City of Bellingham

Technical Memorandum

Biosolids Conversion Technology Evaluation

CDM

CDM

Memorandum

To: *Larry Bateman, City of Bellingham*

From: *Dave Parry, CDM*

Date: *December 31, 2008*

Subject: *Biosolids Conversion Technology Evaluation*

In keeping with sustainable practices currently in place and those planned for the future, the City of Bellingham seeks a long term solution for thermal processing of their biosolids. At the Post Point Wastewater Treatment Plant (Post Point Plant), the City currently processes its biosolids in two Multiple Hearth Furnaces (MHFs). With projected population growth, the increasing cost of energy, and new technologies available for biosolids processing, the City has recognized the need to evaluate the efficiency of its current processing system. This evaluation compares retrofitting the current system for increased efficiency, to the installation of the latest Fluidized Bed Incinerator (FBI) technology. The comparison, based on economic, environmental, operational, and social objectives, will ultimately help the City of Bellingham gain insight into selecting a technology, a timeline for implementation of the selected technology, and a recommended action plan.

Objectives and Evaluation Criteria

Modifications to the current MHF system and a new FBI offer advantages to the City in various ways. To establish a basis for comparison, we will evaluate the biosolids processing options against four objectives. The objectives are as follows.

- **Economic:** Evaluations will be conducted to determine the costs (capital, annual, and life cycle) of each alternative. Implementation schedules for each option will be developed, which will allow financial planning and identify periods that may require temporary solids handling procedures as new or improved facilities are brought on line. These evaluations will allow the City to understand the budgetary impacts of each option.

- **Environmental:** Maintaining and increasing its sustainability is important to the City of Bellingham. Thus, each option will be evaluated based on its impact on the environment. This includes steps necessary to comply with air permitting requirements, generating and using green energy, reducing biosolids processing energy demand, and reducing the Post Point WWTP carbon footprint.

- **Operational:** To ensure an acceptable biosolids conversion technology, operations, and maintenance issues for each alternative need to meet City criteria. Understanding City criteria will help select an option that meets the City's expectations with regards to operational ease and staffing requirements. For example, the City is not interested in a biosolids hauling option at this time.

- **Social:** Upgrades at wastewater treatment plants need to be conducted in a manner to maintain the way of life for the residents. Before proceeding with an implementation strategy, it is important to determine the ease in which permits will be obtained or the potential for fugitive odor emissions. The City also wants to maintain good neighbor status in the surrounding community.

Background

Overview of Post Point WWTP

The City of Bellingham and surrounding communities are served by the Post Point WWTP. Wastewater is collected from over 250 miles of sewers and conveyed to Post Point for treatment. Septic tank waste is also collected from private homes throughout rural areas of Whatcom County and brought to the plant for treatment.

Figure 1 is a simplified schematic of the Post Point WWTP treatment facilities. Plant influent first receives preliminary treatment that includes the removal of coarse solids by mechanical bar screens. Screened effluent then flows into grit tanks for the removal of sand and heavy inert solids. The wastewater is then conveyed by gravity to primary clarifiers where organic solids are settled and removed as waste primary sludge (WPS). Primary effluent then flows to the secondary treatment facilities. In the secondary treatment process microorganisms consume dissolved organic material as a food source. Secondary treatment uses pure oxygen to create an aerobic environment in the aeration basins. Bacteria and solids that are produced in the secondary process are separated from the treated wastewater in the secondary clarifiers. A portion of the settled solids are returned back to the aeration basins to seed the process with active biomass called return activated sludge (RAS). The remainder of the secondary solids are wasted from the liquid treatment system as waste activated sludge (WAS). The complete biological secondary treatment system is referred to as the activated sludge process.

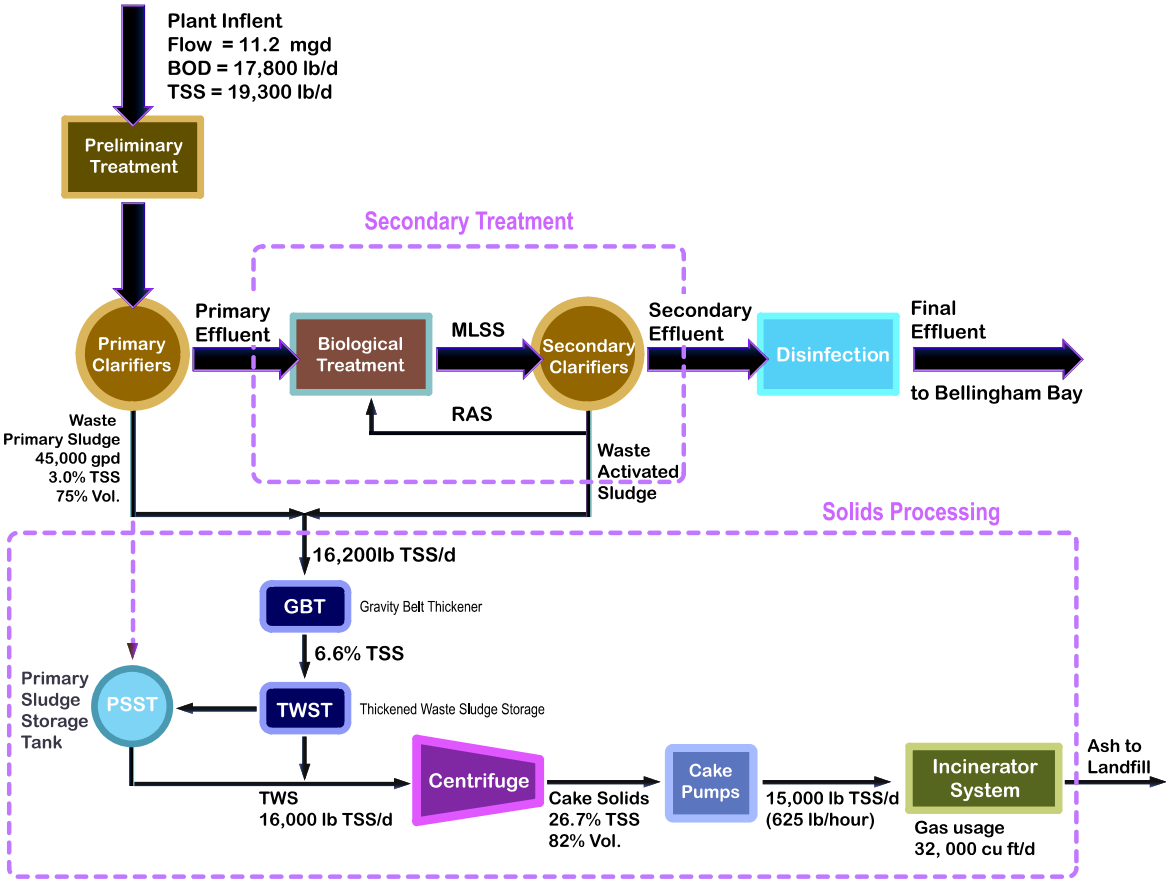

Figure 1: Treatment Plant – Simplified Schematic (1998 – 2003 Data)

Key to this study is "solids processing." Solids processing begins by combining WPS and WAS in a pumping system that feeds the plant's gravity belt thickener's (GBTs). The GBT's produces a product called thickened waste sludge (TWS) that has a total solids concentration greater than 6 percent. Thickened waste sludge is stored in either the thickened waste sludge tanks (TWST) or in the primary sludge storage tank (PSST). Additional water is removed from the TWS by a centrifuge dewatering process. The resultant cake solids have solids concentrations averaging 26 percent and are then fed to one of two multiple hearth furnaces (MHF) for incineration.

According to the 2004 "Re Rating of the Post Point Wastewater Treatment Plant", Post Point WWTP has a rated capacity of 20 MGD, 25,530 lb BOD/day and 47,000 lb TSS/day. However, normal flow during the study period (1998 – 2003) was 11.2 mgd that is comfortably less than its rated capacity. Information on Figure 1 is median (50 percentile) data for a 5- year period from 1998-2003. During this period of time, the incinerator system was fed an average of 15,000 lb TSS/d (625 lb TSS/hr). Normal operation during this period

was to use one MHF at a time and to operate it for approximately 4 days per week. The actual average loading rate to MHF's during 1998 to 2003 was 1100 lb TSS/hour based on a 4 day per week operating schedule.

Existing MHFs

The Post Point WWTP has two multiple hearth furnaces (MHFs) that are used for incineration of the plant's residual solids. Both units have 9 hearths and an outside diameter of 14 ft - 3 inches. The older unit, hereafter referred to as Incinerator 1, was supplied by Envirotech, BSP Division, in 1973. The more recent unit, Incinerator 2, was supplied by Enviroquip in 1993. Incinerator 1 has been modified to include a top hearth afterburner. The dewatered solids are fed to Hearth 2 and the burners on Hearth 1 are used to maintain an exit flue gas temperature of 1200°F. Incinerator 2 has a separate, downflow afterburner chamber with a gas fired burner mounted on top of the chamber. Both incinerators have the same type of air pollution control systems consisting of a venturi scrubber, followed by a tray scrubber, and a wet electrostatic precipitator (WESP). Downstream of each WESP is an induction fan which conveys the combustion gases through the system and maintains draft in the MHFs.

MHFs were formerly the workhorses of the industry and there are approximately 200 of them still operating in the USA. MHFs consist of a series of stacked hearths. The biosolids are fed at the top of the furnace and move down the unit first drying on the upper hearths, followed by burning on the middle hearths and finally cooling of the resultant ash on the bottom hearths. The stacked hearth configuration is thermally efficient in that it allows the hot combustion gases to rise through the furnace and greatly assist in drying of the biosolids on the upper hearths. However, the upper drying hearths emit high levels of VOCs, CO, particulates, and odors that are not acceptable to current air regulatory boards. Because of these pollutant emissions, most MHFs require the use of a fuel-fired afterburner to meet present day air emission standards. The use of a fired afterburner made MHFs quite inefficient since a significant quantity of fuel is required to raise the flue gas temperature exiting the drying hearths to a minimum afterburner outlet temperature of 1200 deg F. At the Post Point WWTP, the afterburners are used to ensure that the furnaces meet their opacity (visible emission) requirement. In addition to their reliance on a fired afterburner, MHFs are more mechanically complex and therefore more difficult to maintain and operate at higher excess air levels than fluidized bed incinerators.

Evaluation of Incineration and Energy Recovery Alternatives

As a first step in evaluating incineration and energy recovery alternatives, it is important to review the existing operations of the incinerators at the Post Point Plant. The following section reviews the incinerator operations over the period of 2006 to Nov of 2008 and focuses on incinerator capacity, operating hours, fuel usage and maintenance issues.

Existing Incinerator Operations

Slag and clinker formation are two of the more frequently occurring operating problems that can restrict the performance of a MHF. Slag is the accumulation of molten or fused ash which sticks to the walls, rabble arms and center shaft of a MHF. Clinkers are hard or soft clumps of fused ash that can jam rabble arms or a unit's ash conveying system. Although both incinerators had some slagging and clinker problems in the past, modifications and improvements to the incinerators have largely eliminated these problems. Adjustments were made on Incinerator 1 to the rabble arms and rabble pattern. On Incinerator 2 the drop holes were enlarged and the burners were modified to operate at very high combustion air levels with very low to zero fuel usage. Introducing a high air flow into the furnace provided better mixing of combustion gases and lowered hearth temperatures that greatly reduced slag and clinker formation. Presently, Incinerators 1 and 2 are both operating reliably and require only normal scheduled maintenance.

The air permit for the incinerators does not require the afterburners on each unit to be fired. However, the air permit does require that both units achieve a 5% opacity on their stack emissions that essentially means that there can be no visible emissions from the incinerator stacks. To ensure compliance with the opacity requirement, the afterburners on each unit are continuously fired to achieve an afterburner exit temperature of 1200 degrees F. Fuel usage in the afterburners is the majority of fuel used by the incinerators.

An annual summary of the incinerator operating records from 2006 through November 2008 is presented in Table 1. Presently, only one incinerator is operated at a time for approximately 5 days per week, 24 hours per day. During weekends the operating incinerator is kept on hot standby by firing natural gas in the furnace burners. As shown in Table 1, the total annual quantity of biosolids burned in 2006 and 2007 was approximately 8.5 million dry lb/yr or 4,250 dry tons/yr. Note that if the total solids for 2008 (through November 23) is proportionately increased for the remaining days in 2008, the total annual quantity of solids burned will be approximately 8.6 million dry lb/yr.

Year		Incinerator 1	Incinerator 2	Total
2006	Operating , Hrs[1]	2,200	3,645	5,845
	Sludge Burned			
	Dry lb/yr	3,088,028	5,285,953	8,373,981
	Dry tons/yr	1,544	2,643	4,187
	Avg Feed Rate, Dry lb/hr	1,404	1,450	
	Total Gas Usage, Cuft/yr	5,508,898	16,360,428	21,869,326
	Standby Gas Usage[2], Cuft/yr	1,391,720	5,836,792	7,228,512
	Standby Gas as % of Total Gas	25.3%	35.7%	33.1%
	Solids Processing Gas per Dry Ton[3], Cuft/ton	2,667	3,982	3,497
2007	Operating Hrs	4,622	1,273	5,895
	Sludge Burned			
	Dry lb/yr	6,657,415	1,923,927	8,581,342
	Dry tons/yr	3,329	962	4,291
	Avg Feed Rate, Dry lb/hr	1,440	1,511	
	Total Gas Usage, Cuft/yr	11,691,097	5,397,597	17,088,694
	Standby Gas Usage, Cuft/yr	2,974,027	2,012,894	4,986,921
	Standby Gas as % of Total Gas	25.4%	37.3%	29.2%
	Solids Processing Gas per Dry Ton, Cuft/ton	2,619	3,518	2,820
2008	Operating Hrs[4]	2,852	2,317	5,169
	Sludge Burned			
	Dry lb/yr	4,359,793	3,371,324	7,731,117
	Dry tons/yr	2,180	1,686	3,866
	Avg Feed Rate, Dry lb/hr	1,529	1,455	
	Total Gas Usage, Cuft/yr	3,994,917	7,323,535	11,318,452
	Standby Gas Usage, Cuft/yr	1,201,908	2,786,468	3,988,376
	Standby Gas as % of Total Gas	30.1%	38.0%	35.2%
	Solids Processing Gas per Dry Ton, Cuft/ton	1,281	2,691	1,896

Notes

1. Operating hours are hours of sludge burning; standby hours are not included.
2. Standby gas usage is gas required to maintain incinerator in hot standby mode.
3. Solids processing gas per dry ton is the total gas minus the standby gas divided by the dry tons of sludge burned.
4. Operating hours for 2008 are Jan. 1 - Nov. 23, 2008.

Table 1: Summary of Incineration Operating Records for 2006 - 2008

The annual incinerator operating hours (which do not include the hours on hot standby) are also very consistent at approximately 5,850 hours per year. The actual operating hours reveal a 94% availability based on the nominal 5 day per week operating schedule (5850/(5 x 24 x 52) = 0.94). The average annual feed rates for both incinerators were about the same, ranging from 1,404 to 1,529 dry lb/hr per incinerator with an overall average for the three years of 1,465 dry lb/hr. This average annual feed rate is very similar to the rated design capacity of the MHFs. The capacity of a MHF is dependent on the percent solids content of the feed biosolids. In general at higher percent feed solids, greater throughput capacity can be achieved. Using the present annual average percent solids, 26.3 percent, the design capacity of a 9 hearth MHF with 14 ft - 3 inch outside diameter is approximately 1,460 dry lb/hr. Thus, both incinerators are operated at their design capacity and the incinerators have been able to process the solids generated at the plant with one MHF operating on a nominal 5 day per week schedule. The second MHF has served as a standby unit that has insured incineration capability during maintenance periods or unexpected downtime of the operating unit. Unless the Post Point WWTP has alternative means of solids disposal, it is recommended that the plant always have one incinerator available as a standby unit except for the future maximum month solids loading when two incinerators in operation would be acceptable.

The natural gas usage for each incinerator from 2006 to 2008 is also shown in Table 1. The following observations are made regarding gas usage in the incinerators:

1. The annual total gas usage appears to be decreasing since 2006 that had the highest gas usage at 21.8 million cubic feet per year (cu ft/yr). Gas usage was less in 2007 at 17.1 million cu ft/yr, and gas usage in 2008 will likely be approximately 13 million cu ft/yr assuming the same rate of gas usage continues for the remainder of the year.

2. Standby gas usage, is defined as the gas used to maintain the incinerators on hot standby. Standby gas usage is a significant percentage of the total gas usage ranging from 25% to a high of 38% of the total gas usage for incineration.

3. The solids processing gas is the total gas used minus the standby gas and it represents the gas used when the incinerator is actually burning solids. For all 3 years the solids processing gas per dry ton of solids processed for Incinerator 2 is significantly greater than for Incinerator 1. Since the major portion of the gas usage in each incinerator is for the afterburner, the significantly higher gas usage for Incinerator 2 is thought to be attributable to the less efficient configuration of the large, separate chamber afterburner on Incinerator 2 versus the top hearth afterburner on Incinerator 1.

Incinerator Capacity Assessment

At this time there is no engineering study or projection of what the solids loading to the incinerators will be in 15 to 20 years. A facilities plan for the Post Point WWTP is presently underway and will eventually supply these future design loads. Presently the only quantitative projection of future loadings to the plant is a recent graphic showing the future growth of average annual BOD loading presented in a White Paper prepared by Carollo Engineering to the Bellingham City Counsel. The projected BOD loading graph is shown in Figure 2. Since this is the only future loading projection available, it was used to estimate the future solids loading to the incinerators. Figure 2 shows an average influent BOD loading of 21,000 lb of BOD /day in 2007 and an average BOD loading of 32,000 lb of BOD /day in 2026. If the present solids quantity of 4,250 dry tons /yr (DTPY) is increased proportionately to the BOD loading, the projected annual average solids quantity to incineration in 2026 will be 6,480 DTPY (32,000/21,000 x 4,250 = 6,480). If the same 5 day/week operating schedule is used (5,850 hr/yr), then the average annual loading to the incinerators would be 2,215 dry lb/hr (6480 X 2000/5850 = 2,215) versus the existing 1,465 dry lb/hr loading. Since this future loading is significantly greater than the design capacity of each incinerator, operation of two incinerators would be required for a considerable portion of the year and the plant would be without a standby unit during that time.

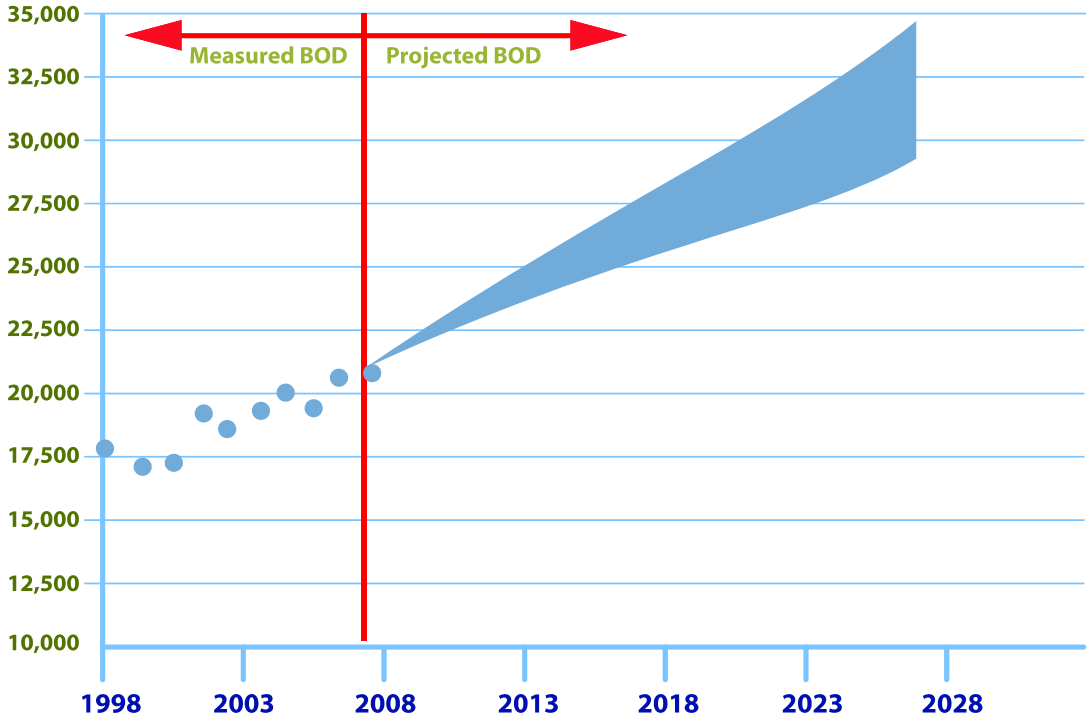

Figure 2. BOD Loadings

To ensure reliable solids incineration capability, it is recommended that one incinerator be able to process the future annual average solids production (6,480 DTPD) while the second incinerator serves as a standby unit. If the plant adopts a nominal 7 day/week operating schedule (8,234 hr/yr assuming a 94% availability), then the average annual incinerator loading would be 1,574 dry lb/hr per incinerator (6,480 x 2000/8,234), which is greater than the rated design capacity of each of the MHFs (1460 dry/lb/hr). Thus, based on the above operating schedule, operation of both incinerators would be required approximately 8 percent of the time (1574/1460 = 1.08) or about 659 hours/yr (0.08 x 8,234hr/yr) or 2.3 days per month (659 hr/yr x day/24 hr x yr/12 months). Although the incinerators could be operated on such a schedule, it would be a very inefficient operating schedule since every month one MHF would have to be heated up which takes about 2 days and requires a significant amount of fuel. It would be preferable to have a new incinerator with larger capacity available by the year 2020 to comfortably handle the annual average solids production at that time.

Multiple Hearth Improvement Alternatives

Capturing waste heat from a MHF is possible and in some cases the necessary retrofits can pay for themselves in a few years. Waste heat can be recovered in the form of steam or hot water. The challenge of recovering energy from an incinerator is finding a continuous use for the steam or hot water. While it is possible to generate superheated steam and produce electric power, the capital invest for the equipment (i.e. high pressure heat recovery boiler, turbine generator, condenser, pumps, piping, etc.) and the additional operating costs (for licensed boiler operator, boiler water treatment, and equipment maintenance) make this alternative economically unfavorable particularly for small incineration plants and relatively low electricity rates. A more feasible and attractive energy recovery alternative is to recover energy in the form of hot water and use it for building heat. This could be accomplished in two ways at the Post Point plant, namely by using an economizer to recover energy from the incinerator flue gas or by installing a heat exchanger to recover energy from the incinerator scrubber water. These two energy recovery alternatives are described and evaluated below, which are followed by an evaluation of two other alternatives - conserving energy with flue gas recirculation and installing a fluidized bed incinerator.

Alternative 1: Provide Economizer to Recover Energy

A schematic diagram showing how an economizer would be incorporated into the MHF incinerator is shown in Figure 3. The economizer would be installed between the afterburner and the venturi scrubber, and it would recover approximately 4.4 MMBH (million BTUs per hour) of thermal energy as hot water. The hot water would be used for plant heating. At the Post Point Plant, the Solids Handling Facility has a package boiler that supplies hot water to a hot water distribution system for heating of the Administration Building, Maintenance Shop, and the Solids Handling Facility. The estimated natural gas usage in this boiler for heating of the above buildings is 66,000 therms per year. Note that the heating season in Bellingham is approximately 6 months per year and for the remainder of the year the hot water could not be utilized.

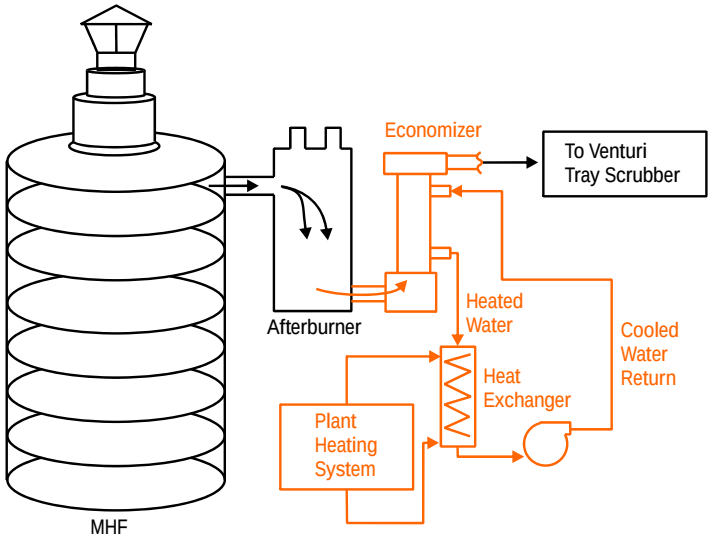

Figure 3. Economizer Diagram

An economic evaluation of the economizer alternative is presented in Table 2. The total estimated construction cost is $1,413,000 and the total annual O&M cost is estimated at $24,000. As stated above, the energy usage for heating of the Solids Handling Facility and adjacent buildings is estimated at 93,000 therms per year. At an average natural gas cost of $0.85/therm, the potential annual energy savings is estimated at $79,000. Subtracting the annual O&M cost from the energy savings yields the net energy savings which amounts to $61,000 per year. The capital investment payback period is obtained by dividing the total estimated construction cost by the net savings which yields a payback period of 23 years to recoup the initial investment. An investment with a payback period of 23 years is not economically justifiable. Another drawback of this alternative is that the economizer, heat exchanger, and pumps all must stay in operation whenever the incinerator is operating. Hence the heat recovery system must be operated throughout the warm weather months

when the recovered energy could not be utilized. Based on the above analysis, energy recovery in a hot water economizer is not recommended.

CONCEPTUAL CONSTRUCTION COSTS		
Economizer (Hot Water Generator) - Uninstalled Vendor's Cost	$260,000	
Heat Recovery Heat Exchanger No. 1	$25,000	
Cooling Heat Exchanger No. 2	$40,000	
Heat Recovery Recirculation Pumps (2)	$40,000	
Nitrogen Expansion Tank and Charging Bottle	$15,000	
Heat Recovery Make-Up Pump	$10,000	
Chemical Water Treatment Bypass Shot Feeder	$15,000	
Air Separator	$5,000	
Subtotal - Purchased Equipment Cost	$410,000	
Installation of Purchased Equipment @ 50%	$205,000	
Mechanical Piping @ 50%	$205,000	
Electrical Equipment and Materials @ 10%	$41,000	
Instrumentation and Controls @10%	$41,000	
Additional Plant Water Supply to Cooling Heat Exchanger No. 2	$50,000	
Construction Subtotal	$952,000	
Contractor Overhead & Profit (12%)	$114,000	
Subtotal	$1,066,000	
Contingency (25%)	$267,000	
Subtotal	$1,333,000	
Escalation to Mid Point of Construction at 4.0% per year ($1.04^{1.5}$)	$80,000	
TOTAL ESTIMATED CONCEPTUAL CONSTRUCTION COST	$1,413,000	
ANNUAL O&M COSTS		
Operating Labor		
No additional staff required	$0	
Maintenance Labor		
No additional staff for required	$0	
Maintenance Materials (3% of Purchased Equipment Cost)		
(0.03 X $410,000)	$12,000	
Power for Heat Recovery System		
(8.2 operating kw x 5850 hr/yr x $0.053/kwhr)	$3,000	
Chemicals for Water Treatment	$3,000	
TOTAL ANNUAL O&M COST		$18,000
ENERGY SAVINGS		
Avoided Natural Gas Heating Cost @ $0.85/therm		
(93,000 therms/yr x $0.85/therm)		$79,000
NET SAVINGS (Energy Savings - Annual O&M Cost)		$61,000
CAPITAL PAYBACK PERIOD IN YEARS		**23**

Table 2: Evaluation of Energy Recovery Using an Economizer

Alternative 2: Recover Energy from Venturi Scrubber Water

Energy can also be recovered from the MHF's venturi scrubber water. This process is shown schematically in Figure 4. The venturi water at approximately 180 deg F would be collected in an insulated tank and then pumped through a spiral heat exchanger and then back to the venturi scrubber for reuse. In the spiral heat exchanger the venturi water would be cooled from 180 to 140 deg F while water from the plant heating system would be heated from 140 to 170 deg F. It does not appear than the scrubber water would have enough energy to fulfill the total heating load of the Solids Handling Building, Administration Building and Maintenance Shop. However, the scrubber water would be able to handle a significant portion of the total heat load.

It should be noted that presently at the Post Point Plant the venturi water and tray scrubber water are mixed that results in a large scrubber water flow at too low a temperature to recover any heat. Therefore, for this alternative to be implemented, the venturi and tray waters would have to be separated. This could be done by routing the venturi water from the sump in the bottom of the tray scrubber to the energy recovery system and sending the tray water by gravity back to the plant headworks. An advantage of this system over the economizer energy recovery system is that it would not have to be operated during the warm weather months. This system could be designed with bypass piping and valves that would send the venturi water to the plant headworks when building heating was not needed.
The disadvantage of this system is that because the venturi water is continually recycled, the water stream continually collects acid gases (sulfur dioxide and hydrogen chloride) from the incinerator flue gas. The acid gases are converted to sulfuric acid and hydrochloric acid in the scrubber water and the scrubber water quite rapidly becomes very acidic. To prevent corrosion of the scrubber recycle loop, caustic is added to neutralize the scrubber water. This process is used at the Upper Blackstone Water Pollution Abatement District in Millbury, MA and the plant continually adds caustic to the scrubber water recycle loop when the system is in operation.

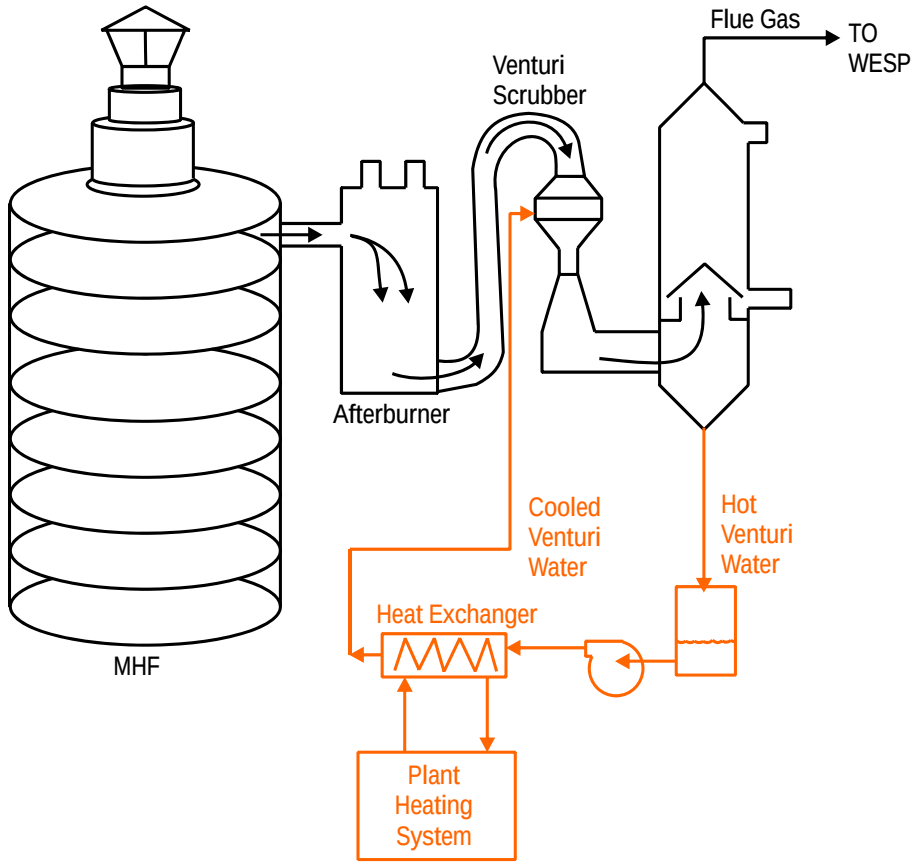

Figure 4. Venturi Scrubber Water Energy Recovery Diagram

An economic evaluation of the scrubber water energy recovery alternative is presented in Table 3. The total estimated construction cost is $530,000, and the total annual O&M cost is $13,000. Note that a significant portion of the annual O&M cost is for caustic to neutralize the recycled scrubber water. Until more precise figures are available for the heating load for the three buildings serviced by the hot water heating system and also the water temperatures throughout this system, we will assume that the scrubber water can fulfill 75 percent of the heating load of the above three buildings. Thus, the energy savings is estimated at 69,750 therms/yr and the potential annual energy savings is $59,000 (93,000 therms/yr x 0.75 x $0.85/therm). The net energy savings is $46,000 ($59,000 - $13,000) which results in a capital payback period of 12 years ($529,500/$46,000/yr). From a private industry's point of view, a payback period of 12 years would not be worth considering. However, from the perspective of a municipality, that has lower interest rate bonds available and the possibility of energy conservation grant funding, a 12 year payback period is worthy of further consideration. However, a 12 year payback corresponds to installing a new incinerator by the year 2020. Installing a venturi heat recovery system should be coordinated with the replacement of the incinerator system to allow for the payback of the investment. For example, the heat recovery system could be installed on the MHF that will remain even with a new incincerator.

CONCEPTUAL CONSTRUCTION COSTS		
Insulated Scrubber Water Tank	$20,000	
Heat Recovery Recirculation Pumps (2)	$30,000	
Spiral Heat Exchanger	$60,000	
Scrubber Modification - separate venturi water from tray water	$10,000	
Caustic Storage Tank - Insulated with Heat Pads	$25,000	
Caustic Metering Pumps (3)	$10,000	
Subtotal - Purchased Equipment Cost	$155,000	
Installation of Purchased Equipment @ 50%	$77,500	
Mechanical Piping @ 50%	$77,500	
Electrical Equipment and Materials @ 10%	$15,500	
Instrumentation and Controls @20%	$31,000	
Construction Subtotal	$356,500	
Contractor Overhead & Profit (12%)	$43,000	
Subtotal	$399,500	
Contingency (25%)	$100,000	
Subtotal	$499,500	
Escalation to Mid Point of Construction at 4.0% per year ($1.04^{1.5}$)	$30,000	
TOTAL ESTIMATED CONCEPTUAL CONSTRUCTION COST	$529,500	
ANNUAL O&M COSTS		
Operating Labor		
No additional staff required	$0	
Maintenance Labor		
No additional staff required	$0	
Maintenance Materials (3% of Purchased Equipment Cost)		
(0.03 X $155,000)	$5,000	
Power for Heat Recovery System		
(4.2 operating kw x 5850 hr/yr x $0.053/kwhr)	$1,000	
Caustic for Scrubber Water Neutralization	$6,500	
TOTAL ANNUAL O&M COST		$13,000
ENERGY SAVINGS		
Avoided Natural Gas Heating Cost @ $0.85/therm		
(93,000 therms/yr x 0.75 x $0.85/therm)		$59,000
NET SAVINGS (Energy Savings - Annual O&M Cost)		$46,000
CAPITAL PAYBACK PERIOD IN YEARS		**12**

Table 3: Evaluation of Energy Recovery Using Incinerator Scrubber Water

Alternative 3: Conserve Energy with Flue Gas Recirculation

Another MHF modification alternative is flue gas recirculation. As shown in Figure 5, flue gas recirculation consists of recycling a portion of the flue gas from the top hearth to lower hearths (Figure 5). Recirculating the flue gas increases turbulence in the furnace, and thus improves mixing, provides a more uniform temperature profile in the furnace, and increases

combustion efficiency. The increased turbulence and lower combustion hearth temperatures also reduce clinker formation and provide increased capacity while decreasing fuel usage. In addition, flue gas recirculation decreases CO, NOX, and VOC emissions. Flue gas recirculation can provide some of the benefits of a FBI, without completely replacing the MHF. Flue gas recirculation systems were installed on two MHFs at the Upper Blackstone Water Pollution Abatement District in Millbury, MA. The units have been in operation for over two years and have achieved the benefits cited above. Since Incinerator 2 uses considerably more fuel than Incinerator 1, installing flue gas recirculation on Incinerator 2 was evaluated for this study. With flue gas recirculation a top hearth temperature of 1200 deg F could be continuously maintained on Incinerator 2. Thus, the need for firing the afterburner would be eliminated and a significant fuel reduction could be achieved. Another benefit would be further reduction in slag and clinker formation in Incinerator 2.

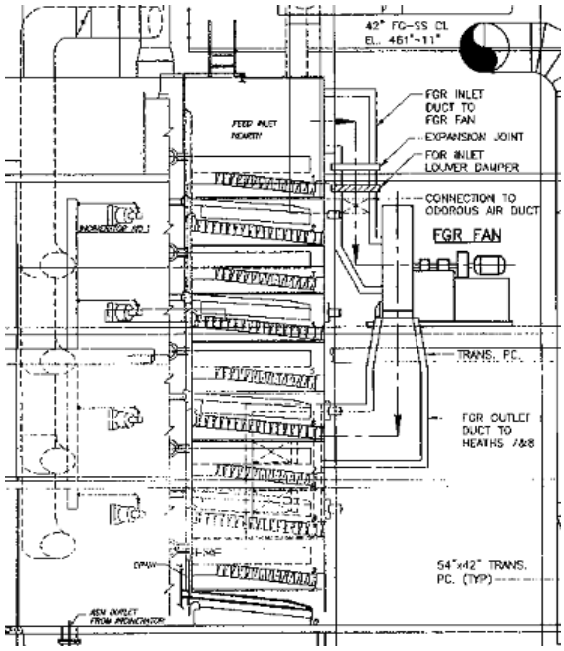

Figure 5. Flue Gas Recirculation Diagram

The CDM computer program, BURN, was used to model Incinerator 2 with flue gas recirculation. The computer output is contained in the Appendix. The resulting fuel use is 0.92 MM Btu/hr which is 65% less than the typical Incinerator 2 fuel use (2.66 MM Btu/hr, excluding the standby fuel). The implementation of flue gas recirculation on Incinerator 2 would require installing a refractory lined duct and an insulated fan capable of operating at 1300 deg F. The ductwork and fan would be quite heavy and a significant amount of structural steel framing would be required to support this equipment. An economic evaluation of flue gas recirculation on Incinerator 2 is presented in Table 4.

The total estimated construction cost is $1,326,800 and the annual O&M cost is estimated at $23,000. The energy savings in avoided natural gas usage is 1.74 MM Btu/hr (2.66 – 0.92 MM Btu/hr). Assuming Incinerator 2 would be operated half of the total incineration hours, the annual reduction in fuel usage would be 50,900 therms/yr (1.74 MM Btu/hr x 10 therms/MM Btu x 5,850 hr/yr x 0.50) and the annual fuel savings would be $43,000. Subtracting the annual O&M cost from the fuel savings yields a net annual savings of $20,000. As shown in Table 4, the capital payback period for this alternative is 66 years, which is not economically justifiable. Therefore, flue gas recirculation is not recommended.

CONCEPTUAL CONSTRUCTION COSTS		
Flue Gas Recirculation System - Uninstalled Vendor's Cost includes:	$524,000	
Refractory lined ducts, insulated fan, expansion joints, dampers		
Subtotal - Purchased Equipment Cost	$524,000	
Installation of Purchased Equipment @ 50%	$262,000	
Electrical Equipment and Materials @ 10%	$52,400	
Instrumentation and Controls @10%	$52,400	
Construction Subtotal	$890,800	
Contractor Overhead & Profit (12%)	$107,000	
Subtotal	$997,800	
Contingency (25%)	$249,000	
Subtotal	$1,246,800	
Escalation to Mid Point of Construction at 4.0% per year ($1.04^{1.5}$)	$80,000	
TOTAL ESTIMATED CONCEPTUAL CONSTRUCTION COST	$1,326,800	
ANNUAL O&M COSTS		
Operating Labor		
No additional staff required	$0	
Maintenance Labor		
No additional staff for required	$0	
Maintenance Materials (3% of Purchased Equipment Cost)		
(0.03 X $524,000)	$16,000	
Power for FGR Fan		
(22 operating kw x 5850 hr/yr x $0.053/kwhr)	$7,000	
TOTAL ANNUAL O&M COST		$23,000
ENERGY SAVINGS		
Avoided Natural Gas Heating Cost @ $0.85/therm		
(50,900 therms/yr x $0.85/therm)		$43,000
NET SAVINGS (Energy Savings - Annual O&M Cost)		$20,000
CAPITAL PAYBACK PERIOD IN YEARS		**66**

Table 4: Evaluation of Energy Recovery Using Flue Gas Recirculation

Alternative 4: Install a Fluidized Bed Incinerator

As previously mentioned, the Post Point Plant can handle its solids production with one MHF operating 5 days per week at its rated capacity of 1460 dry lb/hr. Within 5 years, to meet the plant's increased solids production, one incinerator will have to be operated 6 days per week or alternatively two incinerators could be operated for a few days per month. Both of these options have their drawbacks. The first option will require adding an additional incinerator operator to the staff. The second option will require a costly heat up of an incinerator for only a few days of operation per month. Thus, the plant may want to consider installing a FBI in approximately 5 years from now to handle the increased solids production and still maintain a 5 day per week operating schedule. Since there is limited space in the Solids Handling Facility, this alternative evaluates replacing one of the existing MHFs with a new FBI. The advantages of this alternative are the following:

- Ability to meet the future solids production with just the FBI operating
- Allows for one of the existing MHFs to serve as a standby unit
- Significant reduction in natural gas fuel usage
- Elimination of standby fuel usage on weekend

A process flow diagram of a Fluidized Bed Incineration System (FBIS) is shown in Figure 6. A brief description of a FBIS follows. A FBI can be thought of as a completely mixed process in which drying and combustion take place concurrently and very rapidly, within a few seconds. The combustion related components of a FBIS consist of the following items: the reactor, combustion air heat exchanger or preheater, and the fluidizing air blower. The reactor is a refractory-lined vessel containing three zones: the windbox, the sand bed, and the freeboard. Preheated combustion air supplied by the fluidizing air blower is blown into the windbox and distributed through nozzles or tuyeres to the bottom of the bed. The combustion air fluidizes the sand bed. The sand bed is initially heated to 1400 deg F with a preheat burner. Then fuel (natural gas or fuel oil) and dewatered biosolids are pumped directly into the sand bed to create a hot, turbulent suspension of sand, gases, and burning biosolids. In the 1400°F to 1500 deg F suspension, the water in the biosolids is evaporated and the combustible matter oxidized in a matter of seconds. The combustion gases rise through the sand bed and enter the freeboard where the burnout of volatilized organics is completed. The freeboard is a large open space above the bed which provides a 5 to 6 second gas residence time at temperatures of 1500 to 1600 deg F; hence it acts like an afterburner.

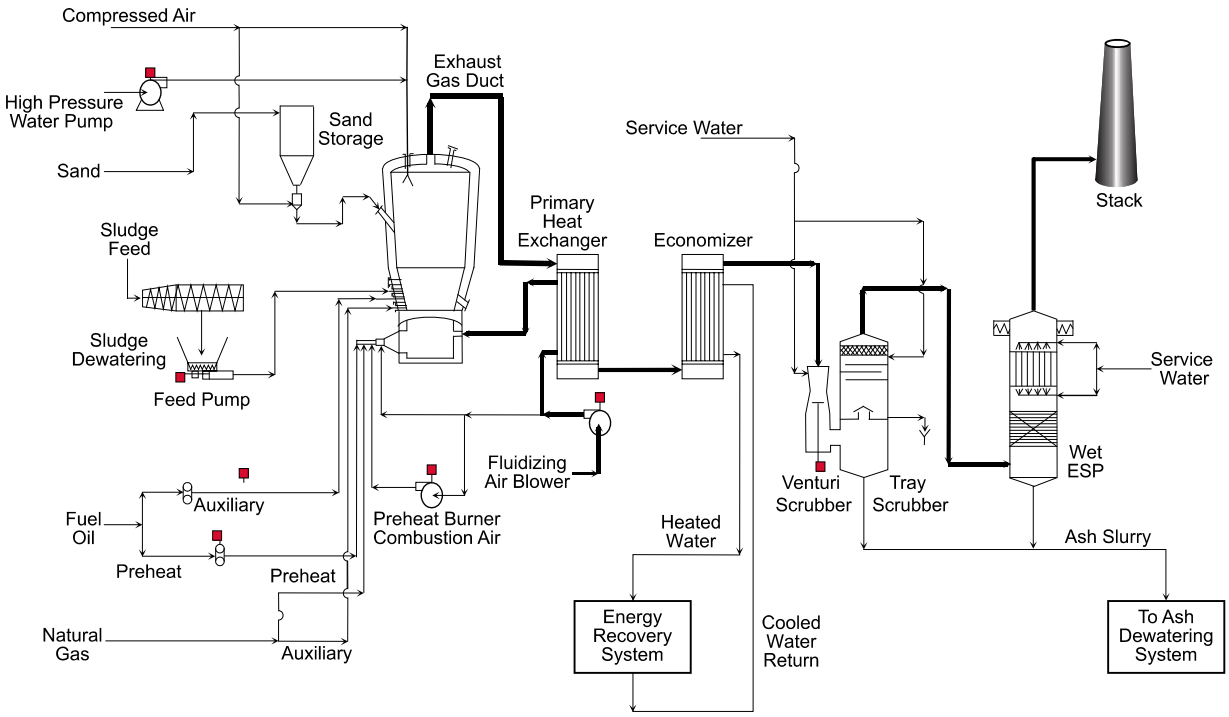

Figure 6: Process Flow Diagram of Fluidized Bed Incineration System

The reactor flue gas then proceeds to the combustion air preheater in which the flue gas flows countercurrent to the combustion air. The combustion air is typically heated to approximately 1200 deg F while the flue gas is cooled to 1000 deg F. The combustion air preheater greatly improves the thermal efficiency of the FBIS by recovering a large portion of the thermal energy in the incineration flue gas and returning it to the reactor. The combustion air preheater makes it possible to burn biosolids with low solids content or low heating value with minimal use of auxiliary fuel.

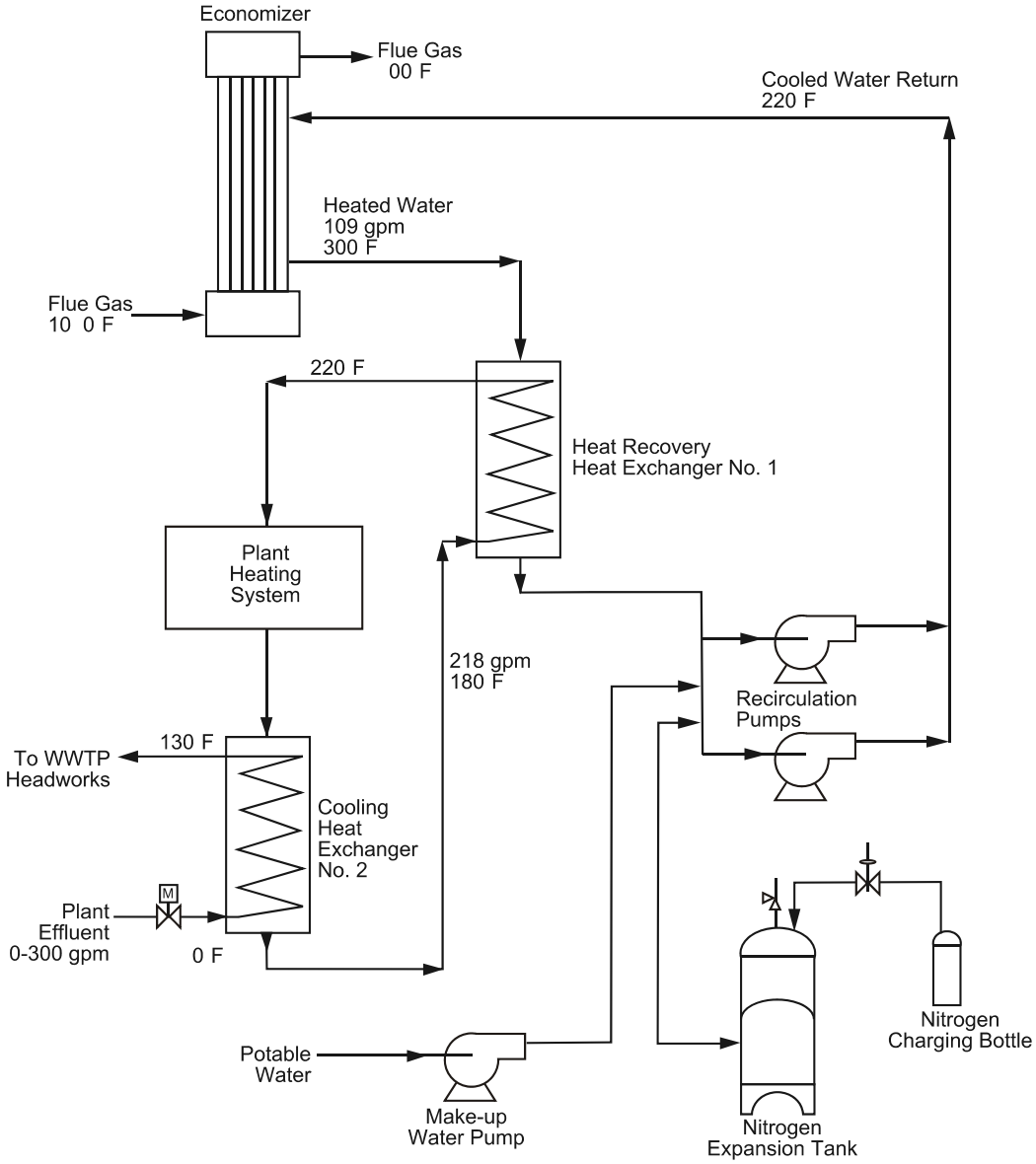

Figure 7: Process Flow Diagram of Energy Recovery System Using an Economizer

Following the air preheater, the flue gas can be sent to an economizer which could generate hot water for building heating. A process flow diagram of an economizer energy recovery system is shown in Figure 7. The merit of including an economizer in the FBIS would have to be evaluated with the overall cost and financing of the FBIS project. Following the economizer the flue gas would enter the air pollution control systems consisting of a venturi scrubber, a tray scrubber, and a wet electrostatic precipitator (WESP). Note that with a FBI all the incineration ash exits the top of the reactor with the flue gas and is subsequently collected in the venturi scrubber water. The resultant ash slurry requires dewatering before it can be

hauled to a landfill. Two types of ash dewatering systems are commonly used: an ash settling basin or a mechanical dewatering system. Given the limited space on the Post Point Plant site, an ash settling basin would not be possible. Therefore, an a mechanical ash system consisting of an ash thickener, ash slurry pumps, a small belt filter press or vacuum filter and a truck loading station would be required. A schematic diagram of a mechanical ash system is shown in Figure 8.

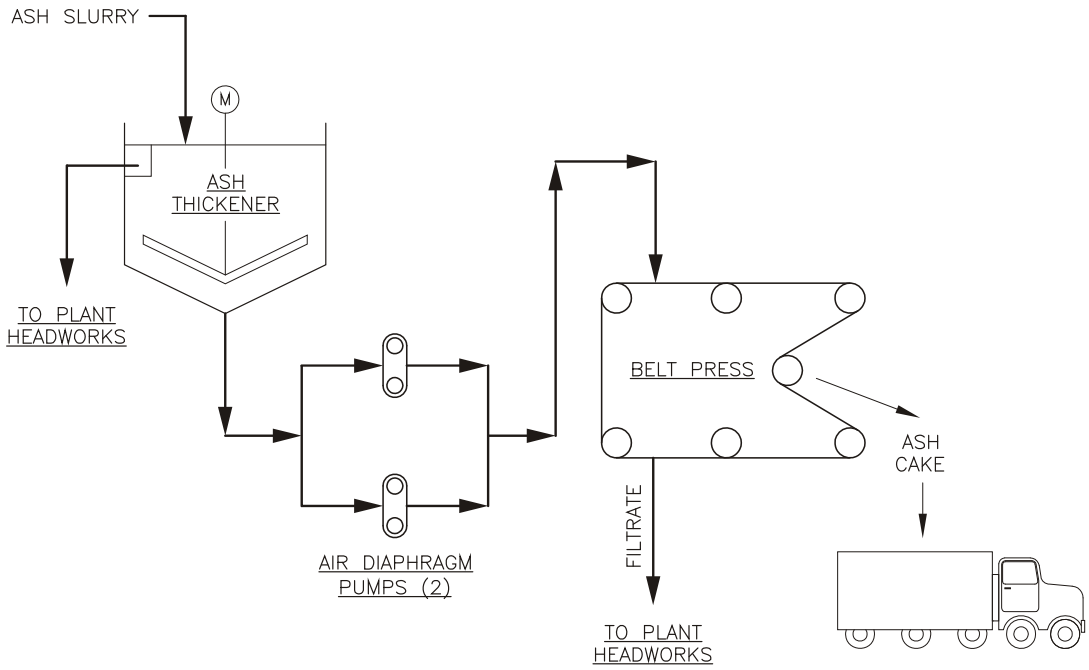

Figure 8: Process Flow Diagram of Ash Dewatering System

A preliminary construction cost estimate of a FBIS to replace one of the MHFs is presented in Table 5. Note that a 2000 dry lb/hr (24 DTPD) FBIS was selected to meet the 2026 solids production with an approximate 5 day per week operating schedule. The estimate construction cost is $16.6 million. Although the FBIS would be capable of operating for short periods without fuel usage, a small amount of fuel is typical used to make the operation of the incinerator easier and more stable (less likely to go through temperature swings). The annual fuel cost is estimated at about $40,000 per year that would result in an annual fuel savings of approximately $130,000 from the present day fuel cost. The FBIS would also enable the plant to maintain the present incinerator operating staff of 3 operators to cover the 5 day/week, 24 hour /day schedule. While a FBIS would result in fuel and labor savings, these savings alone would not justify the capital cost of a new FBI. The primary reason for installing a new FBIS would be to provide assured solids incineration capability in the years to come. This

alternative warrants further investigation. In particular the layout of the FBIS and ash handling system should be further developed at the Post Point WWTP site.

PRELIMINARY CONSTRUCTION COSTS	
Fluidized Bed Incineration System - Uninstalled Vendor's Cost	
Includes: 2000 dry lb/hr hot windbox fluid bed incinerator, refractory, primary heat exchanger, venturi scrubber, tray scrubber, WESP, fluidizing air blower, oil injector purge air blower, oil pumps, preheat burner, preheat burner blower, high pressure water pumps, oxygen analyzer, MCC, instrumentation & controls, PC & PLCs, interconnecting ductwork & piping, start-up & testing	$6,500,000
Ash handling system	
ash slurry pumps, ash thickener, belt filter press, truck loading station	$600,000
Subtotal - Purchased Equipment Cost	$7,100,000
Installation of FBIS (20%)	$1,420,000
Electrical wiring & conduit installation (10%)	$710,000
Mechanical piping installation & piping outside FBIS scope (10%) service water, potable water, scrubber water, ash slurry, oil, fluidizing air, compressed air, sludge feed to FBI, sand	$710,000
Continous emissions monitoring system	$100,000
Demolition of MHF, afterburner, scrubber, emergency bypass stack, fans, ash system, platforms	$700,000
New floor and roof openings, structural steel, equipment pads	$400,000
Construction Subtotal	$11,140,000
Contractor - overhead & profit (12%)	$1,340,000
Subtotal	$12,480,000
Contingency (25%)	$3,120,000
Subtotal	$15,600,000
Escalation to Mid Point of Construction at 4.0% per year ($1.04^{1.5}$)	$950,000
TOTAL ESTIMATED PRELIMINARY CONSTRUCTION COST	$16,550,000

Table 5: Fluidized Bed Incineration System - Preliminary Construction Cost Estimate

Conclusions and Recommendations

The following conclusions and recommendations are made:

1. Adding an economizer on either of the existing MHFs is not economically justifiable and therefore is not recommended.

2. Installing flue gas recirculation on either of the existing MHFs is not economically justifiable and therefore is not recommended.

3. Installing a system to recover energy from the venturi scrubber water of the existing MHFs is possibly economically justifiable and warrants further consideration. A detailed evaluation of this alternative is recommended.

4. Installing a new FBIS would cost approximately $16.6 million. Although a new FBIS would provide fuel and labor cost savings, these savings alone can not justify the capital expenditure. The primary motivation for installing a new FBIS would be to provide assured solids incineration capability in the future. It is recommended that a FBIS and ash handling system be further developed at the Post Point Plant.

5. A new FBIS is recommended at the Post Point Plant in the next 5 to 10 years. Further development of the FBIS at the Post Point Plant should start in 2009 to be prepared for a possible installation in 2013.

6. Energy conservation measures (in addition to venturi heat recovery) compatible with a new FBIS should also be evaluated in 2009.

Appendices

BURN - Version 4.01 COMBUSTION ANALYSIS: RUN ___ FOR _____ __/__/__

DATA FILE USED FOR THIS ANALYSIS: BELL07.in

WASTE FEED STREAMS

NAME	WEIGHT FIRED in Wet LB/Hr	PERCENT (DRY BASIS) Carbon	Hydrogen	Sulfur	Fe(OH)3	Al(OH)3	Oxygen
BELL07	7667.0	41.430	6.120	.930	00.000	00.000	22.540
COMPOSITE (LB)	7667.	762.35	112.61	17.11	.00	.00	414.75
COMPOSITE MOLS	0.	63.48	55.86	.53	.00	.00	12.96
COMPOSITE (% DRY BASIS)		41.43	6.12	.93	.00	.00	22.54

	PERCENT (DRY BASIS) Nitrogen	Chlorine	CaCO3	Inert	Iron	Aluminum	Bromine	Pct.H2O	BTU/LB
# 1	3.680	.300	00.000	25.000	00.000	00.000	00.000	76.000	8200.0
(LB)	67.71	5.52	.00	460.02	.00	.00	.00	5826.92	1968.0
MOLS	2.42	.16	.00	460.02	.00	.00	.00	323.72	
% DRY	3.68	.30	.00	25.00	.00	.00	.00		

	DRY BASIS		WET BASIS	
THE MODIFIED DULONG HEATING VALUE IS:	8532.8	BTU/LB	2047.9	BTU/LB
THE MODIFIED CHANG HEATING VALUE IS:	8428.1	BTU/LB	2022.7	BTU/LB
THE BOIE HEATING VALUE IS:	8325.8	BTU/LB	1998.2	BTU/LB
THE MODIFIED VONDRACEK HEATING VALUE IS:	6007.0	BTU/LB	1441.7	BTU/LB
THE AVERAGE ESTIMATED HEATING VALUE IS:	7823.4	BTU/LB	1877.6	BTU/LB
THE INPUT WASTE HEATING VALUE IS:	8200.0	BTU/LB	1968.0	BTU/LB

DAILY CHARGE RATE EQUALS: 92.0 TONS PER 24-HOUR DAY.

RUN CONDITIONS AS INPUT

AMBIENT AIR: 60.0 DEG. F ; PRESSURE 1.0 ATM; ABSOLUTE HUMIDITY .007500
AVAILABLE PREHEATED AIR 10038.3 ACTUAL CFM AT 1100.0 DEG. F
USING AIR PREHEATER FOLLOWING PRIMARY FURNACE OR AFTERBURNER
OPERATING TEMPERATURES: MINIMUM OF 900.0, MAXIMUM OF 1650.0 DEG. F
FURNACE NOT COOLED, 100.00 % OF AREA; BOILER NOT COOLED, 100.00 % OF AREA
TEMPERATURES MODERATED WITH AIR AND ELEVATED WITH GAS
STEAM CONDITIONS: PRESSURE - 0. PSIA ; TEMPERATURE - 0. DEG. F
ENTHALPY CHANGE FROM FEEDWATER TO STEAM: 80.0 BTU/LB
TEMPERATURE (DEG. F): PROCESS WATER 60. FEEDWATER 60.
FLUE GASES LEAVE THE BOILER AT: 400.0 DEG. F , QUENCHER AT 190.0 DEG. F
FLUE GASES LEAVE THE SUBCOOLER AT: 110.0 DEG. F
MAXIMUM SUBCOOLER WATER DISCHARGE TEMPERATURE IS: 150.0 DEG. F
STACK DIAM. IS 2.0 F, HEIGHT 150.0 F, VELOCITY = .0 FT/SEC
0. BTU/HR IS ABSORBED IN THE PRIMARY COMBUSTION CHAMBER
RESIDUE IS NOT QUENCHED AND LEAVES SYSTEM AT 1550.0 DEG. F
UNBURNED PERCENTAGES OF FEED - CARBON 2.4, IRON 00.0, ALUMINUM 00.0
AFTERBURNER TEMPERATURE: .0 DEG. F ;OPERATING FACTOR: 14.00 % OF DESIGN
ATMOSPHERIC STABILITY CLASS IS: 0; DESIGN % EXCESS AIR IS: 42.0

NOTE: GAS FLOW RATES EXPRESSED IN SCFM ARE AT 60 Deg. F AND 1.0 Atm.

SUMMARY OF FURNACE OPERATIONS

Furnace Flue Gas Sensible Heat Content (SENH) as a Function of Tgas

SENH = A + B*T + C*T*T + D*T*T*T

A = -.3954754E+06 C = .6699598E+00
B = .6551228E+04 D = -.4674763E-04

At Tgas = 1498.32 DEG. F , SENH = .1076715E+08 BTU/HR

GAS ANALYSIS AFTER FURNACE

COMPONENT	VOLUME % DRY BASIS	VOLUME % WET BASIS	MOLS PER MINUTE	LB/HR	
CO2	12.13	6.911	1.033	2726.5	
SO2	.1047	.5965E-01	.8913E-02	34.3	597. PPMV - WET
N2	81.37	46.36	6.926	11642.5	
O2	6.360	3.623	.5414	1039.4	
HCl	.3049E-01	.1737E-01	.2595E-02	5.7	174. PPMV - WET
HBr	.0000	.0000	.0000	.0	0. PPMV - WET
H2O		43.03	6.430	6944.1	
TOTAL	100.0	100.0	14.94	22392.5	

	PERCENT SO2 TO SO3	DEWPOINT DEG. F	EQUIVALENT SO3 ppmw	ppmd
SULFURIC ACID	1	298.35	6.	10.
DEWPOINT FROM	3	316.27	18.	31.
OXIDATION OF	5	324.89	30.	52.
SO2 TO SO3	8	333.00	48.	84.
AT THIS LOCATION	10	336.90	60.	105.
IN THE SYSTEM	15	344.10	89.	157.

EQUILIBRIUM SO3 (USUALLY NOT ATTAINED) AT 1498.3 DEG. F
EQUILIBRIUM CONSTANT FOR SO2+0.5O2-->SO3 IS: .748
EQUILIBRIUM SO3 IS THEN: 267. ppm (wet basis)

PREHEATER ANALYSIS

Allowing 1.5% of transferred heat for heat losses, the combustion & burner air preheater heats ambient air to 1100.0 DEG. F . In turn, the furnace (or afterburner) exit gas temperature drops from
1498.3 DEG. F to 987.5 DEG. F

TOTAL PREHEATED AIR SUPPLIES 4024717. BTU/HR TO THE PRIMARY FURNACE

PREHEATED AIR 10038.24 ACFM (ENTHALPY: 4024717. BTU/HR)
3344.66 SCFM 14159.67 LB/HR

(.0 PERCENT USED WITH FUEL AND/OR TO HOLD EXCESS AIR TARGET)

COMBUSTION AIR .00 ACFM
.00 SCFM .00 LB/HR
COOLING AIR .00 ACFM
.00 SCFM .00 LB/HR

COOLING WATER .00 GAL/MIN .00 LB/HR

WITHOUT COOLING OR FUEL USE BUT USING 10038. ACFM OF PREHEATED AIR, THE

FURNACE TEMPERATURE IS: 1498. DEG. F ; A TEMPERATURE OF 1135. DEG. F WAS USED TO JUDGE POTENTIAL DISSOCIATION OF $CaCO_3$, $Fe(OH)_3$, AND $Al(OH)_3$.

FLUE GAS	15776.95	ACFM	AT	987.5	DEG. F
	5665.46	SCFM	AT	60.0	DEG. F

BURNER FUEL USE .00 CFM (.00 FT3/HR) GAS
EQUAL TO .0 BTU/HR

QUENCH TANK MAKEUP .00 GAL/MIN

RESIDUE ASSUMED TO LEAVE HOT ZONE AT 1550.0 DEG. F
RESIDUE WEIGHT (75.00 % SOLIDS) 637.76 LB/HR
(DRY) 478.32 LB/HR

UNBURNED CARBON IN ASH: 3.822 PERCENT OF TOTAL ASH (INCLUDING CARBON)
HEATING VALUE OF RESIDUE (DRY BASIS): 539.2 BTU/LB OR 257905. BTU/HR

NET HEAT RELEASE (BTU/HR)

1. PRIMARY

FEED	14830750.
FUEL	0.
AIR HEAT	4024717.
TOTAL	18855470.

2. AFTERBURNER

FUEL	0.
AIR HEAT	0.
GRAND TOTAL	18855470.

HEAT LOSSES	BTU/HR	PERCENT OF FEED HEAT CONTENT	PERCENT OF TOTAL HEAT RELEASE
RADIATION	813410.	5.39 PERCENT	4.2 PERCENT
MOISTURE	7366593.	48.82 PERCENT	38.3 PERCENT
DRY GAS	6682065.	44.29 PERCENT	34.7 PERCENT
RESIDUE	296678.	1.97 PERCENT	1.5 PERCENT

DESIGN EXCESS AIR (ON FEED) IS 42.00 PERCENT
ACTUAL EXCESS AIR (ON FEED) IS 42.00 PERCENT
ACTUAL EXCESS AIR (ON TOTAL COMBUSTIBLE) IS 42.00 PERCENT

EQUILIBRIUM THERMAL NOX CONCENTRATION IS 26.3 PPM (VOLUME)
PERCENT FUEL NITROGEN CONVERTED TO NOx= 12.550 PERCENT
FUEL NITROGEN NOx (Estimated by Soete) = 676.715 PPM (VOLUME)

THE EQUILIBRIUM CONSTANT FOR $2HCl+.5O_2 \rightarrow Cl_2+H_2O$ IS: 2.0197
THE EQUILIBRIUM CONSTANT FOR $2HBr+.5O_2 \rightarrow Br_2+H_2O$ IS: 345.3506

EQUILIBRIUM CHLORINE CONCENTRATION AT 987.5 DEG. F IS:
.153 ppm (Wet Basis)
.269 ppm (Dry Basis)

SO_2 UNCONTROLLED EMISSION RATE IS	4.32	GM/SEC	EQUAL TO	34.22	LB/HR
HCl UNCONTROLLED EMISSION RATE IS	.72	GM/SEC	EQUAL TO	5.67	LB/HR
HBr UNCONTROLLED EMISSION RATE IS	.00	GM/SEC	EQUAL TO	.00	LB/HR

WITH ACID GAS CONTROL AT 85.0 PERCENT,

SO_2 CONTROLLED EMISSION RATE IS	.65	GM/SEC	EQUAL TO	5.13	LB/HR
HCl CONTROLLED EMISSION RATE IS	.11	GM/SEC	EQUAL TO	.85	LB/HR
HBr CONTROLLED EMISSION RATE IS	.00	GM/SEC	EQUAL TO	.00	LB/HR

	PERCENT SO2 TO SO3	DEWPOINT DEG. F	EQUIVALENT SO3 ppmw	ppmd
SULFURIC ACID	1	298.35	6.	10.
DEWPOINT FROM	3	316.27	18.	31.
OXIDATION OF	5	324.89	30.	52.
SO2 TO SO3	8	333.00	48.	84.
AT THIS LOCATION	10	336.90	60.	105.
IN THE SYSTEM	15	344.10	89.	157.

EQUILIBRIUM SO3 (USUALLY NOT ATTAINED) AT 987.5 DEG. F
EQUILIBRIUM CONSTANT FOR SO2+0.5O2-->SO3 IS: 29.955
EQUILIBRIUM SO3 IS THEN: 579. ppm (wet basis)

SUMMARY OF BOILER OPERATION CALCULATIONS

BOILER STEAM PRODUCTION 45670.9 LB/HR
PRESSURE .0 PSIA
TEMPERATURE .0 DEG. F

FEEDWATER TEMPERATURE: 60.0 DEG. F
FEEDWATER ENTHALPY: .0 BTU/LB
PRODUCT STEAM ENTHALPY: .0 BTU/LB
ENTHALPY CHANGE: 80.0 BTU/LB

NOTE: THE PERCENT OXIDATION OF FLUE GAS SO2 AT WHICH THE SULFURIC ACID DEWPOINT EQUALS THE FEEDWATER TEMPERATURE IS: .00 PERCENT.

PRODUCT STEAM USE TO HEAT CONDENSATE RETURN FROM 60. DEG. F TO FEEDWATER TEMPERATURE IS: .0 LB/HR

NET STEAM PRODUCTION AFTER FEEDWATER HEATING IS: 45670.9 LB/HR

NOTE!! - IF ACTUAL CONDENSATE RETURN IS ALREADY AT FEEDWATER TEMPERATURE, ADD BACK THE FEEDWATER HEATING STEAM USE TO THE NET STEAMING RATE!!

FLUE GAS TEMPERATURE AT BOILER EXIT 400. DEG. F

RADIATION LOSS 699174. BTU/HR OR 10.46 % OF SENSIBLE HEAT AT BOILER INLET

WITH REFERENCE TO TOTAL ENTHALPY INPUT TO THE COMBUSTION SYSTEM,
THE BOILER EFFICIENCY IS: 19.12 PERCENT

WITH REFERENCE TO FEED HHV ENTHALPY INPUT TO THE COMBUSTION SYSTEM,
THE BOILER EFFICIENCY IS: 23.89 PERCENT

MEAN MOLECULAR WEIGHT OF GASES
(DRY BASIS) 30.25
(WET BASIS) 24.99

TOTAL GAS FLOW RATE	LB/MIN	LB/HR	ACFM
(DRY BASIS)	257.47	15448.40	--
(WET BASIS)	373.31	22398.67	9378.6

EFFLUENT GAS HUMIDITY .4499 (MASS H20/MASS BONE DRY GAS)

GAS DEW POINT IS 172.4 DEG. F

SUMMARY OF QUENCHER OPERATIONS

QUENCHER EXIT TEMPERATURE 190.0 DEG. F

QUENCHER WATER USE 2.6 GAL/MIN

GAS ANALYSIS AFTER QUENCHER

COMPONENT	VOLUME % DRY BASIS	VOLUME % WET BASIS	MOLS PER MINUTE	LB/HR	
CO2	12.13	6.395	1.033	2726.5	
SO2	.1047	.5520E-01	.8913E-02	34.3	552. PPMV - WET
N2	81.37	42.89	6.926	11642.5	
O2	6.360	3.353	.5414	1039.4	
HCl	.3049E-01	.1607E-01	.2595E-02	5.7	161. PPMV - WET
HBr	.0000	.0000	.0000	.0	0. PPMV - WET
H2O		47.29	7.636	8246.5	
TOTAL	100.0	100.0	16.15	23694.9	

GAS WET BULB TEMPERATURE IS: 175.7 DEG. F

HUMIDITY AT WET BULB TEMPERATURE IS: .5406 MASS H20/MASS DRY FLUE GAS

HUMIDITY AT QUENCHER OUTLET IS: .5343 MASS H2O/MASS DRY FLUE GAS
MEAN MOLECULAR WEIGHT OF GASES
(DRY BASIS) 30.25
(WET BASIS) 24.46

TOTAL GAS FLOW RATE	LB/MIN	LB/HR	ACFM
(DRY BASIS)	257.47	15448.40	--
(WET BASIS)	395.04	23702.24	7659.7

EFFLUENT GAS HUMIDITY .5343 (MASS H20/MASS BONE DRY GAS)

GAS DEW POINT IS 176.5 DEG. F

	PERCENT SO2 TO SO3	DEWPOINT DEG. F	EQUIVALENT SO3 ppmw	ppmd
SULFURIC ACID	1	299.03	6.	10.
DEWPOINT FROM	3	316.76	17.	31.
OXIDATION OF	5	325.29	28.	52.
SO2 TO SO3	8	333.31	44.	84.
AT THIS LOCATION	10	337.17	55.	105.
IN THE SYSTEM	15	344.29	83.	157.

EQUILIBRIUM SO3 (USUALLY NOT ATTAINED) AT 987.5 DEG. F
EQUILIBRIUM CONSTANT FOR SO2+0.5O2-->SO3 IS: 29.955
EQUILIBRIUM SO3 IS THEN: 535. ppm (wet basis)

SUMMARY OF SCRUBBER OPERATIONS

GAS ANALYSIS AFTER SCRUBBER

COMPONENT	VOLUME % DRY BASIS	VOLUME % WET BASIS	MOLS PER MINUTE	LB/HR	
CO2	12.15	6.360	1.033	2726.5	
SO2	.1573E-01	.8235E-02	.1337E-02	5.1	82. PPMV - WET
N2	81.47	42.66	6.926	11642.5	
O2	6.368	3.335	.5414	1039.4	
HCl	.4578E-02	.2397E-02	.3892E-03	.9	24. PPMV - WET
HBr	.0000	.0000	.0000	.0	0. PPMV - WET
H2O		47.63	7.733	8352.1	
TOTAL	100.0	100.0	16.24	23766.6	

SCRUBBER EXIT TEMPERATURE 175.74 DEG. F (SATURATED)

SCRUBBER DISCHARGE ABSOLUTE HUMIDITY .5406 MASS H20/MASS DRY FLUE GAS

SCRUBBER NEUTRALIZATION REQUIREMENTS AT: 85.0 % COLLECTION EFFICIENCY
HYDRATED LIME (Ca(OH)2) 38.53 LB/HR
OR 100 PERCENT CAUSTIC SODA 41.66 LB/HR

SCRUBBER EVAPORATION RATE .2112 GAL/MIN

SUMMARY OF SUBCOOLER OPERATIONS

SUBCOOLER REDUCES GAS TEMPERATURE TO: 110. DEG. F

7446.92 LB/HR OF WATER IS CONDENSED

GAS ANALYSIS AFTER SUBCOOLR

COMPONENT	VOLUME % DRY BASIS	VOLUME % WET BASIS	MOLS PER MINUTE	LB/HR	
CO2	12.15	11.05	1.033	2726.5	
SO2	.1573E-01	.1430E-01	.1337E-02	5.1	143. PPMV - WET
N2	81.47	74.11	6.926	11642.5	
O2	6.368	5.793	.5414	1039.4	
HCl	.4578E-02	.4165E-02	.3892E-03	.9	42. PPMV - WET
HBr	.0000	.0000	.0000	.0	0. PPMV - WET
H2O		9.033	.8443	911.8	
TOTAL	100.0	100.0	9.346	16326.3	

FOR MAKEUP WATER AT 60.0 DEG. F AND MAXIMUM DISCHARGE AT 150.0 DEG. F
WATER NEEDED FOR THE SUBCOOLER IS: 175. GAL/MIN

	PERCENT SO2 TO SO3	DEWPOINT DEG. F	EQUIVALENT SO3 ppmw	ppmd
SULFURIC ACID	1	244.02	1.	2.
DEWPOINT FROM	3	262.50	4.	5.
OXIDATION OF	5	271.43	7.	8.
SO2 TO SO3	8	279.84	11.	13.
AT THIS LOCATION	10	283.90	14.	16.
IN THE SYSTEM	15	291.40	21.	24.

EQUILIBRIUM SO3 (USUALLY NOT ATTAINED) AT 987.5 DEG. F
EQUILIBRIUM CONSTANT FOR SO2+0.5O2-->SO3 IS: 29.955
EQUILIBRIUM SO3 IS THEN: 140. ppm (wet basis)

SUMMARY OF STACK REHEATING OPERATION
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TARGET STACK TEMPERATURE IS: .0 DEG. F

NO STACK REHEAT ANALYSIS REQUESTED.

SUMMARY OF STACK CALCULATIONS AFTER SYSTEM

STACK DIAMETER OF 2.00 FEET USED FOR CALCULATIONS

NATURAL DRAFT 1.918E-01 IN H2O
FRICTION LOSS 4.471E-01 IN H2O
VELOCITY HEAD 8.944E-03 IN H2O
MINIMUM FAN PRESSURE 2.643E-01 IN H2O
EXIT VELOCITY 20.6 FT/SEC

TOTAL FLOW @ STACK CONDITIONS 3878.8 CFM

STACK TEMPERATURE IS: 109.1 DEG. F

FLOW CORRECTED TO 12% CO2 (DRY, 1 ATM, 68 F/20 C)) 3315.0 CFM
FLOW CORRECTED TO 7% O2 (DRY, 1 ATM, 68 F/20 C)) 3422.5 CFM

MEAN MOLECULAR WEIGHT OF GASES
(DRY BASIS) 30.22
(WET BASIS) 29.12

TOTAL GAS FLOW RATE	LB/MIN	LB/HR	ACFM
(DRY BASIS)	256.91	15414.45	--
(WET BASIS)	272.12	16327.08	3887.5

EFFLUENT GAS HUMIDITY .0592 (MASS H20/MASS BONE DRY GAS)

GAS DEW POINT IS 111.3 DEG. F

STEAM TEMPERATURE & PRESSURE VALUES ARE NEEDED TO EVALUATE REHEAT USING
MAIN BOILER STEAM. AUXILIARY BOILER (Sat @ 21 Atm.) WILL BE USED.

CALCULATIONS COMPLETE

City of Bellingham

CDM