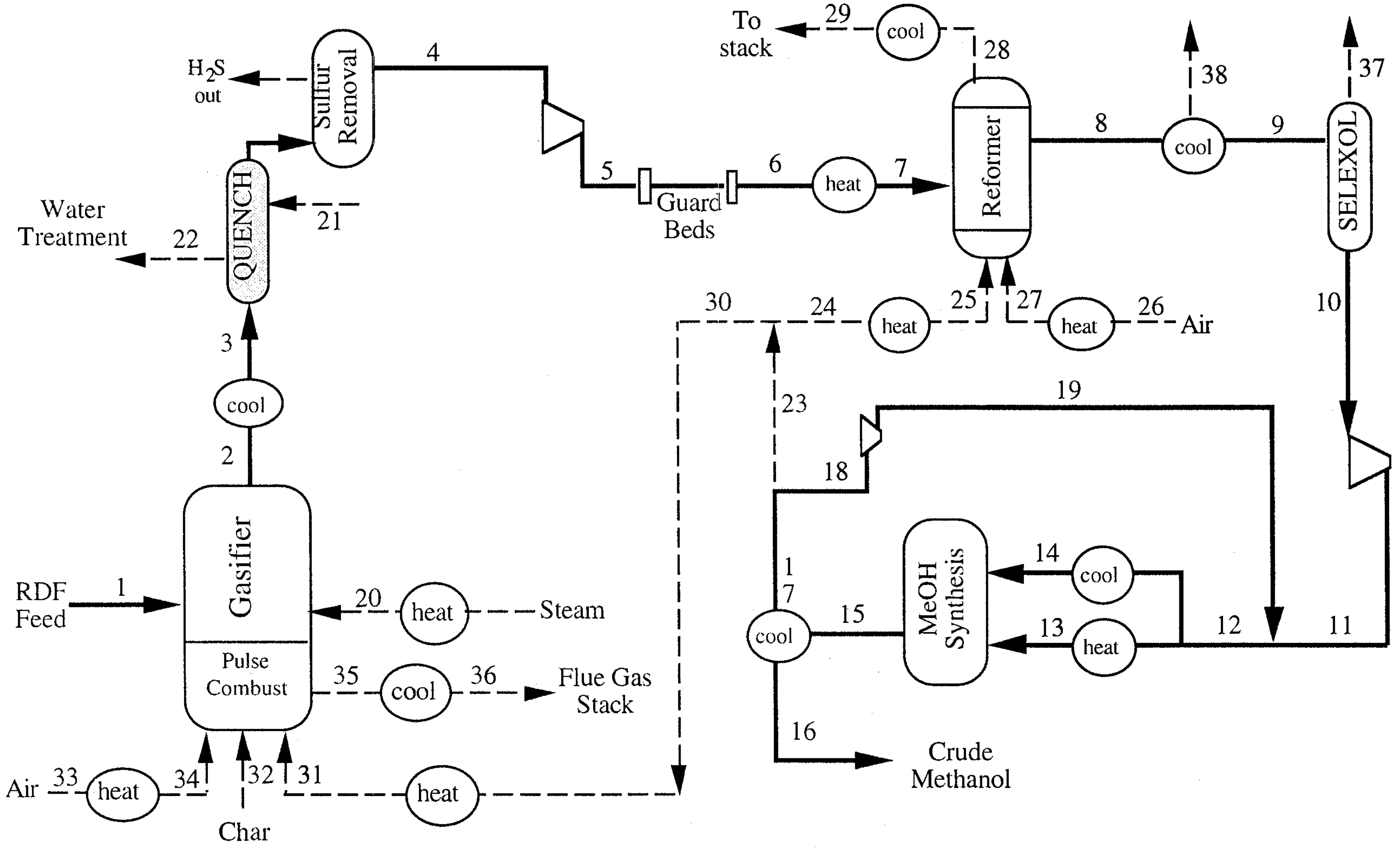


Figure 5A.5 - Flow Diagram for Methanol Production from the MTCI Gasifier

Table 5A.5 - Mass flows for Methanol Production from the MTCI gasifier

Stream	T (K)	P(bars)	kg/h	Total	H2	CO	CO2	CH4	C2H4	C2H6	C3H6	C3H8	C4H10	MeOH	H2O	N2	O2
			Molar Flowrates (kmol/hr)														
1 RDF feed			872.75 dry tonnes/day														
2 gasifier product gas	1061	1.5	73554	3963	882	269	468	153	82	11	9	9	8	0	2073	0	0
3 cooled gasifier product gas	673	1.5	73554	3963	882	269	468	153	82	11	9	9	8	0	2073	0	0
4 cleaned gas	370	1.5	76280	4115	882	269	468	153	82	11	9	9	8	0	2225	0	0
5 compressed gas	554	15.0	76280	4115	882	269	468	153	82	11	9	9	8	0	2225	0	0
6 cool reformer feed gas	638	14.5	76280	4024	783	269	468	153	0	118	0	0	8	0	2225	0	0
7 hot reformer feed gas	850	14.0	76280	4024	783	269	468	153	0	118	0	0	8	0	2225	0	0
8 reformer product gas	1059	13.5	76280	4643	1671	475	571	114	0	0	0	0	0	0	1812	0	0
9 selexol feed	400	13.5	55526	3491	1671	475	571	114	0	0	0	0	0	0	660	0	0
10 selexol output	400	13.5	20523	2306	1671	475	46	114	0	0	0	0	0	0	0	0	0
11 makeup	499	105.9	20523	2306	1671	475	46	114	0	0	0	0	0	0	0	0	0
12 synfeed	347	105.9	62921	11529	9354	567	73	1514	0	0	0	0	0	21	0	0	0
13 hotfeed	523	105.4	23834	4367	3543	215	28	573	0	0	0	0	0	8	0	0	0
14 coldfeed	320	105.4	39086	7162	5811	352	45	940	0	0	0	0	0	13	0	0	0
15 MeOH synthesis product	533	97.3	62921	10508	8290	100	30	1514	0	0	0	0	0	531	44	0	0
16 crude MeOH product	300	96.8	17181	558	1	0	1	4	0	0	0	0	0	509	43	0	0
17 recycled synprod	300	96.8	45739	9951	8288	100	29	1510	0	0	0	0	0	23	0	0	0
18 recycled synprod2	300	96.8	42397	9223	7683	93	27	1400	0	0	0	0	0	21	0	0	0
19 compressed recycled synprod	309	105.9	42397	9223	7683	93	27	1400	0	0	0	0	0	21	0	0	0
20 gasifier steam	700	1.5	36390	2020	0	0	0	0	0	0	0	0	0	0	2020	0	0
21 quench water	293	1.5	196650	10916	0	0	0	0	0	0	0	0	0	0	10916	0	0
22 condensate	370	1.5	193920	10764	0	0	0	0	0	0	0	0	0	0	10764	0	0
23 purge stream	300	96.8	3342	727	606	7	2	110	0	0	0	0	0	2	0	0	0
24 cool reformer comb. feed	300	96.8	1391	303	252	3	1	46	0	0	0	0	0	1	0	0	0
25 hot reformer comb. feed	850	96.8	1391	303	252	3	1	46	0	0	0	0	0	1	0	0	0
26 cool reformer comb. air	293	1.0	34656	1201	0	0	0	0	0	0	0	0	0	0	0	949	252
27 hot reformer comb. air	1000	1.0	34656	1201	0	0	0	0	0	0	0	0	0	0	0	949	252
28 reformer hot flue	1079	1.0	36047	1376	0	0	50	0	0	0	0	0	0	1	344	949	33
29 reformer cool flue	393	1.0	36047	1376	0	0	50	0	0	0	0	0	0	1	344	949	33
30 cool pulse feed	300	96.8	1951	424	354	4	1	64	0	0	0	0	0	1	0	0	0
31 hot pulse feed	850	96.8	1951	424	354	4	1	64	0	0	0	0	0	1	0	0	0
32 char to pulse	293	1.0	535	44.6 kmole C/h													
33 cool pulse air	298	1.0	50801	1761	0	0	0	0	0	0	0	0	0	0	0	1391	370
34 hot pulse air	1000	1.0	50801	1761	0	0	0	0	0	0	0	0	0	0	0	1391	370
35 pulse hot flue	1150	1.0	53287	2007	0	0	115	0	0	0	0	0	0	0	484	1391	16
36 pulse cool flue	393	1.0	53287	2007	0	0	115	0	0	0	0	0	0	0	484	1391	16
37 CO2 out	400	13.5	35003	1185	0	0	525	0	0	0	0	0	0	0	660	0	0
38 condensed water	400	13.5	20754	1152	0	0	0	0	0	0	0	0	0	0	1152	0	0

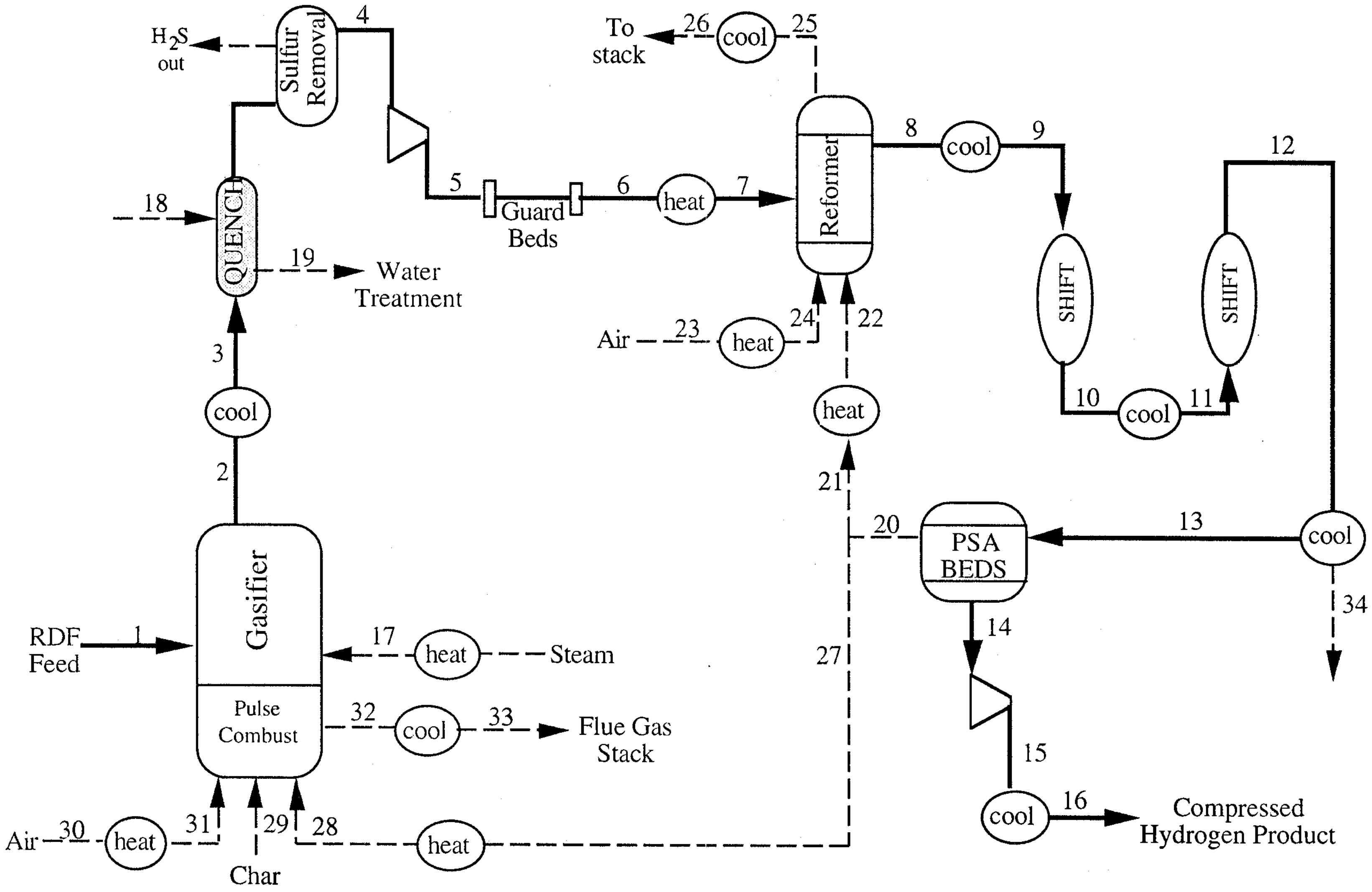


Figure 5A.6 - Flow Diagram for Hydrogen Production from the MTCI Gasifier

Table 5A.6 - Mass flows for Hydrogen Production from RDF using the MTCI gasifier

				Molar Flowrates (kmol/hr)												
Stream	T (K)	P (bars)	kg/h	Total	H_2	CO	CO_2	CH_4	C_2H_4	C_2H_6	C_3H_6	C_3H_8	C_4H_{10}	H_2O	N_2	O_2
1 RDF feed			872.75 dry tonnes/day													
2 gasifier product gas	1061	1.5	73554	3963	882	269	468	153	82	11	9	9	8	2073	0	0
3 cooled gasifier product gas	673	1.5	73554	3963	882	269	468	153	82	11	9	9	8	2073	0	0
4 cleaned gas	370	1.5	76280	4115	882	269	468	153	82	11	9	9	8	2225	0	0
5 compressed gas	556	15.5	76280	4115	882	269	468	153	82	11	9	9	8	2225	0	0
6 cool reformer feed gas	637	15.4	76280	4024	791	269	468	153	0	92	0	17	8	2225	0	0
7 hot reformer feed gas	850	14.9	76280	4024	791	269	468	153	0	92	0	17	8	2225	0	0
8 reformer product gas	1038	14.4	76280	4554	1554	413	588	159	0	0	0	0	0	1839	0	0
9 hi temp shift feed	623	13.9	76280	4554	1554	413	588	159	0	0	0	0	0	1839	0	0
10 hi-temp shift product	690	13.4	76280	4554	1847	120	882	159	0	0	0	0	0	1546	0	0
11 low-T shift feed	500	12.9	76280	4554	1847	120	882	159	0	0	0	0	0	1546	0	0
12 low-T shift product	500	12.4	76280	4554	1952	15	986	159	0	0	0	0	0	1441	0	0
13 cooled PSA feed	313	11.9	50590	3128	1952	15	986	159	0	0	0	0	0	16	0	0
14 hydrogen	313	11.6	3056	1497	1494	0	0	3	0	0	0	0	0	0	0	0
15 compressed hydrogen	408	75.0	3056	1497	1494	0	0	3	0	0	0	0	0	0	0	0
16 hydrogen product	313	75.0	3056	1497	1494	0	0	3	0	0	0	0	0	0	0	0
17 gasifier steam	700	1.3	36390	2020	0	0	0	0	0	0	0	0	0	2020	0	0
18 quench water	293	1.5	196650	10916	0	0	0	0	0	0	0	0	0	10916	0	0
19 condensate	370	1.5	193920	10764	0	0	0	0	0	0	0	0	0	10764	0	0
20 purge stream	313	1.3	47534	1631	458	15	986	156	0	0	0	0	0	16	0	0
21 cool reformer comb. fuel	313	1.3	17112	587	165	6	355	56	0	0	0	0	0	6	0	0
22 hot reformer comb. fuel	850	1.3	17112	587	165	6	355	56	0	0	0	0	0	6	0	0
23 cool reformer comb. air	293	1.0	31177	1081	0	0	0	0	0	0	0	0	0	0	854	227
24 hot reformer comb. air	1000	1.0	31177	1081	0	0	0	0	0	0	0	0	0	0	854	227
25 reformer hot flue	1058	1.0	48289	1583	0	0	417	0	0	0	0	0	0	283	854	30
26 reformer cool flue	393	1.0	48289	1583	0	0	417	0	0	0	0	0	0	283	854	30
27 cool pulse comb. feed	313	1.3	30422	1044	293	10	631	100	0	0	0	0	0	10	0	0
28 hot pulse comb. feed	800	1.3	30422	1044	293	10	631	100	0	0	0	0	0	10	0	0
29 char to pulse	293	1.0	535	44.6 kmole C/h												
30 cool pulse air	298	1.0	57033	1977	0	0	0	0	0	0	0	0	0	0	1562	415
31 hot pulse air	1000	1.0	57033	1977	0	0	0	0	0	0	0	0	0	0	1562	415
32 pulse hot flue	1150	1.0	87990	2869	0	0	785	0	0	0	0	0	0	503	1562	20
33 pulse cool flue	393	1.0	87990	2869	0	0	785	0	0	0	0	0	0	503	1562	20
34 condensed water	313	11.9	25690	1426	0	0	0	0	0	0	0	0	0	1426	0	0

Table 5A.7 - Pinch analysis for methanol production with the MTCI gasifier

stream	tsource K	ttarget K	hin GJ/hr	Hout GJ/hr	delta T K	delta H kw	CP kw/K	Tsource corrected	Ttarget corrected
Pprod (hot)	1150.0	393.0	-105.0	-156.8	-757.0	-14397	19.0	1145.0	388.0
comb1 h	1079.4	393.0	-67.3	-99.0	-686.4	-8803	12.8	1074.4	388.0
gfrprod h	1060.8	673.0	-608.2	-672.0	-387.8	-17728	45.7	1055.8	668.0
refproda h	1059.4	427.0	0.0	-107.7	-632.4	-29917	47.3	1054.4	422.0
synproda h	533.0	381.9	0.0	-53.3	-151.1	-14800	97.9	528.0	376.9
refprodb h	427.0	400.0	-107.7	-159.1	-27.0	-14287	529.1	422.0	395.0
synprodb h	381.9	300.0	-53.3	-102.4	-81.9	-13637	166.6	376.9	295.0
cold1 h	346.7	320.0	-119.1	-125.0	-26.7	-1642	61.5	341.7	315.0
Pair1 (cold)	298.0	1000.0	0.0	38.3	702.0	10639	15.2	303.0	1005.0
refair c	293.0	1000.0	-0.2	26.1	707.0	7307	10.3	298.0	1005.0
reffeed2 c	637.7	850.0	-726.1	-691.2	212.3	9708	45.7	642.7	855.0
comb-fd c	300.0	850.0	-4.2	1.3	550.0	1528	2.8	305.0	855.0
puls-fd c	300.0	850.0	-5.9	1.8	550.0	2143	3.9	305.0	855.0
gfrsteamc c	380.4	700.0	94.7	118.3	319.6	6535	20.4	385.4	705.0
hotfeed1 c	346.7	523.0	-72.6	-48.3	176.3	6749	38.3	351.7	528.0
gfrsteamb c	380.3	380.4	13.3	94.7	0.1	22617	178367	385.3	385.4
gfrsteama c	293.0	380.3	0.0	13.3	87.3	3696	42.4	298.0	385.3
						-44291			

Interval	T(i) K	T(i+1) K	del T K	CPhot kw/K	CPcold kw/K	diff kw/K	delta H kw	Cascade kw	Corrected cascade
0	1145.0	1074.4	70.6	19.0	0.0	19.0	1343.5	1343.5	1343.5
1	1074.4	1055.8	18.6	31.8	0.0	31.8	591.0	1934.5	1934.5
2	1055.8	1054.4	1.4	77.6	0.0	77.6	111.8	2046.3	2046.3
3	1054.4	1005.0	49.4	124.9	0.0	124.9	6163.3	8209.6	8209.6
4	1005.0	855.0	150.0	124.9	25.5	99.4	14906.8	23116.5	23116.5
5	855.0	705.0	150.0	124.9	77.9	47.0	7044.9	30161.3	30161.3
6	705.0	668.0	37.0	124.9	98.3	26.5	981.2	31142.6	31142.6
7	668.0	642.7	25.3	79.2	98.3	-19.2	-484.7	30657.9	30657.9
8	642.7	528.0	114.7	79.2	52.6	26.5	3046.0	33703.8	33703.8
9	528.0	422.0	106.0	177.1	90.9	86.2	9136.2	42840.0	42840.0
10	422.0	395.0	27.0	658.9	90.9	568.0	15337.5	58177.5	58177.5
11	395.0	388.0	7.0	129.8	90.9	38.9	272.2	58449.7	58449.7
12	388.0	385.4	2.6	97.9	90.9	7.0	18.3	58468.0	58468.0
13	385.4	385.3	0.1	97.9	178438	-178340	-22613.5	35854.5	35854.5
14	385.3	376.9	8.4	97.9	112.8	-14.9	-125.0	35729.5	35729.5
15	376.9	351.7	25.2	166.6	112.8	53.8	1353.1	37082.6	37082.6
16	351.7	341.7	10.0	166.6	74.5	92.1	920.5	38003.2	38003.2
17	341.7	315.0	26.7	228.1	74.5	153.6	4100.6	42103.8	42103.8
18	315.0	305.0	10.0	166.6	74.5	92.1	920.5	43024.3	43024.3
19	305.0	303.0	2.0	166.6	67.8	98.7	197.5	43221.8	43221.8
20	303.0	298.0	5.0	166.6	52.7	113.9	569.4	43791.2	43791.2
21	298.0	295.0	3.0	166.6	0.0	166.6	499.7	44290.9	44290.9
22	295.0	293.0	2.0	0.0	0.0	0.0	0.0	44290.9	44290.9

Figure 5A.7 - Grand Composite Curve for Methanol Production from the MTCI Gasifier

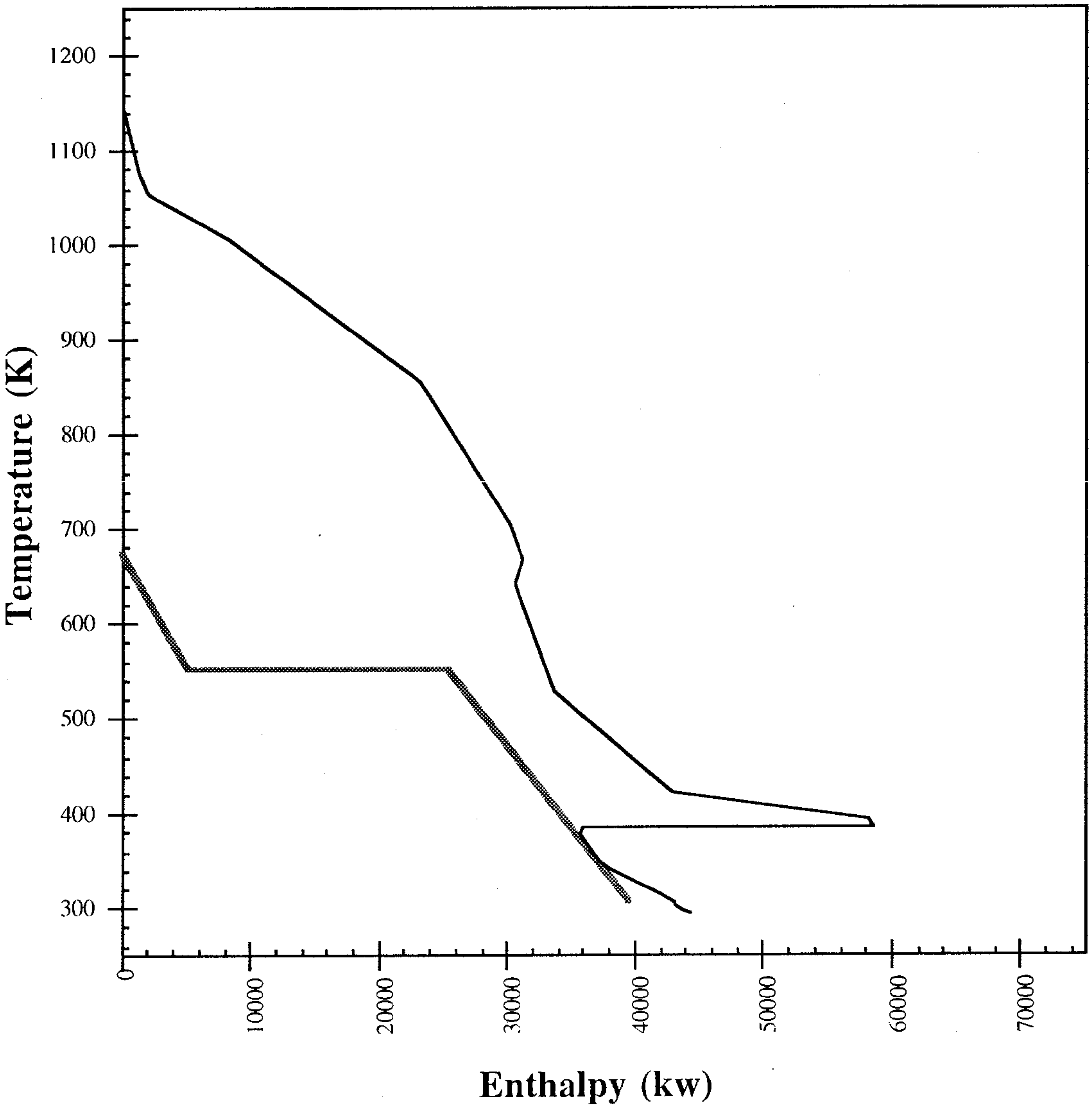
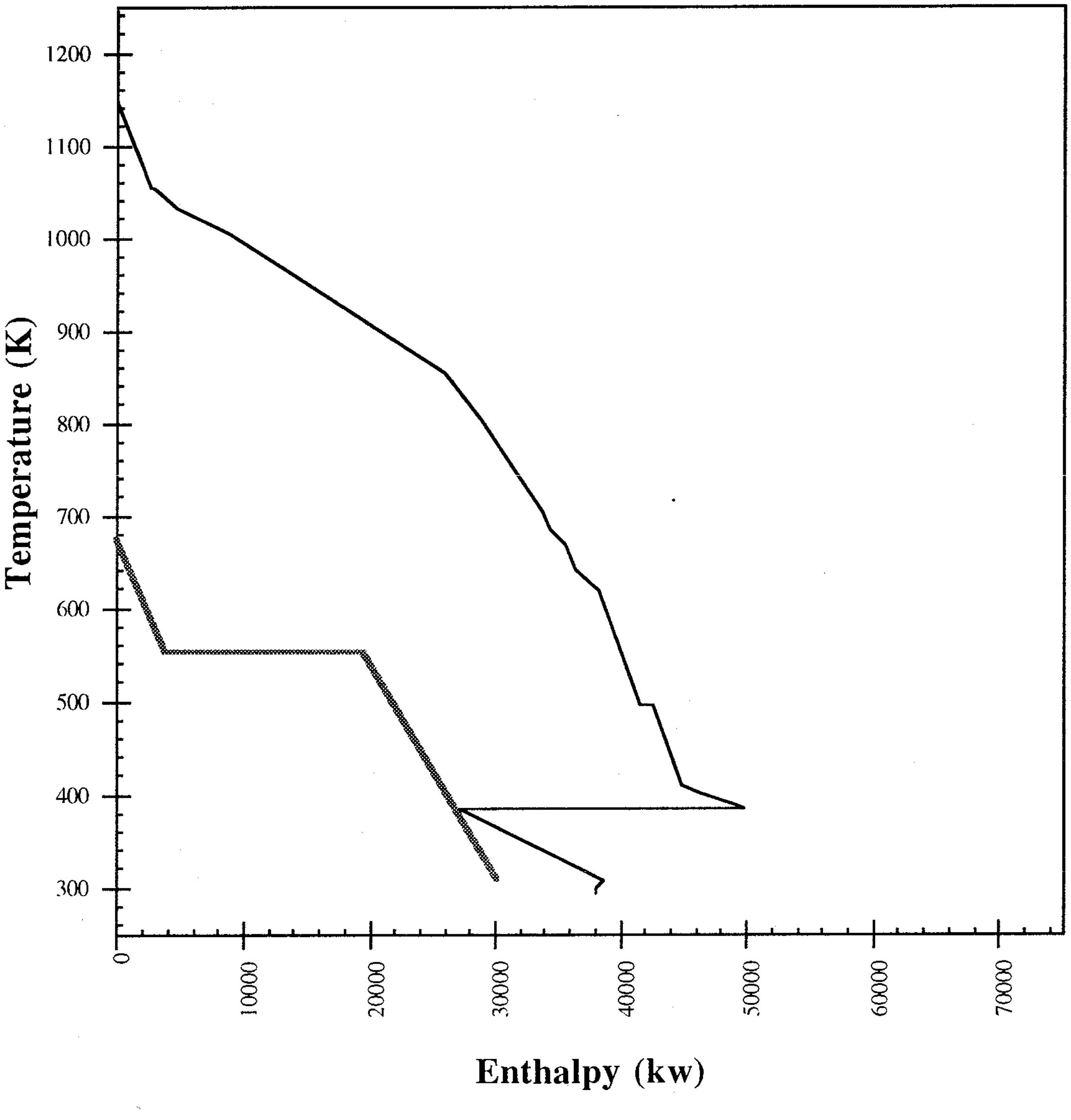


Table 5A.8 - Pinch analysis for hydrogen production with the MTCI gasifier

stream	tsource K	ttarget K	hin GJ/hr	Hout GJ/hr	delta T K	delta H kw	CP kw/K	Tsource corrected	Ttarget corrected
Pprod (hot)	1150.0	393.0	-340.0	-422.0	-757.0	-22767.2	30.1	1145.0	388.0
gfrprod h	1060.8	673.0	-608.2	-672.0	-387.8	-17728.1	45.7	1055.8	668.0
comb1 h	1058.3	393.0	-188.5	-227.6	-665.3	-10858.8	16.3	1053.3	388.0
refprod h	1038.3	623.0	-612.8	-684.7	-415.3	-19988.5	48.1	1033.3	618.0
shift3 h	689.8	500.1	-684.7	-716.1	-189.7	-8727.7	46.0	684.8	495.1
shift4 h	500.1	500.0	0.0	-4.1	-0.1	-1143.5	11434.7	495.1	495.0
shift5a h	500.0	415.1	0.0	-13.6	-84.9	-3765.5	44.3	495.0	410.1
shift5b h	415.1	313.0	-13.6	-94.7	-102.1	-22535.9	220.8	410.1	308.0
h22 h	407.5	313.0	4.7	0.6	-94.5	-1138.4	12.0	402.5	308.0
Pair1 (cold)	298.0	1000.0	0.0	43.0	702.0	11944.1	17.0	303.0	1005.0
refair c	293.0	1000.0	-0.2	23.5	707.0	6573.1	9.3	298.0	1005.0
reffeed4 c	636.5	850.0	-725.9	-690.7	213.5	9777.4	45.8	641.5	855.0
comb-fd c	312.9	850.0	-145.7	-132.5	537.1	3661.6	6.8	317.9	855.0
puls-fd c	313.0	800.0	-259.0	-238.0	487.0	5834.3	12.0	318.0	805.0
gfrsteamc	380.4	700.0	94.7	118.3	319.6	6534.6	20.4	385.4	705.0
gfrsteamb	380.3	380.4	13.3	94.7	0.1	22616.9	178367	385.3	385.4
gfrsteama c	293.0	380.3	0.0	13.3	87.3	3696.4	42.4	298.0	385.3
						-38015			

Interval	T(i) K	T(i+1) K	del T K	CPhot kw/K	CPcold kw/K	diff kw/K	delta H kw	Cascade kw	Corrected cascade
0	1145.0	1055.8	89.2	30.1	0.0	30.1	2682.7	2682.7	2682.7
1	1055.8	1053.3	2.5	75.8	0.0	75.8	189.9	2872.6	2872.6
2	1053.3	1033.3	20.0	92.1	0.0	92.1	1842.2	4714.9	4714.9
3	1033.3	1005.0	28.3	140.2	0.0	140.2	3968.1	8683.0	8683.0
4	1005.0	855.0	150.0	140.2	26.3	113.9	17089.7	25772.7	25772.7
5	855.0	805.0	50.0	140.2	78.9	61.3	3065.5	28838.1	28838.1
6	805.0	705.0	100.0	140.2	90.9	49.3	4932.9	33771.0	33771.0
7	705.0	684.8	20.2	140.2	111.4	28.9	583.2	34354.3	34354.3
8	684.8	668.0	16.8	186.2	111.4	74.9	1258.7	35612.9	35612.9
9	668.0	641.5	26.5	140.5	111.4	29.2	772.0	36384.9	36384.9
10	641.5	618.0	23.5	140.5	65.6	75.0	1765.0	38149.9	38149.9
11	618.0	495.1	122.9	92.4	65.6	26.8	3299.6	41449.5	41449.5
12	495.1	495.0	0.1	11481.1	65.6	11415.6	1141.6	42591.1	42591.1
13	495.0	410.1	84.9	90.7	65.6	25.2	2138.7	44729.8	44729.8
14	410.1	402.5	7.6	267.2	65.6	201.6	1523.2	46252.9	46252.9
15	402.5	388.0	14.5	279.2	65.6	213.6	3103.9	49356.8	49356.8
16	388.0	385.4	2.6	232.8	65.6	167.2	434.8	49791.7	49791.7
17	385.4	385.3	0.1	232.8	178412	-178179	-22593.1	27198.5	27198.5
18	385.3	318.0	67.3	232.8	87.5	145.3	9777.4	36976.0	36976.0
19	318.0	317.9	0.1	232.8	75.5	157.3	10.1	36986.0	36986.0
20	317.9	308.0	9.9	232.8	68.7	164.1	1630.9	38616.9	38616.9
21	308.0	303.0	5.0	0.0	68.7	-68.7	-343.3	38273.6	38273.6
22	303.0	298.0	5.0	0.0	51.7	-51.7	-258.3	38015.3	38015.3
23	298.0	293.0	5.0	0.0	0.0	0.0	0.0	38015.3	38015.3

Figure 5A.8 - Grand Composite Curve for Hydrogen Production from the MTCI Gasifier



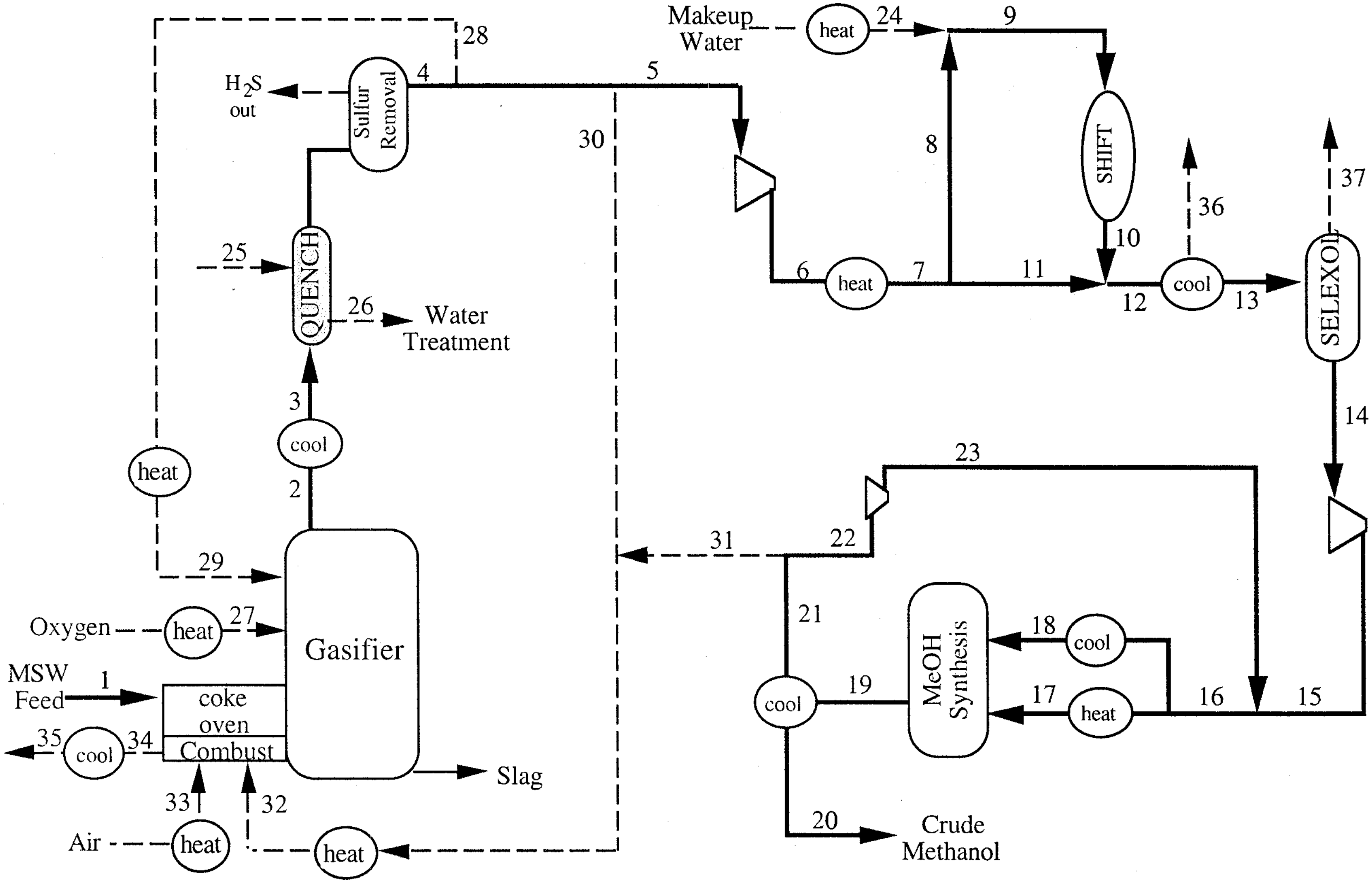


Figure 5A.9 - Flow Diagram for Methanol Production from the Thermoselect Gasifier

Table 5A.9 - Mass flows for Methanol Production from MSW using the Thermoselect Gasifier

Stream	T (K)	P (bars)	kg/h	Molar Flowrates (kmol/hr) Total	H_2	CO	CO_2	CH_4	MeOH	H_2O	N_2	O_2
1 MSW feed			1052 dtpd									
2 gasifier product gas	1473	1.0	58984	3035	990	1028	361	0	0	607	49	0
3 cooled gasifier product gas	673	1.0	58984	3035	990	1028	361	0	0	607	49	0
4 cleaned gas	348	1.0	68930	3587	990	1028	361	0	0	1159	49	0
5 reformer compressor feed	348	1.0	61313	3191	881	914	321	0	0	1031	43	0
6 compressed gas	589	15.5	61313	3191	881	914	321	0	0	1031	43	0
7 hot compressed gas	623	15.0	61313	3191	881	914	321	0	0	1031	43	0
8 feed split to shift	623	15.0	30351	1579	436	453	159	0	0	510	21	0
9 shift feed	621	15.0	45619	2427	436	453	159	0	0	1358	21	0
10 shift product	773	14.5	45619	2427	795	93	519	0	0	998	21	0
11 shift bypass	623	15.0	30961	1611	445	462	162	0	0	521	22	0
12 recombined gas	716	14.5	76581	4038	1240	555	681	0	0	1519	43	0
13 selexol feed	313	14.0	49405	2530	1240	555	681	0	0	11	43	0
14 selexol output	313	13.5	20904	1876	1240	555	38	0	0	0	43	0
15 makeup	501	105.9	20904	1876	1240	555	38	0	0	0	43	0
16 synfeed	347	105.9	132240	9379	5119	840	157	2	23	0	3238	0
17 hotfeed	523	105.4	7052	500	273	45	8	0	1	0	173	0
18 coldfeed	320	105.4	125190	8879	4846	795	148	2	22	0	3066	0
19 MeOH synthesis product	533	97.3	132240	8215	3924	289	126	2	605	31	3238	0
20 crude MeOH product	300	96.8	19630	626	1	0	5	0	582	31	7	0
21 recycled synprod	300	96.8	112610	7589	3922	289	121	2	23	0	3232	0
22 recycled synprod2	300	96.8	111340	7503	3878	285	119	2	23	0	3195	0
23 compressed recycled synprod	309	105.9	111340	7503	3878	285	119	2	23	0	3195	0
24 shift steam	623	15.0	15268	848	0	0	0	0	0	848	0	0
25 quench water	293	1.5	89894	4990	0	0	0	0	0	4990	0	0
26 condensate	348	1.0	79949	4438	0	0	0	0	0	4438	0	0
27 gasifier oxygen		1.0	14465	452	0	0	0	0	0	0	0	452
28 cool gasifier fuel	348	1.0	2618	136	38	39	14	0	0	44	2	0
29 hot gasifier fuel	523	1.0	2618	136	38	39	14	0	0	44	2	0
30 fuel split gas	348	1.0	4999	260	72	75	26	0	0	84	4	0
31 purge gas	300	96.8	1269	86	44	3	1	0	0	0	36	0
32 coke oven fuel	373	1.0	6268	346	116	78	28	0	0	84	40	0
33 coke oven air	373	1.0	14640	507	0	0	0	0	0	0	401	107
34 coke hot flue	1200	1.0	20908	756	0	0	106	0	0	201	441	9
35 coke cool flue	393	1.0	20908	756	0	0	106	0	0	201	441	9
36 condensed water	313	14.0	27176	1508	0	0	0	0	0	1508	0	0
37 CO_2 out	313	13.5	28501	654	0	0	643	0	0	11	0	0

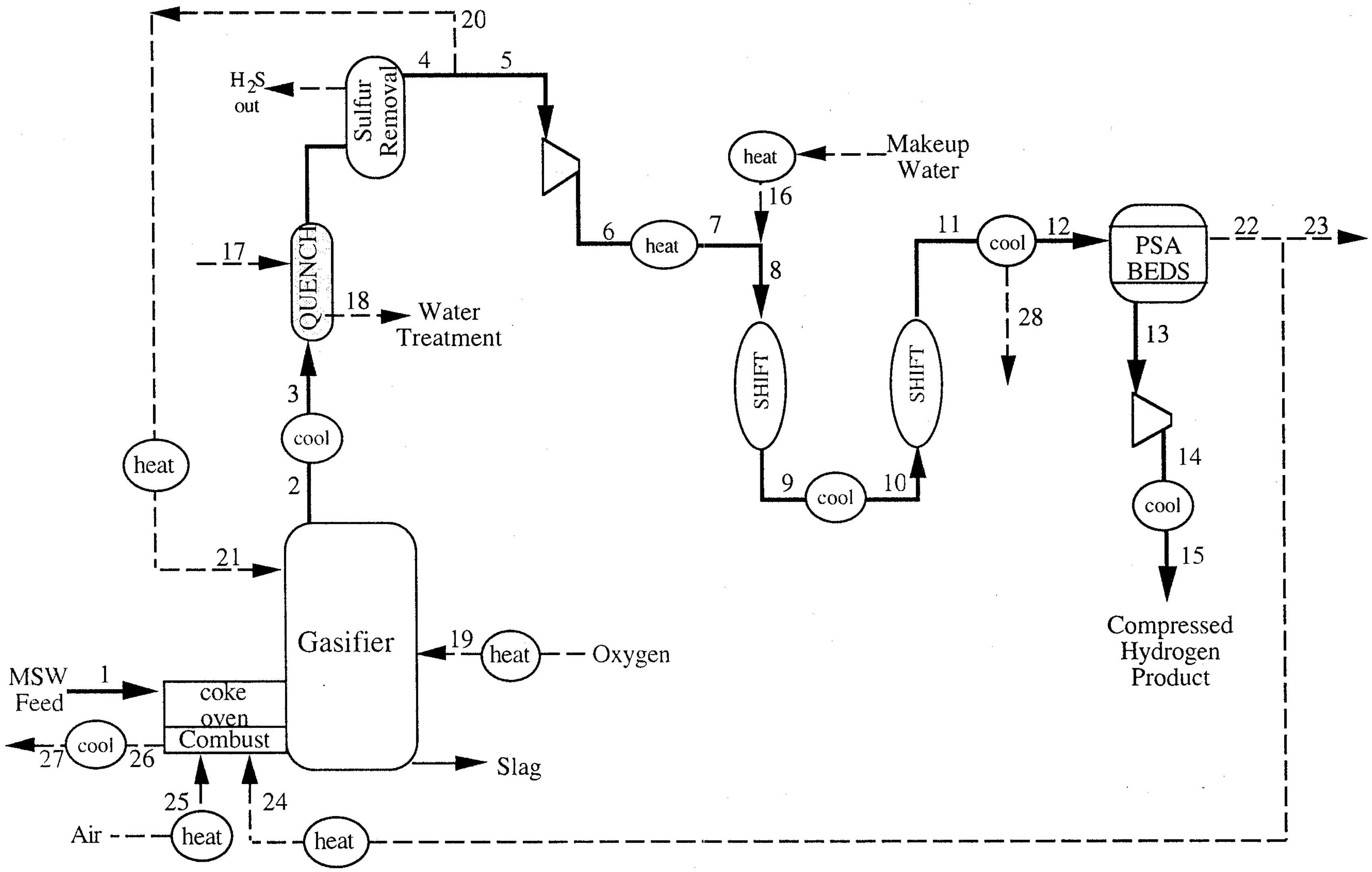


Figure 5A.10 - Flow Diagram for Hydrogen Production from the Thermoselect Gasifier

Table 5A.10 - Mass flows for Hydrogen Production from MSW using the Thermoselect gasifier

Stream	T (K)	P (bars)	kg/h	Molar Flowrates (kmol/hr) Total	H_2	CO	CO_2	CH_4	H_2O	N_2	O_2
1 MSW feed			1052 dtpd								
2 gasifier product gas	1473	1.0	58984	3035	990	1028	361	0	607	49	0
3 cooled gasifier product gas	673	1.0	58984	3035	990	1028	361	0	607	49	0
4 cleaned gas	348	1.0	68930	3587	990	1028	361	0	1159	49	0
5 reformer compressor feed	348	1.0	66312	3451	953	989	347	0	1115	47	0
6 compressed gas	589	15.5	66312	3451	953	989	347	0	1115	47	0
7 hot compressed gas	623	15.0	66312	3451	953	989	347	0	1115	47	0
8 hi-T shift feed	621	15.0	99669	5303	953	989	347	0	2967	47	0
9 hi-T shift product	773	14.5	99669	5302	1738	203	1133	0	2181	47	0
10 low-T shift feed	500	14.0	99669	5302	1738	203	1133	0	2181	47	0
11 low-T shift product	500	13.5	99669	5302	1927	15	1322	0	1993	47	0
12 cooled PSA feed	313	13.0	64038	3325	1927	15	1321	0	15	47	0
13 hydrogen	313	12.7	3476	1724	1724	0	0	0	0	0	0
14 compressed hydrogen	403	75.0	3476	1724	1724	0	0	0	0	0	0
15 hydrogen product	313	75.0	3476	1724	1724	0	0	0	0	0	0
16 shift steam	623	15.0	33357	1852	0	0	0	0	1852	0	0
17 quench water	293	1.5	89894	4990	0	0	0	0	4990	0	0
18 condensate	348	1.0	79949	4438	0	0	0	0	4438	0	0
19 gasifier oxygen		1.0	14465	452	0	0	0	0	0	0	452
20 cool gasifier fuel	348	1.0	2618	136	38	39	14	0	44	2	0
21 hot gasifier fuel	523	1.0	2618	136	38	39	14	0	44	2	0
22 PSA purge gas	313	1.3	60562	1601	202	15	1321	0	15	47	0
23 purge gas to be burned	313	1.3	2111	56	7	1	46	0	1	2	0
24 coke oven fuel	373	1.3	58451	1545	195	14	1275	0	15	45	0
25 coke oven air	373	1.0	15820	548	0	0	0	0	0	433	115
26 coking hot flue	1000	1.0	74272	1988	0	0	1290	0	210	478	10
27 coking cool flue	393	1.0	74272	1988	0	0	1290	0	210	478	10
28 condensed water	313	13.0	35632	1978	0	0	0	0	1977	0	0

Table 5A.11 - Pinch analysis for methanol production with the Thermoselect gasifier

stream	tsource K	ttarget K	hin GJ/hr	Hout GJ/hr	delta T K	delta H kw	CP kw/K	tsource corrected	t target corrected
gfrprod (hot)	1473.0	673.0	-277.0	-366.0	-800.0	-24735	30.9	1468.0	668.0
shift5a h	716.3	426.9	0.0	-41.4	-289.5	-11497	39.7	711.3	421.9
synproda h	533.0	388.1	0.0	-39.4	-144.9	-10945	75.5	528.0	383.1
shift5b h	426.9	313.0	-41.4	-125.8	-113.9	-23449	205.9	421.9	308.0
synprodb h	388.1	300.0	-39.4	-85.6	-88.1	-12833	145.7	383.1	295.0
cold1 h	346.8	320.0	-138.4	-145.7	-26.8	-2029	75.6	341.8	315.0
shiftfd2 (cold)	589.2	623.0	-447.3	-443.6	33.8	1027	30.4	594.2	628.0
shsteamc c	471.5	623.0	41.3	46.7	151.5	1520	10.0	476.5	628.0
hotfeed1 c	346.8	523.0	-7.8	-5.1	176.2	743	4.2	351.8	528.0
shsteamb c	471.4	471.5	11.6	41.3	0.1	8248	147821	476.4	476.5
shsteama c	293.0	471.4	0.0	11.6	178.4	3216	18.0	298.0	476.4
						-70735			

Interval	T(i) K	T(i+1) K	del T K	CPhot kw/K	CPcold kw/K	diff kw/K	delta H kw	Cascade kw	Corrected cascade
0	1468.0	711.3	756.7	30.9	0.0	30.9	23395.2	23395.2	23395.2
1	711.3	668.0	43.3	70.6	0.0	70.6	3060.8	26456.0	26456.0
2	668.0	628.0	40.0	39.7	0.0	39.7	1588.7	28044.7	28044.7
3	628.0	594.2	33.8	39.7	40.4	-0.7	-22.9	28021.8	28021.8
4	594.2	528.0	66.2	39.7	10.0	29.7	1964.2	29986.0	29986.0
5	528.0	476.5	51.5	115.2	14.2	101.0	5200.6	35186.6	35186.6
6	476.5	476.4	0.1	115.2	147825	-147710	-8242.2	26944.4	26944.4
7	476.4	421.9	54.6	115.2	22.2	93.0	5075.0	32019.5	32019.5
8	421.9	383.1	38.8	281.4	22.2	259.2	10061.9	42081.3	42081.3
9	383.1	351.8	31.2	351.7	22.2	329.4	10285.1	52366.4	52366.4
10	351.8	341.8	10.0	351.7	18.0	333.7	3336.6	55703.1	55703.1
11	341.8	315.0	26.8	427.3	18.0	409.3	10982.2	66685.3	66685.3
12	315.0	308.0	7.0	351.7	18.0	333.7	2335.6	69020.8	69020.8
13	308.0	298.0	10.0	145.7	18.0	127.7	1277.2	70298.1	70298.1
14	298.0	295.0	3.0	145.7	0.0	145.7	437.2	70735.3	70735.3
15	295.0	293.0	2.0	0.0	0.0	0.0	0.0	70735.3	70735.3

Figure 5A.11 - Grand Composite Curve for MeOH Production from the ThermoSelect Gasifier

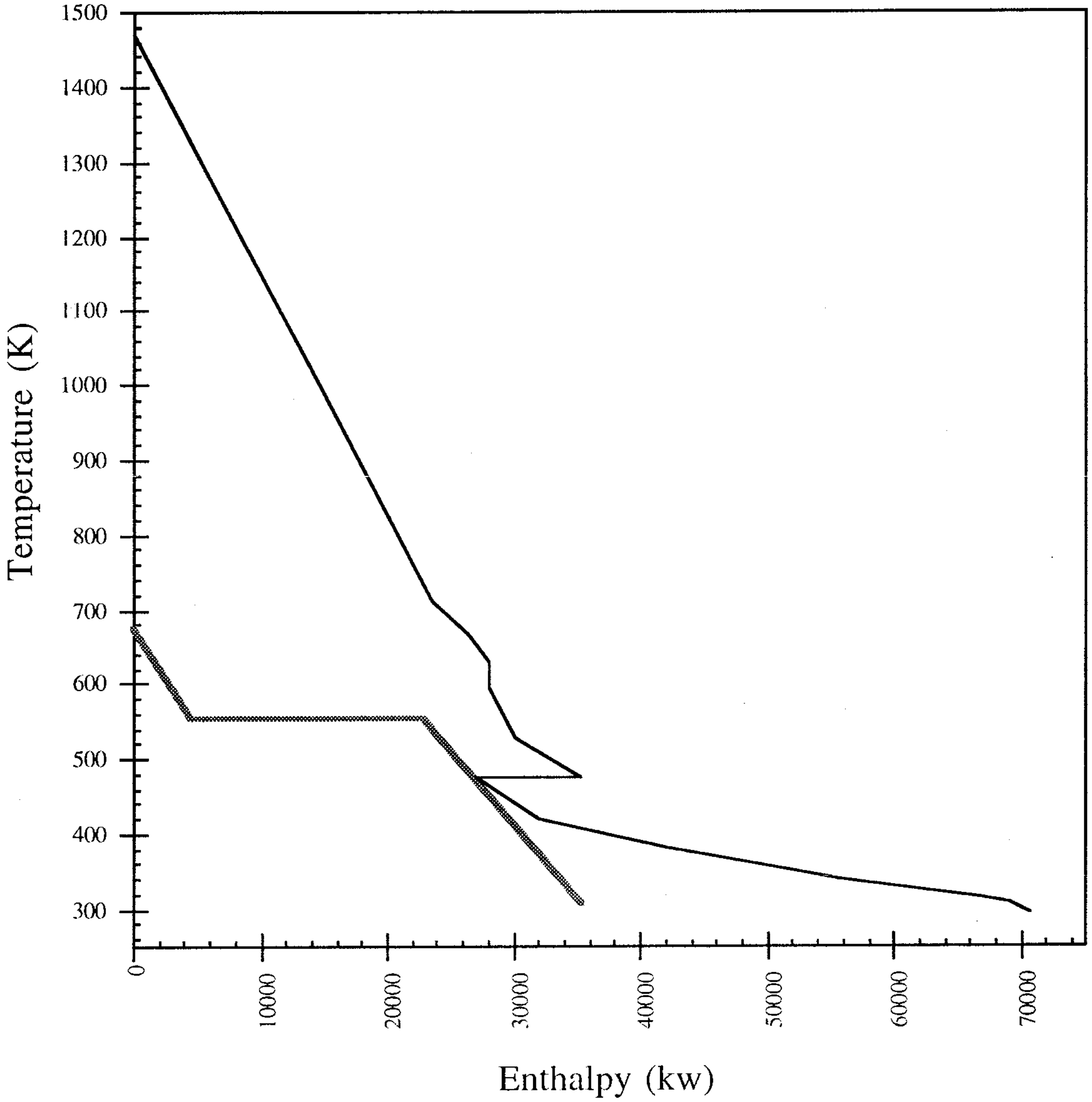
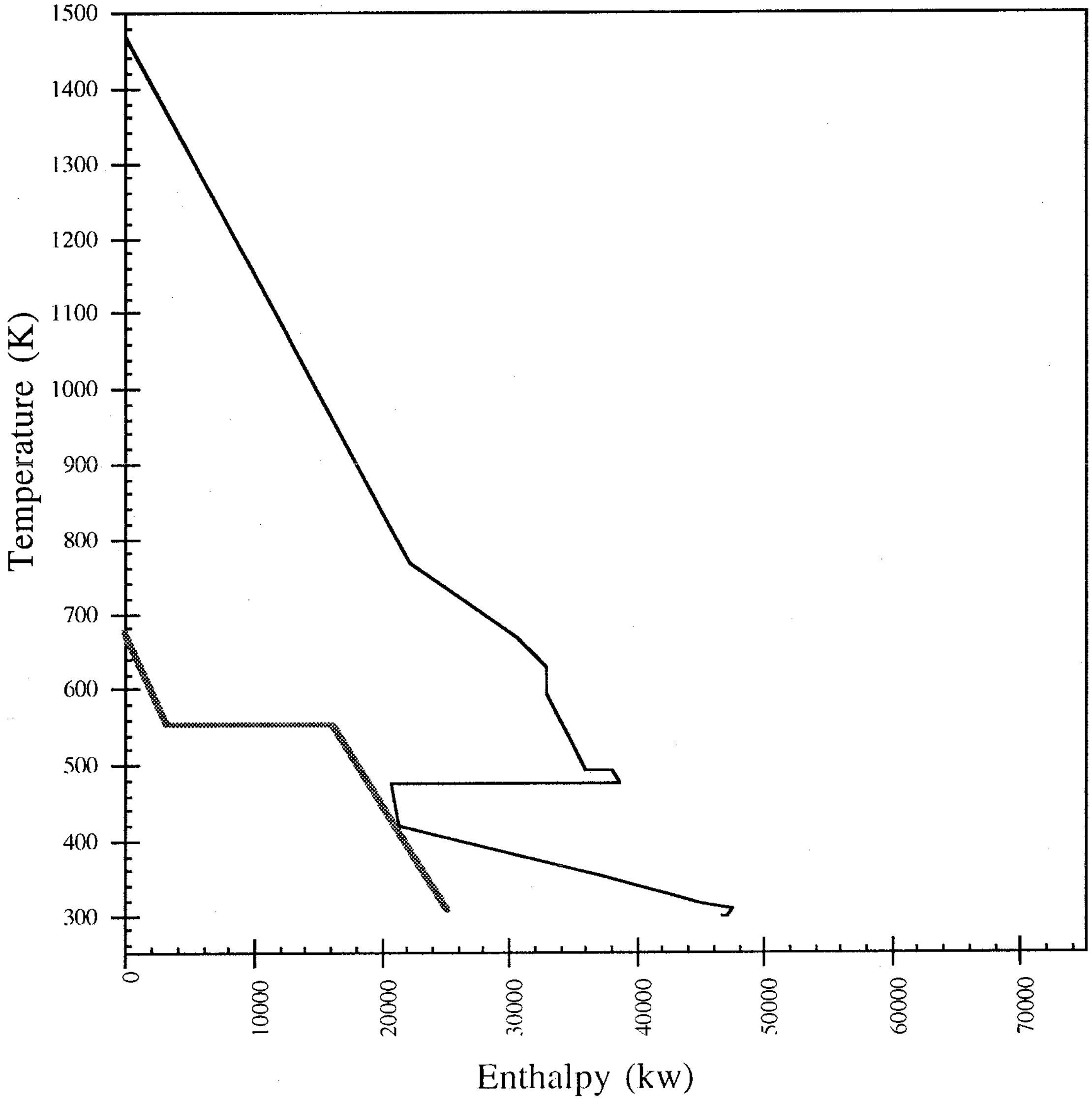


Table 5A.12 - Pinch analysis for hydrogen production with the Thermoselect gasifier

stream	tsource K	ttarget K	hin GJ/hr	Hout GJ/hr	delta T K	delta H kw	CP kw/K	Tsource corrected	Ttarget corrected
gfrprod (hot)	1473.0	673.0	-277.0	-366.0	-800.0	-24734.9	30.9	1468.0	668.0
shift3 h	773.0	500.1	-907.8	-961.3	-272.9	-14858.1	54.4	768.0	495.1
shift4 h	500.1	500.0	-961.3	-968.7	-0.1	-2058.5	20585.3	495.1	495.0
shift5a h	500.0	424.2	0.0	-14.3	-75.8	-3971.6	52.4	495.0	419.2
shift5b h	424.2	313.0	-14.3	-125.1	-111.2	-30773.1	276.8	419.2	308.0
h22 h	403.5	313.0	5.4	0.9	-90.5	-1253.4	13.9	398.5	308.0
shftfd2 (cold)	589.2	623.0	-483.8	-479.8	33.8	1110.9	32.8	594.2	628.0
shsteama c	471.5	623.0	90.2	102.1	151.5	3321.2	21.9	476.5	628.0
hom-fd c	348.0	523.0	-20.2	-19.4	175.0	214.5	1.2	353.0	528.0
shsteamb c	471.4	471.5	25.3	90.2	0.1	18021.1	322959	476.4	476.5
shsteamc c	293.3	471.4	0.0	25.3	178.2	7030.4	39.5	298.3	476.4
hom-oxy c	293.0	400.0	0.0	0.1	107.0	33.9	0.3	298.0	405.0
coke-fd c	312.9	373.0	-506.6	-503.1	60.1	966.6	16.1	317.9	378.0
cokeair c	293.0	373.0	-0.1	1.2	80.0	357.2	4.5	298.0	378.0
						-46593.8			

Interval	T(i) K	T(i+1) K	del T K	CPhot kw/K	CPcold kw/K	diff kw/K	delta H kw	Cascade kw	Corrected cascade
1	1468.0	768.0	700.0	30.9	0.0	30.9	21642.5	21642.5	21642.5
2	768.0	668.0	100.0	85.4	0.0	85.4	8537.6	30180.1	30180.1
3	668.0	628.0	40.0	54.4	0.0	54.4	2177.7	32357.7	32357.7
4	628.0	594.2	33.8	54.4	54.8	-0.3	-10.8	32347.0	32347.0
5	594.2	528.0	66.2	54.4	21.9	32.5	2151.8	34498.8	34498.8
6	528.0	495.1	32.9	54.4	23.1	31.3	1029.6	35528.4	35528.4
7	495.1	495.0	0.1	20585.3	23.1	20562.1	2056.2	37584.6	37584.6
8	495.0	476.5	18.5	52.4	23.1	29.2	540.8	38125.4	38125.4
9	476.5	476.4	0.1	52.4	322961	-322908	-18018.3	20107.1	20107.1
10	476.4	419.2	57.3	52.4	40.7	11.7	669.8	20776.9	20776.9
11	419.2	405.0	14.2	276.8	40.7	236.1	3347.1	24124.0	24124.0
12	405.0	398.5	6.5	276.8	41.0	235.8	1543.7	25667.7	25667.7
13	398.5	378.0	20.5	290.7	41.0	249.7	5106.2	30773.9	30773.9
14	378.0	353.0	25.0	290.7	61.6	229.1	5719.3	36493.2	36493.2
15	353.0	317.9	35.1	290.7	60.3	230.3	8088.5	44581.7	44581.7
16	317.9	308.0	9.9	290.7	44.2	246.4	2443.9	47025.6	47025.6
17	308.0	298.3	9.7	0.0	44.2	-44.2	-430.5	46595.1	46595.1
18	298.3	298.0	0.3	0.0	4.8	-4.8	-1.3	46593.8	46593.8
19	298.0	293.0	5.0	0.0	0.0	0.0	0.0	46593.8	46593.8

Figure 5A.12 - Grand Composite Curve for Hydrogen Production Based on the Thermoselect Gasifier



Chapter 6: Emissions from MSW Gasifiers

6.1 Introduction

Environmental considerations are important in MSW disposal. Public perception of incineration is generally negative, even with advances in air pollution control. Even if the MSW can be economically converted into fuels, the process will still not be a viable option if gasification does not dispose of waste in an environmentally sound manner.

In this chapter comparisons of emissions from gasifiers and incinerators are briefly discussed. Also, available emission characteristic data of the BCL, MTCI, and Thermoselect gasifiers are presented. Though the data from the indirectly heated gasifiers are sparse, the Thermoselect gasifier has extensive emissions measurements which suggest that it would emit fewer pollutants than incinerators. Since the emissions characteristics of these fuel production processes have significant impact on the process' feasibility, emission control systems that can be employed to clean out the contaminants in the gas stream are also discussed in this chapter.

6.2 Qualitative Comparison of Emissions from Gasifiers and Incinerators

Preliminary comparisons of emissions characteristics can be made between fuel production and incineration by considering fundamental characteristics of each process. Because gasification takes place in an oxygen deficient environment, while incinerators operate with excess air, gasification produces lower volumes of gas than incineration. None of the three gasification systems that are examined for fuel production are air-blown, so there is no nitrogen dilution of the product gas. This further reduces the gas volumes that will need cleanup treatment.[1] Since the costs of most pollution control devices scale with volume flow, the gas clean up for gasifiers should be less expensive than for incinerators, all other factors being equal. Also, the

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[1] Nitrogen constitutes around 70% of an incinerator's flue gas volume on a wet basis.

smaller gas volume will lead to a higher concentration of contaminants, which might facilitate the removal of some potential pollutants.

Table 6.1 compares the volume flows from the three gasifiers to that of an incinerator. Treating the product and flue gases as ideal gases, the molar flow rates can be directly translated into volume flow rates. This comparison shows that the gas volumes from the incinerator are three to five times greater than the volumes from the gasifiers.

Table 6.1 - Gas Volumes of Gasifiers and Incinerators (in kmol/dry tonne MSW)

	Incinerator[a]	BCL[b]	MTCI[b,c]	Thermoselect[b]
Product Gas	---	56	91	69
Flue Gas	361	64	30	---
Total Gas Flow	361	120	121	69
Relative to incinerator gas volume	1.00	0.33	0.33	0.19

Notes:

a) Incinerator flue gas calculated assuming 22.5% moisture and a MSW organic fraction of $CH_{1.6}O_{0.53}$. 80% excess air is assumed, as recommended by the EPA (US EPA, 1987a)

b) Gasifier product and flue gas flowrates are taken from chapters 3 and 5.

c) The MTCI flue gas should not need extensive cleaning since it mainly burns cleaned product gas.

It has been conjectured that gasification would produce fewer polychlorinated dibenzo-p-dioxin (PCDD) and polychlorinated dibenzofuran (PCDF)[2] than combustion (Rovere, 1994; Durai-Swamy, 1991a, Riegel et al., 1993), but available empirical evidence is too limited to confirm this. PCDDs/PCDFs are toxic halogenated aromatic compounds that are of considerable public concern (Figure 6.1; more information on PCDD/PCDF is in Appendix 6A). MSW incinerators have been determined to be by far the largest sources of PCDD/PCDF emissions (Thomas and Spiro, 1994), so an alternative MSW technology that could produce fewer PCDD/Fs would be attractive. Although PCDD/PCDF formation is not fully understood, oxygen is needed for the compound to form. One reason why it has been postulated that gasifiers might emit fewer PCDD/PCDFs than combustors is that gasifiers have no or

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[2] In current literature, these compounds are often referred to as dioxin and furan.

at least sub-stoichiometric molecular oxygen inputs. However, oxygen contained in the MSW or steam inputs (with the indirectly heated gasifiers) will be present, so PCDD/PCDFs may still be produced in the gasifier or product gas. Clearly, more experimental work and in-depth analyses of the reaction kinetics are required to fully understand PCDD/F formation and destruction in gasifiers and combustors.

The previous discussion has focused on the gasifier itself. Downstream reactors needed for fuels production from gasified MSW also provide environmental advantages over incineration. Reactor catalysts require much more stringent restrictions on pollutants than do public health-based environmental regulations. For example, a minute concentration of sulfur (0.5 ppm) in the synthesis gas will damage the reformer catalyst, while the air quality restriction for SO_2 in flue gases @ 7% O_2 is over 30 ppm (80 mg/Nm^3). Thus the product gas must be thoroughly cleaned to prevent any poisoning of the processing equipment. In other words, the air emissions from the fuel production plant will be very low or non-existent because the equipment tolerances dictate so. This is not the same as cleaning the flue gas merely to meet air emission regulations. This argument does not ensure that liquid and solid effluents from fuels production will be uncontaminated, however.

6.3 Analysis of Gasifier Emissions Data

The previous discussion has outlined why fuel production could conceivably produce less pollution than incineration. Unfortunately, there are only limited data available on emissions from gasifiers. Of the three gasifiers examined, only Thermoselect has been thoroughly tested for its emissions characteristics. For the two indirectly heated gasifiers, BCL and MTCI, complete analyses of all the effluent streams have not yet been made (Table 6.2). Though gasifier emissions data are scant, preliminary analyses are promising.

When discussing emissions, it is important to consider all mass flows exiting a process. Designers of incinerators have focused their attention on meeting air emission regulations, and incinerator emissions have been reduced remarkably over the last twenty years. These improvements have come about through better combustion practices and better post-combustion pollution control equipment. Since the emphasis has mainly been on air pollution control (APC) devices, there is now increased regulatory pressure on the solid emissions from incinerators, i.e. - the hazardous nature of the ash. The trace metals and organic compounds that were scrubbed and captured from the flue are now present as a solid effluent. The same reasoning applies for the fuel processing system.

Table 6.2 - Available Emissions Data for the Three Gasifiers

Effluent Stream	BCL	MTCI	Thermoselect
Product Gas	No	No	Yes
Flue Gas	No	No	N/A
Liquid	Yes	No	Yes
Solid	No	Yes	Yes

N/A = not applicable

6.3.1 Emissions Data for the Three Gasifiers

6.3.1.1 Battelle Columbus Laboratories Emissions Data

BCL's pilot gasifier setup included the RDF feeder and transport system, the BCL gasifier, a spray tower to scrub the combustor flue gas, and a simple water quench to wash the product gas (as was modeled in the reference cases). The waste water from the quench was cleaned through a simple treatment process (settling chamber, sand filter, and charcoal filter). Instruments continuously measured the temperature, pressure, and the flowrates of air, steam, and RDF. The flue gas and product gas were continuously sampled to quantify major components (Paisley et al., 1989).

The product gas was not analyzed for any pollutants. Also, the solid residues were not analyzed. Samples of the waste water were taken from only the outlet of the water treatment process. Tests of these samples showed that the water was well within the US Environmental Protection Agency (EPA) Safe Drinking Water Act standards for every pollutant (Paisley et al., 1989). Thus, it appears that the simple water treatment process is adequate to clean the dirty quench water, but since no samples of the waste water were taken before it was treated, it is not clear that the quench water was contaminated. Since the product gas also was not tested, it is possible that many pollutants remained in the gas stream. Also, the RDF used in the test runs was relatively clean, in that negligible amounts of chlorine were in the feed. This may have been a reason for the low concentration of chlorinated compounds in the water samples.

Like an incinerator, the BCL gasifier burns a fraction of its char, but the resulting flue gas volumes are much less than the volumes from an incinerator (Table 6.1). However, no analysis of the char composition or the pollutant content of the combustor flue gas has been made.

6.3.1.2 MTCI Emissions Data

MTCI's experimental setup for RDF feeds consisted of the gasifier, cyclone, a quench scrubber, and a flare to burn the product gas. Like BCL, MTCI measured only the major constituents of the product gas but did not test it for pollutants. MTCI also did not analyze the waste water from the scrubber (MTCI, 1990; Durai-Swamy et al., 1991a). The char caught by the cyclone was not recirculated to the gasifier for these experiments. Samples of the solid residue were taken to analyze the ash's toxicity characteristics using the EPA's toxicity characteristics leachate procedure (TCLP).

As can be seen in Table 6.3, the tests show that the solid residue would not be classified as a hazardous substance.

Table 6.3 - TCLP Metal Concentrations (mg/ltr) from the Cyclone Caught Ash from the MTCI Gasifier Using RDF Feed[a]

Metal	Detection Limit	Test Result	STLC[b]
As	0.0125	ND	5.0
Ba	0.1	1.3	100.0
Cd	0.1	ND	1.0
Cr	0.1	ND	5.0
Pb	0.1	ND	5.0
Hg	0.0025	ND	0.2
Se	0.0125	0.021	1.0
Au	0.1	ND	5.0

Notes: from MTCI (1990) ND = none detected

a) TCLP is the Toxicity Characteristics Leachate Procedure adopted by the EPA

b) The STLC is the Soluble Threshold Limit Concentration. If any heavy metal concentrations in the leachate of the tested substance are greater than the STLC, then the substance is classified as a hazardous material.

Though no analysis of PCDD/F has yet been run for RDF feeds, MTCI has measured PCDD concentrations in the wastewater and solid residues from gasifying paper mill sludge (Table 6.4). Far fewer PCDDs were found in the solid and liquid output streams than in the sludge inputs, but the PCDD concentrations were not measured in the product gas.

Table 6.4 - PCDD Levels in Papermill Sludge Feed and MTCI Gasifier Effluents (ppt)

	Total PCDD	2,3,7,8 TCDD
Feed sludge	1543.00	33.00
Bed material	10.10	ND
Cyclone ash	100.20	27.00
Condensate	0.33	0.07

Notes: ND = none detected
from Durai-Swamy et al., 1991b

No tests were run on the flue gas because natural gas was used as the pulse combustor fuel in the experiment. In actual operation, the cleaned product gas would be used as the pulse combustor fuel, which should burn at least as cleanly as natural gas. For the base case analysis of Chapter 5, it was assumed that the unreacted char would be cleaned and also burned as fuel for the pulse combustor. As is the case for the BCL gasifier, it is uncertain how cleanly the char from RDF burns. With little thermodynamic penalty, the MTCI gasifier could burn only clean product gas and dispose of the char as solid residue (see Chapter 5).

6.3.1.3 Thermoselect Emissions Data

The Thermoselect plant in Italy has an extensive clean up process that follows the gasifier. This includes a water quench, alkaline scrubbers, sulfur removal equipment, activated carbon beds (to remove trace amounts of acid gases), and a fabric filter. Waste water is treated in a water treatment system and then recycled back to the quench. As shown in Table 6.1, Thermoselect is the only gasifier of the three examined here that has analyzed its solid, liquid, and gas effluent streams for pollutants.

RWTUV, the German federal technology testing agency, ran 18 hours of comprehensive tests on the demonstration plant in Italy to measure pollutants in the cleaned synthesis gas, the waste water from the quench (before and after treatment), the granulate in the slag, and the leachate from the granulate (Lutzke and Wehde, 1993). PCDD levels were measured in the waste water and product gas.[3]

The tests demonstrated that the Thermoselect process produces very few emissions. Each effluent stream analyzed easily passes current emissions standards.

The gas emissions from the Thermoselect gasifier are displayed in Table 6.5.

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[3] It is safe to assume that no dioxin would be in the granulate since the slag acts as an internal vitrification system. Vitrification is a high temperature thermal treatment that is considered as one suitable method for stabilizing MSW incinerator ash.

Air emissions from a new US incinerator equipped with state-of-the-art air pollution control equipment and the current European and US regulations/guidelines for incinerators are also shown for comparison. For every pollutant, the incinerator emits more than the gasifier. There are especially prominent differences between the Thermoselect and incinerator emissions for mercury, SO_2, HCl, and PCDDs. The 60-fold difference in the mercury emissions between incineration and Thermoselect may be a result of Italian MSW (used in the testing) having much less mercury than US MSW. Incinerators have a much higher SO_2 concentration because the sulfur is oxidized in incinerators, whereas in the gasifier it is reduced to H_2S. As can be seen in Table 6.5, the H_2S can be efficiently removed (see Appendix 6B). Thermoselect has

Table 6.5 - Comparison of Gas Emissions from MSW Processing: Gasification vs Incineration of MSW

	Gasifier	Incinerators →		
	Thermoselect[a]	U.S. guidelines[b]	European Standard '93[c]	New Incinerator[d]
mg/Nm^3				
PM	0.25	36	10	0.48
Cd	0.001		0.1	0.001
Hg	0.006		0.1	0.111
Pb	0.005			0.015
H_2S	<0.1			
SO_2	1.35	78	50	20.8
HCl	0.5	37	10	33.2
HF	<0.06			n.d.
ng/Nm^3				
PCDDs/PCDFs (total)	0.193	30.0		n.a.
TEQ[e]	0.0035			0.08

Notes: n.d. = none detected n.a. = not available

a) Thermoselect data are for the cleaned synthesis gas (Lutzke and Wehde, 1993).

b) EPA, 40 CFR Part 60 - NSPS Emissions Guidelines for Large Municipal Waste Combustors permitted after 12/89.

c) from Riegel and Runyon (1993)

d) Data for the Kent County, MI incinerator: a 625 tpd mass-burn facility that started operation in 1990. It uses a spray-dry scrubber and baghouse filter to control emissions (Sullivan, 1993).

e) PCDD/PCDF Toxicity Equivalent using NATO standards (see Appendix 6A) low HCl emissions most likely because Thermoselect has extensive gas cleaning equipment and the product gas volumes are small, which makes gas cleaning easier. Emissions from both the incinerator and Thermoselect are well within US air emission guidelines. Except for the mercury emissions from the incinerator, both technologies can also meet European standards.

Analyses of the original granulate substance from the slag and the leachate from the granulate show that the granulate would be classified as non-hazardous in both Germany and the US. The waste water from the quench was analyzed before and after the water treatment plant. RWTUV determined that the treated waste water could be discharged to a standard water treatment plant (Lutzke and Wehde, 1993).

6.3.2 Material Balance Estimates (for Thermoselect)

Pollutants can leave the Thermoselect gasifier in three different effluent streams: the solid residue (slag), waste water, or product gas. Boiling points of the various compounds help predict if the compound will vaporize and leave with the

Table 6.6 - Melting and Boiling Points (in °C) of Some Metals Found in MSW

Metal	T_m	T_b
Hg	-39	357
As		613
K	63	759
Cd	321	767
Na	98	883
Zn	420	907
Mg	650	1090
Ca	842	1484
Pb	328	1749
Al	660	2519
Cu	1085	2562
Sn	232	2602
Cr	1907	2671
Fe	1538	2861
Ni	1455	2913

from CRC (1972)

product gas (Table 6.6). Clearly, almost all of the nickel should be in the slag, while most of the arsenic will be in the waste water or product gas. To corroborate this simple theory, material balances of some key pollutants for the Thermoselect gasifier have been estimated to quantify how various pollutants leave the gasifier.

Using emissions data (Lutzke and Wehde, 1993), which are given as concentrations, and multiplying them by appropriate flowrates, the mass flows can be calculated. The flowrates are the key assumptions here. For 3 tonnes MSW/hour input (the module size tested by RWTUV), the effluent streams are assumed to have the following flowrates adapted from a set of plant specifications (Runyon, 1993): 3000 Nm^3/h of cleaned product gas, 15000 ltrs/hour of water, 690 kg/hour of granulate minerals, and 87 kg/h of metals.

The results of these material balances are tabulated in Table 6.7, showing where some major pollutants end up. It is clear that most of the mercury is volatilized in the gasifier but is later condensed into the waste water stream by the scrubber.

Table 6.7 - Mass Flowrates (in grams/hr) of Selected Compounds Emitted from the Thermoselect Gasifier Categorized by Effluent Stream[a]

Stream	**Pollutant**				
	Hg	Pb	Fe[b]	S[c]	Cl[d]
gas	0.02	0.02		2.04	0.75
water	0.38	165.00	240.00	2640.00	99519.78
granulate	0.03	690.00	82800.00	897.00	
metals	0.01	69.60	78300.00	348.00	
Total (g/hr)	0.45	924.62	161340.00	3887.04	99520.52

Notes:
a) mass flows calculated by multiplying the compound concentration (from Lutzke and Wehde, 1993) with an assumed volume flowrate based on Runyon (1993). The assumed flowrates: 3000 Nm^3/h (product gas), 15000 ltrs/hour (water based on holding tank size), 690 kg/hour (granulate), and 87 kg/h (metals). Estimate is for MSW throughput of 3 tonne/day.
b) Fe in granulate is both Fe and Fe_2O_3
c) S in gas is mainly SO_2. In water it is mainly SO_3
d) Cl is in the form of chlorides in the water. For simplicity, it is assumed to be NaCl. Negligible amounts of chlorides were found in the granulate leachate.

The iron, on the other hand, almost all ends up in the granulate (in both its reduced and oxidized form), or it is separated out as a metal product. Lead has a boiling point in between the other two metals (1749°C). Most of it ends up in the granulate, but quite a bit of lead gets to the waste water stream. The majority of sulfur is in the waste water stream, most likely from the acid gas scrubber. About a quarter of it gets into the slag. Clearly, most of the chlorine is scrubbed out of the gas and ends up in the waste water stream.

The results of these mass balance estimates follow the same trend that would be predicted from the boiling points of each pollutant. A similar analysis for theindirectly heated gasifiers (especially BCL) would be valuable because it is important to know which compounds enter the product gas stream and which are burned or volatilized with the char and enter the flue gas.

6.3.3 Estimates of Total PCDD/PCDF Emissions from the Thermoselect Process

The PCDD/PCDF emissions are extremely low in the Thermoselect process. Recall from Table 6.5 that the Thermoselect gasifier only emits 0.193 ng/Nm^3 of total PCDD/PCDF in the gas stream, two orders of magnitude lower than the US guidelines for incinerators. The toxicity equivalent (TEQ), based on NATO standards, of the gasifier PCDD emissions is only 0.004 ng/Nm^3.

It is conceivable that the PCDD/PCDFs are just scrubbed out of the product gas, transfering the air emissions problem into a waste water problem. But the emissions data show the contrary. The untreated quench water has 0.00122 ng/liter of 2,3,7,8 TCDD[4], which is lower than the maximum contaminant level standard (0.03 ng/l) set in the EPA's safe drinking water act. Though the PCDD/F concentrations are low in the waste water, water could have been added merely to dilute the concentrations.

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[4] the most toxic of dioxin compounds (See Appendix 6A)

To eliminate these uncertainties, mass balances of the PCDD/F output from Thermoselect have been estimated using the same assumptions as in Table 6.7. Table 6.8 shows concentrations of the PCDD/F (ng/Nm^3) found by RWTUV and the total PCDD/F output (ng/hour) estimated from the mass balances. The waste water does contain more PCDD/PCDFs (seven times more) than the gas stream, but only a threefold higher level of toxicity equivalents. The mass flow estimates show that even if the PCDD/PCDFs were not scrubbed, the product gas would pass the U.S. guidelines and the most strict state regulations (3 ng/Nm^3 on TEQ basis).

Table 6.8 - PCDD/PCDF Emissions from the Thermoselect Gasifier[a]

	2,3,7,8 TCDD	Total	TEQ[d]
product gas			
(ng/Nm^3)[a]	0.0003	0.193	0.004
(ng/hour)[b]	0.9	579	12
waste water			
(ng/ltr)[a]	0.00122	0.2883	0.0026
(ng/hour)[c]	18.3	4325	39

Notes:
a) from Lutzke and Wehde (1993)
b) multiplying concentration by product gas flowrate of 3000 Nm3/hr assumed for 3 tonne/hr plant
c) multiplying concentration by waste water flowrate of 15000 ltrs/hr assumed for 3 tonne/hr plant
d) Toxicity equivalent using NATO standards

This mass flow estimate of PCDD/Fs heavily depends on the flowrate assumptions. If the water flowrate were actually an order of magnitude greater, the mass flow estimate of PCDD/Fs in the water would increase accordingly. The flowrate assumptions can be checked by examining the mass balances of chlorine estimated in Table 6.7. The total chlorine output is estimated to be almost 100 kg, which is equivalent to an enormous 3% chlorine content in the MSW feed (3 tonnes were processed). Since this is much higher than the average chlorine concentration in MSW (~0.5%), it is most likely that the water flowrate is overestimated rather than underestimated. Thus, the mass flows for PCDD/F shown in Table 6.8 are conservative estimates and very likely to be too high.

To put things into perspective, the Elk River, MN 1360 tonne/day RDF-fired plant is a new incinerator that began commercial operation in 1989. The facility uses state-of-the-art air pollution control equipment and has low emission levels of 0.793-7.27 ng total PCDDs/Nm^3. The Elk river plant emits only 0.0042 - 0.037 mg total PCDD/F per tonne raw MSW (Sullivan et al., 1993). Thermoselect produces 0.0016 mg total PCDD/Fs per tonne MSW, three to twenty times less than the Elk River incinerator (Table 6.9). The Thermoselect figures include the PCDD/Fs in the water stream, whereas the Elk River numbers do not include any PCDD/Fs that may be attached to the fly ash. On a toxicity equivalent basis, the incinerator emits 74-1030 ng/tonne MSW to the air, four to sixty times more than is emitted by Thermoselect (17 ng/tonne TEQs).

Table 6.9 - Comparison of PCDD/PCDF Emissions from Gasifier to Incinerators[a]

	Gasifier	Incinerators →		
	Thermoselect[b]	Polk County, MN	Elk River, MN	Indianapolis, IN
Facility Type	gasifier	modular burn	RDF burn	mass burn
MSW throughput (tonnes/hour)	3	3.4	57	83
Total PCDD/PCDF				
ng/Nm^3		269.7	7.27	19.5
mg/tonne MSW	0.0016	1.3	0.0384	0.0305
Toxicity Equivalent				
ng/Nm^3		9	0.195	0.0844
ng/tonne MSW	17	33700	1030	132

Notes:
a) Incinerator emissions data taken from Sullivan et al. (1993). Thermoselect data is calculated from Lutzke and Wehde (1993).
b) The incinerator emissions only account for the air emissions. Thermoselect emissions include PCDD in the waste water stream.

The Polk County, MN modular mass burn unit began operating in 1988 and combusts 80 tonnes/day, which is about the same size as the Thermoselect Italian plant. The Polk County plant uses an ESP to clean the flue gas. Measurements of the flue gas were taken before and after the ESP. The analysis showed that the flue gas contains 359 ng/Nm^3 before it was cleaned (Sullivan et al, 1993), which has been calculated to be 1.3 mg total PCDD/Fs per tonne MSW. This is three orders of magnitude more than Thermoselect emits. It should be noted that the ESP only has a 29% removal efficiency so the plant would not pass current federal guidelines for large facilities (>230 tons/day). The PCDD/PCDF emissions from this facility may be as high as they are because there were no PCDD/PCDF guidelines in the original permit.

Based on the mass flow estimates, the Thermoselect gasifier produces and emits fewer PCDD/PCDFs than the MSW combustors. Why the gasifier produces so much less is not clear. More research on PCDD/PCDF formation and destruction is needed.

The results of this analysis suggest that the Thermoselect gasifier is a potentially attractive technology for MSW disposal and fuel production from an emissions perspective. The same cannot be claimed for the indirectly heated gasifiers since they still require more emissions tests.

6.4 Discussions of Emissions Control Systems

Though the gasifier might not produce as much PCDD/F as incinerators and the product gas volumes are much lower than flue gas volumes, pollution control equipment will still be required to clean out all the heavy metals, particulate matter, organics, and acid gases in the product gas.

6.4.1 Effects of Air Pollution Control Equipment on Efficiencies

Regardless of what cleanup equipment is used, the process performances determined in Chapter 5 will not show significant variation. As mentioned in the models of Chapter 4, the scrubber is assumed to have negligible pressure losses. The wet scrubber may incur a higher pressure drop to remove fine PM, but this would only slightly alter the system's energy requirements.[5] If the system requires a better PM removal system, a fabric filter can be added downstream. This will cause an increased pressure drop (<0.05 bar), for which fans and blowers can compensate with negligible energy requirements compared to the other process energy inputs. ESPs also have modest energy requirements. The different types of scrubbers have similar energy requirements, since the largest energy inputs are associated with pumps to inject the sorbents and blowers to compensate the pressure drops. Thus, the efficiencies determined in the last chapter would not vary noticeably if different clean up options are used.

How would new clean up technologies affect the process performances? Recent developments have been made on hot gas clean up (HGCU) systems, which can clean gas streams at very high temperatures. Catalytic crackers that operate between 700-900 °C can break down tars and oils. Ceramic candle filters can remove alkali gases and PM at temperatures of 600 °C (Katofsky, 1993). These HGCU systems are extremely attractive for gas turbine (GT) applications: by keeping the gas hot instead of cooling it down for clean up, a large fraction of the sensible enthalpy of the gas is saved, which boosts the plant efficiency.

For fuel production, however, HGCU will not have too much effect on the system efficiency. Cleanup requirements for the fuel production process are more stringent than for GT applications. A scrubber would probably still be needed to remove acidic halide gases and the fine PM. Even if the HGCU could remove tars and oils, fine PM, and corrosive gases, the product gas would still need to be cooled to around 100°C to remove the H_2S.[6] Also, the various guard beds for the downstream reactors operate at moderate temperatures of 400 °C. Extremely high temperatures

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[5] Ordinarily, a 15 in H_2O (0.04 bar) pressure drop can be expected to reach a PM concentration of 48 mg/Nm3. To remove 99.9% of the >1 micron PM, a 0.25 bar drop would be required (Cheng, 1994).

[6] Hot gas desulfurization systems that can remove H_2S at high temperatures (~500°C) using limestone injection or metal oxides are under development for coal gasification power generation (IGCC) applications, but require extensive testing and development before being commercially ready (Coal Gasification Guidebook, 1993)

will cause these beds to sinter. Another reason that the product gas will be cooled to low temperatures even if HGCU were employed is that it is preferable to compress the gas (before entering the reformer) at lower temperatures to reduce the compression load.

Even if HGCU technology were developed sufficiently to eliminate the condensable gases in the product gas, the thermodynamic efficiency of the fuel production process would not increase significantly.[7] As an example, H_2 production with the BCL gasifier is used to make the comparison between the baseline case and

Table 6.10 - Energy Balances for Hydrogen Production from MSW using a Hot Gas Clean Up System with the BCL Gasifier

	Base case[a]	with HGCU[b]
Temperature (K)	1077	1077
Feed throughput (dry tonnes MSW/day)	1052	1052
Gasifier Input (MW)	182.8	182.8
H_2 output (MW)	110.9	110.9
Total Electricity Inputs (MW)	13.0	12.9
Waste Heat Suitable to Raise Steam (MW)	53.6	60.3
Electricity from Waste Heat (MW)[c]	14.8	16.7
Electricity from Purge Gases (MW)	0.0	0.0
Net Electricity Needs (MW)	-1.9	-3.8
Cooling Needs (MW)	1.0	0.2
Heating Needs (MW)	0.0	0.0
Energy Ratio	0.582	0.582
Thermal Efficiency[d]	0.599	0.617

Notes:

a) taken from BCL H_2 case in Chapter 5

b) Assumes that the hot gas clean up removes all tars and oils so the gas can be cooled to low temperatures (120°C) before being quenched.

c) Using a steam rankine cycle with a thermal efficiency of 27.7%

d) Both of these are net exporters of electricity so their TEs are greater than their ERs

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[7] The HGCU should crack the tars and oils so that the product gas energy content increases. But for this analysis, it is assumed that the product gas does not increase in energy content. Since the tars/oils content is small in the BCL and MTCI cases (< 1% of the feed carbon), this assumption has little effect on the outcome.

the case with HGCU.[8] The only difference between the two systems is that for the case with HGCU, the product gas is cooled with heat exchangers down to 120°C before being flooded by the scrubber, instead of 400 °C as it is in the base case. Table 6.10 summarizes the results. The energy ratio remains the same for the HGCU case at 0.582. The HGCU case, however, has more waste heat and is therefore able to export two more MWe than the base case system. The increase in the TE for the HGCU case is only 3%.

Though various pollution control systems can be employed to clean the product gas, none seem to have notable effects on the thermal efficiency. Using the same assumptions of chapter 5, only slight variations on the efficiencies determined by the reference models should be expected if the clean up systems are changed. The cost of the process, however, depends on the type of clean up used. This issue is examined in chapter 7.

6.4.2 Air Pollution Control (APC) Equipment

In the fuels production systems, it is important for the product gas to be cleaned adequately. Various combinations of pollution control technology can be used to clean out tars/oils, heavy metals, particulate matter (PM), organics, and halide gases (HCl and HF). Appendix 6B briefly describes the various air pollution control (APC) equipment.

One variation to the base case is to use a spray dry scrubber/fabric filter (SD/FF) system, which is considered to be state-of-the-art APC equipment for US incinerators, in place of the wet scrubber. The two types of scrubbers are comparable cleaning devices, but the spray dry scrubber avoids waste water treatment. In terms of acid gas removal efficiencies, the two are comparable, with >95% HCl removal and

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[8] The BCL H_2 case was chosen because it is the simplest configuration to manipulate. Similar results could be expected for the other cases.

70-90% SO_2 removal (WTE Permit, 1992). The SD/FF has a higher PM removal efficiency, mainly because the fabric filter is a better PM removal device than a wet scrubber.

A disadvantage to the SD/FF is that tars and oils are not removed as effectively. Also, the spray dry scrubber does not cool the gas as easily as the wet scrubber so it cannot ensure the removal of some pollutants, like mercury. With this in mind, the wet scrubber seems to be a simpler device that ensures easy removal of tars and oils from the product gas while also scrubbing out halide acid gases, metals, and particulate matter. Nevertheless, the SD/FF should be considered as a possible cleanup system for fuels production.

Other options are also possible. A fabric filter could be added downstream of the wet scrubber if better PM removal is required. Since the fabric filter can be fouled by moisture, the gas must not be saturated, which may require a condenser for the product gas. Should the fabric filter be too difficult to use in conjunction with the wet scrubber, an ESP could be used instead.

6.4.3 Sulfur Removal

Unlike biomass, MSW will generally contain non-negligible levels of sulfur. The source of sulfur is primarily rubber products, but various other materials also contain sulfur. For fuel production schemes, it is crucial to remove sulfur compounds because many of the catalysts in the downstream fuel processing system (the reformer, MeOH synthesis reactor, and the fuel cell itself) are poisoned by minute amounts (0.5 ppm). Almost all of the sulfur will be in the form of H_2S in the product gas. Other sulfur compounds that will be found are carbonyl sulfide (COS), carbon disulfide (CS_2) and mercaptans (RCH_2SH).

For biomass feeds, it is assumed that the sulfur content is low enough (0.01-0.1% by mass) that simple metal oxide (e.g. - ZnO) guard beds suffice as the sulfur removal system. Metal oxide or activated carbon beds can completely absorb H_2S from a gas stream up to concentrations of about 75 ppm. Any more H_2S in the stream will force the guard beds to be changed too often (Fenner, 1994). When over 100 kg/day of sulfur has to be removed, it is generally believed that simple scavenging will not be adequate.

As seen in Chapter 2, the average sulfur concentrations in MSW range from 0.1% - 0.5% by mass (wet basis). Assuming that MSW is 0.5% sulfur, a gasifier throughput rate of 1400 tonnes/day would require the removal of 7 tonnes sulfur per day. Clearly, simple guard beds do not suffice, and bulk acid gas removal systems are required. The metal oxide beds are still used in the system to remove the trace amounts of sulfur compounds that are not recovered by the bulk sulfur removal systems.

There is much experience with acid gas clean up in the natural gas, petroleum refining, and chemical processing industry. The Lo Cat liquid oxidation process uses an iron oxide solution to directly recover elemental sulfur from H_2S. This process is assumed to be the sulfur removal system used in the base case analyses, but there are many other acid gas washing processes that could be used. Appendix 6C describes various sulfur removal processes.

The most common system used for sulfur removal is the amine wash system combined with a Claus unit. The amine wash system removes the H_2S and the Claus unit converts the H_2S into raw sulfur. To convert more of the H_2S into sulfur, a SCOT process can be added. The main reason for not choosing the amine/Claus/SCOT system for the present analysis is that it is more suited for large scale facilities that produce over 30 tonnes sulfur/day to be economical. Liquid oxidation units are suited for smaller scales -- less than 25 tonnes S/day (Leppin, 1994). This is about the right capacity for the MSW-to-fuel facility, which produces about 7 tonnes sulfur/day. Other reasons for selecting the liquid oxidation system for the fuel production application are that it is more selective for H_2S and does not need to operate at high pressures.

Since sulfur removal systems are very costly, the costs of fuel production depend on what removal system is being used. The costs of these sulfur removal processes and their effects on the system costs are examined in the next chapter.

6.5 Concluding Remarks

Gasifiers produce significantly lower volumes of gas than the incinerators, so the contaminants are more concentrated in the product gas. In theory, the greater concentration of pollutants will be easier to clean out. Also, the lower volumes of gas require less pollution control equipment, which reduces costs. The air emissions from the fuel production system must be clean because the downstream reactor catalysts have tolerances more stringent than air emission regulations.

There are limited emissions data from gasifiers against which these hypotheses can be checked. Both BCL and MTCI have not yet run extensive emission tests. But the measurements from Thermoselect gasifier suggest that it indeed is a cleaner process than incineration. Analyses of the PCDD/PCDF output show that it produces far less PCDD/PCDFs than modern incinerators. Though emissions from gasifiers must be investigated further and more basic research of PCDD/F formation is required to make conclusive arguments, preliminary analyses have shown promising emissions characteristics of gasification.

Discussion of the emissions control equipment and sulfur recovery systems reveal that the performances of base case systems would vary only marginally if different gas clean up systems are used. The results of the analysis in this chapter are used in Chapter 7, when the economics of fuel production from MSW is evaluated.

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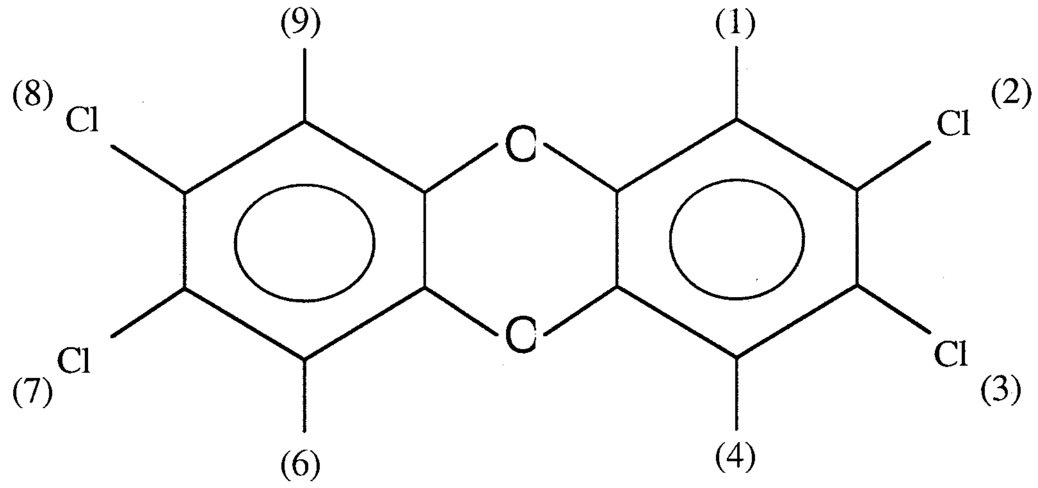
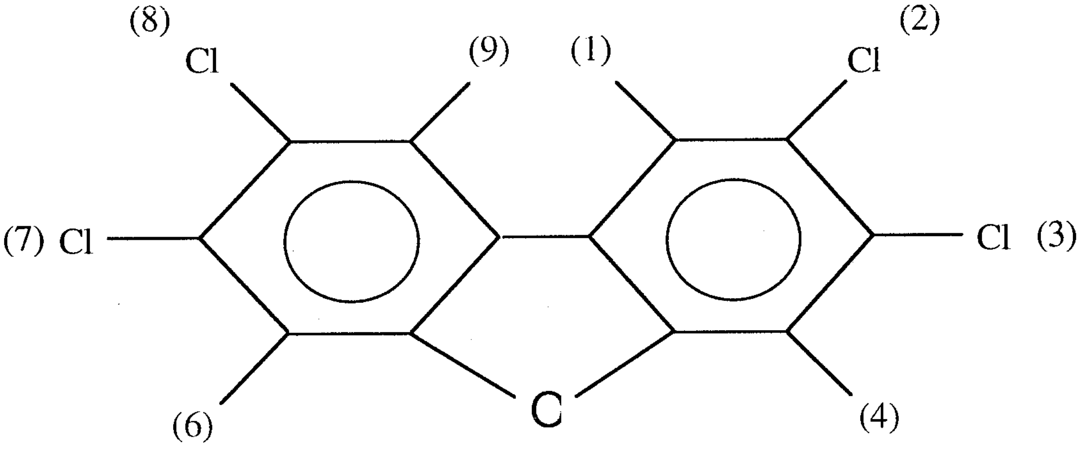


2,3,7,8 Tetrachlorodibenzo-para-dioxin (2,3,7,8 TCDD)



2,3,7,8 Tetrachlorodibenzofuran (2,3,7,8 TCDF)

Figure 6.1 - Chemical structure of dioxins and furans

Appendix 6A PCDD/PCDF

Since there has been considerable public attention given to polychlorinated dibenzo-p-dioxin (PCDD) and polychlorinated dibenzofuran (PCDF), these compounds are discussed here in greater detail. PCDD/Fs' health effects on humans are still not completely understood. They were considered to be one of the most toxic substances known to humans in the late 1970s, when extremely low doses to rats led to tumor formation (Kociba, 1978). Many reassessments made in recent years have supported earlier findings, although others have contradicted them. A consensus is growing that PCDD/F is dangerous as a carcinogen and has even more harmful non-carcinogenic effects, e.g. - harming reproductive systems (Hileman, 1993). To give an idea of how dangerous the EPA believes PCDD/F is, the safe drinking water act restricts PCDD/F concentrations to be less than one millionth the concentration allowed for chloroform. Regardless of the debate over its dangers, PCDD/F is a public health concern, and its production from MSW disposal must be addressed.

PCDD/PCDFs are chemically classified as halogenated aromatic hydrocarbons. The chemical structure of PCDD and PCDF is shown in figure 6.1. PCDDs consist of two benzene rings connected by two oxygen bridges. PCDFs have one oxygen bridge. These compounds have 1 to 8 chlorine atoms. PCDD/PCDFs with a given number of chlorine atoms are homologues of the PCDD or PCDF family.[9] The PCDD family has 75 different positional congeners and the PCDF family has 135 positional congeners.

The most toxic congener of both families is the 2,3,7,8-tetrachlorodibenzo-p-dioxin (2,3,7,8 TCDD). It is the most potent compound evaluated to date for carcinogenicity by the EPA. There are other compounds more toxic on an acute basis. Still 2,3,7,8 TCDD is considered three orders of magnitude more toxic than strychnine (WTE Permit, 1992). Since other congeners are not as toxic as 2,3,7,8 TCDD, there

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[9] i.e. - there are eight homologues of dioxins and furans (mono through octa).

is something known as Toxicity Equivalency Factors, which ranks other PCDD/PCDF congeners in comparison to 2,3,7,8 TCDD (Table 6A.1).

Table 6A.1 - NATO Toxicity Equivalency Factors (TEF) for Dioxin and Furan Congeners (adopted by the US EPA)

Dioxins		Furans	
Congener	TEF	Congener	TEF
2,3,7,8 TCDD	1.0	2,3,7,8 TCDF	0.1
other TCDDs	0	other TCDFs	0
1,2,3,7,8 PeCDD	0.5	1,2,3,7,8 PeCDF	0.05
other PeCDDs	0	2,3,4,7,8 PeCDF	0.5
		other PeCDDs	0
1,2,3,4,7,8 HxCDD	0.1	1,2,3,4,7,8 HxCDF	0.1
other HxCDDs	0	other HxCDFs	0
1,2,3,4,6,7,8 HpCDD	0.01	1,2,3,4,6,7,8 HpCDF	0.01
other HpCDDs	0	other HpCDFs	0
OCDD	0.001	OCDF	0.001

from WTE Permitting Handbook (1992)

Four pathways have been proposed for PCDD/F formation in MSW incinerators. First, PCDD/Fs already present in the MSW feed may pass through the combustor without being destroyed. This proposed pathway is no longer accepted as a source of PCDD/F from combustors. Most combustors operate around 1000°C, while PCDD/Fs are destroyed in two seconds at 870°C (Schoner, 1992). Various researchers have measured more PCDD/Fs in the incinerator flue gases than in the feed, implying that other pathways must exist (WTE Permit, 1992; Barton et al., 1990).

The second proposed pathway involve the formation of PCDD/Fs in the combustor from precursors (e.g., polychlorinated biphenyls) and chlorine donors, which can occur through relatively simple reactions (US EPA, 1987a). These precursors are more stable at high temperatures than PCDD/F. Nevertheless, 99.9% destruction of

PCBs will occur at 815°C with a residence time of one second (Clarke, 1988). Thus, good combustion practice should destroy these precursors so that no PCDD/F would form.

The third proposed pathway involves PCDD/F formation in the reactor from nonprecursor organics (e.g., aliphatics) and a chlorine donor. According to this theory, PCDD/F forms from its building blocks. The simplest mechanisms involve species similar in structure to PCDD/F. Various intermediate chemical reactions could explain precursor formation which would then lead to PCDD/F formation. Many scientists, however, believe that the formation of PCDD/F from long chain hydrocarbons is unlikely due to the slow reaction mechanisms involved and unfavorable equilibrium conditions (Helble, 1993).

The fourth pathway suggests that the PCDD/Fs are catalytically formed downstream of the reactor on fly ash particles. This theory of fly ash catalyzing PCDD/F formation is gaining widespread acceptance and is supported by ongoing research (Gullett et al., 1993). Experiments simulating combustor flue gases have shown that PCDD/Fs form on fly ash from precursors at temperatures between 200 °C and 400 °C, peaking around 300 °C (WTE Permit., 1992). Other studies have shown that nonprecursor organics and chlorine donors in the presence of fly ash also form PCDD/F in the same temperature window. These experiments suggest, however, that extremely long residence times (2 hours) are required (Vogg et al., 1986). PCDD/F precursors in the presence of the fly ash catalyst will have reaction rates up to 100 times faster than nonprecursors. This pathway seems consistent with empirical evidence that PCDD/F levels were high when an electrostatic precipitator was operated above 240 °C but that PCDD/F levels greatly decrease below 240 °C (Brna and Kilgroe, 1990).

Copper chloride ($CuCl_2$) is believed to be the primary catalyst in the fly ash that promotes PCDD/F formation (Acharya et al., 1991). The mechanism proposed for Cl_2 formation is known as the Deacon process:

$$2CuCl_2 + \frac{1}{2}O_2 \leftrightarrow Cu_2OCl_2 + Cl_2$$

$$Cu_2OCl_2 + 2HCl \leftrightarrow 2CuCl_2 + H_2O$$

This mechanism helps explain experiments where decreasing the HCl or O_2 levels reduces the amount of PCDD/F formed. The $CuCl_2$ is believed both to promote Cl_2 formation and act as the catalyst to form the dual ring structure of PCDD/F out of monoaromatics (Born et al, 1993). Though this fourth pathway of fly ash catalysis in the flue gas is now getting much attention, PCDD/F formation is not fully understood.

Appendix 6B: Brief Descriptions of Air Pollution Control (APC) Equipment

Feed preparation

Though pre-processing the feed is not considered a pollution control because it is not downstream of the reactor, it is included here because it is an effective measure to control emissions. Pre-processing the feed separates and removes many of the non-combustible materials in the feed. This reduces the ash content of the feed which reduces the particulate matter (PM) that gets emitted. Also, pre-processing can help reduce volatile metal emissions, which can include source separation measures, i.e. - battery recycling programs help reduce the mercury concentration in MSW. As seen in Table 6.5, the much lower mercury emissions for the Thermoselect gasifier were probably the result of a source separation program.

Cyclones

A cyclone is a simple device that removes particulate matter (PM) by forcing the gas stream to flow in a spiral pattern. Due to centrifugal force, the larger particles move outwards and impact the wall of the cyclone. These particles slide to the bottom of the cyclone, while the gas continues its spiral path out the top. Cyclones have no moving parts and thus can withstand high temperatures, pressures, and enormous particle loads. They are very inexpensive. The tradeoff against the cyclone's simplicity is that it cannot remove very small particulate matter (PM) efficiently. Typically, cyclones achieve 90% removal efficiencies for particles larger than 15-20 microns.[10] The best cyclones available today can remove up to 98% of the PM greater than 5 microns in diameter (Cooper and Alley, 1986). Improved performances are accomplished at the cost of a greater pressure drop because the collection efficiency rises by increasing the number of spirals the gas stream has to make, which can be

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[10] For particulate removal devices, the overall efficiency is defined as the total mass collected as a fraction of the total mass entering the first device.

accomplished by increasing the length of the unit, reducing the inlet duct width, etc. Cyclones are mainly used as a bulk PM removal device. Other types of air pollution control (APC) equipment are needed to clean out very fine PM. Oftentimes, two or more cyclones are placed in series. The first would be a conventional cyclone that is better suited for higher throughputs. The second would be a cyclone that removes the finer particles more efficiently.

Electrostatic Precipitators (ESP)

These devices remove particulate matter (PM) by first ionizing the gas stream flowing between electrodes. PM in the gas are charged by the electric field and then migrate to oppositely charged plates. The dust stays on the plates, while the gas travels through the device. Occasional rapping will knock particles from the plates into a hopper below. ESPs can handle large volume flows and low pressures. They have very high efficiencies; 99+% is possible for submicron particles (Flagan and Seinfeld, 1988). They are often used in conjunction with scrubbers to remove pollutants other than PM, like acid gases. An advantage of ESPs is that they can collect both dry and wet particles. Their drawbacks are that they are expensive and cannot handle wide variations in operating conditions.

Fabric Filters (FF)

Baghouse (fabric) filters are the most efficient dust collection systems for fine submicron particulate matter (PM). The PM accumulates on the cloth filter as the gas passes through. Fabric filters are shaped like long bags, thus the name baghouse filters. Dirty gas flows from the outside into the baghouses and up through the top. The PM is therefore collected on the outside of these tubular filters. Many filters usually operate in parallel. Shaking them or reversing the gas flows will knock off the dust to be collected below. Fabric filters have efficiencies greater than 99% for submicron particles (Flagan and Seinfeld, 1988). They are often used in conjunction with semi-dry scrubbers to clean the flue gas of most MSW incinerators. There are a few negative aspects to fabric filters. Besides being very expensive, the FFs cannot tolerate strong acids and alkalis, high temperatures, and wet environments.

Scrubbers

Scrubbers are used to wash out gaseous pollutants like acid gases in the gas stream. Some sort of reagent is sprayed into the gas stream to react with and wash out the corrosive gases. Scrubbers also cool the gas stream, forcing volatilized metals and organics to condense. There are three types of scrubbers: wet, dry and semi-dry.

Wet scrubbers have been used since the 1970s for multipollutant control, especially in Japan and Europe (US EPA, 1987b). Wet scrubbers can remove PM as well as gaseous pollutants. They operate on the principles of impaction, intercepting particles with liquid droplets. The droplets are separated from the gas by dropping out of the stream or hitting a wall. The liquid is sprayed into the gas stream as a fine mist, optimized for particle contact. Depending on the scrubber configuration, the spray may be counterflow or perpendicular to the dirty gas. Venturi scrubbers inject a liquid spray at the throat of a venturi duct through which the dirty gas passes through. This increases the velocity difference between the dirty gas and liquid spray, thereby allowing venturi scrubbers to achieve removal efficiencies of 99% for >0.5 micron sized PM (Flagan and Seinfeld, 1988). Greater pressure drops are incurred for higher efficiencies. Wet scrubbers can absorb acid gases, like HCl, HF, and HCN. Ammonia (NH3), which is often in the product gas stream, will also be removed by the scrubber. For simple particulate removal, only water is sprayed into the gas stream. To efficiently remove acid gases and prevent corrosion of the scrubber, an alkali solution, like sodium hydroxide or soda ash, is used as the liquid spray.

As indicated by the name, dry scrubbers inject dry alkaline sorbents as a powder into the flue gas stream. The lime or sodium bicarbonate powder reacts with the acid gases and is collected along with PM and other condensed matter by either an ESP or FF. The advantage of dry scrubbers is that no liquid waste stream is generated.

Spray dry (or semi-dry) scrubbers in combination with a fabric filter (SD/FF) are considered the state of the art in air pollution control technology for incinerators in the US. The spray dry scrubber injects the sorbent as an atomized liquid slurry, but any water present evaporates in the hot flue gas so only dry solid particles are collected. This takes advantage of the rapid absorption of acid gases by liquid spray particles, but still produces no waste water stream. Like the other scrubbers, this cools the gas stream so that volatiles tend to condense on the particulates. The solid sorbent is collected with the particulate matter by a filter or ESP.

Appendix 6C: Sulfur Removal Processes

Brief descriptions of common acid gas removal systems are given here, but the reader is referred to the reference book by Kohl (1985) for details on the various gas removal processes.

In general, acid gases can be removed by absorption, adsorption, and chemical conversion. There are also some newer techniques like membrane separation and cryogenic separation that are not yet as commonly used as the others.

Physical or chemical absorption into a liquid solvent is the dominant technology used today for acid gas removal. Physical solvent processes use an organic solvent that absorbs the acid gases at low temperatures or high pressures or both. Chemical solvent systems are better suited for low and medium pressure environments, but still require sufficiently low temperatures and high pressures to absorb the acid gases. Liquid absorption processes operate on a counterflow basis, with the gas stream flowing upwards through the absorber tower contacting the solvent flowing downwards. The solvent rich with acid gases is pumped to a flash tank and heat exchanger. The high temperatures and low pressures either strip the gases out of solution or reverse the absorber's chemical reactions. The acid gases are released, and the solvent is regenerated and recycled to the absorber.

The workhorse for the industry has been the amine (R_3N) chemical wash in combination with a Claus plant. The amine washing system scrubs out H_2S and CO_2 from the gas stream. The generic reaction to take out H_2S is

$$H_2S + R_3N \rightarrow R_3N.H_2S \text{ (aqueous)}$$

Alkanolamines have been used as solvents for H_2S and CO_2 removal since 1930 (Kohl, 1985). Monethanolamine (MEA) and diethanolamine (DEA) are the most common because they are effective and low cost solvents. MEA must be used in low concentrations because of its corrosive properties. DEA is not as strong a base as MEA, but can be used in higher concentrations. It also reacts slightly with carbonyl sulfide and carbon disulfide. Though amines are chemical solvents, they usually operate at fairly high pressures. Below 200 psia, it becomes increasingly difficult to meet pipeline gas restrictions of 4 ppm H_2S (Echterhoff and Quinlan, 1993). To regenerate the amine containing H_2S and CO_2, the pressure is reduced to 75 psia and the temperature is raised to 93°C. Other commonly used amine solvents include DGA (diethylene glycolamine), MDEA (methyl diethanolamine), and DIPA (diisopropanolamine). MDEA and DIPA are often used for selective acid gas removal because they react much faster with H_2S than with CO_2. Other chemical solvent processes include the Benfield potassium carbonate process and the Giammarco Vetrocoke arsenic trioxide processes.

Physical solvents must operate at more extreme temperatures and pressures. The Selexol process, which is modeled for CO_2 removal in methanol production, uses a dimethyl ether of polyethylene glycol. It is highly selective for sulfur compounds over CO_2, so is often used when selective acid gas removal is required. The Selexol process is appropriate to be modeled for CO_2 removal because the upgraded gas stream is already compressed to high pressures, and there is abundant industrial experience with selexol solvent. The Rectisol process is another physical solvent process that uses a methanol solvent which has extremely high sulfur compound removal efficiencies while letting CO_2 slip by. The process needs to operate around -40°C to be effective. Fluor solvent, an anhydrous propylene carbonate, is good for combined CO_2 and H_2S removal when their concentrations are very high.

The Claus process converts a concentrated H_2S stream into elemental sulfur. When the amine solvent is regenerated, it releases H_2S into a concentrated acid gas stream. This stream is fed to a Claus plant, which operates in two stages, the thermal and catalytic stage. The key reactions are

$$H_2S + \frac{3}{2}O_2 \rightarrow H_2O + SO_2 \qquad \Delta H = -520 \text{ MJ/kmol}$$

$$2H_2S + SO_2 \leftrightarrow 2H_2O + 3S(g) \qquad \Delta H = -88 \text{ MJ/kmol}$$

Generally, a third of the H_2S is burned to SO_2 in the thermal stage at atmospheric pressure. Within the furnace, 60% of the H_2S is converted to sulfur gas through the second reaction. The sulfur is condensed out by cooling the gases, and the flue gases are passed to the catalytic stage, where an aluminum oxide catalyst recovers more sulfur. To increase the conversion rates, there are usually a set of catalytic reactors, operating at progressively lower temperatures, with condensing and reheating steps in between each pair. The last reactor usually operates around 200 °C.

The tail gas will have some sulfur compounds remaining. A standard Claus plant will usually have 97% H_2S removal efficiencies. For higher removal efficiencies, the tail gas from the Claus plant can be recycled. The most common tail gas recycling unit is the Shell Claus Offgas Treating (SCOT) process. This can increase the overall sulfur recovery to over 99.9% for large scale operations (Echterhoff and Quinlan, 1993).

The SCOT process has a reduction step and an amine step. In the reduction step, all the sulfur compounds are converted to H_2S through various hydrogenation and hydrolysis reactions. Usually the tail gas has enough hydrogen and steam to carry out these reactions. The amine section involves the selective absorption of H_2S. The H_2S is then recirculated to the Claus unit. Since the amine must be selective for H_2S, DIPA is often used instead of MEA or DEA as the amine solvent. This step operates at atmospheric pressures.

This full combination of amine wash/Claus/SCOT typically is best for large scale facilities that produce over 20 tonnes/day of sulfur. Normal capacities that employ SCOT units recover around 50 tonnes/day. Large sour natural gas facilities processing 100 MMSCFD or coal gasification plants processing 5000 dry tonnes/day have the economies of scale to make the Claus and SCOT plant worth building. Very few Claus plants are built for applications removing less than 5 tonnes/day. In fact, it is questionable whether there are any SCOT facilities that produce less than 5 tonnes/day Sulfur (Echterhoff and Quinlan, 1994). With this level of sulfur output, the amine/Claus/SCOT process becomes very expensive. Other removal systems would probably be more applicable.

For smaller scale operations, adding subdewpoint or selective catalytic reactors to the Claus plant will improve sulfur recovery. Amoco's Cold Bed Absorption (CBA) process is a tail gas cleanup process that extends the Claus process by using catalytic reactors that operate below the dew point of sulfur. The CBA is more appropriate for <30 tonnes S/day applications but will not be quite as efficient as the SCOT process. Sulfur recovery is between 97-99% (Tannehill et al., 1994).

The liquid oxidation (also known as liquid redox) sulfur recovery method is a newer technology that converts H_2S directly into sulfur. For applications that remove less than 25 tonnes S/day, liquid redox processes are increasingly being used to meet stringent sulfur recovery requirements (Leppin, 1994). The Lo-Cat process designed by ARI Technologies[11] is a popular liquid oxidation method and uses a dilute aqueous iron oxide solution. Iron ions oxidize the H_2S into elemental sulfur through the reaction

$$H_2S + 2Fe^{+3} \leftrightarrow S + 2H^+ + 2Fe^{+2}$$

The iron ions can be regenerated in the oxidizer vessel by air via

$$1/2O_2 + 2Fe^{+2} + 2H^+ \leftrightarrow 2Fe^{+3} + H_2O$$

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[11] now known as Wheelabrator Clean Air Systems, Inc.

Besides recovering sulfur in one stage, liquid redox is not as dependent on pressure as the amine solutions. It also has extremely high sulfur recovery rates of 99.8% or higher (Echterhoff and Quinlan, 1993). Another attractive feature is that it is able to handle highly variable sulfur contents in the gas stream, in contrast to the amine wash systems, which need fairly consistent sulfur feeds to operate effectively. This attractive property of liquid redox systems may be necessary for the MSW gasifier product gas which will have a fluctuating H_2S content. Also, the liquid redox processes are highly selective for the removal of H_2S over CO_2. Some of the more expensive solvents like MDEA or Selexol are selective, but not as much so as iron oxide solutions.

The main drawback to the liquid redox process is that it has a very low solution loading for H_2S. This means that more solution is required for a given amount of H_2S in comparison to any amine solution. This is why the liquid redox process makes sense for only small scale usage (< 20 tonnes S/day). A minor drawback is that the sulfur produced is not as high grade as sulfur from a Claus unit. Still, the Lo-Cat process is more conducive to small scales, selective for H_2S, and does not need to operate at high pressures.

Chapter 7: Economics of Fuel from MSW

In this chapter, base case costs of fuel production from MSW are estimated for the three gasifiers. The cost estimates are compared to the costs of producing fuel from biomass, natural gas, and coal feedstocks. A few parameters in the base case analysis are varied to examine their effects on the costs of production. These include the feedstock cost (tipping fee), the scale of the facility, and the amount of pollution control equipment required for the plant. The chapter concludes with an economic comparison between fuel cell vehicles (FCVs) that use MeOH and H_2 from MSW and conventional vehicles fueled with gasoline.

7.1 Base Case Cost Estimates

7.1.1 Underlying Assumptions

Based on previous calculations by Williams et al. (1994) for biomass, the total levelized costs of producing methanol and hydrogen from MSW have been estimated here for the six base case systems. These are long-term fuel production cost estimates for commercial systems (not demonstration plants). The base case analysis considers plants with 1050 dry tonnes per day (dtpd) capacities (see section 5.2). Assuming a 25% moisture content, this would correspond to a 1400 wet tonne/day MSW treatment facility. The capital costs of major equipment with an estimated accuracy of ±25% are developed from various sources including vendors, consultants, and other studies. This costing analysis estimates the costs of a completely operational plant, accounting for contingencies, start up fees, overhead, auxiliary equipment, etc. by including costing factors that have been used in other recent analyses (OPPA, 1990; Wyman et al., 1992; Williams et al., 1994).

For the base case analysis, the negative feed cost (tipping fee) is assumed to be fixed at $22/tonne, which is the average landfill fee in Los Angeles (Sanitation Department of LA, 1993). This corresponds to a raw feedstock cost of about -$2/GJ.

7.1.2 Costs of Producing Fuel from MSW

The base case cost estimates of MeOH production show that the BCL system is the most cost effective for MSW feeds, with a levelized cost of production (COP) of \$14.94/GJ (Table 7.1). The Thermoselect system is the most costly, at \$18.07/GJ, even though it has the best thermodynamic performance characteristics. The same trend is true for hydrogen production (Table 7.2), with the BCL system having the lowest production cost (\$10.16/GJ) and Thermoselect having the highest (\$13.26/GJ). Producing methanol is more costly than hydrogen for two reasons: 1) hydrogen production is more efficient than methanol production and 2) the combination of the methanol synthesis reactor and CO_2 removal system is far more costly than the PSA that separates hydrogen from the shift reactor exhaust gases. (Other equipment are common to H_2 and MeOH production.)

The main reason that the BCL system has low production costs is that it does not require external electricity inputs. Recall that one of the key findings in Chapter 5 is that the BCL systems export electricity, while the other configurations require external electricity inputs. The high fuel production costs for the Thermoselect cases are mainly due to the large external electricity requirements (double that of MTCI cases). The difference in the levelized external energy costs between the Thermoselect and BCL cases is over \$2/GJ, accounting for nearly 70% of the differences in the total production costs.

The low COP for the BCL process and high COP for Thermoselect are also the results of the BCL process being the least capital-intensive, while the Thermoselect system is the most. The BCL gasifier itself is the simplest of the designs considered, consisting of two simple vessels. The MTCI gasifier includes a pulse combustor heat exchanger. The Thermoselect process operates at extremely high temperatures and requires a slag removal system. For fuel production, the BCL gasifier only makes up

Table 7.1 - Baseline production cost estimates (in 1991$) for methanol from MSW

FEEDSTOCK	MSW		
PROCESS	BCL	MTCI	ThS
Input			
dry tonnes feed/day	1052	1052	1050
dry RDF tonnes/day	873	873	n.a.
GJ/hr	686	662	656
MeOH Output[a]			
kmol/hr	461	509	582
tonnes/day	354	391	447
GJ/hr	335	370	423
Annual Feed and Products[b]			
1000 tonnes dry feed	345.42	345.42	344.93
Feed Input (10^6 GJ/yr)	5.41	5.22	5.17
Product Output (10^6 GJ/yr)	2.64	2.91	3.33
Installed Equipment Costs (10^6 $)			
Feed Pre-processing[c]	46.17	46.17	0.00
Gasifier[d]	8.12	21.91	68.54
High Temperature gas cooling[e]	0.00	0.00	24.88
Oxygen plant[f]	0.00	0.00	19.60
Sulfur removal[g]	9.67	9.67	9.66
Reformer Feed Compressor[h]	8.69	10.41	9.76
Reformer[i]	12.89	15.45	0.00
Shift Reactor[j]	0.00	0.00	0.00
CO2 removal[k]	7.91	10.56	12.17
MeOH synthesis & purification[l]	21.05	22.48	24.56
Steam turbine plant[n]	22.69	16.94	15.81
Balance of Plant Equipment[o]	34.30	38.40	46.24
Subtotal	171.49	191.98	231.22
Contingencies[p]	34.30	38.40	46.24
Owner's fees, costs, profits[p]	17.15	19.20	23.12
Startup[p]	8.57	9.60	11.56
Total Capital Requirement	231.51	259.18	312.14
Working capital[p]	17.15	19.20	23.12
Land[q]	1.24	1.24	1.24
Total Capital Costs (10^6 $)	249.90	279.61	336.50
Variable Operating Costs (10^6 $/year)			
Feed[r]	-9.81	-9.16	-9.49
Catalysts and chemicals[s]	0.92	1.02	0.31
Energy[t]	-0.46	3.48	6.90
Subtotal	-9.34	-4.66	-2.27
Fixed Operating Costs (10^6 $/year)			
RDF O&M[u]	4.28	4.28	0.00
Labor[v]	0.70	0.70	0.70
Maintenance[w]	3.76	4.37	6.94
General Overhead	2.90	3.30	4.96
Direct Overhead	0.32	0.32	0.32
Subtotal	11.96	12.98	12.92
Total Operating Costs (10^6 $/year)	2.62	8.32	10.64
Levelized Costs ($/GJ of MeOH)			
Capital[x]	13.94	14.13	14.87
Labor & Maintenance	4.88	4.80	3.97
Energy	-0.17	1.19	2.07
Feed	-3.72	-3.14	-2.85
TOTAL COST OF PRODUCTION	14.94	16.98	18.07

n.a. not applicable

Table 7.2 - Baseline production cost estimates (in 1991$) for hydrogen from MSW

FEEDSTOCK	MSW		
PROCESS	BCL	MTCI	ThS
Input			
dry tonnes feed/day	1052	1052	1050
dry RDF tonnes/day	873	873	n.a.
GJ/hr	686	662	656
Hydrogen Output[a]			
kmol/hr	1397	1494	1724
million Nm^3/day	0.75	0.80	0.93
GJ/hr	399	427	493
Annual Feed and Products[b]			
1000 tonnes dry feed	345.42	345.42	344.93
Feed Input (10^6 GJ/yr)	5.41	5.22	5.17
Product Output (10^6 GJ/yr)	3.15	3.37	3.89
Installed Equipment Costs (10^6 $)			
Feed Pre-processing[c]	46.17	46.17	0.00
Gasifier[d]	8.12	21.91	68.54
High Temperature gas cooling[e]	0.00	0.00	24.88
Oxygen plant[f]	0.00	0.00	19.60
Sulfur removal[g]	9.67	9.67	9.66
Reformer Feed Compressor[h]	7.86	10.57	10.56
Reformer[i]	13.05	13.07	0.00
Shift Reactor[j]	2.95	3.40	2.26
PSA[m]	4.83	5.06	5.60
Hydrogen Compressor[h]	2.34	2.34	2.74
Steam turbine plant[n]	20.48	14.33	12.79
Balance of Plant Equipment[o]	28.87	31.63	39.15
Subtotal	144.34	158.16	195.77
Contingencies[p]	28.87	31.63	39.15
Owner's fees, costs, profits[p]	14.43	15.82	19.58
Startup[p]	7.22	7.91	9.79
Total Capital Requirement	194.85	213.51	264.29
Working capital[p]	14.43	15.82	19.58
Land[q]	1.24	1.24	1.24
Total Capital Costs (10^6 $)	210.52	230.56	285.11
Variable Operating Costs (10^6 $/year)			
Feed[r]	-9.81	-9.16	-9.49
Catalysts and chemicals[s]	0.92	1.02	0.31
Energy[t]	-0.74	3.06	7.57
Subtotal	-9.62	-5.08	-1.60
Fixed Operating Costs (10^6 $/year)			
RDF O&M[u]	4.28	4.28	0.00
Labor[v]	0.70	0.70	0.70
Maintenance[w]	2.94	3.36	5.87
General Overhead	2.37	2.64	4.27
Direct Overhead	0.32	0.32	0.32
Subtotal	10.62	11.30	11.16
Total Operating Costs (10^6 $/year)	1.00	6.23	9.56
Levelized Costs ($/GJ of H_2)			
Capital[x]	9.84	10.08	10.80
Labor & Maintenance	3.67	3.66	2.95
Energy	-0.24	0.91	1.95
Feed	-3.12	-2.72	-2.44
TOTAL COST OF PRODUCTION	10.16	11.93	13.26

n.a. not applicable

Notes to Table 7.1 and 7.2 (all costs in 1991$)

a) Based on thermodynamic analysis of Chapter 5.

b) Assumes a capacity factor of 0.9.

c) The feed pre-processing costs are scaled by dry MSW feed rates using a 0.7 power factor. Baseline costs are estimated from the SPSA (1993b), Mid-Connecticut (Mazzaccaro, 1993), W. Palm Beach (Herrmann, 1993) and NSP (Shurtz, 1993) Resource Recovery plants as $55 million for 1350 tonnes/day. The Thermoselect pre-processing system (coke oven) cost is wrapped into the gasifier cost.

d) The BCL and MTCI gasifier costs are scaled by 0.7 factor from other estimates by dry RDF feed capacities, BCL: $7.25 million for 740 dtpd (Breault and Morgan, 1992); MTCI: $3.78 million for 72.5 dtpd (MTCI, 1994). Being a modular design, the Thermoselect gasifier is scaled linearly by dry MSW feed capacities. It costs $90 million for 383 dtpd including the pyrolysis oven, waste water cleanup, hot gas cleaning, acid gas removal, oxygen production, auxiliaries, fees, startup, working capital, and contingencies (Runyon, 1993). Subtracting out all extra costs, the base cost is calculated to $25 million for 383 dtpd for the gasifier & coke-oven.

e) High temperature gas cooling is only necessary for the Thermoselect case because the product gas temperatures are too high for conventional HX's (Moore, 1995). The BCL and MTCI product gases can be cooled by water with carbon steel HX's (Moore, 1995), which are incorporated in the steam turbine costs. The high temperature gas cooler cost is based on the coal plant in the OPPA study, which estimated $157.1 million for 7982 dtpd (OPPA, 1989). Costs are scaled by 0.7 factor based on heat removal rate (215.6 MW in coal case). This includes the shift reactor for the MeOH production case. The cost of the additional shift reactor capacity required for hydrogen production is included in the shift reactor costs (Moore, 1994).

f) Oxygen is assumed to cost in millions of dollars 0.26* (tonnes O_2/day)$^{0.712}$. This is based on Brown's estimates of 95% pure oxygen and Klosek's cost scale for higher purity (Brown, 1987 and Klosek, 1986).

g) Sulfur removal assumes the LO-CAT direct sulfur conversion process is used. Costs are scaled by sulfur removal rate using a 0.7 factor. Base costs for 2 tonnes S/day are $4.11 million (Echterhoff and Quinlan, 1993).

h) Compressor costs assumed to be $900/kW (Williams et al, 1994).

i) The reformer costs include the bfw pumps, steam drum, fans, all internal HX's including those to cool the reformate to ambient temperatures, desulfurization and hydrogenation guard beds, piping, controls, instruments, analyzers, initial catalyst charge and water treatment equipment. Costs are based on an estimate from Air Products for a steam reformer with 750 MW of natural gas feed . Costs are scaled by heat removal rate using a 0.57 power factor. The natural gas case had a 767.8 MW heat duty (Moore, 1994).

j) Shift reactor costs are scaled by volume flow of H_2 and CO using a 0.65 scaling factor , based on 9.02 million for 8819 kmol/hour (Moore, 1994).

k) CO_2 removal assumes a SELEXOL process. Baseline estimates are $14.3 million for 810 kmol/hr of CO_2 removal. Costs are scaled by CO_2 removal rate using 0.7 power (Epps, 1991).

l) Methanol synthesis costs are based on the ICI process, including the make-up and recycle compressors and synthesis loop equipment. The estimate for a facility producing 2012 tonnes/day of MeOH is $66.25 million (Moore, 1994). Costs are scaled to this estimate by MeOH production rates and scaled by a 0.66 power factor (Mansfield, 1991).

notes to Tables 7.1 and 7.2 continued

m) PSA assumed to be a one pass system that uses 2 parallel sets of 10 beds to remove 3951 kmols/hour of H_2. The baseline estimate for this system is $10 million. Costs are scaled by H_2 production raised to the 0.7 power (Solomon, 1993).

n) Steam Turbine costs (in $/kW) calculated with the following algorithm from Consonni and Larson (1994): 3785 * $\text{Output(MW)}^{-0.374}$. This estimate includes the entire power production plant: turbine, generator, pumps, HRSG, all HX's, etc.

o) Balance of Plant (BOP) is assumed to be 25% of the sum of other hardware costs (Wyman et al., 1993 and Moore, 1994).

p) Multipliers of the installed hardware costs are used (Wyman et al., 1993). contingencies 20%; owner's fees, costs, and profits 10%; working capital 10%, startup 5%.

q) Based on Wyman's estimates for biomass, land costs are assumed to be $423*(\text{tpd})^{1.147}$ (Wyman et al., 1993)

r) Assuming a negative feed cost based on the average tip fee for LA, $22/tonne of raw MSW (Sanitation Dept. of LA, 1993)

s) Catalysts and chemicals are scaled linearly by production rate from biomass cases. $2.9 million for 1110 tonnes MeOH/day in the BCL and MTCI cases and $55 million for 790 tonnes/day in the high temperature cases. The Thermoselect case has no reformer so has smaller catalyst costs (Wyman et al., 1993).

t) Energy costs are taken as $0.05/kWh of electricity and $4/GJ for heat. The BCL case is a net electricity producer and sells electricity at $0.05/kWh.

u) RDF labor and maintenance is scaled linearly by production size and is based on the SPSA (SPSA, 1993a), Mid-Connecticut (Municipal WTE Tech. Assess., 1992), and W. Palm Beach plants (Arvan, 1993): $5.5 million for 1350 tonne dry MSW/day facility. This cost includes the labor, maintenance, direct and general overhead costs.

v) Labor costs are based on Wyman et al. (1993). This assumes that the costs are scaled by facility size (in million$): $889*(\text{dtpd})^{0.859}$. This does not include the labor required to produce RDF.

w) Based on Wyman et al. (1993), these figures do not include the RDF processing plant: Maintenance is 3% of the installed hardware costs (not including the RDF plant), general overhead is 65% of labor and maintenance, and direct overhead is 45% of labor.

x) Assumed a 15.1% capital charge rate, based on average financial parameters for major US corporations from 1984-1988 (9.91% real rate of return on equity, 6.2% real rate of return on debt, a 30% debt fraction, 44% corporate income tax), an insurance rate of 1.5% per year, and a 25 year plant life. For land and working capital, the annual capital charge rate is taken to be 9.91%, the corporate discount rate (Larson and Katofsky, 1992).

around 7% of the system capital costs. The MTCI gasifier constitutes over 14% of the capital costs. The Thermoselect gasifier and coking oven account for over 37% of the capital costs.

Though the gasifier cost for the Thermoselect case seems to be the highest, the estimates in Tables 7.1 and 7.2 include the coking oven which is essentially the feed preparation costs. The other two gasifiers should be analyzed in conjunction with their feed preparation (RDF processing) costs to make a consistent comparison. For the indirectly heated gasifier cases, pre-processing the MSW feed is the most costly unit operation of the entire process ($46 million), ranging from 30-40% of the total capital costs. The RDF production plant and BCL gasifier cost $54 million, which is 40% and 47% of the total capital costs for methanol and hydrogen production, respectively. The MTCI system requires $68 million for feed preparation and gasification (44% and 54% of the capital costs). The Thermoselect coking oven and gasification unit have a combined cost of $68 million also, so the feed preparation and gasifier costs do not account for the Thermoselect system's high capital requirement.

The major contributors to the high capital costs of the Thermoselect systems are the high temperature cooling equipment and oxygen plant (25-30% of the total capital cost). The Thermoselect system does not require a reformer, which is used in the configurations using indirectly heated gasifiers,[1] but the costs of the O_2 plant and high temperature cooler are more than double that of the reformer.

Aside from the RDF production or gasifier costs, the methanol synthesis reactor is one of the most costly pieces of equipment for methanol production. For the three cases, the costs ranged from $21-25 million, or 13-15% of the total capital costs. The steam turbine costs also represent a significant fraction of the capital costs, especially for the BCL cases which have the most excess heat to generate the power.

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[1] Though the reformer is a significant investment for natural gas processes, it represents only ~10% of the installed hardware costs in the BCL and MTCI cases.

Based on the base case cost analysis, the BCL gasifier is the most economic choice for MSW-to-fuel production systems. These conclusions, however, are made cautiously because only the Thermoselect gasifier has demonstrated that it has much lower emissions than current regulations allow. Since the emissions from these systems are important in determining their feasibility, the BCL and MTCI gasifiers must test all their output streams before more definitive conclusions can be made.

Another important consideration is that the Thermoselect gasifier is a commercialized process, whereas the indirectly heated gasifiers have gasified MSW (RDF) for only limited durations at small scales. Thus, the Thermoselect systems could be set up now, whereas the others still need further development before they will be ready for commercial operation.

7.1.3 The MTCI Systems with No Char Recycling

The base case estimates for the MTCI systems assume that unreacted char is burned in the pulse combustor. By not burning the char and using only clean gas as the pulse combustor fuel, the MTCI systems are potentially more attractive from an emissions standpoint. As discussed in Chapter 5, such configurations reduce the thermodynamic performances of the MTCI systems. As a result, the COPs for these MTCI configurations increase to \$17.5/GJ for MeOH and \$12.5/GJ for H_2 (Table 7.3). The COPs for these MTCI cases are still cheaper than the COPs of the Thermoselect cases, but by only a small margin (3% for MeOH; 6% for H_2). The Thermoselect case would be competitive with the MTCI cases with only a slight change in the costs, e.g., if the Thermoselect gasifier costs were reduced by 10%. The results from this cost estimate suggests that the two systems are competitors and should both be further developed.

Table 7.3 - MTCI System Costs without Recycling Char to the Pulse Combustor[a]

	Base case		No char recycling	
	Methanol	Hydrogen	Methanol	Hydrogen
MSW Input (GJ/hr)	662.13	662.13	662.13	662.13
Fuel output (GJ/hr)	369.59	427.10	355.28	405.55
Energy Ratio	0.56	0.65	0.54	0.61
Thermal Efficiency	0.49	0.58	0.47	0.55
Total Capital Costs (10^6 \$)	279.61	230.56	277.75	230.58
Total Operating Costs (10^6 \$/year)	8.32	6.23	8.08	6.03
Levelized Costs (\$/GJ)				
Capital	14.13	10.08	14.60	10.61
Labor and Maintenance	4.80	3.66	4.96	3.84
Energy	1.19	0.91	1.19	0.91
Feed	-3.14	-2.72	-3.27	-2.86
TOTAL COP:	16.98	11.93	17.48	12.50

Notes: a) based on the same assumptions used in Tables 7.1 and 7.2. Costs in 1991\$

7.2 Comparison to Fuel Production Costs Using Biomass, Natural Gas, and Coal

As discussed in Chapter 4 the fuel production system for biomass is essentially the same as for MSW. Other studies have also used the BCL and MTCI gasifiers to estimate the costs of producing methanol and hydrogen from biomass (Katofsky, 1993; Williams et al., 1994). This provides an opportunity to make a consistent comparison of MSW and biomass feeds for fuel production using these two gasifiers (Tables 7.4 and 7.5). The cost estimates for the biomass cases are taken from Williams et al. (1994)[2] assuming a delivered biomass cost of \$2.5/GJ, a projected cost that is considered to be typical for biomass in the period near 2010 (Williams et al., 1994).

Except for the BCL H_2 case, the levelized cost of fuel production from biomass is cheaper than from MSW. For MeOH production the biomass-fed cases have a levelized cost of production of under \$14/GJ, which is \$1.4/GJ cheaper than the BCL

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[2] The analysis by Williams et al. (1994) assumes a feed rate of 1650 dtpd but has been scaled to 873 dtpd using appropriate scaling factors (see notes to Tables 7.1 and 7.2) for this comparison.

Table 7.4 - Comparison of methanol production costs (in 1991$) from MSW and biomass feeds[a]

FEEDSTOCK	MSW		Biomass[b]	
PROCESS	BCL	MTCI	BCL	MTCI
Input				
dry tonnes feed/day[c]	873	873	873	873
GJ/hr	686	662	708	706
MeOH Output				
kmol/hr	461	509	591	597
tonnes/day	354	391	454	459
GJ/hr	335	370	429	434
Annual Feed and Products				
Feed Input (10^6 GJ/yr)	5.41	5.22	5.58	5.56
Product Output (10^6 GJ/yr)	2.64	2.91	3.38	3.42
Installed Equipment Costs (10^6 $)				
Feed Pre-processing	46.17	46.17	8.43	8.46
Gasifier	8.12	21.91	8.12	21.91
High Temperature gas cooling	0.00	0.00	0.00	0.00
Oxygen plant	0.00	0.00	0.00	0.00
Sulfur removal	9.67	9.67	0.00	0.00
Reformer Feed Compressor	8.69	10.41	6.29	8.43
Reformer	12.89	15.45	11.96	0.00
Shift Reactor	0.00	0.00	1.32	0.00
CO2 removal	7.91	10.56	9.18	9.85
MeOH synthesis & purification	21.05	22.48	24.80	24.97
Steam turbine plant	22.69	16.94	14.84	15.35
Balance of Plant Equipment	34.30	38.40	21.24	22.24
Subtotal	171.49	191.98	106.19	111.20
Contingencies	34.30	38.40	21.24	22.24
Owner's fees, costs, profits	17.15	19.20	10.62	11.12
Startup	8.57	9.60	5.31	5.56
Total Capital Requirement	231.51	259.18	143.36	150.12
Working capital	17.15	19.20	10.62	11.12
Land	1.24	1.24	1.00	1.00
Total Capital Costs (10^6 $)	249.90	279.61	154.97	162.24
Variable Operating Costs (10^6 $/year)				
Feed	-9.81	-9.16	13.95	13.90
Catalysts and chemicals	0.92	1.02	1.19	0.32
Energy	-0.46	3.48	1.53	2.36
Subtotal	-9.34	-4.66	16.66	16.59
Fixed Operating Costs (10^6 $/year)				
RDF O&M	4.28	4.28		
Labor	0.70	0.70	0.59	0.59
Maintenance	3.76	4.37	3.19	3.34
General Overhead	2.90	3.30	2.45	2.55
Direct Overhead	0.32	0.32	0.26	0.26
Subtotal	11.96	12.98	6.49	6.74
Total Operating Costs (10^6 $/year)	2.62	8.32	23.15	23.33
Levelized Costs ($/GJ of MeOH)				
Capital	13.94	14.13	6.74	6.99
Labor & Maintenance	4.88	4.80	2.27	2.07
Energy	-0.17	1.19	0.45	0.69
Feed	-3.72	-3.14	4.12	4.07
TOTAL COST OF PRODUCTION	14.94	16.98	13.59	13.81
Tipping fee ($/tonne) to match biomass case	30.0	44.2	n.a.	n.a.

n.a. not applicable

Table 7.5 - Comparison of hydrogen production costs (in 1991$) for MSW and biomass feeds[a]

FEEDSTOCK	MSW		Biomass[b]	
PROCESS	BCL	MTCI	BCL	MTCI
Input				
dry tonnes feed/day[c]	873	873	873	873
GJ/hr	686	662	708	706
Hydrogen Output				
kmol/hr	1397	1494	1813	1874
million Nm^3/day	0.75	0.80	0.97	1.01
GJ/hr	399	427	518	536
Annual Feed and Products				
Feed Input (10^6 GJ/yr)	5.41	5.22	5.58	5.56
Product Output (10^6 GJ/yr)	3.15	3.37	4.09	4.22
Installed Equipment Costs (10^6 $)				
Feed Pre-processing	46.17	46.17	8.43	8.46
Gasifier	8.12	21.91	8.12	21.91
High Temperature gas cooling	0.00	0.00	0.00	0.00
Oxygen plant	0.00	0.00	0.00	0.00
Sulfur removal	9.67	9.67	0.00	0.00
Reformer Feed Compressor	7.86	10.57	6.29	7.93
Reformer	13.05	13.07	12.78	0.00
Shift Reactor	2.95	3.40	3.31	3.37
PSA[d]	4.83	5.06	1.29	2.55
Hydrogen Compressor	2.34	2.34	12.77	12.75
Steam turbine plant	20.48	14.33	10.65	2.80
Balance of Plant Equipment	28.87	31.63	15.91	14.94
Subtotal	144.34	158.16	79.55	74.71
Contingencies	28.87	31.63	15.91	14.94
Owner's fees, costs, profits	14.43	15.82	7.96	7.47
Startup	7.22	7.91	3.98	3.74
Total Capital Requirement	194.85	213.51	107.40	100.86
Working capital	14.43	15.82	7.96	7.47
Land	1.24	1.24	1.00	1.00
Total Capital Costs (10^6 $)	210.52	230.56	116.35	109.33
Variable Operating Costs (10^6 $/year)				
Feed	-9.81	-9.16	13.95	13.90
Catalysts and chemicals	0.92	1.02	1.19	0.32
Energy	-0.74	3.06	4.43	7.24
Subtotal	-9.62	-5.08	19.57	21.46
Fixed Operating Costs (10^6 $/year)				
RDF O&M	4.28	4.28		
Labor	0.70	0.70	0.59	0.59
Maintenance	2.94	3.36	2.39	2.24
General Overhead	2.37	2.64	1.93	1.84
Direct Overhead	0.32	0.32	0.26	0.26
Subtotal	10.62	11.30	5.17	4.93
Total Operating Costs (10^6 $/year)	1.00	6.23	24.74	26.39
Levelized Costs ($/GJ of H_2)				
Capital	9.84	10.08	4.19	3.81
Labor & Maintenance	3.67	3.66	1.56	1.24
Energy	-0.24	0.91	1.09	1.71
Feed	-3.12	-2.72	3.41	3.29
TOTAL COST OF PRODUCTION	10.16	11.93	10.25	10.06
Tipping fee ($/tonne) to match biomass case	21.4	37.1	n.a.	n.a.

n.a. not applicable

Notes to tables 7.4 and 7.5

a) cost estimates based on the same assumptions used in Table 7.1 and 7.2

b) biomass figures based on estimates by Williams et al. (1994)

c) the dry feed to the gasifiers is kept constant for the two feeds. For the MSW cases, this is the RDF throughput rate. The throughput rate here is the same as the 1050 dry tonnes MSW/day used for the base cases of Tables 7.1 and 7.2.

d) the PSA in the biomass case uses the Gemini-9 system from Air Products, Inc. which removes CO_2 and H_2O in the first bed and produces a high purity H_2 gas in the second bed. Costs are scaled by hydrogen production using a 0.7 power factor based on an estimate of $30.82 million for 8474 kmol/hour (Moore, 1994). Wrapped into this cost is the recycle compressor which is assumed to cost $900/kW of capacity. The costs of the one pass PSA for the MSW case is described in Table 7.2. The PSA cost is singled out here because it is the only unit in the fuel production process that is different for the two feeds.

MSW case and over $3/GJ less than the MTCI MSW case. For hydrogen production the biomass cases have levelized COPs around $10/GJ. The COP for hydrogen from MSW using the MTCI gasifier is nearly $2/GJ more costly. The COP for the BCL MSW case is actually less than for the BCL biomass case for hydrogen production. The hydrogen production costs for the two feeds are much closer than the methanol production costs because the difference in efficiencies for the two feeds is smaller for hydrogen production than for methanol production (see Chapter 5).

The negative feed cost of MSW naturally makes it an attractive feed in comparison to biomass, but the MSW-fed systems require higher capital costs. The main contributor to the high capital costs is the feed pre-processing cost, as shown in the disaggregated cost estimates of Tables 7.4 and 7.5. In addition to the RDF plant, the MSW-fed system using the MTCI gasifier has the extra cost of a reformer. For the BCL cases, the RDF production plant is the only hardware unit that is different between the MSW and biomass-fed systems. Besides the higher capital costs, the MSW-fed processes are not as efficient as the biomass-fed processes, which in turn raises the COP for the MSW cases.

For the BCL H_2 system, the assumed negative feedstock cost for MSW (-$22/tonne) is sufficient to make the MSW case competitive with the biomass case. But for the other systems, the $22/tonne tipping fee does not compensate for the higher capital costs and lower energy efficiencies of the MSW-fed system to compete with biomass-fed systems. The bottom lines of Tables 7.4 and 7.5 show the tipping fee (negative feed cost) that is required for the MSW-fed system to compete with the biomass-fed process (using the same gasifier). As seen in the tables, the tipping fees required for MSW to be competitive with biomass for MeOH production are \$30/tonne and \$44/tonne for the BCL and MTCI cases, respectively. For hydrogen production, the breakeven tipping fees are lower, at \$21/tonne for BCL (which is lower than the base case tipping fee) and \$37/tonne for MTCI. As described below, these tipping fees are lower than the average landfill fees of some states.

A consistent comparison of MSW, biomass, coal, and natural gas feeds for fuels production is given in Tables 7.6 and 7.7. The cost estimates for the coal and natural gas cases are taken from Williams et al. (1994). Both of these systems are assumed to be much larger than the MSW and biomass base case systems. For natural gas, the assumed feed rate is 1.64 million Nm^3/day (typical of modern natural gas methanol facilities), so it has an output capacity five times greater than the MSW and biomass cases. For coal, the assumed feed rate is 5000 dry tonnes/day, so it has an output capacity roughly ten times greater than the MSW and biomass cases.

The feedstock prices assumed for natural gas and coal are \$3.8/GJ and \$1.4/GJ, respectively. These are the average prices projected for industrial users in the US in 2010 (EIA, 1995).

The costs of producing methanol from natural gas (\$9/GJ) and coal (\$9.9/GJ) are much lower than the COP from MSW (\$14.9/GJ). For hydrogen, the COPs from natural gas (\$6.8/GJ) and coal (\$8.6/GJ) are also much lower than the COP from MSW (\$10.2/GJ). To break even, the tipping fee in the MSW case would have to be significantly higher.

Table 7.6 - Comparison of methanol production costs (in 1991$) for different feeds

FEEDSTOCK	MSW	Biomass	N Gas	Coal
PROCESS	BCL	BCL	SMR	Shell
Input				
dry tonnes feed/day	873	873	1224	5000
GJ/hr	686	708	2700	6188
MeOH Output				
kmol/hr	461	591	2616	5530
tonnes/day	354	454	2012	4252
GJ/hr	335	429	1900	4017
Annual Feed and Products				
Feed Input (10^6 GJ/yr)	5.41	5.58	21.28	48.78
Product Output (10^6 GJ/yr)	2.64	3.38	14.98	31.67
Installed Equipment Costs (10^6 $)				
Feed Pre-processing	46.17	8.43	0.00	67.96
Gasifier	8.12	8.12	0.00	120.06
High Temperature gas cooling	0.00	0.00	0.00	113.27
Oxygen plant	0.00	0.00	0.00	95.42
Sulfur removal	9.67	0.00	0.00	36.26
Reformer Feed Compressor	8.69	6.29	0.00	0.00
Reformer	12.89	11.96	50.00	0.00
Shift Reactor	0.00	1.32	0.00	0.00
CO2 removal	7.91	9.18	0.00	59.50
MeOH synthesis & purification	21.05	24.80	66.24	108.56
Steam turbine plant	22.69	14.84	17.11	43.04
Balance of Plant Equipment	34.30	21.24	33.33	161.02
Subtotal	171.49	106.19	166.67	805.08
Contingencies	34.30	21.24	33.33	161.02
Owner's fees, costs, profits	17.15	10.62	16.67	80.51
Startup	8.57	5.31	8.33	40.25
Total Capital Requirement	231.51	143.36	**225.01**	**1086.86**
Working capital	17.15	10.62	16.67	80.51
Land	1.24	1.00	4.26	9.76
Total Capital Costs (10^6 $)	249.90	154.97	245.93	1177.12
Variable Operating Costs (10^6 $/year)				
Feed	-9.81	13.95	80.46	67.81
Catalysts and chemicals	0.92	1.19	2.58	10.87
Energy	-0.46	1.53	5.45	15.47
Subtotal	-9.34	16.66	88.48	94.15
Fixed Operating Costs (10^6 $/year)				
RDF O&M	4.28			
Labor	0.70	0.59	1.00	3.13
Maintenance	3.76	3.19	5.00	24.15
General Overhead	2.90	2.45	3.90	17.74
Direct Overhead	0.32	0.26	0.45	1.41
Subtotal	11.96	6.49	10.35	46.43
Total Operating Costs (10^6 $/year)	2.62	23.15	**98.83**	**140.58**
Levelized Costs ($/GJ of MeOH)				
Capital	13.94	6.74	2.41	5.46
Labor & Maintenance	4.88	2.27	0.86	1.81
Energy	-0.17	0.45	0.36	0.49
Feed	-3.72	4.12	5.37	2.14
TOTAL COST OF PRODUCTION	14.94	13.59	9.00	9.90

Notes: Based on same assumptions used in Table 7.1 and 7.4. Biomass, natural gas, and coal estimates based on Williams et al. (1994). Feedstock prices assumed to be $3.8/GJ for gas and $1.39/GJ for coal. SMR = steam methane reforming. The coal is gasified by the Shell gasifier.

Table 7.7 - Comparison of hydrogen production costs (in 1991$) for different feeds

FEEDSTOCK	MSW	Biomass	N Gas	Coal
PROCESS	BCL	BCL	SMR	Shell
Input				
dry tonnes feed/day	873	873	1224	5000
GJ/hr	686	708	2700	6188
Hydrogen Output				
kmol/hr	1397	1813	8473	16754
million Nm^3/day	0.75	0.97	4.56	9.01
GJ/hr	399	518	2422	4788
Annual Feed and Products				
Feed Input (10^6 GJ/yr)	5.41	5.58	21.28	48.78
Product Output (10^6 GJ/yr)	3.15	4.09	19.09	37.75
Installed Equipment Costs (10^6 $)				
Feed Pre-processing	46.17	8.43	0.00	67.96
Gasifier	8.12	8.12	0.00	120.06
High Temperature gas cooling	0.00	0.00	0.00	113.27
Oxygen plant	0.00	0.00	0.00	95.42
Sulfur removal	9.67	0.00	0.00	36.26
Reformer Feed Compressor	7.86	6.29	0.00	0.00
Reformer	13.05	12.78	43.91	0.00
Shift Reactor	2.95	3.31	9.02	7.28
PSA	4.83	1.29	37.93	67.82
Hydrogen Compressor	2.34	12.77	9.64	19.22
Steam turbine plant	20.48	10.65	11.09	20.48
Balance of Plant Equipment	28.87	15.91	27.90	136.94
Subtotal	144.34	79.55	139.49	684.70
Contingencies	28.87	15.91	27.90	136.94
Owner's fees, costs, profits	14.43	7.96	13.95	68.47
Startup	7.22	3.98	6.97	34.24
Total Capital Requirement	194.85	107.40	188.31	924.35
Working capital	14.43	7.96	13.95	68.47
Land	1.24	1.00	4.26	9.76
Total Capital Costs (10^6 $)	210.52	116.35	206.51	1002.58
Variable Operating Costs (10^6 $/year)				
Feed	-9.81	13.95	80.46	67.81
Catalysts and chemicals	0.92	1.19	2.58	10.87
Energy	-0.74	4.43	7.82	57.18
Subtotal	-9.62	19.57	90.85	135.86
Fixed Operating Costs (10^6 $/year)				
RDF O&M	4.28			
Labor	0.70	0.59	1.00	3.13
Maintenance	2.94	2.39	4.18	20.54
General Overhead	2.37	1.93	3.37	15.39
Direct Overhead	0.32	0.26	0.45	1.41
Subtotal	10.62	5.17	9.00	40.48
Total Operating Costs (10^6 $/year)	1.00	24.74	99.86	176.33
Levelized Costs ($/GJ of H_2)				
Capital	9.84	4.19	1.58	3.90
Labor & Maintenance	3.67	1.56	0.61	1.36
Energy	-0.24	1.09	0.41	1.51
Feed	-3.12	3.41	4.21	1.80
TOTAL COST OF PRODUCTION	10.16	10.25	6.81	8.57

Notes: Based on same assumptions used in Table 7.2 and 7.5. Biomass, natural gas, and coal estimates based on Williams et al. (1994). Feedstock prices assumed to be $3.8/GJ for gas and $1.39/GJ for coal. SMR = steam methane reforming. The coal is gasified by the Shell gasifier.

7.3 Effects of Varying Cost Parameters

7.3.1 The Feedstock Cost

This section examines the impact on cost of different *tipping fees*. The *tipping fee* is the cost that the MSW disposal facility charges its customers per tonne of raw refuse. Basically, the tipping fee reflects the capital and operating costs of the disposal facility. The base case cost analysis assumes that the feedstock costs are -$22/tonne of raw waste, which is the average tipping fee for the four landfills in Los Angeles county owned by the Sanitation District of Los Angeles (Pomona, Whittier, Calabasas, and Glendale).[3] The average tipping fee for all eleven landfills in the county is around $23.70/tonne (Gaw, 1993). Southern California actually has a lower tipping fee in comparison to northern California because there is much less precipitation in southern California; therefore, a less advanced leachate collection system is required. Also, since there are more canyons in southern California, the landfills can be deeper, taking up less land (Tellus, 1991). In 1990, the Western region of the US (CA, NE, OR, WA, ID) had an average tipping fee of $28.25/tonne which was about the same as the nation's average (NSWMA, 1991). This is higher than the central part of the country (Montana to Texas), which has average tipping fees of $12-$13/tonne. Still, it is dwarfed by the more populous Northeast, which has an average tipping fee of $71/tonne. Not surprisingly, the Northeast region has the fewest years of remaining landfill capacity (20 years maximum) and the highest percentage of landfills that have leachate collection systems (91%) (NSWMA, 1991). The Mercer County region of NJ (Princeton University is in Mercer County) has especially high costs for waste disposal because MSW is exported. Including taxes, commercial and residential wastes cost over $120/tonne to deliver to the Ewing transfer station (Bauer, 1994). From there, the MSW is delivered to the Geological

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[3] In 1993, the lowest of these four was $17.50/tonne and the highest was $26.60/tonne (Sanitation District of LA, 1993). Some of the privately owned landfills charge more. In Sun Valley, WMX Technologies, Inc. (formerly Waste Management Inc.) charges about $37/tonne (Gaw, 1993)

Reclamation Operation and Waste Systems (GROWS) landfill in Morrisville, Pennsylvania.

Tipping fees also apply to incinerators. The average tipping fee for the three incinerators in California is $60/tonne (Tellus, 1991), which is much higher than the landfill tipping fees in California. The large additional cost of disposing of MSW in an incinerator rather than a landfill explains in part why there are so few incinerators in California. (Air pollution concerns may be another reason.) In the Northeast, where much of the MSW is combusted,[4] incinerator tipping fees are competitive with landfill fees. The Hartford, CT plant, which sometimes accepts MSW from New York City has a tipping fee of $56/tonne (Pilgrim, 1993; Mazzaccaro, 1994). On average, New Yorkers pay about a $70/tonne fee to send their garbage to the Fresh Kills landfill on Staten Island (Wurster, 1994).

With increased environmental regulations, MSW disposal costs will certainly rise. A recent study examined the average disposal fee in California that would reflect all costs, including environmental externalities, and estimated that the average cost of disposal should be $55/tonne (Tellus, 1991).

Methanol Production Costs

Figure 7.1 shows the effect of the tipping fee on the cost of methanol production for the three cases examined. At a tipping fee of $30/tonne, the BCL case is competitive with a biomass-to-methanol system of the same scale using the BCL gasifier (assuming $2.5/GJ biomass feed prices). Using the MTCI system, MSW is economically preferred to biomass at tipping fees over $40/tonne. The MSW-fed Thermoselect system becomes competitive with a biomass system when the tipping fee rises to nearly $60/tonne. All three MSW-fed systems are competitive with

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[4] In New England, 44% of the disposed MSW was burned in 1992 (Glenn, 1992).

biomass-fed systems at tipping fees lower than the current tipping fees in the Northeast.

The MSW-fed systems become competitive with coal-fed systems (assuming a coal price of \$1.4/GJ) at slightly higher tipping fees. For BCL, the breakeven tipping fee is \$52/tonne. At \$71/tonne, the MTCI case competes with coal. At \$85/tonne, the Thermoselect system breaks even with the coal-fed system.

For the base case feed cost of \$3.8/GJ, the costs of producing MeOH from natural gas would be \$9/GJ. The tipping fees required for the MSW cases to become competitive are \$57/tonne for the BCL system; \$78/tonne for MTCI; and \$92/tonne for Thermoselect.

At higher tipping fees, the three cases can be competitive with MeOH production from natural gas even at lower gas prices. Assuming a natural gas feed cost of \$2.50/GJ (which is roughly the current US utility gas price),[5] the cost of producing MeOH is \$7.2/GJ (based on Williams et al., 1994). The BCL case is competitive with the natural gas system with a \$68/tonne tipping fee, which is still lower than the average in the Northeast. The MTCI case requires a tipping fee of \$91/tonne, and the Thermoselect case requires a tipping fee of \$106/tonne to compete with methanol produced from natural gas today.

Hydrogen Production Costs

Figure 7.2 shows how hydrogen production costs are affected by the tipping fee. In all three cases, the breakeven tipping fees for MSW to be competitive with biomass feeds are less than the Northeast average. For BCL, the \$21/tonne breakeven fee is lower than average landfill fee for L.A. (\$22/tonne).

For the base case coal price of \$1.4/GJ, the COP for hydrogen is \$8.57/GJ. The tipping fees required for the MSW cases to become competitive with the coal cases

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[5] The 1990 delivered natural gas price for utilities was \$2.39/GJ (EIA, 1994)

are lower than the average Northeast landfill fee: \$33/tonne for the BCL system; \$49/tonne for MTCI; and \$64/tonne for Thermoselect.

For a natural gas price of \$3.8/GJ, the cost of hydrogen production is \$6.81/GJ. Under this situation, both the BCL and MTCI systems would be economically favored at tipping fees well below the Northeast average (\$45.6/tonne and \$63/tonne, respectively), and the Thermoselect case becomes competitive with a tipping fee of only \$80/tonne.

Hydrogen produced from natural gas (at \$2.50/GJ) has a levelized production cost of \$5.39/GJ. The breakeven tipping fees for hydrogen production from MSW are \$56/tonne for the BCL case, \$75/tonne for MTCI and \$93/tonne for Thermoselect. Only the BCL system can compete at tipping fees lower than the Northeast average, but the MTCI case is also close.

Fuels vs. Electricity Generation

Figure 7.3 displays the effects of the tipping fee on the cost of electricity production using conventional waste-to-electricity plants and the fuel production costs using the BCL gasifier. As can be seen on the graph, the cost of electricity production from the incinerator is more sensitive to the tipping fee than the cost of fuel production. This is a result of the higher efficiency of fuel production in comparison to electricity production.

The costs of MeOH and H_2 production from MSW are competitive with conventional fuel production technologies at lower tipping fees than when incineration becomes economically competitive with new coal-fired utility power generation. As shown on Figure 7.3 the incinerator becomes competitive with coal-fed power generation plants when the tipping fee is nearly \$70/tonne. Methanol produced from MSW becomes competitive with methanol produced from natural gas (assuming a gas price of \$3.8/GJ) when the tipping fee is \$57/tonne. Hydrogen production from MSW shows favorable economics to hydrogen from natural gas at $46/tonne. Thus, for the assumed natural gas price a fuel production facility would be competitive as a waste disposal facility with conventional waste-to-electricity plants whose output must compete with coal-fired electric power.

Fuels from MSW vs. Gasoline

The fuel production costs can also be compared to conventional fuel prices. Compared to the projected reformulated gasoline production cost of $7.34/GJ (Ogden et al., 1994), methanol and hydrogen production from MSW is competitive on a $/GJ basis at tipping fees of $67/tonne and $42/tonne, respectively (Figure 7.3). Comparisons on a $/km basis depend on the vehicle technology under consideration as discussed in Section 7.4.

7.3.2 Scale Effects on Costs

Fuel production plants for the base case analyses have fuel input rates (1050 tonnes/day of dry MSW) that are comparable to those for biomass based fuel conversion facilities, as well as modern mass burn incinerators. Biomass plants cannot be too large because transportation costs become excessive. But such plants produce small amounts of methanol (400 tonnes/day) compared to state-of-the-art natural gas methanol facilities operating today (over 2000 tonnes/day production (Katofsky, 1993)). Moreover, Katofsky (1993) and Williams et al. (1994) suggest that fuel production plants based on use of coal feeds would be very large (5000 dry tonnes/day input) to take advantage of scale economies of the capital equipment. Coal feedstock delivery costs are comparatively insensitive to scale because of coal's higher volumetric energy density.

MSW transportation costs are sensitive to scale, but many crowded cities still haul their waste over long distances to find cheap disposal options. Infrastructures are already set up for hauling MSW from various regions to a given site, whether it be a transfer station, landfill, or incinerator. Many solid waste management authorities serve a group of counties and municipalities. If these management districts or an entire city delivered MSW to one site, the scale of the fuel production plant could be greater than 1050 dry tonnes/day. To be the same size as a large coal plant (e.g., 5000 dry tonnes/day), the facility would handle the MSW disposed by over 4 million people.[6] Most waste management authorities serve fewer than one million people, but a plant this size might be plausible for large cities. The Bradley landfill in Sun Valley (in LA county) receives 5-6 thousand raw tonnes/day. The largest landfill in the country, the Fresh Kills landfill in Staten Island, NY, takes in about 13 thousand raw tonnes/day on average and has accepted over 25 thousand tonnes in one day (Wurster, 1994).

Figures 7.4 and 7.5 show the scale effects on the costs of methanol and hydrogen production for all three cases. The tipping fee is fixed at $22/tonne, the same as that of the base case. In the case of the Thermoselect gasifier it is assumed that, except for the gasifier, the unit costs for all installed hardware are scale sensitive (see notes to table 7.1 and 7.2). The Thermoselect gasifier and coke-oven have been pre-designed for modular use, so that specific costs are conservatively assumed to be independent of scale. This is not the case for the MTCI and BCL gasifiers which are assumed to take advantage of scale economies. Therefore, the difference in production costs between the MTCI and BCL systems on the one hand, and the Thermoselect system on the other, increases as the size increases.

The reverse is true for the smaller sized plants. Since the Thermoselect plant is specifically designed for smaller scales (480 tonnes/day), it competes favorably with the indirectly heated cases at the small plant sizes. Even though the oxygen plant and high temperature cooling system become more expensive in relative terms,

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[6] Assuming that the average MSW disposal rate is 1.6 kg/capita-day.

the feed preparation and gasification plant become cost competitive with the other two cases. The plant sizes must be smaller than 300 tonnes/day for the Thermoselect system to produce fuel competitively with the BCL system. At 500 dry tonnes/day, both methanol and hydrogen are produced at lower costs with the Thermoselect gasifier than with the MTCI gasifier. But at these scales the cost of methanol production is almost $21/GJ and hydrogen is $15/GJ. These COPs would not be competitive with biomass or coal, let alone natural gas. The tipping fee, however, is assumed to be fixed at $22/tonne raw MSW. If the tipping fee rose to $100/tonne, the Thermoselect system at this small scale would be competitive with large scale biomass- and coal-fed methanol plants, i.e., have a levelized COP of around $11/GJ of MeOH.[7]

Naturally, the cost of fuel production decreases as the scale of the system increases. At sufficiently large scales, the fuel produced from MSW can be competitive with fuels derived from other feeds even at low tipping fees. For example, at 3000 dtpd the BCL system becomes nearly competitive with hydrogen produced from natural gas at the projected gas cost for 2010 of $3.8/GJ. The large scale MSW-based systems, however, cannot compete with fuel produced from natural gas at current gas costs unless the plant scale is greater than 5000 dtpd.

At feedstock input rates greater than 3000 tonnes/day, plants might become too large for typical existing waste handling infrastructures. Moreover, the systems would begin to lose the benefits of the scale economies at a certain size. The RDF plant would have to increase the number of process lines. Multiple gasifiers operating in parallel would be used instead of one.

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[7] Based on the study by Williams et al. (1994). Biomass facility throughput assumed to be 1650 dtpd; coal assumed to be 5000 dtpd. Feedstock costs assumed to be $2.5/GJ for biomass and $1.75/GJ for coal.

7.3.3 Effects of Cleanup Equipment on Costs

The amount of pollution control equipment that is necessary to achieve the emission levels discussed in the previous chapter remains uncertain. Will cyclones in conjunction with a simple water scrubber be sufficient to clean the gas of tars, oils, particulate matter, HCl, HF, and other pollutants, so that the downstream catalysts will not be contaminated? The base case cost estimates (Tables 7.1 and 7.2) assume that these simple devices are used. Costs could be different for alternative cleanup configurations.

(i) Hot Gas Cleanup Costs

Consider first hot gas cleanup. The BCL hydrogen production configuration is used as an example of how installing a hot gas cleanup (HGCU) system affects the production costs. The simplest case is to assume that a HGCU system works so that the gas can be adequately cooled for cleanup without fouling any equipment. The thermodynamic analysis of the HGCU case analyzed in Chapter 6 reveals that the energy ratio does not change from the base case. More waste heat is available to

Table 7.8 - Costs of Producing H_2 with a Hot Gas Clean Up System[a]

	BCL w/ HGCU	Baseline
MSW Input (GJ/hr)	685.62	685.62
H_2 output (GJ/hr)	399.09	399.26
Energy Ratio	0.58	0.58
Total Capital Costs (10^6 \$)[b]	213.14	210.52
Total Operating Costs (10^6 \$/year)	0.31	1.00
Levelized Costs (\$/GJ)		
Capital	9.97	9.84
Labor and Maintenance	3.70	3.67
Energy	-0.48	-0.24
Feed	-3.12	-3.12
TOTAL COP:	**10.07**	**10.16**

Notes: a) based on the same assumptions used in Table 7.2. Costs in 1991$
b) HGCU case assumes no extra cost for the HGCU, but the capital costs are higher due to the added steam turbine costs

generate electricity for export, however, so the thermal efficiency of the process rises. The higher electricity exports are reflected in the lower COP for the HGCU case in comparison to the base case.

Table 7.8 compares the costs of the HGCU and base case scenarios. The primary savings in the HGCU case come from the added electricity sales. The production cost of the HGCU scenario is only 3% less than that of the base case. The analysis assumes that the cost of a HGCU system is the same as for the base case cleanup hardware. If the installed cost of a HGCU were $1.0 million more than the base case cleanup system, then the costs of producing hydrogen with the HGCU would become greater than the base case COP. Thus, adding a hot gas cleanup system in these configurations has a very modest beneficial impact at best on the efficiencies and costs of fuel production.

(ii) Further Air Pollution Equipment

The quench system in the base case analysis is assumed to be a simple scrubber and is included in the gasifier costs. If the scrubber needs complex materials, packed towers, and alkali scrubbing solutions, it would become more costly. If further particulate matter cleanup is required, a baghouse filter could be added downstream of the scrubber.

To show the effects of added pollution controls on the production costs, hydrogen production based on the BCL gasifier is again used as an example (Table 7.9). Adding all these various clean up systems increases the cost of hydrogen production by only 6% to $10.73/GJ. Since the fuel production systems are capital intensive projects, the costs of adding various clean up equipment does not have substantial effects on the fuel production costs.

Table 7.9 - The Effect of Extra Air Pollution Control on the Costs of the BCL Baseline Case for H_2 Production

	Base case	With extra APC equipment
MSW Input (GJ/hr)	685.62	685.62
dry MSW tonnes/day	1051.50	1051.50
H_2 Output (GJ/hr)	399.26	399.26
CAPITAL COSTS (10^6 \$)		
Advanced scrubber and tower[a]		0.61
Baghouse filter[b]		2.05
SD/FF[c]		2.80
Other installed hardware	144.34	145.71
Subtotal	144.34	151.17
Total Capital Costs (10^6 \$)	210.52	220.43
Total O&M Costs (10^6 \$/year)	1.00	1.34
Total Levelized COP (\$/GJ)	**10.16**	**10.73**

Notes: based on the same assumptions used in Table 7.2. Costs in 1991\$

a) Costs include venturi scrubber and spray tower absorber made of reinforced fiberglass, NaOH solution tank, controls, ductwork, fans, and installation. Based on gas volume flow of 80,000 Nm^3/hour, the cost is estimated to be \$1 million (Kerchinsky, 1994). The cost of the ordinary scrubber included in gasifier costs - 5% of gasifier cost (\$385,000) (Breault and Morgan, 1992) - is subtracted out.

b) \$2.05 million for baghouse estimated by the US EPA (1987) for 94500 Nm^3/hr (4220 kmol/hr) flue from a fluidized bed combustor with 21.3 g/Nm^3 PM loading.

c) This is the state-of-the-art spray dry/fabric filter system used to clean the combustor flue. Costs are scaled by volume flow with 0.62 power. The reference case is \$4 million for 45650 Nm^3/hour (2038 kmol/hr) (USEPA, 1987). The stack gas clean up, which is included the BCL gasifier cost, is subtracted out. Stack gas cleaning is assumed to be 25.3% of the BCL gasifier cost (Breault and Morgan, 1992).

(iii) Effects of Alternative Sulfur Cleanup Technologies on the Costs

The sulfur in the stream can be cleaned through a variety of methods, which vary widely in costs. As mentioned earlier, the liquid oxidation-type plant that is used in the base case is a \$10 million investment. If simple guard beds can be used instead for sulfur removal, the savings in capital costs translate into a savings of \$1/GJ H_2 in the COP for the BCL H_2 case (a 10% reduction). Some of the other sulfur removal methods mentioned in Chapter 6 could be applied to reduce the costs. Most other gas cleaning configurations require selective amine washes, which would operate under high pressures (1.4 MPa). If the gas stream were cooled after being compressed and then stripped of sulfur compounds, the amine wash system would become a viable alternative to the liquid oxidation plant. Though the compression load would be greater, the sulfur cleaning costs could be reduced.

Table 7.10 shows some quick cost estimates of various alternative sulfur removal systems for the BCL H_2 plant and their effects on the total production costs. Recall that the cold bed adsorption (CBA) used in conjunction with the Claus process is not as efficient as the liquid oxidation process. Also, the amine+Claus systems do not respond that well to variable sulfur concentrations in the gas stream. Even though the amine/Claus/CBA system could reduce the overall production costs to $9.50/GJ (a 6% reduction), the liquid oxidation direct conversion system is likely to be more reliable and efficient.

Table 7.10 - Sulfur Cleanup Effects on Costs for H_2 Production from the BCL Gasifier[a]

Sulfur removal system type	none	Amine+ Claus/CBA[b]	Amine+ Lo Cat[c]	Base Case Lo Cat[d]	Amine+ Claus/SCOT[e]
Input Capacity (dry MSW tonnes/day)	1051.50	1051.50	1051.50	1051.50	1051.50
Capital costs of sulfur removal (10^6 \$)	0.00	3.39	5.67	9.67	12.60
Total Capital Costs (10^6 \$)	193.01	199.15	203.28	210.52	215.85
Total O&M Costs (10^6 \$/year)	0.40	0.61	0.75	1.00	1.18
Total Levelized Costs (\$/GJ)	9.15	9.50	9.74	10.16	10.46

Notes:

a) Costs for sulfur removal assume a 0.5% sulfur content in MSW on a wet basis. This amounts to a sulfur removal rate of 7 tonnes/day.

b) Amine bath costs scaled by sulfur removal with 0.7 power factor. The base estimates are $0.8 million for 3.8 tonnes/day (Tannehill et al., 1994). Claus/CBA costs uses the function $0.447*(\text{tonne S/day})^{0.574}$ (Tannehill et al., 1994). A 34.4% multiplier is added for engineering and contingencies (Echterhoff and Quinlan, 1993).

c) Amine wash calculated as above. Lo Cat costs calculated by the function $0.32*(\text{tonne S/day})^{1.176}$ (Tannehill et al., 1994). Again, a multiplier is used to account for engineering and contingencies.

d) From table 7.1

e) Amine wash with Claus/SCOT is scaled by daily tonnes of sulfur removal with 0.7 power factor. Reference costs are $5.36 million for a sulfur removal rate of 2 tonnes/day (Echterhoff and Quinlan, 1993).

7.4 Fuel Cell Vehicle Costs with Fuels from MSW

Using the base case cost estimates, the costs for methanol or hydrogen fueled fuel cell vehicles (FCVs) are compared here to a gasoline fueled internal combustion engine vehicle (ICEV). Detailed models of vehicle performance and costs for various vehicle configurations have been developed by DeLuchi (1992). The performance

Table 7.11 - Comparison of Vehicle Characteristics[a]

Vehicle Type	ICEV[b]	FCV[b]	FCV
Fuel	Gasoline	Methanol	Hydrogen
Storage system	metal tank	metal tank	carbon/Al tank
Driving range (km)	640	560	400
Battery Type	n.a.	bipolar Li/S	bipolar Li/S
Gross power of fuel cell (kW)	n.a.	25	25
Max. power at wheels (kW)	101	76	76
Vehicle Life (km)[c]	193,080	241,350	241,350
km driven per year[d]	17,800	23,800	23,800
Volume of fuel storage and fuel cell system (ltrs)[e]	63	301	310
Weight of whole vehicle (kg)	1415	1328	1311
Drag coefficient[f]	0.28	0.23	0.23
Fuel economy (HHV, gasoline equivalent mpg)[g]	25.8	61.5	71.6
Fuel economy (gasoline equivalent ltrs/100km)	9.1	3.8	3.3

Notes:

a) Table taken from Ogden et al. (1994). The gasoline vehicle is based on a year 2000 version of the Ford Taurus. Other vehicles are hypothetically designed from the ICEV. The FCV uses the PEMFC.
b) ICEV = Internal combustion engine vehicle; FCV = Fuel cell vehicle.
c) Assumed that Electric Vehicles will have a longer life than ICEVs.
d) It is assumed that the FCV is driven more because it has a lower operating cost.
e) This includes energy storage (battery, MeOH tank, hydrogen tank, or gasoline tank), the fuel cell, and MeOH reformer.
f) Assumed to be lower for the FCVs because the energy storage is more costly, so there is more value in improving the efficiency.
g) Calculated as the mile/GJ fuel economy of the FCV multiplied by the HHV of gasoline (0.131875 GJ/gallon). PEM FCs have 45% efficiencies, so the FCV has three times the efficiency of the ICE. The fuel economy is calculated with the model takes and into account a detailed set of parameters including the power source efficiency, weight, drag, powertrain efficiency, city vs. highway driving and other factors.

characteristics from these models are used to analyze the lifecycle costs of competing vehicle technologies.

Table 7.11 shows key characteristics of the ICEV and the FCV based on DeLuchi's models. An important attribute of the FCV is that it has a much higher fuel economy than the ICEV. The ICEV is assumed to be a year-2000 version of the Ford Taurus fueled by reformulated gasoline, with a fuel economy of 26 miles/gallon. It should be mentioned that the conservative fuel economy assumed for the ICEV could be improved.

DeLuchi assumes for his model that FCVs would have lower travel ranges than the ICEVs because hydrogen has a lower energy density than gasoline. The lower energy density of hydrogen and methanol, however, are compensated by the assumed much greater efficiency of the fuel cell relative to the IC engine. DeLuchi also assumes that the FCV's maximum power to the wheels is less than that of the ICEV. Since the FCV has an electric drive train, it does better for urban or highway driving needs. Another important note is that the FCV electrochemical engine will probably take up more room than the IC engine (Ogden et al., 1994).[8]

Lifecycle costs for the gasoline fueled ICEV and the FCV fueled by methanol and hydrogen from MSW are summarized in Table 7.12. The reformulated gasoline is assumed to have a production cost of $0.97/gallon, which corresponds to a crude oil price of $26.4/bbl. This is slightly below the crude oil price projected by the US DOE (reference case) for 2010 (EIA, 1994). The analysis shows that on a lifecycle basis the FCV is cheaper than the ICEV (in cents per kilometer traveled) for every fuel production configuration, assuming that the FCV meets the performance and cost characteristics outlined in Tables 7.9 and 7.10. Even the Thermoselect system, which would produce fuel at the highest cost, would produce methanol and hydrogen at a low

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[8] The large volume of the FCV electrochemical engine should not be problematic for buses or trucks.

Table 7.12 - Baseline Cost Comparison (in 1991$) of ICEV and FCVs Fueled by MSW Derived Fuels[a]

Vehicle Type	ICEV	FCV	FCV	FCV	FCV	FCV	FCV
Fuel	Gasoline	MeOH	H_2	MeOH	H_2	MeOH	H_2
Gasifier producing fuel	n.a.	BCL	BCL	MTCI	MTCI	ThS	ThS
Full retail price of vehicle, including taxes ($)	17,976	21,709	25,091	21,709	25,091	21,709	25,091
Fuel retail price, excl. taxes							
($/GJ)[b]	8.95	18.03	15.09	20.07	16.86	21.16	18.19
($/gal. gasoline equiv.)[c]	1.18	2.38	1.99	2.65	2.22	2.79	2.40
Lifecycle Costs of Alternative Vehicles (in cents/km)							
Base vehicle	11.60	7.76	7.82	7.76	7.82	7.76	7.82
Battery and auxiliaries	0.00	1.89	2.06	1.89	2.06	1.89	2.06
Fuel cell & reformer system	0.00	1.86	1.53	1.86	1.53	1.86	1.53
Fuel storage	0.00	0.02	1.00	0.02	1.00	0.02	1.00
Misc. O&M[d]	5.80	5.24	5.23	5.24	5.23	5.24	5.23
Fuel taxes	0.78	0.78	0.78	0.78	0.78	0.78	0.78
Fuel	2.84	2.01	1.49	2.40	1.78	2.59	1.97
Total lifecycle cost	21.01	19.94	20.16	20.21	20.36	20.35	20.51
Breakeven gasoline price[c] ($/gallon)		1.06	1.15	1.17	1.23	1.23	1.29

Notes:

a) These costs are adapted from analyses by DeLuchi (1992) and Ogden et al. (1994)

b) These MeOH and H_2 costs are taken from the analysis done earlier in this chapter. Delivery and refueling station costs are assumed to be $3.09/GJ for MeOH and $4.93/GJ for hydrogen (Ogden et al., 1994). The gasoline price assumes that it is reformulated gasoline (15 cents/gallon more than conventional gasoline) and the oil price is $26.40/bbl. The distribution and refueling costs for gasoline are assumed to be $1.60/GJ.

c) $/gasoline equivalent is calculated by multiplying the fuel retail price by the volumetric HHV of gasoline, 0.125 MMBTU/gallon (131.875 MJ/gallon)

d) Includes repairs, insurance, oil, tires, etc.

e) The retail price of gasoline, including taxes, that will make the lifecycle cost (cents/km basis) for the gasoline ICEV equal to the lifecycle cost of the alternative fuel FCV. Taxes are assumed to be equivalent for all fuels on a cent/km basis at 0.78 c/km. For gasoline, this is $0.31/gallon.

enough cost for the FCV to compete favorably with the gasoline fueled ICEV modeled in this analysis.

Table 7.12 shows that the lifecycle costs of the FCV are not sensitive to the fuel costs. Since the FCVs have such high efficiencies, the impacts of fuel costs on the lifecycle costs are modest. For example, the variations in COP for fuels derived using Thermoselect are 20% greater than for producing fuels using BCL, but the lifecycle costs for the FCV (in cents/km) vary by only 2% for the two cases.

The bottom line of Table 7.12 shows the retail gasoline price, including taxes, that would be competitive in the modeled ICEV with the methanol and hydrogen fueled FCVs. These breakeven prices range from $1.06/gallon to $1.29/gallon, which are all lower than the projected retail price of reformulated gasoline of $1.49/gallon in 2010.[9] Therefore the price of gasoline would have to be dramatically reduced (by 13%-30%) for the ICEV to compete with the FCV using methanol and hydrogen from MSW, using the performance and cost assumptions from DeLuchi (1992).

Figures 7.6 and 7.7 show the effects of the tipping fee on the breakeven price of gasoline using the same assumptions from Tables 7.9 and 7.10. As the tipping fee increases and the production cost of methanol and hydrogen falls, the gasoline price must decrease to stay in the market. At the Northeast average tipping fee of $71/tonne, gasoline prices must be $0.86/gallon to remain competitive with methanol produced from MSW using Thermoselect. At these high tipping fees, the breakeven cost compared to the BCL case study would be $0.60/gallon gasoline. Even with no tipping fee, gasoline must be sold at $1.23/gallon to compete with methanol fuel produced by the BCL system. To compete with the Thermoselect methanol case with no tipping fee, gasoline must be sold at $1.35/gallon, still $0.14/gallon less than the projected gasoline price.

For hydrogen production from MSW, all three cases again require no tipping fee for the FCV to be favored over the gasoline-fueled ICEV.

7.5 Concluding Remarks

In general, this economic analysis shows favorable prospects for MSW feeds for fuel production. For the base case analysis of a 1050 dtpd facility with a $22/tonne tipping fee, the systems examined are nearly competitive with the biomass systems

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[9] Based on the projected crude oil price of $26.40/bbl, reformulated gasoline production costs are $0.97/gallon. Fuel distribution and filling station costs are $0.21/gallon. The gasoline tax is assumed to be $0.31/gallon (Ogden et al, 1994)

operating at the same scale. If the scale of the facility or the tipping fee were increased, the economics for MSW-fed systems begin to look favorable. The tipping fee assumed in the base case is lower than the national average. At higher tipping fees, like that in the Northeast ($71/tonne), the BCL MSW-to-fuel system can compete with conventional fuel production facilities that use natural gas feeds (assuming a $3.8/GJ gas price).

The baseline economic analysis reveals that using the high temperature gasifier to produce both methanol and hydrogen is more costly than using the indirectly heated gasifiers. The BCL system seems to be the economic choice for fuels production, but it would be premature to advocate the development of only the BCL gasifier for two reasons: 1) the emissions from the indirectly heated gasifiers must be tested, and 2) the Thermoselect gasifier has proven commercial experience with MSW, while the indirectly heated gasifiers have run only limited experimental tests with RDF. All three systems (and perhaps others) should be developed further to learn more about emissions and scale up issues.

The final section of this chapter shows that an FCV fueled by methanol or hydrogen from MSW would be less costly than a gasoline fueled ICEV (with gasoline prices projected for the 2010 time frame) on a lifecycle basis for all six fuel production systems regardless of tipping fee and at modest scale of fuels production.

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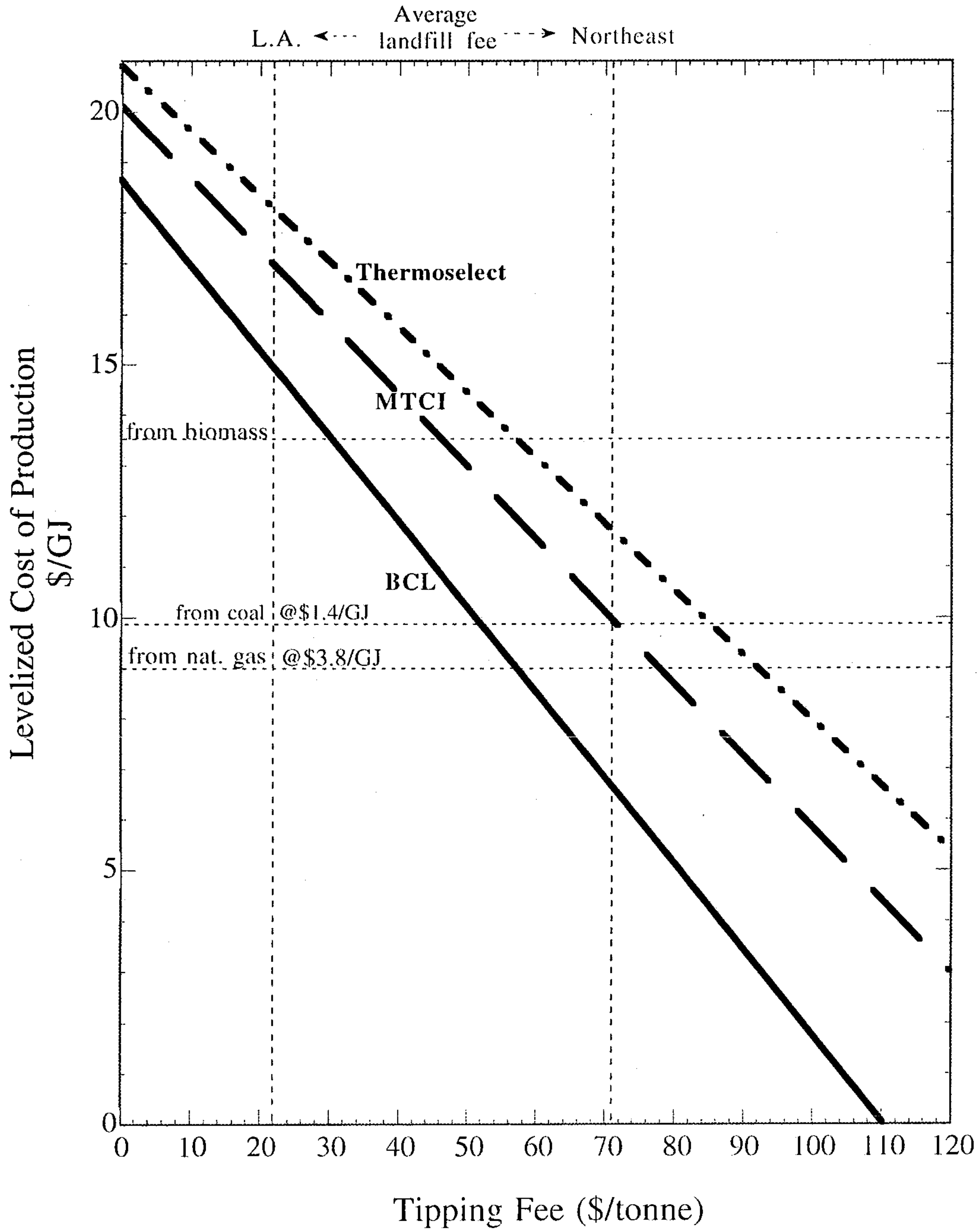

Figure 7.1 - Cost Estimates for Methanol Production from MSW as a Function of the Feedstock Cost

Notes: All costs in 1991$. Methanol costs from natural gas and biomass are based on analysis by Williams et al. (1994). Costs of fuel from biomass assume a facility the same size as the MSW facility (see notes to Tables 7.4, 7.5). The natural gas and coal facilities are assumed to be much larger (see Tables 7.6, 7.7). The capital charge rate is assumed to be the same as that for the MSW systems (see notes to Table 7.1 and 7.2). Landfill fees taken from the Sanitation District of L.A (1993) and NSWMA (1991).

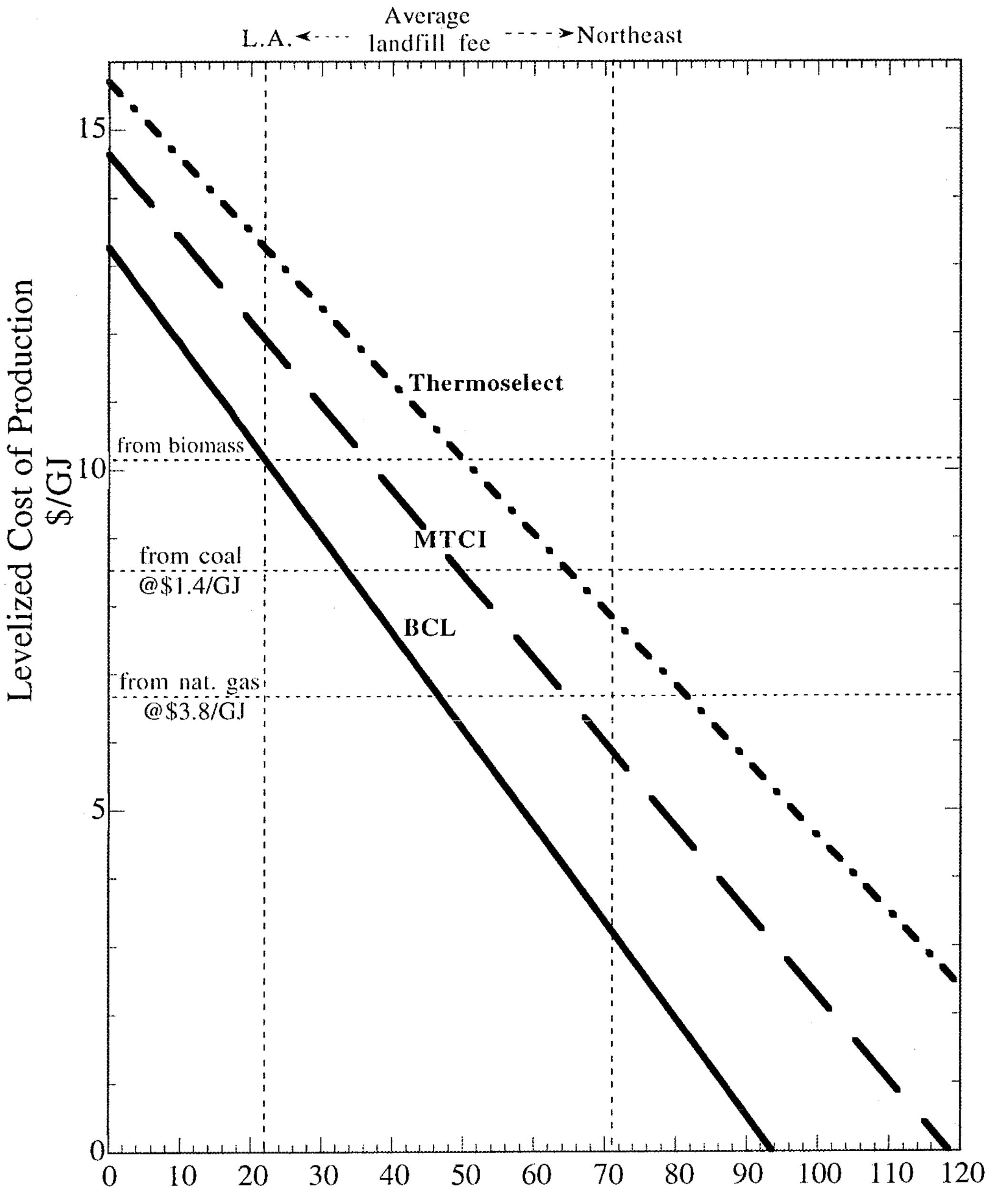

Figure 7.2 - Cost Estimates for Hydrogen Production from MSW as a Function of the Feedstock Cost

Notes: Same assumptions used in Figure 7.1

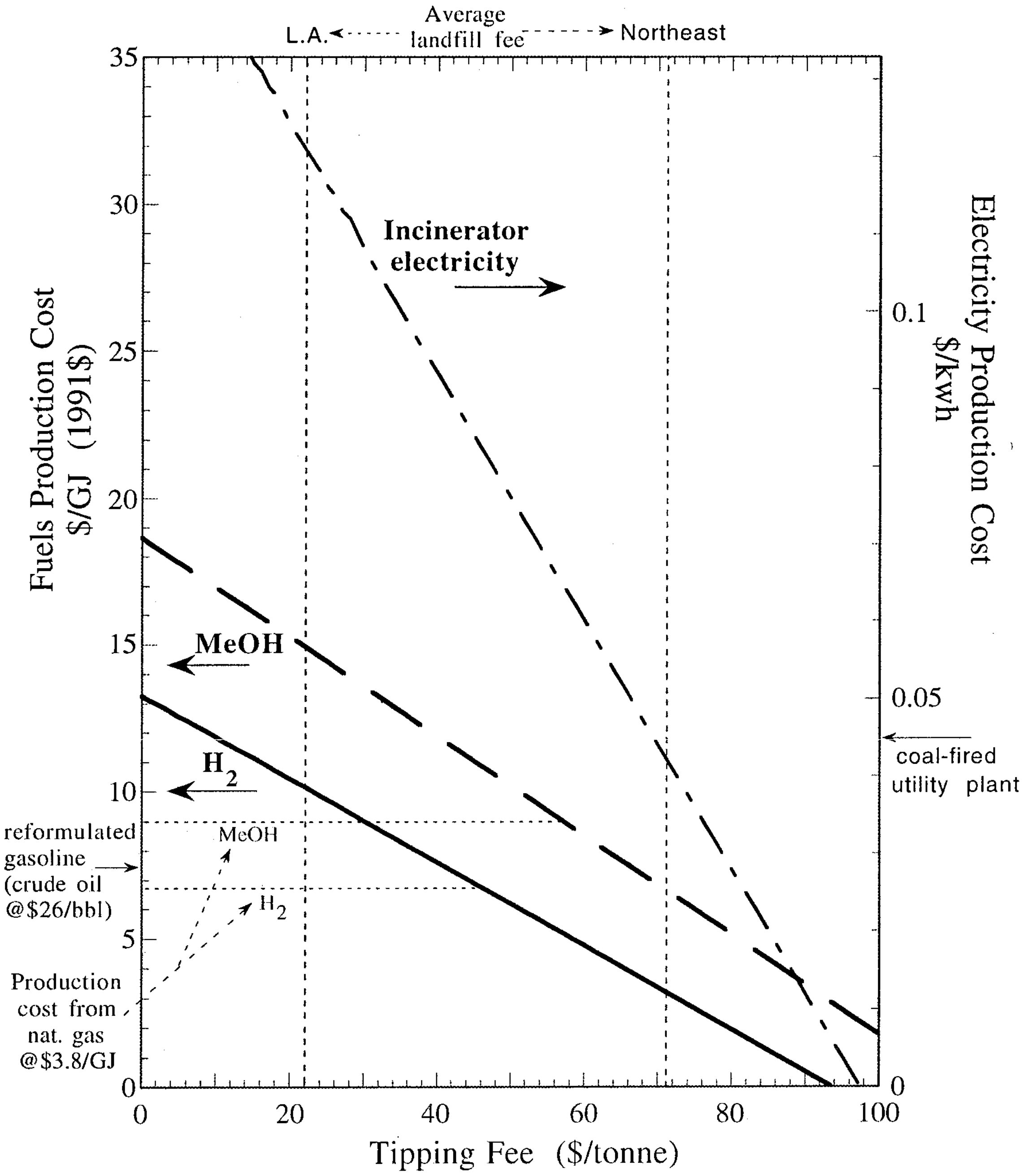

Figure 7.3 Economics of Energy Production from MSW (with the BCL gasifier)

Notes: Costs in 1991$. Steam turbine electricity (incinerator) costs are taken from EPRI's TAG^{TM} Technical Assessment Guide - Electricity Supply (1989), assuming CRF of 15.1% and a wet HHV of 11.2 MJ/kg for MSW. The methanol and hydrogen costs from natural gas are based on Williams et al. (1994). Coal-fired electricity costs taken from EPRI's TAG (1993), assuming a feed cost of $1.5/GJ and a capital recovery factor of 0.117. Landfill fees taken from the Sanitation District of L.A (1993) and NSWMA (1991).

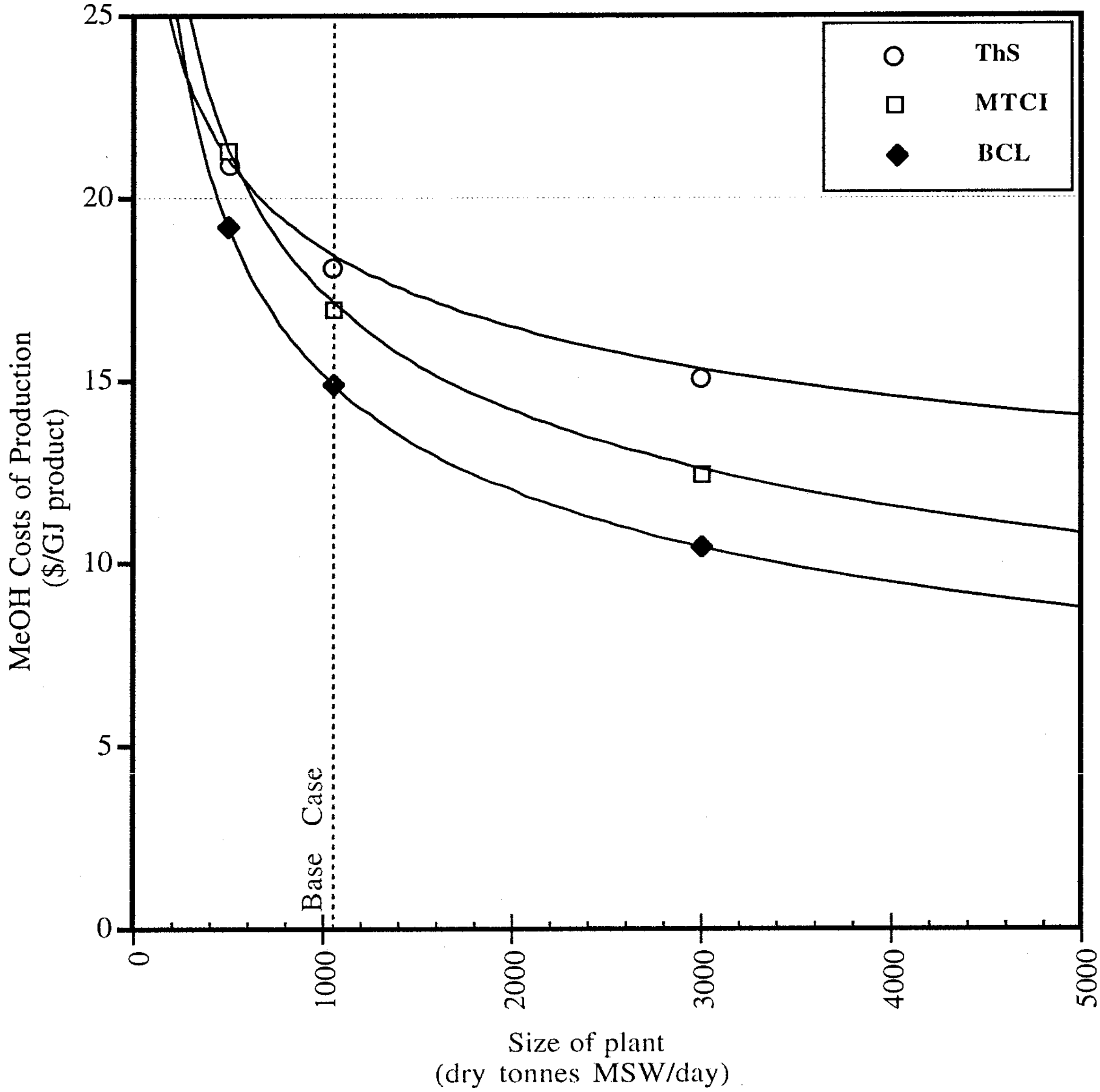

Figure 7.4 - Effects of Scale on Methanol Production Costs

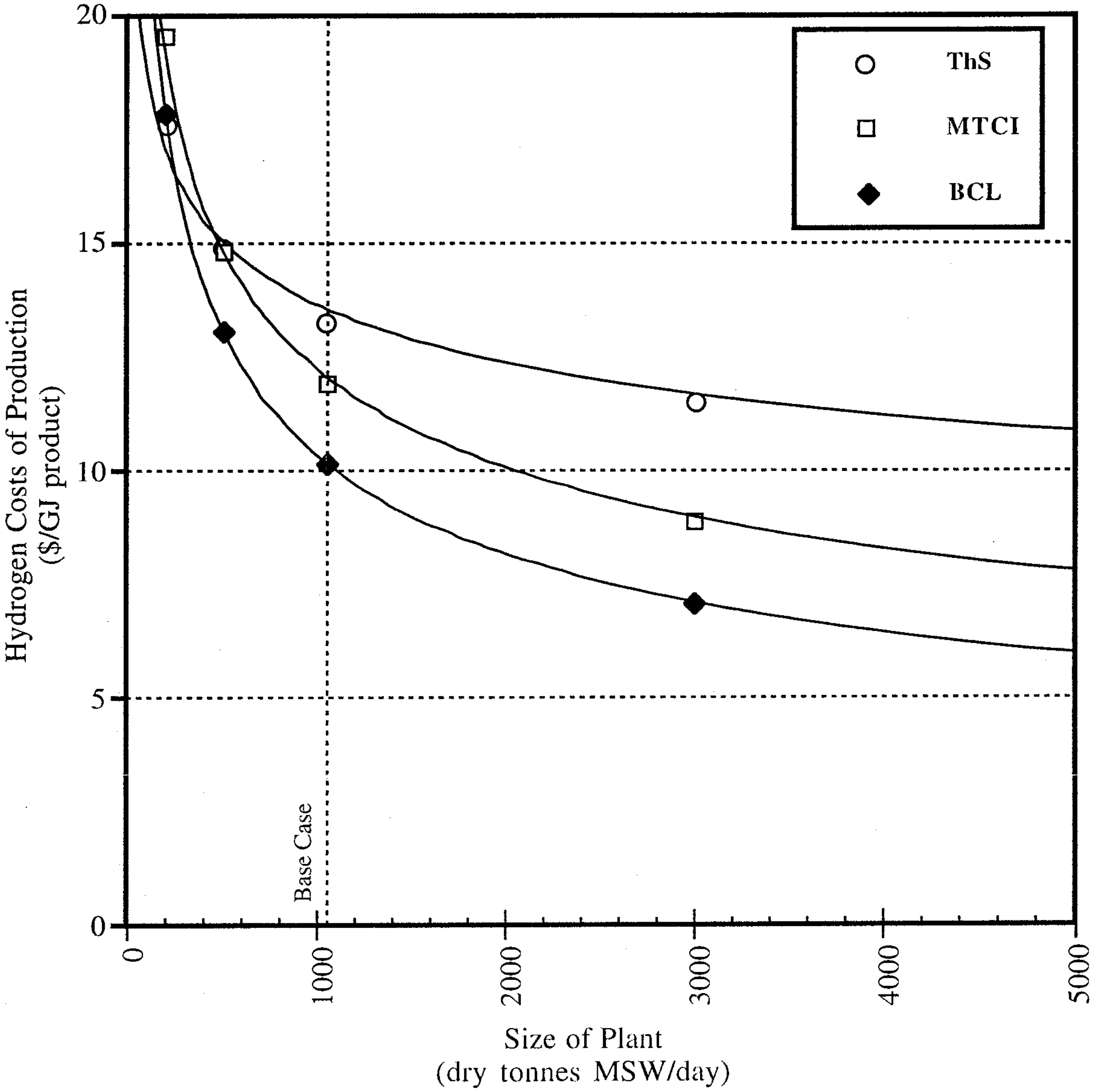

Figure 7.5 - Effects of Scale on Hydrogen Production Costs

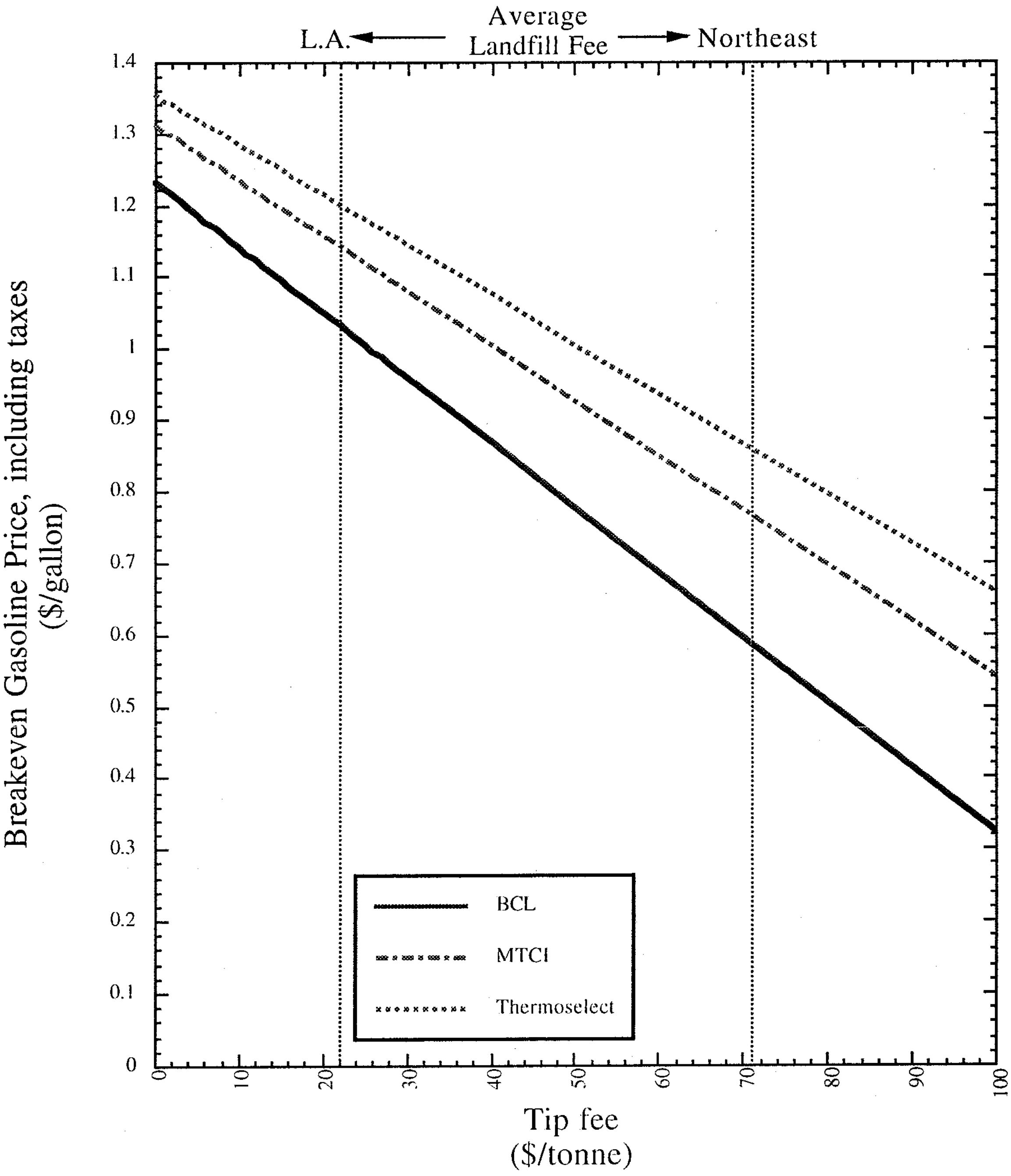

Figure 7.6 - Effect of Tipping Fee on the Breakeven Cost of Gasoline to MSW derived Methanol for FCVs

Notes: Breakeven costs based on economic models of ICEVs and FCVs by Ogden et al. (1994). Landfill fees taken from the Sanitation District of L.A (1993) and NSWMA (1991).

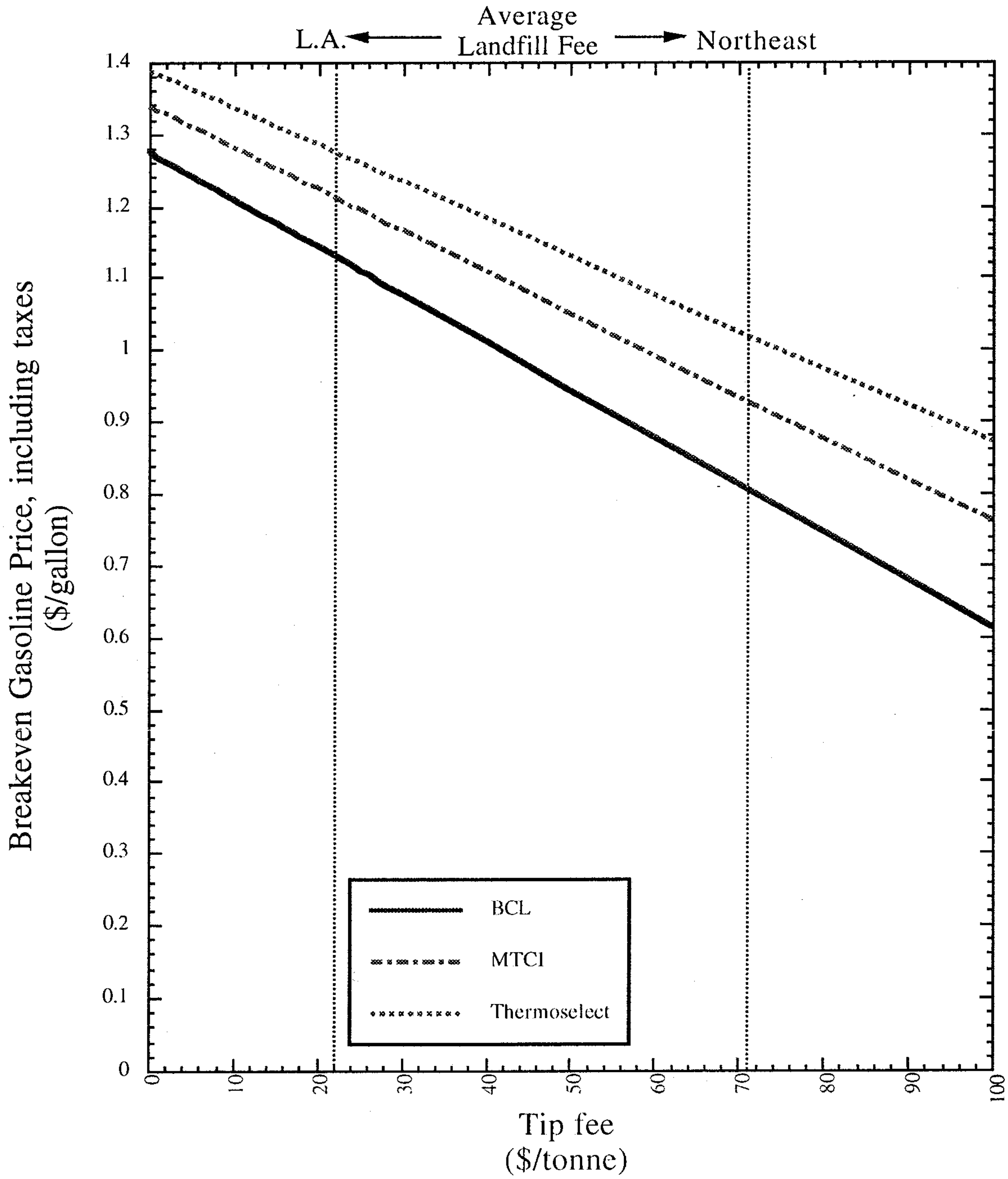

Figure 7.7 - Effect of Tipping Fee on the Breakeven Cost of Gasoline to MSW derived Hydrogen for FCVs

Chapter 8: Conclusions

Due to growing public opposition and more strict environmental regulations, it is becoming more difficult to establish new landfills and build new incinerators. An alternative to these conventional municipal solid waste (MSW) disposal technologies is to produce methanol or hydrogen from MSW for use as transportation fuels.

Both methanol and hydrogen can be used as fuels for a fuel cell vehicle (FCV), a low polluting and highly efficient alternative to the conventional gasoline fueled internal combustion engine vehicle (ICEV). The FCV is being developed to address the environmental and security concerns currently facing the US transportation sector.

Though natural gas will be the first feedstock used to produce methanol and hydrogen, other feedstocks, like biomass, MSW, and coal, can also be converted into methanol and hydrogen.

Biomass is an especially attractive renewable feedstock for fuels production because no net carbon emissions would result from its production and use if it is grown sustainably in energy plantations. Other studies have examined biomass as a possible resource for fuel production and have shown promising results.

MSW is very similar to biomass once the inorganic fraction of MSW (glass, metals, dirt, etc.) is removed. Producing fuel from MSW is therefore very similar to producing fuel from biomass. As with biomass feeds, the gasifier is the only non-commercial technology in the entire process. The main difference between MSW and biomass feeds is the feed preparation step, which may involve producing refuse-derived fuel (RDF). Since people are willing to pay to dispose of MSW (through a tipping fee), MSW has the attractive property of having a negative feedstock cost.

Based on the previous analyses of biomass, this thesis has assessed the feasibility of producing methanol and hydrogen from municipal solid waste (MSW) for

FCVs. The assessment includes examining various MSW (or RDF) gasification technologies, calculating energy efficiencies of alternative fuel production systems, comparing environmental characteristics of gasifiers and incinerators, and estimating the costs of producing fuels from MSW.

Three different gasifier technologies that have experience with MSW (or RDF) feeds have been characterized and analyzed for fuel production -- BCL, MTCI, and Thermoselect. The BCL and MTCI processes are both low temperature indirectly heated gasifiers, whereas the Thermoselect gasifier is a directly heated high temperature gasifier. The two indirectly heated gasifiers have run tests at the pilot or laboratory scales using RDF feeds. Thermoselect has one commercially operating gasifier in Italy that operates on raw MSW.

The specific results and conclusions from this study are the following:

- MSW would be a significant fuel resource when converted to methanol or hydrogen and used in FCVs: present MSW levels would provide enough methanol and hydrogen to fuel 20% or more of the nation's current light duty vehicle transportation demands.

- Though the effort to commercialize MSW and RDF gasification technologies had early setbacks, a few gasifiers are now operating reliably at a commercial-scale for gas production (not methanol or hydrogen production). New gasifier designs that look especially promising for fuels production from biomass (i.e. - indirectly heated gasifiers) are under development.

- The thermal efficiency calculated for producing methanol from MSW is about 50% for the three systems examined in this thesis. Hydrogen production systems are calculated to have a thermal efficiency of roughly 60%.

- The efficiencies of fuel production from MSW are only slightly lower than from biomass. The main added loss with MSW is in the feed preparation step.

- Gasifiers produce lower volumes of gas than incineration, and pollutants in the product gas are thus likely to be more concentrated. The higher concentrations will facilitate cleanup of many pollutants, and the lower volumes should reduce cleanup equipment costs compared to incinerator flue gas cleanup.

- Though emissions data from the three gasifiers is limited, the available empirical evidence suggests that gasifiers emit fewer PCDD/Fs and other pollutants than incinerators. This is based on comprehensive measurements conducted on the Thermoselect facility. BCL and MTCI still need to test all four effluent streams from their gasifiers: product gas, flue gas, waste water, and solid residue.

- Estimates show in particular that the Thermoselect gasifier produces much less PCDD/PCDF than modern incinerators. This characteristic makes the process especially attractive as a MSW disposal technology. It is uncertain why the gasifier has lower PCDD/PCDF emissions.

- For a baseline set of assumptions, the BCL systems appear to be the least costly option. For the same assumptions, the high temperature Thermoselect case is the most costly, mainly due to the O_2 requirement and high temperature of the process.

- Variations in the pollution control equipment and sulfur removal systems do not significantly affect the process efficiencies. Additional clean up equipment increases the fuel production costs, but these changes are relatively small because the system's total capital costs are much larger in comparison.

- For the base case tipping fee of $22/tonne, only hydrogen production from MSW using the BCL gasifier is economically competitive with biomass at expected future biomass prices of $2.5/GJ. The other MSW-to-fuel production systems become competitive with biomass feeds at tipping fees of $30-$60/tonne. For comparison, an average tipping fee in the Northeast is around $70/tonne.

- At tipping fees below $70/tonne, the MSW-to-fuel production systems using the BCL gasifier is competitive with natural gas at future gas prices ($3.8/GJ). The MTCI and Thermoselect systems become competitive with natural gas at tipping fees below $95/tonne.

- On a lifecycle basis, the FCV fueled by methanol and hydrogen derived from MSW is less costly than a gasoline fueled ICEV. This is true even if there is no tipping fee for the MSW.

This report shows that producing methanol or hydrogen from MSW is a promising alternative MSW disposal technology. The analysis, however, is only a preliminary assessment. More detailed research of MSW gasification and fuels production from MSW and further development of process components, such as the gasifier, are required to bring the fuel production system to commercially readiness. Some suggested areas of future research are as follows:

- More detailed design for these systems are needed. From a cost perspective, a significant additional analysis needed is a detailed set of estimates for the heat exchangers that would be needed to achieve the thermal integration levels implied by the pinch analysis.

- More experience with RDF feeds is required for the indirectly heated gasifiers. Longer durations and larger scales of operation must be tested. Comprehensive emissions measurements need to be made for all three gasifiers, especially for the indirectly heated gasifiers.

- A better understanding of PCDD/PCDF formation and destruction in gasification processes is required.

- Leachability of slags in comparison to ashes should be investigated. The composition and leachability of the char and ashes from the low temperature gasifiers need to be measured.

- Several interesting alternative configurations should be examined more closely. These include not using reformers and co-producing electricity to lower the costs of fuel production. Other advanced technologies, such as autothermal reforming or one pass methanol synthesis, should also be evaluated for possible cost reductions.