CHAPTER 6

PROCESS OPTIMIZATION

6.1 Introduction

The chapter focuses on the optimization study of the model after it had been reconciled with actual plant data. The optimization objective function is presented later on in this chapter. A sensitivity analysis would then be carried out to check the result, and examine the sensitivity of changes to the model. The optimization study in this research was made to focus on the gain of ethylene recovered from acetylene, and also to find and achieve the optimum condition.

6.2 Optimization

The steps used to solve optimization problems are as follows:

Step 1. Analyze the process and list all the variables in the process

The acetylene hydrogenation unit is a unit in which acetylene hydrogenation to ethylene occurred, catalyzed by Palladium-based on Al_2O_3, the reaction yield ample than necessary.

The variables in the reactor unit are listed as follows:

For feed stream:

Feed flow rate, F_{in} (kg/hr)

Feed temperature, T_{in} (kelvin)

Feed pressure, P_{in} (kg/cm^2)

Feed composition X_{in}, (mole) (x refers to each component)

However, since all the parameters used in this model had been fine tuned using data-fit and data reconciliation already, the variables in the reactor would not be varied in the optimization model. As for all 2 heat exchangers the hot stream outlet temperature is varied with a heat duty maximum constrain for each unit.

Step 2. Determination of the criterion for optimization and specifying the objective function in terms of the above variables together with coefficient.

Formulating of the specific optimal control problem is related to the overall economic optimization of the acetylene reactor performance. A performance index is developed addressing the objectives of the acetylene reactor control problem. Based on the economics of the system, the cost function should address the ethylene loss and the degree of deactivation. A study carried out by Michael et al, (1991) has shown that the loss of ethylene increases dramatically over time. Also, a sensitivity analysis proved that the catalyst activity is relatively insensitive to changes in the manipulated variables, the deactivation term can be neglected and the cost function is defined solely in terms of the net ethylene conversion. In this research, 3 objective function was carried out to test the model.

The first objective function was to maximize ethylene gain.

$$\text{Maximize } F(t_1, t_2, t_3) = \text{Net ethylene gain} \quad (6.1)$$

$$\text{Net ethylene gain} = (C_2H_2 \rightarrow C_2H_4) - (C_2H_4 \rightarrow C_2H_6) \quad (6.2)$$

The second objective function was to minimize ethane gain.

$$\text{Minimize } F(t_1, t_2, t_3) = \text{Net ethane lost} \quad (6.3)$$

$$\text{Net ethane gain} = (C_2H_4 \rightarrow C_2H_6) \quad (6.4)$$

Step 3. Development via mathematical expression a valid process or equipment model related to the input-output variables of the process and associated coefficients.

A mathematical expression of the first objective function was described as follows:

Maximize net ethylene gain = Mass flow rate of C_2H_4 in (product-feed) (kg/hr) (6.5)

Constrains of the objective function included:

1. Overall mass balance
2. Overall energy balance
3. Maximum heat exchange in the heat exchangers
4. Output temperature of each reactor $\leq$ 100 °C
5. Minimum flow in pipe $\leq F_i \leq$ Maximum flow in pipe
6. $P_{in} \leq$ Maximum pressure allowed in pipe and units
7. C_2H_2 outlet $\leq C_2H_2$ max allowed

A mathematical expression of the second objective function was described as follows:

Minimize net ethane gain = Mass flow rate of C_2H_6 in (product-feed) (kg/hr) (6.6)

Constrains of the objective function included:

1. Overall mass balance
2. Overall energy balance
3. Maximum heat exchange in the heat exchangers
4. Output temperature of each reactor $\leq$ 100 °C
5. Minimum flow in pipe $\leq F_i \leq$ Maximum flow in pipe
6. $P_{in} \leq$ Maximum pressure allowed in pipe and units
7. C_2H_2 outlet $\leq C_2H_2$ max allowed

Table 6.1 Values used in the simulation (TOC technical department)

Ethylene Price	US$480/Ton
Cost for cooling water	1 Baht/m^3

Step 4. Determine if the problem formulation is too large in scope.

In this research case, the problem formulation can be reduced once consider the facts of operation and nature hood of each variable.

When normally operated, the unit feed (F_{in}) is sent from the cracker, which as a whole provide a rather steady and un-interrupted amount of feed. Hence, the F_{in} variable is varied in a very narrow range, or sometime left unchanged.

Similar to F_{in} , the composition of the feed stream is an un-controlled variable, we can not change the X_{in} of each component, it is also left unchanged. The X_{in} used in the calculation of the optimization run is based on the result from the data-fit and data-reconciliation run. Also, P_{in} is considered stable (standard deviation over 2 months is only 0.27 kg/cm^2) so this variable is also keep constant.

In Summary the only variables we have in the feed stream is the Feed temperature (T_{in}), and variables in the heat exchanger.

Step 5. Apply a suitable optimization technique to the mathematical statement of the problem.

The algorithm used is the SQP method. This method is a state-of-the-art, quasi-Newton nonlinear programming algorithm. It can converge tear streams, equality constraints, and inequality constraints simultaneously with the optimization problem. The SQP method usually converges in only a few iterations but requires numerical derivatives for all decision and tear variables at each iteration.

The SQP method as implemented in Aspen Plus includes a novel feature: the tear streams can be partially converged using Wegstein, each optimization iteration and during line searches. This usually stabilizes convergence, and can reduce the overall number of iterations.

The SQP method is used as a black-box or partial black-box method, by converging tear streams and design specifications as an inside loop to the optimization problem (using separate Convergence blocks). This reduces the number of decision variables. The trade-off is the number of derivative evaluations, versus the time required per derivative evaluation. Whether SQP is the method of choice depends on your optimization problem. The default optimization convergence procedure in Aspen Plus is to converge tear streams and the optimization problem simultaneously, using the SQP method.

Table 6.2 Operating condition as of April 1999

	FEEDRX1	FEEDRX2	FEEDRX3	PRODRX1	PRODRX2	PRODRX3
Temperature ^{0}C	68.6	71.2	54.5	83.7	81	71.8
Pressure KG/SQCM	35.97	35.72	35.47	35.72	35.47	35.47
Vapor Frac	1	1	1	1	1	1
Mole Flow KMOL/HR	5882.401	5855.662	5836.754	5855.662	5836.754	5800.147
Mass Flow KG/HR	135007.5	135007.5	135007.5	135007.5	135007.5	135007.99
Volume Flow L/MIN	71731.59	72592.826	68007.955	76154.842	75651.622	72459.999
Enthalpy MMKCAL/HR	-2.824	-3.733	-5.63	-2.824	-3.733	-5.63
Mass Flow KG/HR						
H2	1866.578	1812.675	1774.559	1812.675	1774.559	1700.741
CO	62.450	62.450	62.450	62.450	62.450	62.450
CH4	23903.332	23903.332	23903.332	23903.332	23903.332	23903.332
C2H6	14881.283	14993.586	15202.702	14993.586	15202.702	16200.356
C2H4	60503.654	61044.238	60927.148	61044.238	60927.148	59999.092
C3H8	1037.886	1037.886	1037.886	1037.886	1037.886	1037.886
C3H6	31300.192	31300.192	31686.188	31300.192	31686.188	31826.899
C2H2	674.115	75.131	2.730	75.131	2.730	0.210
PD	424.476	424.476	346.657	424.476	346.657	277.023
MA	353.533	353.533	63.848	353.533	63.848	0.000

From the simulation results, shows that the current ethylene loss after the acetylene hydrogenation unit was 504.56 kg/hr. The acetylene hydrogenation in the reactors were 88.88%, 10.74% and 0.37% respectively. After applying the two optimization schemes, the results were shown in table 6.3 and 6.4.

Table 6.3 Result of optimization using maximize ethylene gain scheme

	FEEDRX1	FEEDRX2	FEEDRX3	PRODRX1	PRODRX2	PRODRX3
Temperature C	69.2	67.2	37.2	84.6	75.1	44.6
Pressure KG/SQCM	35.97	35.72	35.47	35.72	35.47	35.47
Vapor Frac	1	1	1	1	1	1
Mole Flow KMOL/HR	5882.401	5855.056	5839.795	5855.056	5839.795	5824
Mass Flow KG/HR	135007.5	135007.5	135007.5	135007.5	135007.5	135007.75
Volume Flow L/MIN	71882.824	71439.172	62798.258	76388.148	74021.272	64840.515
Enthalpy MMKCAL/HR	-2.786	-4.045	-6.748	-2.786	-4.045	-6.748
Mass Flow KG/HR						
H2	1866.578	1811.456	1780.691	1811.456	1780.691	1748.83
CO	62.450	62.450	62.450	62.450	62.450	62.450
CH4	23903.332	23903.332	23903.332	23903.332	23903.332	23903.332
C2H6	14881.283	14996.073	15174.211	14996.073	15174.211	15615.441
C2H4	60503.654	61056.57	60956.353	61056.57	60956.353	60545.291
C3H8	1037.886	1037.886	1037.886	1037.886	1037.886	1037.886
C3H6	31300.192	31300.192	31594.124	31300.192	31594.124	31640.868
C2H2	674.115	61.531	0.294	61.531	0.294	0
PD	424.476	424.476	365.218	424.476	365.218	347.999
MA	353.533	353.533	132.941	353.533	132.941	105.656

Table 6.4 Result of optimization using minimizing ethane gain scheme

	FEEDRX1	FEEDRX2	FEEDRX3	PRODRX1	PRODRX2	PRODRX3
Temperature C	69.2	67.2	37.2	84.6	75.1	44.6
Pressure KG/SQCM	35.97	35.72	35.47	35.72	35.47	35.47
Vapor Frac	1	1	1	1	1	1
Mole Flow KMOL/HR	5882.401	5855.056	5839.795	5855.056	5839.795	5824
Mass Flow KG/HR	135007.5	135007.5	135007.5	135007.5	135007.5	135007.75
Volume Flow L/MIN	71882.824	71439.172	62798.258	76388.148	74021.272	64840.515
Enthalpy MMKCAL/HR	-2.786	-4.045	-6.748	-2.786	-4.045	-6.748
Mass Flow KG/HR						
H2	1866.578	1811.456	1780.691	1811.456	1780.691	1748.83
CO	62.45	62.45	62.45	62.45	62.45	62.45
CH4	23903.332	23903.332	23903.332	23903.332	23903.332	23903.332
C2H6	14881.283	14996.073	15174.211	14996.073	15174.211	15615.441
C2H4	60503.654	61056.57	60956.353	61056.57	60956.353	60545.291
C3H8	1037.886	1037.886	1037.886	1037.886	1037.886	1037.886
C3H6	31300.192	31300.192	31594.124	31300.192	31594.124	31640.868
C2H2	674.115	61.531	0.294	61.531	0.294	0
PD	424.476	424.476	365.218	424.476	365.218	347.999
MA	353.533	353.533	132.941	353.533	132.941	105.656

Table 6.5 Values of objective function

Optimization Scheme	Gain (KG/HR)
Maximization of Ethylene gain	43.655
Minimization of Ethane gain	41.637

The result shows that by increasing the inlet feed temperature in the lead reactor from 68.6 ^{0}C to 69.2 ^{0}C, the lead reactor removes nearly 91% of the acetylene in the

feed stream and almost 8% of the rest is removed by the intermediate reactor. Also, what was discovered was that the optimal inlet temperature of the lead reactor is always at the upper bound of which constrains were the effluent temperature of the lead reactor. However, this optimum temperature for the feed stream to the lead reactor was within the normal operation range since it was found from operation data that the highest inlet temperature to the lead reactor was 82.5 ^{0}C.

The intermediate reactor result shows that by lowering the inlet temperature from 71.2 ^{0}C to 67.2 ^{0}C could lead to a lower ethylene dehydrogenation and effective acetylene removal. Along with that, the guard reactor result shows that by lowering the inlet temperature from 54.5 ^{0}C to 37.2 ^{0}C could lead to both a lower ethylene dehydrogenation and effluent acetylene specification. The study shows that in the normal operation case, the lost of ethylene happens mostly in the intermediate and the lead reactor. In initiative of high inlet temperature in both reactors leads to an exothermic reaction of ethylene, MA and PD.

The optimization model was then tested to verify the consistency of the model with other sets of data. Results are shown as followed.

Table 6.6 Result of optimization using maximize ethylene gain scheme (May 1999)

	FEEDRX1	FEEDRX2	FEEDRX3	PRODRX1	PRODRX2	PRODRX3
Temperature C	70	67.5	37.2	85.4	75.4	44.5
Pressure KG/SQCM	35.97	35.72	35.47	35.72	35.47	35.47
Vapor Frac	1	1	1	1	1	1
Mole Flow KMOL/HR	5923.294	5895.671	5880.267	5895.671	5880.267	5864.696
Mass Flow KG/HR	136014.5	136014.5	136014.5	136014.5	136014.5	136014.74
Volume Flow L/MIN	72574.021	71973.943	63173.266	77110.798	74576.99	65186.311
Enthalpy MMKCAL/HR	-5.487	-6.797	-9.547	-5.487	-6.797	-9.547
Mass Flow KG/HR						
H2	1845.238	1789.554	1758.502	1789.554	1758.502	1727.092
CO	55.133	55.133	55.133	55.133	55.133	55.133
CH4	24119.525	24119.525	24119.525	24119.525	24119.525	24119.525
C2H6	17737.466	17854.079	18031.882	17854.079	18031.882	18466.306
C2H4	59823.521	60380.844	60280.303	60380.844	60280.303	59875.586
C3H8	1359.398	1359.398	1359.398	1359.398	1359.398	1359.398
C3H6	29587.437	29587.437	29888.806	29587.437	29888.806	29935.640
C2H2	679.198	60.946	0.299	60.946	0.299	0.000
PD	451.303	451.303	390.545	451.303	390.545	373.293
MA	356.282	356.281	130.107	356.281	130.107	102.769

Table 6.7 Result of optimization using minimizing ethane gain scheme (May 1999)

	FEEDRX1	FEEDRX2	FEEDRX3	PRODRX1	PRODRX2	PRODRX3
Temperature C	70	67.5	37.2	85.4	75.4	44.5
Pressure KG/SQCM	35.97	35.72	35.47	35.72	35.47	35.47
Vapor Frac	1	1	1	1	1	1
Mole Flow KMOL/HR	5923.294	5895.671	5880.267	5895.671	5880.267	5864.696
Mass Flow KG/HR	136014.5	136014.5	136014.5	136014.5	136014.5	136014.74
Volume Flow L/MIN	72574.021	71974.188	63172.71	77110.798	74577.321	65185.576
Enthalpy MMKCAL/HR	-5.487	-6.797	-9.547	-5.487	-6.797	-9.547
Mass Flow KG/HR						
H2	1845.238	1789.554	1758.5	1789.554	1758.5	1727.094
CO	55.133	55.133	55.133	55.133	55.133	55.133
CH4	24119.525	24119.525	24119.525	24119.525	24119.525	24119.525
C2H6	17737.466	17854.079	18031.888	17854.079	18031.888	18466.276
C2H4	59823.521	60380.844	60280.299	60380.844	60280.299	59875.616
C3H8	1359.398	1359.398	1359.398	1359.398	1359.398	1359.398
C3H6	29587.437	29587.437	29888.823	29587.437	29888.823	29935.65
C2H2	679.198	60.946	0.297	60.946	0.297	0
PD	451.303	451.303	390.542	451.303	390.542	373.292
MA	356.282	356.281	130.095	356.281	130.095	102.761

Table 6.8 Values of objective function (May 1999)

Optimization Scheme	Gain (KG/HR)
Maximization of Ethylene gain	52.065
Minimization of Ethane gain	52.095

Table 6.9 Result of optimization using maximize ethylene gain scheme (June 1999)

	FEEDRX1	FEEDRX2	FEEDRX3	PRODRX1	PRODRX2	PRODRX3
Temperature C	72.7	36.2	37.3	88.9	37.3	44.8
Pressure KG/SQCM	35.97	35.72	35.47	35.72	35.47	35.47
Vapor Frac	1	1	1	1	1	1
Mole Flow KMOL/HR	5926.271	5896.451	5893.951	5896.451	5893.951	5878.009
Mass Flow KG/HR	136014.5	136014.965	136014.965	136014.965	136014.965	136014.97
Volume Flow L/MIN	73395.168	62621.757	63425.451	78100.625	63425.451	65484.488
Enthalpy MMKCAL/HR	-4.962	-8.804	-8.804	-4.991	-8.804	-8.804
Mass Flow KG/HR						
H2	1880.067	1819.919	1814.878	1819.919	1814.878	1782.740
CO	56.970	56.970	56.970	56.970	56.970	56.970
CH4	23493.945	23493.945	23493.945	23493.945	23493.945	23493.945
C2H6	18380.897	18510.936	18561.147	18510.936	18561.147	19007.033
C2H4	60228.164	60822.569	60775.725	60822.569	60775.725	60359.731
C3H8	1411.824	1411.824	1411.824	1411.824	1411.824	1411.824
C3H6	29091.145	29091.145	29126.089	29091.145	29126.089	29172.969
C2H2	663.831	0.000	0.000	0.000	0.000	0.000
PD	451.339	451.339	444.294	451.339	444.294	427.025
MA	356.318	356.318	330.092	356.318	330.092	302.728

Table 6.10 Result of optimization using minimizing ethane gain scheme (June 1999)

	FEEDRX1	FEEDRX2	FEEDRX3	PRODRX1	PRODRX2	PRODRX3
Temperature C	72.7	36.2	37.3	88.9	37.3	44.8
Pressure KG/SQCM	35.97	35.72	35.47	35.72	35.47	35.47
Vapor Frac	1	1	1	1	1	1
Mole Flow KMOL/HR	5926.271	5896.451	5893.951	5896.451	5893.951	5878.009
Mass Flow KG/HR	136014.5	136014.965	136014.965	136014.965	136014.965	136014.97
Volume Flow L/MIN	73395.168	62621.757	63425.706	78100.625	63425.451	65484.488
Enthalpy MMKCAL/HR	-4.962	-8.804	-8.804	-4.991	-8.804	-8.804
Mass Flow KG/HR						
H2	1880.067	1819.919	1814.878	1819.919	1814.878	1782.740
CO	56.970	56.970	56.970	56.970	56.970	56.970
CH4	23493.945	23493.945	23493.945	23493.945	23493.945	23493.945
C2H6	18380.897	18510.936	18561.147	18510.936	18561.147	19007.033
C2H4	60228.164	60822.569	60775.725	60822.569	60775.725	60359.731
C3H8	1411.824	1411.824	1411.824	1411.824	1411.824	1411.824
C3H6	29091.145	29091.145	29126.089	29091.145	29126.089	29172.969
C2H2	663.831	0.000	0.000	0.000	0.000	0.000
PD	451.339	451.339	444.294	451.339	444.294	427.025
MA	356.318	356.318	330.092	356.318	330.092	302.728

Table 6.11 Values of objective function (June 1999)

Optimization Scheme	Gain (KG/HR)
Maximization of Ethylene gain	131.567
Minimization of Ethane gain	131.567

The result shows that the optimum lead inlet temperature was always at upper bound, slightly from the first optimal point. This was from the consistent amount of

acetylene entering the lead reactor between 4000-5000ppm. The optimum intermediate inlet temperature changes due to that it has to meet the constraint of concentration of acetylene in the effluent. The optimum intermediate inlet temperature remains the same due to almost all of the acetylene was removed and ethylene hydrogenation was keep to minimum.

Step 6. Check the answers, and examine the sensitivity of the result to changes in the coefficient in the problem and the assumptions.

This study is carried out in section 6.3

6.3 Sensitivity Analysis and Case Studies

The model has been develop with a mathematical model that was developed earlier and then was adjust to fit the actually plant data. The reconciled model is then used to find the optimal condition for the operation. Once done, sensitivity analysis brings out the procedures of check the answers, and examines the sensitivity of the result to changes in the coefficient in the problem and the assumptions. The major benefit of the reactor model is its ability to predict the reactors' performance when the feed composition or condition is changed.

Case 1. A 10% increase of acetylene concentration in the feed.

With a 10% increase in acetylene in the feed. If there were no adjustment to any of the bed inlet temperature, the acetylene in the reactor effluent will go off specification from 0 ppmv to 672 ppmv. To remain on specification, varying the inlet temperature to each of the beds to maintain the 1ppmv at the reactor effluent from the guard reactor.

In table 6.14, by holding the intermediate and the guard bed inlet temperature constant, we have to increase the lead bed inlet temperature from 69.20 ^{0}C to 72.66 ^{0}C to meet the acetylene specification at the guard bed outlet.

In table 6.15, by holding the lead and the guard bed inlet temperature constant, we have to increase the intermediate bed inlet temperature from 66.8 ^{0}C to 84.64 ^{0}C.

The outlet temperature of the intermediate bed of 100.1 ^{0}C was considerably high, of which still couldn't keep the acetylene effluent of the guard reactor on specification.

In table 6.16, by holding the lead and the intermediate bed inlet temperature constant, we have to increase the guard bed inlet temperature from 37.2 ^{0}C to 74.55 ^{0}C. The outlet temperature of the guard bed of 125.62 ^{0}C was considerably high, of which still couldn't keep the acetylene effluent of the guard reactor on specification.

Showing that the lead reactor removes most of the acetylene from the feed stream. The major reactor to focus of these three series of reactor is the lead reactor. Acetylene fluctuation and product off specification could be avoid and controlled by the lead reactor.

Table 6.12 Reactor effluent in normal case

	Lead		**Intermediate**		**Guard**	
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Temp, C	69.20	84.66	66.80	74.53	37.20	44.69
C2H2, PPMV	4996.55	452.92	452.92	5.38	5.38	0.00

Table 6.13 Reactor effluent with 10% increase of acetylene inlet concentration

	Lead		**Intermediate**		**Guard**	
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Temp, C	69.20	84.65	66.80	74.55	37.20	44.70
C2H2,PPMV	5668.90	1128.54	1128.54	680.25	680.25	672.14

Table 6.14 Reactor effluent with 10% increase of acetylene inlet concentration with lead bed inlet temperature adjustment

	Lead		**Intermediate**		**Guard**	
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Temp, C	72.66	90.52	66.80	74.39	37.20	44.65
C2H2,PPMV	5668.90	406.82	406.82	0.00	0.00	0.00

Table 6.15 Reactor effluent with 10% increase of acetylene inlet concentration with intermediate bed inlet temperature adjustment

	Lead		**Intermediate**		**Guard**	
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Temp, C	69.2	84.64	84.64	100.10	37.20	44.19
C2H2,PPMV	5668.90	1128.54	1128.54	210.66	210.66	202.74

Table 6.16 Reactor effluent with 10% increase of acetylene inlet concentration with guard bed inlet temperature adjustment

	Lead		**Intermediate**		**Guard**	
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Temp, C	69.20	84.65	66.80	74.55	74.55	125.62
C2H2,PPMV	5668.90	1128.54	1128.54	680.25	680.25	624.26

Case 2 A 10% increase of the feed flow rate.

Increase feed rate to the reactor changed the space velocity of the reactor. To compensate for the changes, slight increase or decrease in the inlet temperature should be made to obtain optimal operation. The result is shown in table 6.17.

The ethylene gain was reported 239.251 kg/hr, compared to the normal case of 43.65 kg/hr.

Table 6.17 Reactor effluent with 10% increase of feed flow rate

	Lead		Intermediate		Guard	
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Temp, C	73.80	38.20	35.80	90.30	39.70	42.10
C2H2,PPMV	4993.16	126.01	126.01	3.00	3.00	0.00

Case 3 A combination of 10% increase of the feed flow rate and 10% increase of acetylene in the feed.

This case is a combination of the above two disturbances happening simultaneously. It is difficult enough for operators to try to predict how to react to a single disturbance, let alone one facing such a combination would be more. The model computes and optimizes the required changed to negate the disturbances shown in table 6.18.

Table 6.18 Reactor effluent with a combination of 10% increase of the feed flow rate and 10% increase of acetylene in the feed

	Lead		Intermediate		Guard	
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Temp, C	76.52	95.00	50.79	53.75	36.76	43.33
C2H2,PPMV	5668.90	208.81	208.81	2.97	2.97	0.00

Case 4 Lead inlet temperature change

This is shown in figure 6.1 and figure 6.2. The lead inlet temperature has a significant impact on process. During the length of the operation, higher temperatures are required to ensure the required acetylene conversion. The lead inlet temperature affects the rate of C_2H_6 production in all of the reactors.

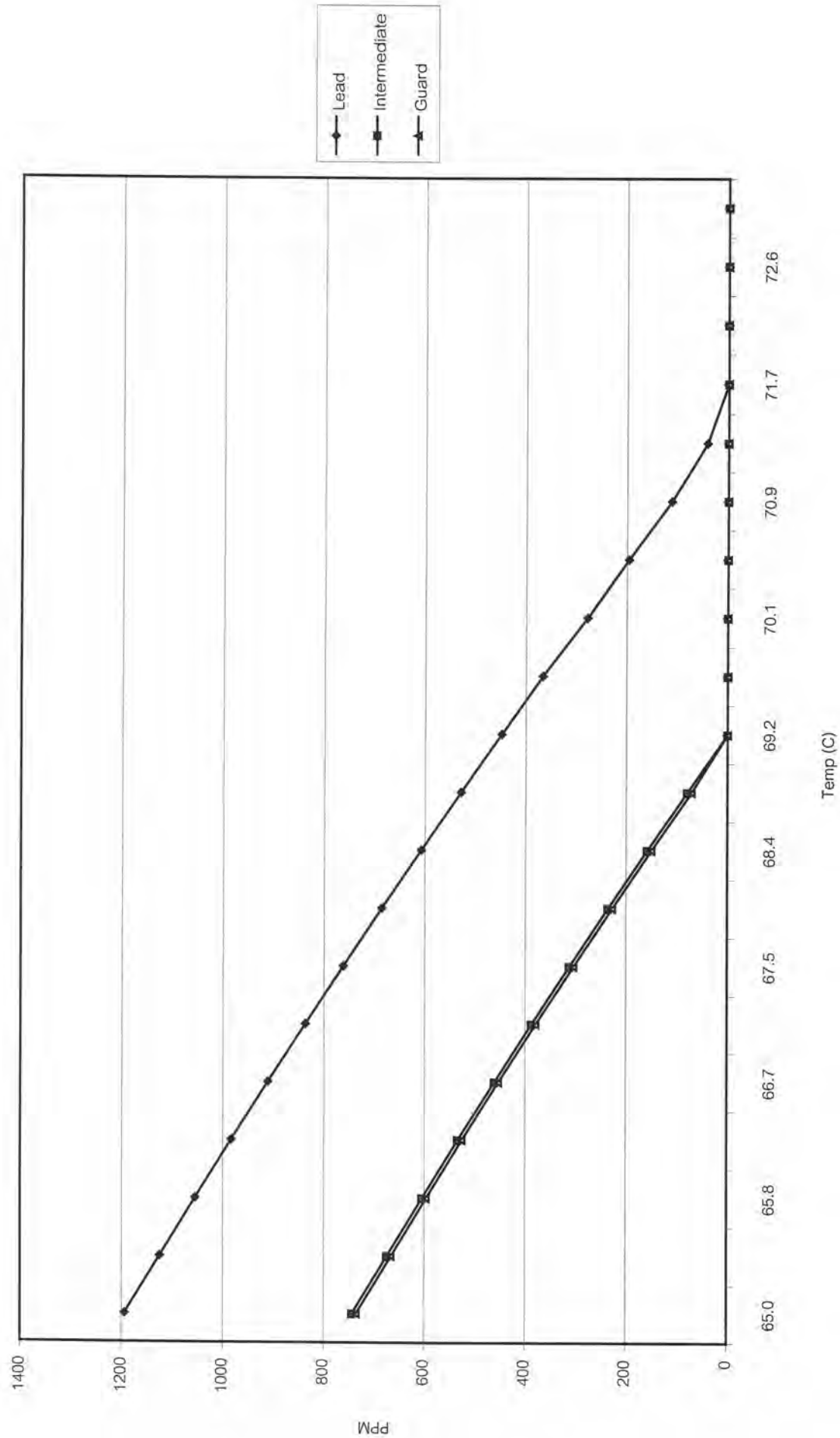

Figure 6.1 Sensitivity of acetylene concentration to lead inet temperature

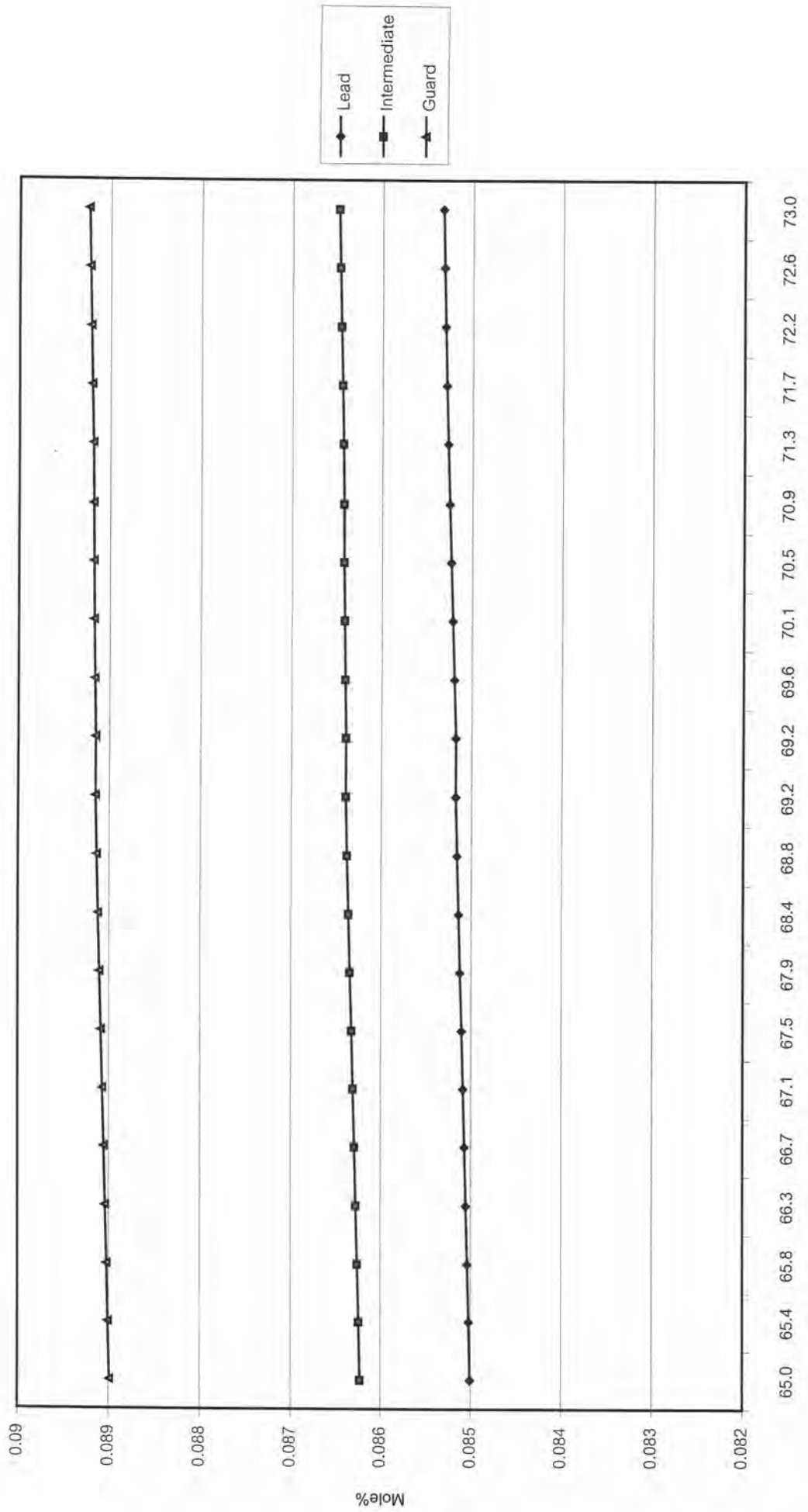

Figure 6.2 Sensitivity of ethane concentration to lead inet temperature

6.4 Summary

In this chapter, an optimization and sensitivity study was carried out to observed the optimal operating condition for the process. It was found that recently the process was expecting an ethylene loss of over 504.56 kg/hr. Most of the ethylene lost occurred from the intermediate and guard bed due to high inlet temperature causing a domino effect to other exothermic reactor within the reactors.

By raising the lead inlet temperature, the acetylene conversion in the lead reactor increases. Then a lower inlet temperature in the intermediate and guard reactor could result to a lower ethylene conversion. Result shows that by employing the optimum condition, an ethylene gain of 43.6 kg/hr was achieved.

The study employs three objective functions to verify the optimum condition for each objective function. Results shows that the optimum condition for all three objective functions was nearly the same. Weighting on the objective function was also employed and the result shows that minimizing the ethane production has more effect to the optimum condition than maximizing the ethylene production.

The optimization scheme was then tested with other months of data, the results were that the lead inlet temperature is always at the upper bound constraint of the effluent temperature. The ethylene hydrogenation in the lead reactor was trace compared to the acetylene hydrogenation resulting the trial or maximum conversion of acetylene in the lead reactor, leading to an upper bound constrain limit. As for the intermediate and the guard reactor, opposite to the lead reactor, a minimum ethylene conversion was prevented, leading to a lower inlet temperature for both reactors.

A sensitivity analysis was made to study the effect of disturbance to the optimum condition, how the reactors will response and how to encounter the situation if such events occur.

A 10% increment of acetylene in the feed stream leads to a higher inlet temperature for the lead reactor. However, a share of the load and temperature could be shared to the intermediate and guard reactor as well, but little could be achieve since the catalyst activity of both reactors were lesser than the lead reactor. A result shows that nearly 85%-95% of the acetylene removal happens in the lead reactor.

A 10% increment of feed stream flowrate shows the impact lesser residence time of the feed stream in the acetylene hydrogenation reactor unit. Still a higher lead inlet temperature was employed to ensure the product specification with little adjustments of the intermediate and guard inlet temperature.

A 10% increment of both the acetylene and feed flowrate was introduced to the process resulting a similar result, that is a higher lead inlet temperature. A notice of the effluent temperature of the lead reactor should be considered since it almost reaches 100C. An increment of inlet temperature from the intermediate reactor was employed to reduce this warning.

A study of the effect of the acetylene conversion and ethylene conversion was also made. Results verified that the decrease of the lead inlet could be done, however, the acetylene concentration outlet of the guard reactor should be carefully monitored when employing a lead inlet temperature of lower than 69.2 ^{0}C.

An increment of lead inlet temperature ensures a complete removal of the acetylene at the lead reactor, however will shorten the catalyst life if catalyst life was taken in concern.